

PROCESS FOR INCREASED PRODUCTION OF FCC GASOLINE

FIELD OF THE INVENTION

[0001] This invention relates to methods and processes for increasing the production of FCC (Fluid Catalytic Cracking) gasoline products, and optionally distillate products, from refinery feedstocks.

BACKGROUND OF THE INVENTION

[0002] An important process to the overall gasoline production in the world is the refining Fluid Catalytic Cracking (“FCC”) related processes. FCCs utilize very small particulate catalysts which are raised to very high temperatures and subsequently fluidized. These fluidized particles contact high molecular weight petroleum feeds and catalytically “crack” these larger hydrocarbon molecules to lower boiling products which are more valuable products. Most FCC processes contact heavy feed oils (such as vacuum gas oils, atmospheric gas oils, and often petroleum resids) with the fluidized catalysts typically with the goal to maximize naphtha production volumes.

[0003] In the FCC process these low-value, high boiling point hydrocarbon feedstocks are catalytically converted into more valuable products by contacting the feeds with fluidized catalyst particles in the process. In modern “short contact time” fluidized catalytic cracking (FCC) units, the hydrocarbon feedstocks are typically contacted with the fluidized catalyst particles in the riser section of the FCC reactor. 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, such as naphthas, and minimize the formation of less desirable products such as light gases and coke.

[0004] The FCC naphthas derived from such processes are very valuable products as they are used as a component in final gasoline production. FCC naphthas can often account for about 50% or more of the overall “gasoline blending feedstock” in a refinery. Additionally FCC naphthas typically have a relatively high octane value as compared to “straight run” naphthas that are typically produced by a refinery’s crude unit. This high octane value of the FCC naphthas is in large part due to the high olefin content of the FCC naphthas. As such, maximizing the total of production of FCC naphthas suitable for gasoline blending is of significant importance to the commercial operations and economics of any petroleum refinery.

[0005] While FCC units may target the maximization of other hydrocarbon products, such as distillates used in diesel production, or much smaller quantities of petrochemical production, such as propylene, most FCC units, particularly in the United States, target to maximize the overall naphtha production. The overall naphtha production (i.e., light cat naphtha “LCN” and heavy cat naphtha “HCN”) can typically exceed over 50 vol% of the overall products obtained from an FCC unit. As such, even small percentages of increases in the naphtha (or gasoline) production from an FCC unit result in significant cost improvements for the average refinery. For example, with a 1% increase in naphtha production, an FCC unit operating at 100,000 barrels per day (bbl/day) would see an approximate increase in gasoline production of about 500 bbl/day or about 20,000 gallons per day of increase in gasoline production. As can be seen, even small improvements in naphtha/gasoline yields from the FCC unit result in significant financial benefit.

[0006] As such, what is needed in the industry are new methods and processes for improving the naphtha yields from a refinery FCC unit that do not require significantly changing the overall FCC process.

SUMMARY OF THE INVENTION

[0007] The processes of present invention are designed to increase the overall naphtha production from a Fluid Catalytic Cracking (“FCC”) unit for maximizing gasoline production. The processes herein may also be utilized to optionally increase distillate production, preferably for increased diesel and/or jet fuel production. The processes herein are aimed at modifying the properties of a typical FCC feedstream in order to increase the naphtha, or optionally distillate, production obtained from the FCC process.

[0008] In the processes herein, the iso-paraffin content of the feed to the FCC unit is increased resulting in higher naphtha (i.e., gasoline) production from the FCC unit. It has also been found that distillate (i.e., diesel) production can also be increased with a corresponding decrease in heavier cat bottoms products. Preferably, at least 50 wt%, more preferably at least 75 wt%, even more preferably at least 90 wt%, and most preferably substantially all of the normal paraffins in the feed to the hydroisomerization unit are converted to isoparaffins within the process. At least a portion of the hydroisomerized product is sent to an FCC unit for further processing into naphtha and distillate products.

[0009] A first non-limiting embodiment of the invention relates to a process for increasing Fluid Catalytic Cracking (“FCC”) gasoline production comprising:

a) contacting a hydrocarbon-containing hydroisomerization feedstream with a hydroisomerization catalyst under hydroisomerization conditions to produce at least one hydroisomerized liquid product stream that has a higher iso-paraffin content than the hydroisomerization feedstream;

b) contacting in the reaction zone of an FCC reactor riser an FCC feedstream comprising at least a portion of the hydroisomerized liquid product stream of step a) with a fluid catalytic cracking catalyst thereby catalytically

cracking the FCC feedstream into an FCC product that has an average lower boiling point than the FCC feedstream, and producing a spent catalyst;

c) separating the FCC product from the spent catalyst;

d) cooling the FCC product; and

e) fractionating the FCC product into multiple FCC product streams, wherein at least one of the FCC product streams is a naphtha boiling-range product stream; and

f) utilizing at least a portion of the naphtha boiling-range product stream for gasoline production.

[0010] A second non-limiting embodiment of the invention relates to the first embodiment, further comprising:

- contacting a hydrocarbon-containing hydrotreater feedstream containing at least 250 ppmw of sulfur with a hydrotreating catalyst under hydrotreating conditions to produce at least one hydrotreated liquid product stream and one hydrotreated vapor stream, wherein the hydrotreated liquid product stream has a lower sulfur content than the sulfur-containing hydrocarbon feedstream;

- separating the hydrotreated liquid product stream from the hydrotreated vapor stream; and

- utilizing at least a portion of the hydrotreated liquid product stream as the hydroisomerization feedstream in step a).

[0011] A third non-limiting embodiment of the invention relates to the first embodiment, further comprising:

- contacting a hydrocarbon-containing hydrotreater feedstream containing at least 250 ppmw of sulfur with a hydrotreating catalyst under hydrotreating conditions to produce the hydroisomerization feedstream.

[0012] A fourth non-limiting embodiment of the invention relates to a process for increasing Fluid Catalytic Cracking (“FCC”) gasoline production comprising:

a) contacting a hydrocarbon-containing hydroisomerization feedstream with a hydroisomerization catalyst under hydroisomerization conditions to produce at least one hydroisomerized product stream that has a higher iso-paraffin content than the hydroisomerization feedstream;

b) contacting at least a portion of the hydroisomerized product stream with a hydrotreating catalyst under hydrotreating conditions to produce at least one hydrotreated liquid product stream and one hydrotreated vapor stream, wherein the hydrotreated liquid product stream has a lower sulfur content than the sulfur-containing hydrocarbon feedstream;

c) separating the hydrotreated liquid product stream from the hydrotreated vapor stream;

d) contacting in the reaction zone of an FCC reactor riser an FCC feedstream comprising at least a portion of the hydrotreated liquid product stream of step c) with a fluid catalytic cracking catalyst thereby catalytically cracking the FCC feedstream into an FCC product that has an average lower boiling point than the FCC feedstream, and producing a spent catalyst;

e) separating the FCC product from the spent catalyst;

f) cooling the FCC product; and

e) fractionating the FCC product into multiple FCC product streams, wherein at least one of the FCC product streams is a naphtha boiling-range product stream; and

g) utilizing at least a portion of the naphtha boiling-range product stream for gasoline production.

[0013] In other more preferred embodiments of the first through fourth embodiments, at least 50 wt% of the normal paraffins in the hydroisomerization feedstream are converted to iso-paraffins in the hydroisomerized liquid product

stream in step a). In other more preferred embodiments, the hydroisomerization catalyst comprises at least one Group VIIIA metal, and further comprises a zeolite selected from EU-1, ZSM-35, ZSM-11, ZSM-57, NU-87, SAPO-11, ZSM-22, and ZSM-48. In other preferred embodiments, the Group VIIIA metal of the hydroisomerization catalyst is selected from Pt and Pd. In yet other more preferred embodiments, the hydroisomerization catalyst further comprises at least one Group VIA metal, wherein the Group VIA of the hydroisomerization catalyst is selected from Mo and W, and the Group VIIIA metal of the hydroisomerization catalyst is selected from Ni and Co; even more preferably, the Group VIA of the hydroisomerization catalyst is W, the Group VIIIA metal of the hydroisomerization catalyst is Ni, and the zeolite in the hydroisomerization catalyst is ZSM-48.

[0014] In other more preferred embodiments of the first through fourth embodiments, the hydroisomerization feedstream contains over 300 ppmw of sulfur.

[0015] In other more preferred embodiments of the first through third embodiments, the at least one hydroisomerized liquid product stream of step a) is sent to a distillation column of to produce the at least a portion of the hydroisomerized liquid product stream of step b), as well as producing a distillation column overhead vapor stream and at least a first distillate product stream from the distillation column, wherein the distillation column overhead vapor stream and the first distillate product stream are not sent to the reaction zone of the FCC reactor riser.

[0016] In another more preferred embodiment of the fourth embodiment, the separation of step c) is performed in a distillation column to produce the hydrotreated liquid product stream and the hydrotreated vapor stream, as well as producing a distillation column overhead vapor stream and at least a first

distillate product stream from the distillation column, wherein the distillation column overhead vapor stream and the distillate product stream are not sent to the reaction zone of the FCC reactor riser.

DETAILED DESCRIPTION OF PREFERRED EMBODIMENTS

[0017] An object of the present processes of invention is to increase the overall naphtha production from a Fluid Catalytic Cracking (“FCC”) unit for maximizing gasoline production. The processes herein may also be utilized to optionally increase distillate production, preferably for increased diesel and/or jet fuel production. The processes herein are aimed at modifying the properties of a typical FCC feedstream in order to increase the naphtha, or optionally distillate, production obtained from the FCC process. The present invention does not significantly alter the FCC process and as such, it can be used with most existing FCC units and hardware. Additionally, since the overall FCC process is not significantly changed, the FCC pretreatment processes herein can be used to prepare any amount of the FCC feedstock from a small portion of the overall FCC feedstream (e.g., < 25wt% or < 10 wt% of the overall feedstream) to preferably a significant portion of the overall FCC feedstream (e.g., > 75wt% or > 85 wt% of the overall feedstream) without significantly altering the overall effects of the FCC unit processing on the remainder of the overall feedstream to FCC unit.

[0018] In an embodiment of the present invention, the composition of a typical FCC feedstream can be utilized as a starting feed to the present processes. Preferably, the feedstream utilized has a T5 boiling point of at least 400°F, more preferably of at least 450°F, and a T95 boiling point of less than about 1150°F, more preferably less than about 1100°F. The “T5 boiling point” is defined as the temperature under atmospheric pressure at which 5 wt% of the product sample boils. Similarly, “T95 boiling point” is defined as the

temperature under atmospheric pressure at which 95 wt% of the product sample boils. Such feeds may be derived from many sources within the refinery but typically are comprised of an atmospheric gas oil (“AGO”), vacuum gas oil (“VGO”) or both. These feeds may also contain hydroprocessed feed components such as a product stream from a hydrocracking unit that falls within the boiling points noted above.

[0019] It should also be noted that the term “naphtha” as used herein shall mean a hydrocarbon-based stream that has a T5 boiling point of at least 80°F (27°C) and a T95 boiling point of less than 450°F (232°C). The term “distillate” as used herein shall mean a hydrocarbon-based stream that has a T5 boiling point of at least 350°F (177°C) and a T95 boiling point of less than 650°F (343°C). Both naphthas and distillates typically refer to intermediate product streams in a petroleum or petrochemical refinery that may also be utilized for final product blending.

[0020] For the processes herein, the feedstream to the overall process embodiments as described preferably is comprised substantially of a hydrocarbon feedstream derived from a fossil-based oil material such as a crude oil, tar sands, or bitumens. In preferred embodiments, the feedstream is comprised of at least 75 wt%, more preferably at least 85 wt%, of a hydrocarbon feedstream derived from a fossil-based oil material. The processes herein may also be utilized to process hydrocarbon streams that are derived from renewable materials (i.e., “biofuel sources”). However, in preferred embodiments of the processes herein, from 5 to 25 wt%, more preferably from 10 to 20 wt%, of the overall feedstream is derived from renewable biofuel sources. Such biofuel sources include, but are not limited to, vegetables, animal, fish, and/or algae materials. If such renewable biofuel sources are utilized as a portion of the feedstream herein, it is preferred that the materials have been hydroprocessed

and deoxygenated prior to incorporating them with the fossil-based oil material being supplied to the present processes.

[0021] The process herein has the ability to process feedstreams with high sulfur contents. In preferred embodiments herein the feedstream has a sulfur content of at least 250 ppmw sulfur, or at least 500 ppm sulfur, or at least 1000 ppmw sulfur, or at least 3000 ppmw sulfur. The feedstreams may also contain nitrogen, but it is preferred that the nitrogen content of the feed be kept to less than about, 5000 ppm w, more preferably less than 2000 ppmw, even more preferably less than 1500 ppmw, and most preferably less than 100 ppmw of nitrogen.

[0022] In a preferred embodiment of the processes of invention herein, a hydrocarbon feedstream is sent to a hydroisomerization unit. In this step of the process, at least a portion of the hydrocarbon feedstream is hydroisomerized. The increase in the iso-paraffin content of the feed to the FCC unit is shown to increase the much desired naphtha (i.e., gasoline) production from the FCC unit. It has also been found that distillate (i.e., diesel) production can also be increased with a corresponding decrease in heavier cat bottoms products. In this step, at least a portion, preferably most, of the normal paraffins are converted to isoparaffinic hydrocarbon species. Preferably, at least 50 wt%, more preferably at least 75 wt%, even more preferably at least 90 wt%, and most preferably substantially all of the normal paraffins in the feed to the hydroisomerization unit are converted to isoparaffins within the process. Here, it is preferred that the hydroisomerization unit is run under conditions to maximize normal paraffin to isoparaffin conversion, while minimizing the conversion of naphthalenes and aromatics in the hydroisomerization feed, as these latter components are valuable to gasoline production in the FCC unit. With the disclosures herein, one of skill in the art will be able to make such adjustments to the system so as to achieve these results. In more preferred embodiments, the liquid product from

the hydroisomerization unit contains at least 10 wt%, more preferably at least 15 wt%, and even more preferably at least 20 wt% of isoparaffinic species. In all configurations herein, the liquid product from the hydroisomerization unit will have a higher isoparaffin content (by wt%) than the hydrocarbon feedstream to the hydroisomerization unit. This hydroisomerization will also result in some additional beneficial isomerization in alkyl side-chains of the ringed molecules (such as for example, aromatics, naphtheno-aromatics, and naphthenes) in the feed which can be present in significant amounts. Such additional alkyl side-chain isomerization will also improve the final naphtha and/or distillate production in the FCC stage of the present processes which are to be further discussed herein.

[0023] Preferred operating conditions in the hydroisomerization reaction unit include contacting the hydroisomerization feed obtained from the first hydrotreating step described above with an isomerization catalyst at a temperature of from 400 to 850°F (204 to 454°C), preferably 525 to 750°F (274 to 399°C), a hydrogen partial pressure of from 1.8 to 34.6 mPa (250 to 5000 psi), preferably 4.8 to 20.8 mPa, a liquid hourly space velocity of from 0.2 to 10 v/v/hr, preferably 0.5 to 3.0, and a hydrogen circulation rate of from 35.6 to 1781 m³/m³ (200 to 10,000 scf/B), preferably 178 to 890.6 m³/m³ (1000 to 5000 scf/B).

[0024] Preferably, the hydroisomerization catalysts utilized in the processes herein are comprised of at least one zeolite. More preferably, the zeolites have a unidimensional pore structure. Preferred catalysts include 10 member ring pore zeolites, such as EU-1, ZSM-35 (or ferrierite), ZSM-11, ZSM-57, NU-87, SAPO-11, ZSM-22, and ZSM-48. Other suitable materials are EU-2, EU-11, ZBM-30, MCM-48, and ZSM-23. Most preferably, the hydroisomerization catalyst is comprised of ZSM-48. Note that a zeolite having the ZSM-23 structure with a silica to alumina ratio of from about 20:1 to about 40:1 can

sometimes be referred to as SSZ-32. Other molecular sieves that are isostructural with the above materials include Theta-1, NU-10, EU-13, KZ-1, and NU-23.

[0025] In various embodiments, the hydroisomerization catalyst further comprises a metal hydrogenation component. The metal hydrogenation component is typically a Group VIA and/or a Group VIIIA metal. Preferably, the hydroisomerization catalyst includes at least one Group VIIIA metal. More preferably, the hydroisomerization catalyst includes at least one Group VIIIA metal and at least one Group VIA metal. In some preferred embodiments, the metal hydrogenation component of the hydroisomerization catalyst is a Group VIIIA noble metal. Preferably, the metal hydrogenation component is Pt, Pd, or a mixture thereof. In an alternative preferred embodiment, the metal hydrogenation component can be at least one non-noble Group VIIIA metal optionally coupled with at least one Group VIA metal. Suitable combinations of this alternative preferred embodiment can include Ni, Co, or Fe with Mo or W, preferably Ni with Mo or W. The alternative hydroisomerization catalysts are particularly preferred in embodiments wherein the feedstream to the hydroisomerization catalyst is not first subjected to a desulfurization step with H₂S removal.

[0026] Please note that the designation of Group VIA and Group VIIIA herein corresponds to the older IUPAC designations such as shown in the Periodic Table of Elements, published by the *Sargent-Welch Scientific Company*, 1979, wherein the Group VIA elements include the column from the periodic table of elements containing Cr, Mo, and W, and the Group VIIIA elements include the columns from the periodic table of elements containing Fe, Co, Ni, Ru, Rh, Pd, Os, Ir, and Pt.

[0027] The amount of metal in the hydroisomerization catalyst can be at least 0.1 wt% based on catalyst, or at least 0.15 wt%, or at least 0.2 wt%, or at least 0.25 wt%, or at least 0.3 wt%, or at least 0.5 wt% based on catalyst. The amount of metal in the catalyst can be 20 wt% or less based on catalyst, or 10 wt% or less, or 5 wt% or less, or 2.5 wt% or less, or 1 wt% or less. For embodiments where the metal is Pt, Pd, another Group VIIIA noble metal, or a combination thereof, the amount of metal can be from 0.1 to 5 wt%, preferably from 0.1 to 2 wt%, or 0.25 to 1.8 wt%, or 0.4 to 1.5 wt%. For embodiments where the metal is a combination of a non-noble Group VIIIA metal with a Group VIA metal, the combined amount of metal can be from 0.5 wt% to 20 wt%, or 1 wt% to 15 wt%, or 2.5 wt% to 10 wt%.

[0028] Preferably, the hydroisomerization catalysts used in processes according to the invention are catalysts with a low ratio of silica to alumina. For example, for ZSM-48, the ratio of silica to alumina in the zeolite can be less than 200:1, or less than 110:1, or less than 100:1, or less than 90:1, or less than 80:1. In various embodiments, the ratio of silica to alumina can be from 30:1 to 200:1, 60:1 to 110:1, or 70:1 to 100:1.

[0029] The hydroisomerization catalysts useful in processes according to the invention can also include a binder. In some embodiments, the hydroisomerization catalysts used in process according to the invention are formulated using a low surface area binder, a low surface area binder represents a binder with a surface area of 100 m²/g or less, or 80 m²/g or less, or 70 m²/g or less.

[0030] As noted in the alternative preferred embodiment of the hydroisomerization catalysts used in the processes herein, the metal hydrogenation component is selected to be a combination of a non-noble Group VIIIA metal with a Group VIA metal. Such suitable combinations can include

Ni, Co, or Fe with Mo or W, preferably Ni with Mo or W. In this preferred embodiment, low limits of sulfur content do not need to be maintained in the feed to the hydroisomerization unit. In these embodiment, the feedstream to the hydroisomerization unit can contain over 100 ppmw sulfur or even over 300 ppmw, or even over 500 ppmw as these catalysts will be resistant to substantial loss of hydroisomerization activity. These catalysts should also have some amount of amount of hydrodesulfurization activity.

[0031] As such, in further preferred configurations of these embodiments, some sulfur is removed from the hydroisomerization feedstream in this portion of the overall process. The sulfur may be removed from the hydroisomerization liquid product stream by sending the hydroisomerization liquid product stream to a vapor separator, which will remove the sulfur primarily in the form of H_2S . This vapor separation step is preferred to be included particularly when there is no other additional hydrotreating or substantial fractionation of the hydroisomerization unit product prior to the hydroisomerization product being sent to an FCC unit as described in the following steps. Preferably, this hydroisomerization stage also removes some portion of the nitrogen from the hydroisomerization feedstream by additionally removing NH_3 from the hydroisomerization liquid product stream in the vapor separator.

[0032] In preferred embodiments of the processes herein wherein there are no additional substantial intermediate treating steps between the hydroisomerization unit and the FCC unit (except for post-isomerization vapor separation as described), at least a portion of the hydroisomerized product stream is sent for further processing in a Fluid Catalytic Cracking ("FCC") unit. Here, the hydroisomerized product stream is preferably mixed with at least one other heavy hydrocarbon feedstream (although the addition of the heavy hydrocarbon feedstream is not required for the invention embodiments herein) to make an FCC feedstream, which is then injected through one or more feed nozzles into

the feed zone of an FCC reactor riser. Such heavy hydrocarbon feedstreams can include heavy hydrocarbon feeds boiling in the range of about 430°F to about 1050°F (221 to 566°C), such as gas oils, heavy hydrocarbon oils comprising materials boiling above 1050°F (566°C); 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; and mixtures thereof. The FCC feed may also comprise recycled hydrocarbons, such as light or heavy cycle oils. Preferred heavy hydrocarbon feedstreams for use in the present process are vacuum gas oils boiling in the range above about 650°F (343°C).

[0033] Within this reactor riser, the FCC feedstream, containing at least a portion of the hydroisomerization products is contacted with a catalytic cracking catalyst under cracking conditions thereby resulting in spent catalyst particles containing carbon deposited thereon and a lower boiling product stream. The cracking conditions will typically include: temperatures from about 900 to about 1060°F (482 to 571°C), preferably about 950 to about 1040°F (510 to 560°C); hydrocarbon partial pressures from about 10 to 50 psia (70-345 kPa), preferably from about 20 to 40 psia (140-275 kPa); and a catalyst to feed (wt/wt) ratio from about 3 to 8, preferably about 5 to 6, where the catalyst weight is total weight of the catalyst composite. Steam may be concurrently introduced with the feed into the reaction zone. The steam may comprise up to about 5 wt% of the feed. Preferably, the FCC feed residence time in the reaction zone is less than about 5 seconds, more preferably from about 3 to 5 seconds, and even more preferably from about 2 to 3 seconds.

[0034] Catalysts suitable for use within the FCC reactor herein are fluid cracking catalysts comprising either a large-pore molecular sieve or a mixture of

at least one large-pore molecular sieve catalyst and at least one medium-pore molecular sieve catalyst. Large-pore molecular sieves suitable for use herein can be any molecular sieve catalyst having an average pore diameter greater than 0.7 nm which are typically used to catalytically "crack" hydrocarbon feeds. It is preferred that both the large-pore molecular sieves and the medium-pore molecular sieves used herein be selected from those molecular sieves having a crystalline tetrahedral framework oxide component. Preferably, the crystalline tetrahedral framework oxide component is selected from the group consisting of zeolites, tectosilicates, tetrahedral aluminophosphates (ALPOs) and tetrahedral silicoaluminophosphates (SAPOs). More preferably, the crystalline framework oxide component of both the large-pore and medium-pore catalyst is a zeolite. It should be noted that when the cracking catalyst comprises a mixture of at least one large-pore molecular sieve catalyst and at least one medium-pore molecular sieve, the large-pore component is typically used to catalyze the breakdown of primary products from the catalytic cracking reaction into clean products such as naphtha and distillates for fuels and olefins for chemical feedstocks.

[0035] Large pore molecular sieves that are typically used in commercial FCC process units are also suitable for use herein. FCC units used commercially generally employ conventional cracking catalysts which include large-pore zeolites such as USY or REY. Additional large pore molecular sieves that can be employed in accordance with the present invention include both natural and synthetic large pore zeolites. Non-limiting examples of natural large-pore zeolites include gmelinite, chabazite, dachiardite, clinoptilolite, faujasite, heulandite, analcite, levynite, erionite, sodalite, cancrinite, nepheline, lazurite, scolecite, natrolite, offretite, mesolite, mordenite, brewsterite, and ferrierite. Non-limiting examples of synthetic large pore zeolites are zeolites X, Y, A, L, ZK-4, ZK-5, B, E, F, H, J, M, Q, T, W, Z, alpha and beta, omega, REY and USY zeolites. It is preferred that the large pore molecular sieves used herein be

selected from large pore zeolites. The more preferred large-pore zeolites for use herein are the faujasites, particularly zeolite Y, USY, and REY.

[0036] Medium-pore size molecular sieves that are suitable for use herein include both medium pore zeolites and silicoaluminophosphates (SAPOs). Medium pore zeolites suitable for use in the practice of the present invention are described in "Atlas of Zeolite Structure Types", eds. W. H. Meier and D. H. Olson, Butterworth-Heinemann, Third Edition, 1992, which is hereby incorporated by reference. The medium-pore size zeolites generally have an average pore diameter less than about 0.7 nm, typically from about 0.5 to about 0.7 nm and includes for example, MFI, MFS, MEL, MTW, EUO, MTT, HEU, FER, and TON structure type zeolites (IUPAC Commission of Zeolite Nomenclature). Non-limiting examples of such medium-pore size zeolites, include ZSM-5, ZSM-12, ZSM-22, ZSM-23, ZSM-34, ZSM-35, ZSM-38, ZSM-48, ZSM-50, silicalite, and silicalite 2. The most preferred medium pore zeolite used in the present invention is ZSM-5, which is described in U.S. Pat. Nos. 3,702,886 and 3,770,614. ZSM-11 is described in U.S. Pat. No. 3,709,979; ZSM-12 in U.S. Pat. No. 3,832,449; ZSM-21 and ZSM-38 in U.S. Pat. No. 3,948,758; ZSM-23 in U.S. Pat. No. 4,076,842; and ZSM-35 in U.S. Pat. No. 4,016,245. As mentioned above SAPOs, such as SAPO-11, SAPO-34, SAPO-41, and SAPO-42, which are described in U.S. Pat. No. 4,440,871 can also be used herein. Non-limiting examples of other medium pore molecular sieves that can be used herein are chromosilicates; gallium silicates; iron silicates; aluminum phosphates (ALPO), such as ALPO-11 described in U.S. Pat. No. 4,310,440; titanium aluminosilicates (TASO), such as TASO-45 described in EP-A No. 229,295; boron silicates, described in U.S. Pat. No. 4,254,297; titanium aluminophosphates (TAPO), such as TAPO-11 described in U.S. Pat. No. 4,500,651; and iron aluminosilicates. All of the above patents are incorporated herein by reference.

[0037] The medium-pore size zeolites used herein can also include "crystalline admixtures" which are thought to be the result of faults occurring within the crystal or crystalline area during the synthesis of the zeolites. Examples of crystalline admixtures of ZSM-5 and ZSM-11 are disclosed in U.S. Pat. No. 4,229,424 which is incorporated herein by reference. The crystalline admixtures are themselves medium-pore size zeolites and are not to be confused with physical admixtures of zeolites in which distinct crystals of crystallites of different zeolites are physically present in the same catalyst composite or hydrothermal reaction mixtures.

[0038] The large-pore and medium-pore catalysts of the present invention will typically be present in an inorganic oxide matrix component that binds the catalyst components together so that the catalyst product is hard enough to survive inter-particle and reactor wall collisions. The inorganic oxide matrix can be made from an inorganic oxide sol or gel which is dried to "glue" the catalyst components together. Preferably, the inorganic oxide matrix will be comprised of oxides of silicon and aluminum. It is also preferred that separate alumina phases be incorporated into the inorganic oxide matrix. Species of aluminum oxyhydroxides- γ -alumina, boehmite, diasporite, and transitional aluminas such as α -alumina, β -alumina, γ -alumina, δ -alumina, ϵ -alumina, κ -alumina, and ρ -alumina can be employed. Preferably, the alumina species is an aluminum trihydroxide such as gibbsite, bayerite, nordstrandite, or doyleite. The matrix material may also contain phosphorous or aluminum phosphate. It is within the scope of this invention that the large-pore catalysts and medium-pore catalysts be present in the same or different catalyst particles, in the aforesaid inorganic oxide matrix.

[0039] In the FCC reactor, the cracked FCC product is removed from the fluidized catalyst particles. Preferably this is done with mechanical separation

devices, such as an FCC cyclone. The FCC product is removed from the reactor via an overhead line, cooled and sent to a fractionator tower for separation into various cracked hydrocarbon product streams. These product streams may include, but are not limited to, a light gas stream (generally comprising C₄ and lighter hydrocarbon materials), a naphtha (gasoline) stream, a distillate (diesel and/or jet fuel) stream, and other various heavier gas oil product streams. In the present invention, the gasoline production is increased (maximized) due to the series of pretreatment steps described herein to at least one component stream of the combined FCC feedstream to the FCC reactor.

[0040] In the FCC reactor, after removing most of the cracked FCC product through mechanical means, the majority of, and preferably substantially all of, the spent catalyst particles are conducted to a stripping zone within the FCC reactor. The stripping zone will typically contain a dense bed (or "dense phase") of catalyst particles where stripping of volatiles takes place by use of a stripping agent such as steam. There will also be space above the stripping zone wherein the catalyst density is substantially lower and which space can be referred to as a "dilute phase". This dilute phase can be thought of as either a dilute phase of the reactor or stripper in that it will typically be at the bottom of the reactor leading to the stripper.

[0041] The majority of, and preferably substantially all of, the stripped catalyst particles are subsequently conducted to a regeneration zone wherein the spent catalyst particles are regenerated by burning coke from the spent catalyst particles in the presence of an oxygen containing gas, preferably air thus producing regenerated catalyst particles. This regeneration step restores catalyst activity and simultaneously heats the catalyst to a temperature from about 1200°F to about 1400°F (649 to 760°C). The majority of, and preferably substantially all of, the hot regenerated catalyst particles are then recycled to the FCC reaction zone where they contact injected FCC feed.

[0042] In additional preferred embodiments of the processes of invention, a hydrotreating step is included in the overall process either prior to or following the hydroisomerization step. In the hydrotreating step, a feedstream is contacted with a hydrotreating catalyst under hydrotreating conditions which include temperatures in the range 450°F to 750°F (232°C to 399°C), preferably 550°F to 700°F (288°C to 371°C) at pressures in the range of 1480 to 20786 kPa (200 to 3000 psig), preferably 2859 to 13891 kPa (400 to 2000 psig), a space velocity of from 0.1 to 10 LHSV, preferably 0.1 to 5 LHSV, and a hydrogen treat gas rate of from 18 to 890 m³/m³ (100 to 5000 scf/B), preferably 44 to 178 m³/m³ (250 to 1000 scf/B).

[0043] Hydrotreating catalysts suitable for use herein are those containing at least one Group VIA metal and at least one of a Group VIIIA metal, including mixtures thereof. Preferred metals include Ni, W, Mo, Co and mixtures thereof, with CoMo, NiMoW, or NiW being preferred. These metals or mixtures of metals are typically present as oxides or sulfides on refractory metal oxide supports. The mixture of metals may also be present as bulk metal catalysts wherein the amount of metal is 30 wt% or greater, based on catalyst.

[0044] Suitable metal oxide supports for the hydrotreating catalysts include oxides such as silica, alumina, silica-alumina, titania, or zirconia; preferably alumina. Preferred aluminas are porous aluminas such as gamma or eta. When a porous metal oxide support is utilized, the catalyst has an average pore size (as measured by nitrogen adsorption) of preferably in the range of about 100 Å to about 1000 Å, more preferably from about 200 Å to about 500 Å; and the catalyst has a surface area (as measured by the BET method) of about 100 to 350 m²/g, more preferably about 150 to 250 m²/g. The amount of metals for supported hydrotreating catalysts, either individually or in mixtures, ranges from 0.5 to 35 wt%, based on catalyst. In the case of preferred mixtures of Group VIA and Group VIIIA metals, the Group VIIIA metals are present in amounts of

from 0.5 to 5 wt% based on catalyst, and the Group VIA metals are present in amounts of from 5 to 30 wt% based on the catalyst.

[0045] In an embodiment, the hydrotreating step may comprise a unit separate from the hydroisomerization step, such unit comprising at least one hydrotreating reactor, and in an alternate embodiment comprising two hydrotreating reactors arranged in series flow. Here, a vapor separation drum is preferably oriented after each hydrotreating reactor and removes the vapor phase reaction products from the reactor effluent(s). This vapor phase is primarily comprised of hydrogen, H_2S , NH_3 , and hydrocarbons containing four (4) or less carbon atoms (i.e., "C₄- hydrocarbons"). In the hydrotreating process, preferably at least 70 wt%, more preferably at least 80 wt%, and even more preferably at least 90 wt% of the sulfur content in the feedstream is removed from the resulting liquid products. Additionally, preferably at least 50 wt%, more preferably at least 75 wt%, of the nitrogen content in the feedstream is removed from the resulting liquid products. Preferably, the final liquid product from the hydrotreating unit has less than about 100 ppmw sulfur, more preferably less than about 50 ppmw sulfur, and most preferably, less than about 30 ppmw sulfur. However, as will be described more fully in select embodiments below, the liquid product from the hydrotreating unit may contain over 100 ppmw sulfur or even over 300 ppmw depending on the catalyst and conditions in the isomerization stage of the process as will be described next.

[0046] In embodiments or the processes herein wherein the hydrocarbon feedstream is first hydrotreated, then hydroisomerized, or wherein the hydrocarbon feedstream is first hydroisomerized and then hydrotreated prior to processing in the FCC unit, an intermediate vapor separator drum, such as described above, may optionally be employed. If employed, the vapor separation step would be utilized to remove at least a portion of the sulfur

species and nitrogen species in the feed as gases (e.g., H_2S and NH_3) prior to the treated product from the former stage being processed in the latter stage.

[0047] In an alternate and improved embodiment, at least a portion, preferably all, of the hydrotreating catalyst and at least a portion of the, preferably all, of the hydroisomerization catalyst are located in the same reactor. Here, it is preferred that the hydrotreating catalyst is located in at least one separate catalyst bed within the reactor and that the hydroisomerization catalyst is located in at least one separate catalyst bed within the same reactor. In this embodiment, the hydrocarbon feedstream can first contact, or flow through, the hydrotreating catalyst bed and then contact, or flow through, the hydroisomerization catalyst bed, or visa versa.

[0048] In this single reactor embodiment of the present processes however, it is preferred that the hydrocarbon feedstream first contacts the hydrotreating catalyst bed prior to the hydroisomerization catalyst bed. In these common reactor embodiments, no intermediate vapor removal is required. However, the introduction of a hydrogen-containing stream between the beds may be optionally employed. In alternative embodiments, the hydroisomerization catalyst and the hydrotreating catalyst which are located in the same reactor do not need to be in separate beds but rather the bed(s) can be comprised of a mixture of the hydroisomerization and hydrotreating catalysts. In these single reactor embodiments, it is preferred that the metal hydrogenation component of the hydroisomerization catalyst is a non-noble Group VIIIA metal optionally coupled with at least one Group VIA metal. Suitable combinations of metals in this embodiment of the hydroisomerization catalyst can include Ni, Co, or Fe with Mo or W, preferably Ni with Mo or W.

[0049] In preferred embodiments, a vapor separation drum or a fractionation stage is employed between the hydroisomerization or combination hydrotreating/hydroisomerization steps described and the processing of the

hydrocarbons in the FCC unit. As described prior, a vapor separation drum may be utilized in this step to remove hydrogen, H_2S , NH_3 and/or light gas products. However, in a preferred embodiment of the processes herein, a distillation column (or “fractionator”) is utilized to separate some of the products from the hydroisomerization or combination hydrotreating/hydroisomerization steps prior to sending the remaining hydrocarbon products to the FCC unit for further processing. In a particularly preferred embodiment, the product from the hydroisomerization or combination hydrotreating/hydroisomerization steps is sent to a distillation column for further separation of components. Here at least one overhead vapor stream is removed, at least one distillate product stream is removed, and at least one other distillation product stream is removed from the distillation column. This at least one distillation product stream is then sent to the FCC unit for further processes as noted in the invention herein. In a preferred embodiment, this at least one distillation product stream comprises higher boiling point hydrocarbon fractions than the distillate product stream. In another preferred embodiment, this at least one distillation product stream comprises naphtha range hydrocarbon fractions. In these embodiments employing this “pre-FCC” distillation column, in a preferred embodiment, the products from the FCC process are further fractionated in an FCC fractionator column from which at least one distillate range product stream is drawn from the FCC fractionator and combined with the distillate product stream from the pre-FCC distillation column. In this manner, overall distillate production can be increased.

[0050] Additionally or alternatively, the present invention can be described according to one or more of the following embodiments.

[0051] Embodiment 1. A process for increasing Fluid Catalytic Cracking (“FCC”) gasoline production comprising:

a) contacting a hydrocarbon-containing hydroisomerization feedstream with a hydroisomerization catalyst under hydroisomerization conditions to produce at least one hydroisomerized liquid product stream that has a higher iso-paraffin content than the hydroisomerization feedstream;

b) contacting in the reaction zone of an FCC reactor riser an FCC feedstream comprising at least a portion of the hydroisomerized liquid product stream of step a) with a fluid catalytic cracking catalyst thereby catalytically cracking the FCC feedstream into an FCC product that has an average lower boiling point than the FCC feedstream, and producing a spent catalyst;

c) separating the FCC product from the spent catalyst;

d) cooling the FCC product; and

e) fractionating the FCC product into multiple FCC product streams, wherein at least one of the FCC product streams is a naphtha boiling-range product stream; and

f) utilizing at least a portion of the naphtha boiling-range product stream for gasoline production.

[0052] Embodiment 2. The process of embodiment 1, further comprising:

- contacting a hydrocarbon-containing hydrotreater feedstream containing at least 250 ppmw of sulfur with a hydrotreating catalyst under hydrotreating conditions to produce at least one hydrotreated liquid product stream and one hydrotreated vapor stream, wherein the hydrotreated liquid product stream has a lower sulfur content than the sulfur-containing hydrocarbon feedstream;

- separating the hydrotreated liquid product stream from the hydrotreated vapor stream; and

- utilizing at least a portion of the hydrotreated liquid product stream as the hydroisomerization feedstream in step a).

[0053] Embodiment 3. The process of embodiment 1, further comprising:

- contacting a hydrocarbon-containing hydrotreater feedstream containing at least 250 ppmw of sulfur with a hydrotreating catalyst under hydrotreating conditions to produce the hydroisomerization feedstream.

[0054] Embodiment 4. A process for increasing Fluid Catalytic Cracking (“FCC”) gasoline production comprising:

a) contacting a hydrocarbon-containing hydroisomerization feedstream with a hydroisomerization catalyst under hydroisomerization conditions to produce at least one hydroisomerized product stream that has a higher iso-paraffin content than the hydroisomerization feedstream;

b) contacting at least a portion of the hydroisomerized product stream with a hydrotreating catalyst under hydrotreating conditions to produce at least one hydrotreated liquid product stream and one hydrotreated vapor stream, wherein the hydrotreated liquid product stream has a lower sulfur content than the sulfur-containing hydrocarbon feedstream;

c) separating the hydrotreated liquid product stream from the hydrotreated vapor stream;

d) contacting in the reaction zone of an FCC reactor riser an FCC feedstream comprising at least a portion of the hydrotreated liquid product stream of step c) with a fluid catalytic cracking catalyst thereby catalytically cracking the FCC feedstream into an FCC product that has an average lower boiling point than the FCC feedstream, and producing a spent catalyst;

e) separating the FCC product from the spent catalyst;

f) cooling the FCC product; and

e) fractionating the FCC product into multiple FCC product streams, wherein at least one of the FCC product streams is a naphtha boiling-range product stream; and

g) utilizing at least a portion of the naphtha boiling-range product stream for gasoline production.

[0055] Embodiment 5. The process of any of embodiments 1-4, wherein at least 50 wt% of the normal paraffins in the hydroisomerization feedstream are converted to iso-paraffins in the hydroisomerized liquid product stream in step a).

[0056] Embodiment 6. The process of any of embodiments 1-5, wherein the hydroisomerization catalyst comprises at least one Group VIIIA metal, and further comprises a zeolite selected from EU-1, ZSM-35, ZSM-11, ZSM-57, NU-87, SAPO-11, ZSM-22, and ZSM-48.

[0057] Embodiment 7. The process of embodiment 6, wherein the Group VIIIA metal of the hydroisomerization catalyst is selected from Pt and Pd.

[0058] Embodiment 8. The process of any of embodiments 6-7, wherein the hydroisomerization catalyst further comprises at least one Group VIA metal, wherein the Group VIA of the hydroisomerization catalyst is selected from Mo and W, and the Group VIIIA metal of the hydroisomerization catalyst is selected from Ni and Co.

[0059] Embodiment 9. The process of embodiment 8, wherein the Group VIA of the hydroisomerization catalyst is W, the Group VIIIA metal of the hydroisomerization catalyst is Ni.

[0060] Embodiment 10. The process of any of embodiments 6-9, wherein the zeolite in the hydroisomerization catalyst is ZSM-48.

[0061] Embodiment 11. The process of any of embodiments 1-10, wherein the hydroisomerization feedstream contains over 300 ppmw of sulfur.

[0062] Embodiment 12. The process of any of embodiments 1-11, wherein the hydroisomerization conditions include a temperature of from 400 to 850°F (204 to 454°C), a hydrogen partial pressure of from 1.8 to 34.6 mPa (250 to

5000 psi), a liquid hourly space velocity of from 0.2 to 10 v/v/hr, and a hydrogen circulation rate of from 35.6 to 1781 m³/m³ (200 to 10,000 scf/B).

[0063] Embodiment 13. The process of any of embodiments 1-12, wherein the conditions in the reaction zone of the FCC reactor include a temperature from about 900 to about 1060°F (482 to 571°C), a hydrocarbon partial pressure from about 10 to 50 psia (70-345 kPa), and a catalyst to feed (wt/wt) ratio from about 3 to 8, where the catalyst weight is total weight of the fluid catalytic cracking catalyst.

[0064] Embodiment 14. The process of any of embodiments 1-13, wherein the fluid catalytic cracking catalyst comprises at least one large-pore size faujasite zeolite and at least one medium-pore size zeolite selected from ZSM-5, ZSM-12, ZSM-22, ZSM-23, ZSM-34, ZSM-35, ZSM-38, ZSM-48, ZSM-50, silicalite, and silicalite 2.

[0065] Embodiment 15. The process of any of embodiments 1-14, wherein the FCC feedstream further comprises a heavy hydrocarbon feedstream boiling in the range of about 430°F to about 1050°F (221 to 566°C).

[0066] Embodiment 16. The process of embodiment 15, wherein the heavy hydrocarbon feedstream is comprised of a hydrocarbon stream selected from gas oil, heavy and reduced petroleum crude oil; petroleum atmospheric distillation bottoms, petroleum vacuum distillation bottoms, pitch, asphalt, bitumen, heavy hydrocarbon residues, tar sand oils, shale oil, and liquid products derived from coal liquefaction processes.

[0067] Embodiment 17. The process of any of embodiments 1-16, wherein feed residence time in the reaction zone of the FCC reactor riser is less than about 5 seconds.

[0068] Embodiment 18. The process of any of embodiments 1-3 as further limited by any of embodiments 5-17, wherein the at least one hydroisomerized liquid product stream of step a) is sent to a distillation column of to produce the at least a portion of the hydroisomerized liquid product stream of step b), as well as producing a distillation column overhead vapor stream and at least a first distillate product stream from the distillation column, wherein the distillation column overhead vapor stream and the first distillate product stream are not sent to the reaction zone of the FCC reactor riser.

[0069] Embodiment 19. The process of embodiment 4 as further limited by any of embodiments 5-17, wherein the separation of step c) is performed in a distillation column to produce the hydrotreated liquid product stream and the hydrotreated vapor stream, as well as producing a distillation column overhead vapor stream and at least a first distillate product stream from the distillation column, wherein the distillation column overhead vapor stream and the distillate product stream are not sent to the reaction zone of the FCC reactor riser.

[0070] Embodiment 20. The process of any of embodiments 18-19, wherein at least one of the FCC product streams is an FCC distillate boiling-range product stream and at least a portion of the first distillate product stream is combined with at least a portion of the FCC distillate boiling-range product stream to form a combined distillate product stream.

[0071] Embodiment 21. The process of embodiment 20, wherein at least a portion of the combined distillate product stream is utilized for diesel product blending.

[0072] Embodiment 22. The process of any of embodiments 2 or 4 as further limited by any of embodiments 5-21, wherein the hydrotreated liquid product stream contains less than 30 ppmw of sulfur.

[0073] Embodiment 23. The process of any of embodiments 2-4 as further limited by any of embodiments 5-21, wherein the hydrotreating catalyst comprises at least one Group VIA metal and at least one Group VIIIA metal on a refractory oxide support, wherein the refractory oxide support comprises silica, alumina, or silica-alumina; and the hydroisomerization catalyst is comprised of at least one Group VIIIA metal, and a zeolite selected from EU-1, ZSM-35, ZSM-11, ZSM-57, NU-87, SAPO-11, ZSM-22, and ZSM-48.

[0074] Embodiment 24. The process of embodiment 23, wherein the hydrotreating catalyst has a has an average pore size of from about 100 Å to about 1000 Å, and a surface area of from about 100 to 350 m²/g.

[0075] Embodiment 25. The process of any of embodiments 2-24, wherein the hydrotreating conditions include a temperature in the range 450°F to 750°F (232°C to 399°C), pressure in the range of 1480 to 20786 kPa (200 to 3000 psig), a space velocity of from 0.1 to 10 LHSV, and a hydrogen treat gas rate of from 18 to 890 m³/m³ (100 to 5000 scf/B).

[0076] Embodiment 26. The process of any of embodiments 2 or 3, as further limited by any of embodiments 5-25, wherein the hydrotreater feedstream has a T5 boiling point of at least 400°F and a T95 boiling point of less than about 1150°F.

[0077] Embodiment 27. The process of embodiment 26, wherein the hydrotreater feedstream is comprised of at least 75 wt% of a hydrocarbon feedstream derived from a fossil-based oil material, and is further comprised of from 5 to 25 wt% of oil derived from renewable biofuel sources.

[0078] Embodiment 28. The process of any of embodiments 3 or 4, as further limited by any of embodiments 5-27, wherein the hydrotreating catalyst and the hydroisomerization catalyst are in a single reactor.

[0079] The principles and modes of operation of this invention have been described above with reference to various exemplary and preferred embodiments. As understood by those of skill in the art, the overall invention, as defined by the claims, encompasses other preferred embodiments not specifically enumerated herein.

EXAMPLE

[0080] In the Example herein, an FCC kinetic research model was utilized to test the effects of isomerizing the normal paraffins in a typical heavy hydrocarbon FCC feedstream composition. This model represents and models the effects of converting all of the normal paraffins to isoparaffins via a hydroisomerization catalytic process and then catalytically cracking the resulting hydroisomerized hydrocarbon material in a fluid catalytic cracking (FCC) process.

[0081] The feed compositions for the non-hydroisomerized FCC feed ("Base Case") and the hydroisomerized FCC feed ("Isomerized Case") of the present invention are shown in Table 1.

Table 1
FCC Feed Compositions

FCC Feed Composition	Base Case	Isomerized Case
Specific Gravity	0.905	0.906
IBP, (°C)	247	247
FBP, (°C)	608	608
Molecular Weight	384	387
Sulfur, (ppm)	380	380
Nitrogen, (ppm)	954	654
N-paraffins, (vol%)	5.8	0
I-paraffins, (vol%)	12.6	18.4
Naphthenes, (vol%)	31.8	31.8
Aromatics , (vol%)	49.8	49.8

[0082] Table 2 shows a comparison of the predicted FCC product compositions utilizing the Base Case and the Isomerized Case FCC feed compositions shown in Table 1.

Table 2
FCC Product Compositions

FCC Product Composition, (vol%)	Base Case	Isomerized Case
C2 and lighter	1.4	1.3
C3	1.5	1.6
C3=	6.8	6.9
C4	6.8	7.0
C4=	9.1	9.4
LCN (light gasoline) (123°C FBP)	31.0	31.8
HCN (heavy gasoline) (167°C FBP)	16.6	16.8
LCO (255°C FBP)	19.7	19.8
HCO (397°C FBP)	16.2	15.4
Cat Bottoms	6.7	6.0

TABLE 2 NOTES:

- 1) "C3=" represents C3 olefins ; "C4=" represents C4 olefins; "LCN" = light cat naphtha, "HCN" = heavy cat naphtha, "LCO" = light cycle oil, and "HCO" = heavy cycle oil.
- 2) Error range in calculated values +/- 0.05% (absolute).
- 3) Total values reflect an increase in volumetric yield of the product from the FCC.

[0083] As can be seen in the product data above, the process of invention unexpectedly resulted in a 2.5% increase in desired FCC light gasoline production (31.0 vol% to 31.8 vol%) or a 2.1% increase in FCC overall gasoline production (i.e., LCN + HCN, 47.6 vol% to 48.6 vol%). As noted prior, in a typical refinery, this can translate into an FCC gasoline production increase of over 40,000 gallons per day in a refinery with a typical size FCC unit. Just as unexpected, and very economically beneficial, is the fact that almost all of the increase in the light gasoline production (most desired product) is offset by an almost corresponding decrease in the cat bottoms production (least desired product). This increases the overall refinery economics of practicing this process configuration.

CLAIMS:

1. A process for increasing Fluid Catalytic Cracking ("FCC") gasoline production comprising:
 - a) contacting a hydrocarbon-containing hydroisomerization feedstream with a hydroisomerization catalyst under hydroisomerization conditions to produce at least one hydroisomerized liquid product stream that has a higher iso-paraffin content than the hydroisomerization feedstream;
 - b) contacting in the reaction zone of an FCC reactor riser an FCC feedstream comprising at least a portion of the hydroisomerized liquid product stream of step a) with a fluid catalytic cracking catalyst thereby catalytically cracking the FCC feedstream into an FCC product that has an average lower boiling point than the FCC feedstream, and producing a spent catalyst;
 - c) separating the FCC product from the spent catalyst;
 - d) cooling the FCC product; and
 - e) fractionating the FCC product into multiple FCC product streams, wherein at least one of the FCC product streams is a naphtha boiling-range product stream; and
 - f) utilizing at least a portion of the naphtha boiling-range product stream for gasoline production.
2. The process of claim 1, further comprising:
 - contacting a hydrocarbon-containing hydrotreater feedstream containing at least 250 ppmw of sulfur with a hydrotreating catalyst under hydrotreating conditions to produce at least one hydrotreated liquid product stream and one hydrotreated vapor stream, wherein the hydrotreated liquid product stream has a

lower sulfur content than the sulfur-containing hydrocarbon feedstream;

- separating the hydrotreated liquid product stream from the hydrotreated vapor stream; and
- utilizing at least a portion of the hydrotreated liquid product stream as the hydroisomerization feedstream in step a).

3. The process of claim 1, further comprising:

- contacting a hydrocarbon-containing hydrotreater feedstream containing at least 250 ppmw of sulfur with a hydrotreating catalyst under hydrotreating conditions to produce the hydroisomerization feedstream.

4. The process of claim 2, wherein the hydrotreated liquid product stream contains less than 30 ppmw of sulfur.

5. The process of claim 3, wherein the hydrotreating catalyst and the hydroisomerization catalyst are in a single reactor.

6. The process of any one of claims 2-5, wherein the hydrotreating catalyst comprises at least one Group VIA metal and at least one Group VIIIA metal on a refractory oxide support, wherein the refractory oxide support comprises silica, alumina, or silica-alumina; and the hydroisomerization catalyst is comprised of at least one Group VIIIA metal, and a zeolite selected from EU-1, ZSM-35, ZSM-11, ZSM-57, NU-87, SAPO-11, ZSM-22, and ZSM-48.

7. The process of claim 6, wherein the hydrotreating catalyst has a has an average pore size of from about 100 Å to about 1000 Å, and a surface area of from about 100 to 350 m²/g.

8. The process of any one of claims 2-7, wherein the hydrotreating conditions include a temperature in the range 450°F to 750°F (232°C to 399°C), pressure in the range of 1480 to 20786 kPa (200 to 3000 psig), a space velocity of from 0.1 to 10 LHSV, and a hydrogen treat gas rate of from 18 to 890 m³/m³ (100 to 5000 scf/B).

9. The process of any one of claims 2-8, wherein the hydrotreater feedstream has a T5 boiling point of at least 400°F and a T95 boiling point of less than about 1150°F.

10. The process of claim 9, wherein the hydrotreater feedstream is comprised of at least 75 wt% of a hydrocarbon feedstream derived from a fossil-based oil material, and is further comprised of from 5 to 25 wt% of oil derived from renewable biofuel sources.

11. The process of any prior claim, wherein the at least one hydroisomerized liquid product stream of step a) is sent to a distillation column of to produce the at least a portion of the hydroisomerized liquid product stream of step b), as well as producing a distillation column overhead vapor stream and at least a first distillate product stream from the distillation column, wherein the distillation column overhead vapor stream and the first distillate product stream are not sent to the reaction zone of the FCC reactor riser.

12. The process of claim 11, wherein at least one of the FCC product streams is an FCC distillate boiling-range product stream and at least a portion of the first distillate product stream is combined with at least a portion of the FCC distillate boiling-range product stream to form a combined distillate product stream; and wherein at least a portion of the combined distillate product stream is utilized for diesel product blending.

13. The process of any prior claim, wherein at least 50 wt% of the normal paraffins in the hydroisomerization feedstream are converted to iso-paraffins in the hydroisomerized liquid product stream in step a).

14. The process of any prior claim, wherein the hydroisomerization catalyst comprises at least one Group VIIIA metal, and further comprises a zeolite selected from EU-1, ZSM-35, ZSM-11, ZSM-57, NU-87, SAPO-11, ZSM-22, and ZSM-48.

15. The process of claim 14, wherein the Group VIIIA metal of the hydroisomerization catalyst is selected from Pt and Pd.

16. The process of any one of claims 14-15, wherein the hydroisomerization catalyst further comprises at least one Group VIA metal, wherein the Group VIA of the hydroisomerization catalyst is selected from Mo and W, and the Group VIIIA metal of the hydroisomerization catalyst is selected from Ni and Co.

17. The process of claim 16, wherein the Group VIA of the hydroisomerization catalyst is W, the Group VIIIA metal of the hydroisomerization catalyst is Ni.

18. The process of any one of claims 14-17, wherein the zeolite in the hydroisomerization catalyst is ZSM-48.

19. The process of any prior claim, wherein the hydroisomerization feedstream contains over 300 ppmw of sulfur.

20. The process of any prior claim, wherein the hydroisomerization conditions include a temperature of from 400 to 850°F (204 to 454°C), a hydrogen partial pressure of from 1.8 to 34.6 mPa (250 to 5000 psi), a liquid

hourly space velocity of from 0.2 to 10 v/v/hr, and a hydrogen circulation rate of from 35.6 to 1781 m³/m³ (200 to 10,000 scf/B).

21. The process of any prior claim, wherein the conditions in the reaction zone of the FCC reactor include a temperature from about 900 to about 1060°F (482 to 571°C), a hydrocarbon partial pressure from about 10 to 50 psia (70-345 kPa), and a catalyst to feed (wt/wt) ratio from about 3 to 8, where the catalyst weight is total weight of the fluid catalytic cracking catalyst.

22. The process of any prior claim, wherein the fluid catalytic cracking catalyst comprises at least one large-pore size faujasite zeolite and at least one medium-pore size zeolite selected from ZSM-5, ZSM-12, ZSM-22, ZSM-23, ZSM-34, ZSM-35, ZSM-38, ZSM-48, ZSM-50, silicalite, and silicalite 2.

23. The process of any prior claim, wherein the FCC feedstream further comprises a heavy hydrocarbon feedstream boiling in the range of about 430°F to about 1050°F (221 to 566°C).

24. The process of embodiment 23, wherein the heavy hydrocarbon feedstream is comprised of a hydrocarbon stream selected from gas oil, heavy and reduced petroleum crude oil; petroleum atmospheric distillation bottoms, petroleum vacuum distillation bottoms, pitch, asphalt, bitumen, heavy hydrocarbon residues, tar sand oils, shale oil, and liquid products derived from coal liquefaction processes.

25. The process of any prior claim, wherein feed residence time in the reaction zone of the FCC reactor riser is less than about 5 seconds.