

- [54] **HYDROTREATING AND HYDROISOMERIZING C₅ AND C₆ HYDROCARBON STREAMS**
- [75] Inventor: **Robert E. Quisenberry**, Port Arthur, Tex.
- [73] Assignee: **Texaco, Inc.**, New York, N.Y.
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Primary Examiner—Delbert E. Gantz
 Assistant Examiner—G. J. Crasanakis
 Attorney—Thomas H. Whaley et al.

[57] **ABSTRACT**

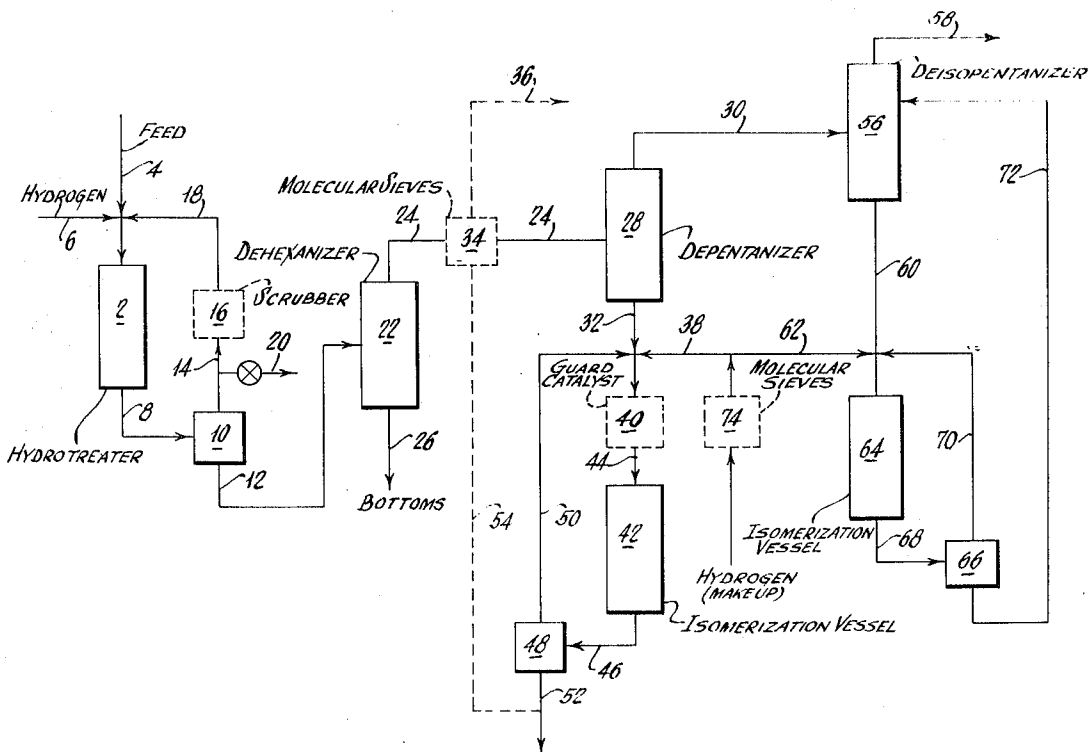
A process is provided for upgrading a light naphtha fraction containing impurities deleterious to isomerization catalyst life and major amount of C₅ and C₆ hydrocarbons into plurality of relatively high octane hydrocarbon streams by hydrotreating the naphtha fraction and recovering a C₅–C₆ isomerizable hydrocarbon stream, separating the C₅ hydrocarbons from the C₆ hydrocarbons and introducing into separate reaction zones the C₅ and C₆ hydrocarbons respectively along with hydrogen under isomerization conversion conditions allowing consistent production of high octane blending components, separating and recovering from each conversion zone the iso C₅ and iso C₆ products and recycling for introduction to the respective isomerization reaction zones unconverted normal C₅ and C₆ components.

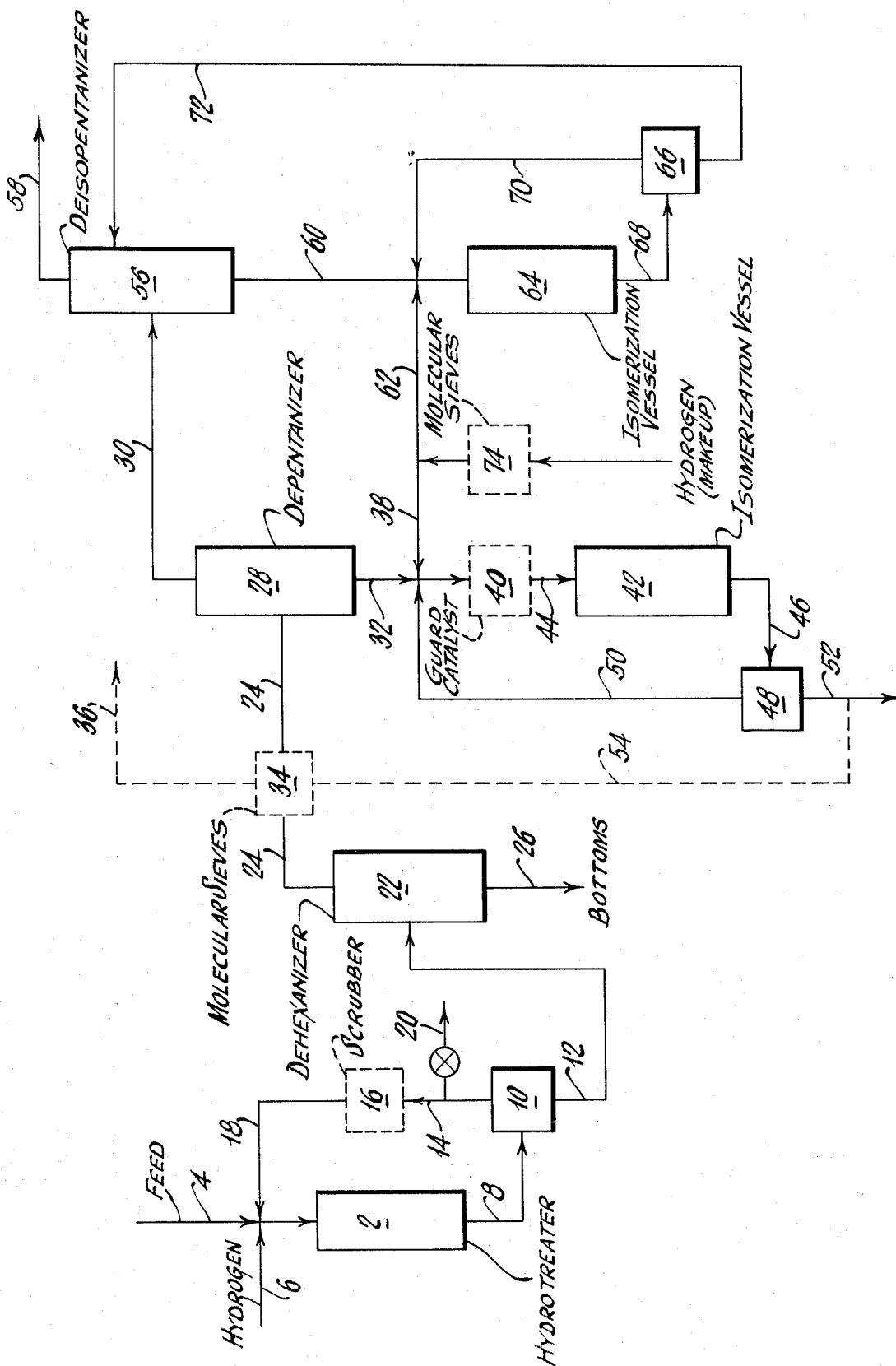
12 Claims, 1 Drawing Figure

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HYDROTREATING AND HYDROISOMERIZING C AND C₆ HYDROCARBON STREAMS

BACKGROUND OF THE INVENTION

This invention relates to a process for the upgrading of hydrocarbon streams. More particularly this invention relates to a process for upgrading hydrocarbon streams such as light naphtha fractions and fractions containing major amounts of C₅ and C₆ hydrocarbons under processing conditions providing consistent production of higher octane blending components.

The isomerization of isomerizable hydrocarbons represents a refining process developed especially for the upgrading of petroleum fractions. By this process the art is enabled to adjust the octane number upwards by converting normal paraffins such as normal pentane and normal hexane to isoparaffins such as 2-methylbutane and 2,2-dimethylbutane. The conversion of normal paraffins to their corresponding isomeric forms provides a pool of gasoline possessing a higher octane number than a gasoline consisting of normal paraffins. Presently, low octane quality hydrocarbons such as light straight run naphtha or other streams containing substantial amounts of C₅ and C₆ hydrocarbons are blended with other high octane components and are marketed as regular gasoline. Recent governmental regulations and potential legislation in the field of emissions from the internal combustion engine limit the use of low octane blending hydrocarbons by requiring reduced or eliminated levels of tetraethyl lead in gasolines. Consequentially the normal hydrocarbons require upgrading before they can be blended with other high octane components to yield a satisfactory gasoline.

Heretofore, isomerization of isomerizable hydrocarbons was performed by passing the hydrocarbon together with hydrogen through a reaction zone containing an isomerization catalyst. The conversion conditions employed while depending to some extent upon specific hydrocarbons being isomerized were, nevertheless, conducted over a single catalyst where the conditions chosen were sufficient to provide the best yield of isomers from a given charge. Operating in such a matter however, provided far less than the optimum yield of high octane blending materials and most importantly a commercially unattractive process.

It is therefore, an object of this invention to provide a process for the upgrading of hydrocarbons to high octane gasoline blending components.

It is another object of this invention to provide a process for upgrading hydrocarbon streams containing a major amount of C₅ and C₆ hydrocarbons which furnishes increased yields and high selectivity towards high octane number blending components.

Yet another object of this invention is to provide an improved process for isomerizing C₅ and C₆ isomerizable hydrocarbons under conversion conditions providing optimum yields of high octane gasoline blending components.

Other objects and advantages will be apparent from a reading of the following detailed description and examples.

SUMMARY OF THE INVENTION

Broadly, this invention contemplates a process for isomerizing isomerizable hydrocarbons, particularly

within the C₅—C₆ range, containing an impurity selected from the group consisting of oxygen compounds, sulfur, sulfur compounds, nitrogen compounds and unsaturated hydrocarbons which comprises:

a. hydrotreating said isomerizable hydrocarbon in the presence of a hydrogenation catalyst resistant to said impurity under hydrotreating conditions and recovering a C₅—C₆ isomerizable hydrocarbon stream;

b. separating said C₅—C₆ isomerizable hydrocarbon stream into a C₅ isomerizable hydrocarbon stream and a C₆ isomerizable hydrocarbon stream;

c. introducing hydrogen and said C₆ isomerizable hydrocarbon stream to an isomerization reactor containing an isomerization catalyst, said reactor maintained at a temperature of about 200° to 400° F.;

d. separating and recovering isohexane from the product of (c) and recycling the remaining product to step (c);

e. introducing hydrogen and said C₅ isomerizable hydrocarbon stream to an isomerization reactor containing an isomerization catalyst, said reactor maintained at a temperature of about 300° to 700° F.; and

f. separating and recovering from the product of (e) isopentane and recycling the remaining product to step (e).

In a particularly desired embodiment, the process of this invention contemplates the upgrading of a light naphtha fraction containing a major amount of C₅ and C₆ hydrocarbons which are converted into a plurality of relatively high octane hydrocarbon streams. The light naphtha fraction may comprise a debutanized naphtha fraction boiling in the range of 90° to 260°F. Distinguishing features of this invention include individual isomerization of separate C₅ and C₆ isomerizable hydrocarbon streams and the recovery of an isopentane rich stream and an isohexane rich stream.

Referring to the drawing, the present invention can be more readily understood by reference thereto.

A feed containing a major amount of C₅ and C₆ isomerizable hydrocarbons, suitably a debutanized light straight run naphtha boiling in the range of 90° to 260° F. containing such impurities as oxygen, sulfur, nitrogen, and unsaturated compounds, is fed to hydrotreater 2 through line 4 along with hydrogen in line 6. Within hydrotreater 2 oxygen compounds in the feed entering through line 4 are converted to water, sulfur or sulfur compounds are converted to H₂S, any nitrogen compounds are converted to ammonia and any unsaturated compounds present are saturated. The so-treated hydrocarbons along with hydrogen and converted impurities such as ammonia, H₂S, and water are withdrawn in line 8 and enter separator 10 where the gaseous effluent is separated from the liquid effluent. The liquid effluent is discharged from separator 10 through line 12. The gaseous effluent is discharged from separator 10 into line 14 and may be passed through a scrubber represented by dotted lines 16 and thence introduced to line 6 via line 18. A bleed line 20 is provided between separator 10 and scrubber 16 for bleeding the gas in line 14 should the concentration of impurities reach high levels. At least a portion of the gaseous effluent from separator 10 is reintroduced to hydrotreater 2 via line 18 and 4. The catalyst employed in hydrotreater 2 is one which is resistant to the degradative effects of any impurity the feed to the

hydrotreater from line 4. Suitable catalysts for this purpose are represented by known commercial materials such as nickel-molybdenum on alumina, cobalt-molybdenum on alumina and nickel sulfide-tungsten sulfide on alumina.

The liquid effluent containing isomerizable hydrocarbons discharged into line 12 from separator 10 are introduced into dehexanizer 22. A stripper tower, not shown, positioned between separator 10 and dehexanizer 22 is beneficially employed to eliminate light hydrocarbons and H_2S formed in hydrotreating and carried along with the liquid effluent in line 22. In dehexanizer 22 an overhead chiefly of pentanes and hexanes is taken through line 24 and a stream of C_7 and heavier bottoms is withdrawn through line 26 and may be routed to a catalytic reforming unit not shown. The dehexanizer overhead in line 24 is routed to depentanizer 28 where pentanes are taken overhead through line 30 and hexanes are removed as a bottom stream through line 32. The depentanizer bottoms in line 32 are preferably dried prior to routing to isomerization vessel 42 by passing the same through a molecular sieve bed, not shown, containing a 4A or 13X sieve wherein the hexane stream in line 32 is purified by the retention of water.

In a particularly desirable embodiment line 24 is provided with an adsorption zone 34 containing a selective adsorbent such as Linde type 5A molecular sieve which selectively adsorbs straight chain hydrocarbons to the substantial exclusion of non-straight chain hydrocarbons, naphthenes and aromatics. The selective adsorption unit or adsorption zone 34 comprises a fixed bed of selective adsorbent maintained at a suitable temperature, such as a temperature in the range of ambient to $800^\circ F.$, preferably $400^\circ F.$ to $500^\circ F.$, to effect selective adsorption of the straight chain hydrocarbons from the remaining dehexanizer effluent introduced thereinto. The adsorption operation is carried out at any convenient pressure, such as a pressure in the range of 0 to 500 p.s.i.g. and higher, preferably 10 to 100 p.s.i.g. Desirably the temperature and pressure within adsorption zone 34 is adjusted so that the effluent undergoing separation therein is maintained in the vapor phase wherein branch chained C_5 and C_6 paraffins along with naphthenes and aromatics are recovered through line 36.

The resulting adsorbent after being substantially saturated with n-pentane and n-hexane is subsequently treated to desorb the normal paraffins therefrom. The desorption of the n- C_5 and C_6 paraffins from the solid adsorbent material may be carried out at any suitable temperature and pressure. For example, the operation may be carried out at a pressure in the range of 0 to 500 p.s.i.g. and at temperature in the range ambient to $800^\circ F.$ It is generally preferred to carry out the operation at a pressure from 10 to 100 p.s.i.g. and a temperature from $400^\circ F.$ to $500^\circ F.$ particularly in the instance wherein a material such as Linde type 5A molecular sieve is employed as the selective adsorbent inasmuch as relatively high temperatures lead to the destruction of adsorbent material. In practice a plurality of adsorption zones 34 may be employed such that adsorptive separation of the stream from line 24 may take place while the other adsorptive zone is undergoing a process of desorption thereby providing a process capable of

continuous and uninterrupted operation. In practice adsorption zone 34 therefore provides a recoverable stream of pentane and hexane isomers and, when present, naphthenes and aromatics through line 36 while simultaneously introducing to depentanizer 28 a stream consisting chiefly of n-pentane and n-hexane.

Into the hexane stream in line 32 there is introduced dry make-up hydrogen suitably from a catalytic reforming unit introduced via line 38. The mixture of hydrogen and hexane pass through line 32 and preferably through guard catalyst bed 40 represented by dotted lines wherein any traces of sulfur present in the hexane feed will be removed and any benzene hydrogenated to cyclohexane and thence to isomerization reactor 42 via line 44. The effluent from isomerization vessel 42 is withdrawn in line 46 and thence passes to high pressure separator 48 wherein hydrogen is separated from the liquid isomerized hydrocarbon effluent through line 50 and re-enters isomerization reactor 42 through lines 32 and 44 joining with make-up hydrogen introduced through line 38 such that the hydrogen to hydrocarbon mole ratio is between 0.05:1 and 5:1, preferably between 0.5:1 and 2:1. The hydrogen and hexane passes to the isomerization vessel 42 maintained under suitable conditions of temperature and pressure for isomerization of hexane. The conversion conditions within reactor 42 include a temperature of from $200^\circ F.$ to $400^\circ F.$, preferably between $280^\circ F.$ to $330^\circ F.$, under hydrogen pressures of from 300 to 750 p.s.i.g. and at liquid hourly space velocities of from 0.5 to 10.0, preferably from 0.75 to 4.0. The hexane stream introduced through line 44 passes over an isomerization catalyst consisting of chlorided platinum-alumina composite activated and stabilized for example in accordance with the processes described in U. S. Pat. No. 3,551,516 and U. S. Pat. No. 3,242,228. The isomerized hydrocarbons discharging from separator 48 pass through line 52 wherein a normal C_6 fraction is separated from an isomerized hexane fraction composed of high octane blending components in accordance with well known procedures. In a preferred embodiment the isomerized hydrocarbons discharging from separator 48 are recycled for reintroduction through adsorption zone 34 via line 54 wherein isohexane, i.e., dimethylbutanes and methylpentanes, is separated and recovered from the isomerized product and the remaining product comprising n-hexane is recycled for reintroduction to isomerization reactor 42 via line 24, depentanizer 28 and lines 32 and 44.

The depentanizer overhead in line 30 is routed to deisopentanizer 56 wherein isopentane is taken overhead through line 58 and routed to gasoline blending. Deisopentanizer bottoms, that is, n-pentane discharging through line 60 is mixed with hydrogen from line 62 and routed to isomerization reactor 64. The effluent from isomerization reactor 64 is routed to separator 66 through line 68 where hydrogen is removed overhead through line 70 for reintroduction to line 60 and thence to isomerization reactor 64 joining with the make-up hydrogen introduced through line 62 such that the hydrogen to hydrocarbon mole ratio is between 0.5:1 and 5:1, preferably between 1:1 and 2:1. Hydrogen and pentane pass through the isomerization vessel maintained under suitable conditions of pressure and temperature for the isomerization of pentanes. The conver-

sion conditions within reactor 64 include a temperature of from 300° to 700° F. and preferably from 600° to 700° F. under hydrogen pressures of from 200 to 500 p.s.i.g. and liquid hourly space velocities of from 0.5 to 4, preferably from 1 to 2. The pentane stream introduced through line 60 passes over an isomerization catalyst suitably one of the type described in connection with isomerization reactor 42 above. The deisopentanizer bottoms in line 60 can be suitably dried in the manner described in connection with the bottoms in line 32. In a highly preferred embodiment the isomerization catalyst consists of a palladium-zeolite Y composite prepared in accordance with the process described in U. S. Pat. No. 3,236,761 and isomerization is permitted to take place at a temperature of from 600° to 700° F. The hydrocarbon effluent in separator 66 is separated via line 72 and introduced to deisopentanizer 56 wherein isopentane is taken overhead and unconverted n-pentane is recycled to isomerization reactor 64 through line 60.

It is seen from the foregoing description of this invention that a highly useful and commercially feasible design for the separate isomerization of C₅ and C₆ hydrocarbons is provided. The invention provides a process in which the isomerization catalyst is kept relatively free of impurities which tend to deactivate the same. The make-up hydrogen suitably introduced to the isomerization reactor through lines 38 and 62 may be purified by passing the same through a molecular sieve bed 74 containing a 4A or 13X sieve wherein the hydrogen is purified by the retention of water, H₂S or any ammonia therein. Make-up hydrogen in line 38 is preferably passed through bed 74.

In order to more fully illustrate the nature of this invention and the manner of practicing the same the following example is presented.

EXAMPLE

A debutanized light straight run naphtha boiling in the temperature range of about 90° to 260° F. and containing about 250 ppm sulfur is fed through line 4 of the accompanying flow diagram into hydrotreater 2. The naphtha contains about three to four weight percent benzene and about nine to ten weight percent C₈ naphthenes along with about 35 to 40 weight percent of C₇ and heavier materials and a minor amount of unsaturated hydrocarbons. The temperature of the naphtha in hydrotreater 2 is about 700° F. and the pressure upon the reaction mixture is about 450 p.s.i.g. where the charge passes through hydrotreater 2 at 4.0 LHSV at a hydrogen to hydrocarbon mole ratio of 0.6. Hydrogen and other gaseous materials containing impurities in the form of water, H₂S, and NH₃ are separated from the liquid hydrocarbon in separator 10 which maintains a pressure on the hydrotreater effluent of about 450 p.s.i.g. and passes through a conventional scrubber 16 which absorbs a substantial amount of the impurities. The gaseous effluent from scrubber 16 continues through line 18 and is recycled to hydrotreater 2 via line 4. The catalyst in hydrotreater 2 is a nickel-molybdenum on alumina catalyst and represents a commercially available material.

The hydrotreated effluent discharging from separator 10 via line 12 is introduced to dehexanizer 22 where an overhead fraction consisting chiefly of pentanes and

hexanes is taken through line 24 at a temperature of between 90° and 200° F. The fraction contains about 3 to 4 percent by weight aromatics and about 20 percent by weight naphthenes, principally cyclohexane and methylcyclopentane. A stream of C₇ and heavier bottoms is withdrawn through line 26 from dehexanizer 22 and is routed to a catalytic reforming unit not shown.

The dehexanizer overhead in line 24 enters adsorption zone 34 containing a bed of 5A molecular sieve. The liquid hourly space velocity of the hydrocarbons through bed 34 is between 0.1 and 10 depending on the normal paraffin content of the feed, sieve capacity and cycle time, and the temperature of the hydrocarbon feed is about 450° F. A stream of isopentane and isohexanes composed of chiefly 2-methylbutane, 2,2-dimethylbutane, 2,3-dimethylbutane, 2-methylpentane and 3-methylpentane, along with naphthenes and aromatics is recovered from bed 34 through line 36 wherein they are pumped to gasoline blending facilities not shown. In practice adsorption zone 34 is composed of a plurality of molecular sieve beds where adsorption of n-paraffins takes place in one bed while desorption of another bed takes place. At least two adsorption beds are provided, one being on-stream absorbing normal paraffins while the other is off-stream being desorbed. It is generally preferred to carry out the adsorption at a pressure from 10 to 100 p.s.i.g. and at a temperature of 400° to 500° F. The selective adsorbent employed is any crystalline aluminosilicate capable of preferentially adsorbing normal paraffins from a mixture of normal and isoparaffins; for example, the selective adsorbent may be 5A molecular sieve. The adsorbing cycle time may be from 5 to 120 minutes, preferably 5 to 30 minutes. At the end of the adsorbing cycle the molecular sieve adsorbent bed is taken off-line and desorbed. The desorbing conditions are preferably the same as the absorbing conditions, i.e., a pressure of 10 to 100 p.s.i.g. and a temperature of 400° to 500° F. The desorbing agent used is normal butane and the agent's liquid hourly space velocity, based on the volume of molecular sieve, is 0.1 to 10, preferably 1 to 2. After cooling and condensing the normal butane and the desorbed normal pentane and normal hexane, the normal butane can be separated by conventional fractionation techniques. Desorbed n-paraffin composed of chiefly n-pentane and n-hexane is thereafter routed to depentanizer 28 through line 24 wherein pentanes are taken overhead through line 30 and hexanes are removed as a bottom stream through line 32. Make-up hydrogen from a catalytic reforming unit enters the process via line 38 and is independently purified by passing through a 4A sieve in bed 74 and is joined by recycle hydrogen in line 50. Hydrogen together with the hexane fraction pass together through a guard catalyst bed 40 containing a nickel refractory supported hydrogenation catalyst wherein traces of sulfur present in the hexane are removed and any benzene hydrogenated to cyclohexane. This hydrogenation catalyst bed is suitably maintained at a temperature of 300° to 400° F., a pressure from 300 to 750 p.s.i.g. and an LHSV of 1 to 16. In operation there is employed a temperature of 300° F., a pressure of 500 p.s.i.g., a hydrogen to hydrocarbon mole ratio of 1.25 and an LHSV of 16. The effluent from guard catalyst bed 40 passes through line 44 to isomerization reactor 42

which contains a chlorided platinum-alumina composite catalyst having a platinum content of 0.5 weight percent and chlorine content of 8 weight percent. The temperature of reactor 42 is maintained at 300° F., a pressure of 500 p.s.i.g., a liquid hourly space velocity of 1 and hydrogen to hydrocarbon mole ratio 1.25. The hydrocarbon isomerization process effluent is discharged into line 46 and enters high pressure separator 48 where relatively pure hydrogen is separated from the isomerized hydrocarbon and leaves the high pressure separator 48 in line 50 for recycle to the isomerization process via lines 32 and 44. The isomerized hydrocarbons are discharged from separator 48 into line 52 and may undergo a distillation step in fractionating towers to separate the hydrocarbons into the fractions designated. Preferably the isomerized hydrocarbons discharged into line 52 are recycled to adsorption zone 34 via line 54 wherein hexane isomers are separated through line 36 and normal hexane recycled to isomerization reactor 42 via line 24, depentanizer 28 and lines 32 and 44.

The depentanizer overhead in line 30 is taken overhead from depentanizer 28 and introduced to deisopentanizer 56 wherein an isopentane fraction is taken overhead in line 58 and an n-pentane fraction is removed as bottoms from the deisopentanizer 56 through line 60 and introduced to isomerization reactor 64. Make-up hydrogen from a catalytic reforming unit enters the process via line 62, and if desired is independently purified by passing through a 4A sieve in bed 74. The liquid hourly space velocity of the n-pentane charge through isomerization vessel 64 is 2.0, and conversion is undertaken in isomerization vessel 64 at 600° F. at a pressure of 500 p.s.i.g. and a hydrogen to hydrocarbon mole ratio of 1 to 2. The hydrocarbon isomerization process effluent is discharged in line 68 and enters high pressure separator 66 where relatively pure hydrogen is separated from the isomerized hydrocarbon and leaves separator 66 for recycle to the isomerization process via lines 70 and 60. The hydrogen to hydrocarbon mole ratio in isomerization vessel 64 is between 0.5 and 5.0 and preferably 1 to 2. Isomerization vessel 64 contains a 0.5 weight percent palladium-zeolite Y composite. Alternatively, the isomerization catalyst may be one described in connection with isomerization vessel 42 and in this instance make-up hydrogen is introduced isomerization vessel 64 via bed 74 and line 62. The C₅ isomerized fraction discharging from separator 66 enters line 72 at a temperature of 85° F. and is reintroduced to deisopentanizer 56 where the isopentane fraction is taken overhead and n-pentane is recycled to isomerization reactor 64 through line 60.

It is thus seen that a significantly useful process design for the isomerization of isomerizable hydrocarbons, particularly those within the C₅ and C₆ range, has been provided by the instant invention. The process enables isomerization of normal paraffins over a long period of time by eliminating from the feed to the isomerization vessel substantial quantities of impurities which tend to deactivate the catalyst. The flexibility of the process affords separate control over the hydrotreating and each isomerization section of the process design and therefore allows for a greater conversion and selectivity over the process. In an impor-

tant embodiment the process permits the advantageous use of separate isomerization reactors each maintained at optimum conditions for the isomerization of C₅ and C₆ streams respectively along with the use of isomerization reactor effluent hydrogen as recycle. As pointed out above and in actual practice of this invention, a pair of molecular sieve beds are used for each bed described herein. As will be evident the operation proceeds continuously such that when one of the beds is used in the line the other is being regenerated. For this purpose gas is used to desorb the sieve bed. Hydrogen can be used to desorb the molecular sieves in vessel 74 and, when used, in the drying of the hexane stream. Normal butane is the preferred desorbant for use in zone 34. Suitably this gas is hydrogen which can be recycled hydrogen from isomerization vessels 42 or 64. When one bed is regenerated it is replaced in the process by the other bed and is in turn regenerated.

I claim:

1. A process for isomerizing a hydrocarbon fraction containing C₅-C₆ isomerizable hydrocarbons, said fraction containing an impurity selected from the group consisting of oxygen compounds, sulfur and sulfur compounds, nitrogen compounds and unsaturated hydrocarbons, said process comprising:

- hydrotreating said hydrocarbon fraction in the presence of a hydrogenation catalyst resistant to said impurity under hydrotreating conditions and separating from said fraction a C₅ - C₆ isomerizable hydrocarbon stream;
- separating said C₅ - C₆ isomerizable hydrocarbon stream into a C₅ isomerizable hydrocarbon stream and a C₆ isomerizable hydrocarbon stream;
- introducing hydrogen and said C₆ isomerizable hydrocarbon stream to a first isomerization reactor containing an isomerization catalyst, said reactor maintained at a temperature of about 200° to 400° F.;
- separating and recovering isohexane from the reaction effluent of step (c) and recycling the remaining effluent to step (c);
- introducing hydrogen and said C₅ isomerizable hydrocarbon stream to a second isomerization reactor containing an isomerization catalyst, said reactor maintained at temperature of about 300° to 700° F.;
- separating and recovering isopentane from the reaction effluent of step (e) and recycling the remaining effluent to step (e).

2. A process according to claim 1 wherein said hydrotreated and separated C₅ - C₆ isomerizable hydrocarbon stream in step (a) passes through a molecular sieve bed which selectivity absorbs normal C₅ - C₆ hydrocarbons and recovering an iso C₅ - C₆ fraction.

3. A process according to claim 1 wherein the conditions of isomerization in step (c) include a pressure within the range of 300 to 750 p.s.i.g., a liquid hourly space velocity of isomerizable hydrocarbon of between about 0.5 and 10 and a mole ratio of hydrogen to hydrocarbon between about 0.05 and 5.

4. A process according to claim 1 wherein the conditions of isomerization in step (e) include a pressure within the range of 200 to 500 p.s.i.g., a liquid hourly space velocity of between about 0.5 and 4 and a mole

ratio of hydrogen to hydrocarbon between about 0.5 to 5.

5. A process according to claim 1 wherein said isomerization catalyst in step (c) is a chlorided platinum-alumina composite.

6. A process according to claim 1 wherein said isomerization catalyst in step (e) is a chlorided platinum-alumina composite.

7. A process according to claim 1 wherein said isomerization catalyst in step (e) is a palladium-zeolite Y composite.

8. A process according to claim 1 wherein said fraction is a debutanized naphtha fraction boiling in the range of about 90° to 260° F.

9. A process according to claim 2 wherein hydrogen is separated from said reaction effluent of step (c) the liquid effluent is passed to said molecular sieve bed

wherein isohexane is separated from said liquid effluent.

10. A process according to claim 1 wherein said C₆ isomerizable hydrocarbon stream in step (b) is passed through a guard hydrogenation catalyst immediately before entering said isomerization reactor in (c).

11. A process according to claim 1 wherein said C₆ isomerizable hydrocarbon stream in step (b) is passed through a molecular sieve bed to selectively separate impurities therein prior to entering said isomerization reactor in (c).

12. A process according to claim 1 wherein hydrogen is passed through a molecular sieve bed to selectively separate impurities therein prior to introduction to said isomerization vessels in steps (c) and (e).

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