

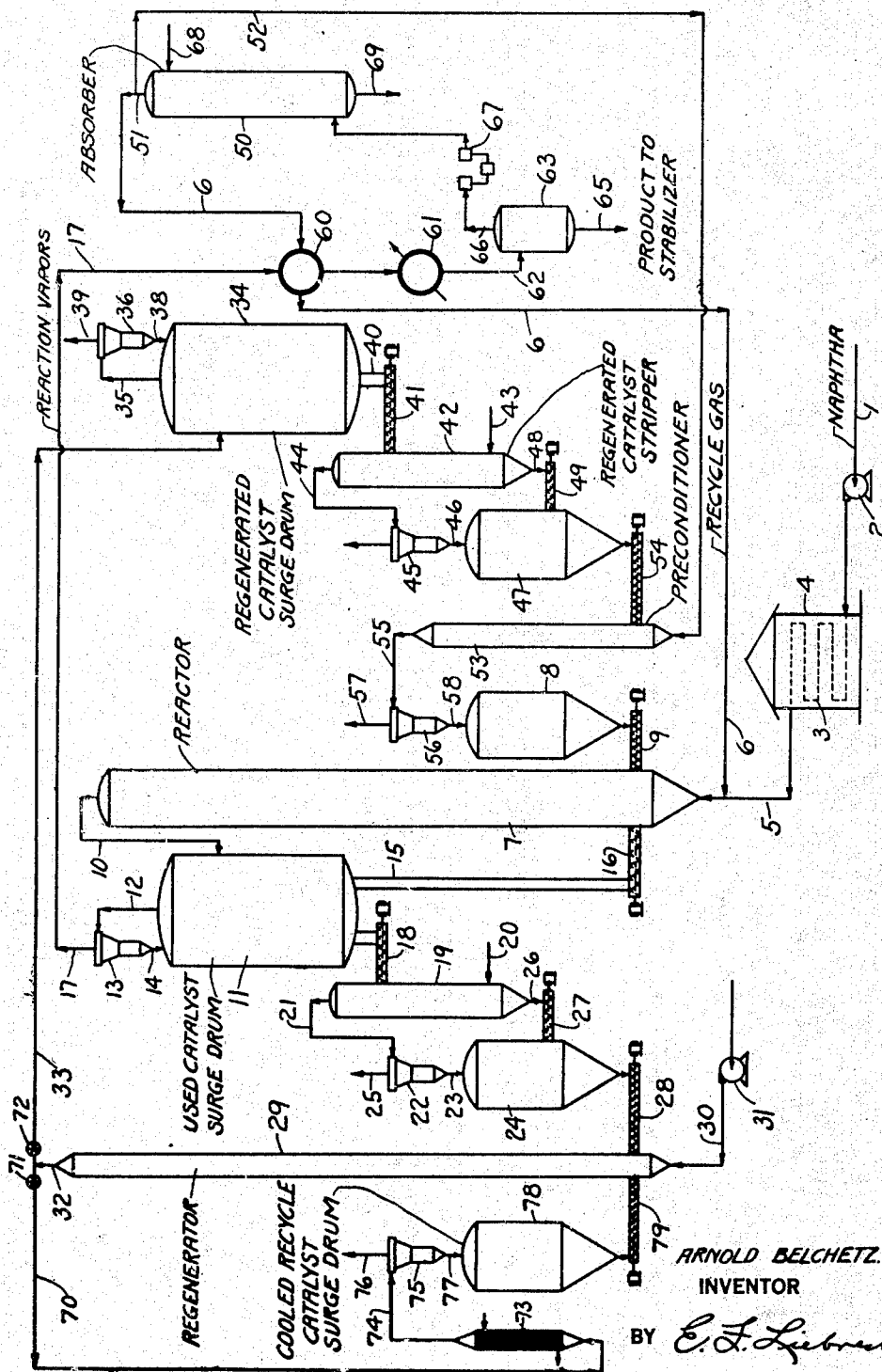
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CATALYTIC CONVERSION OF HYDROCARBONS

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CATALYTIC CONVERSION OF HYDRO-CARBONS

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The present invention relates to the catalytic conversion of hydrocarbons, and more particularly, to the conversion of straight chain hydrocarbons and mixtures thereof to aromatic hydrocarbons by contacting the straight chain compounds under suitable reaction conditions with a catalyst capable of dehydrogenating and cyclicizing them to aromatic compounds. In its preferred aspect, the invention relates particularly to improvements in the treatment of low anti-knock naphtha fractions having a large proportion of straight chain hydrocarbon compounds boiling within the motor fuel range to convert such compounds to aromatic hydrocarbons and thereby considerably increase the anti-knock quality of the naphtha.

Heretofore, a dehydro-aromatization process of considerable commercial interest and merit for the conversion of low anti-knock naphtha fractions boiling within, and slightly above, the motor fuel range, to motor fuel of high aromatic content has been developed. Pursuant to this process, a suitable charging stock, for example a paraffinic naphtha of low anti-knock quality, is mixed with hydrogen and contacted in the vapor phase with a suitable catalyst and under suitable reaction conditions for the desired dehydro-aromatization reaction. By means of this reaction, a large proportion of the straight chain hydrocarbons is converted by dehydrogenation and cyclization to corresponding aromatic compounds and is attended by a net production of hydrogen. Also, it has been proposed to effect this type of conversion by either contacting the hydrocarbons in a reaction zone containing a stationary mass or bed of the catalyst, or by an alternative procedure of suspending the catalyst in the vapors undergoing treatment, passing the suspension through the reaction zone, and thereafter separating the suspended catalyst from the vaporous reaction products.

The latter type of operation, while having certain inherent advantages such as the high degree of intimacy of contact between the vapors and finely divided suspended catalyst, is subject to various difficulties and disadvantages which have precluded its utilization on a practical or commercial scale. A general object of my invention is the provision of a process whereby these difficulties and disadvantages are largely eliminated.

Among the more important obstacles which have been encountered in attempts to practice a dehydro-aromatization process by a procedure wherein the catalyst is suspended in the hydrocarbon vapors passing through the conversion zone are the following:

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(a) Necessity of supplying a considerable endothermic heat of reaction to the conversion zone,

(b) Difficulty of securing satisfactory yields and qualities of conversion products with available catalysts when utilizing equipment of a commercially feasible size and proportions of reactants and catalyst,

(c) Difficulties arising out of the tendency of some of the otherwise more effective catalysts, such as those comprising molybdenum compounds, to undergo reduction during the conversion stage and oxidation during the catalyst regeneration stage.

One of the objects of my invention is the provision of a method for supplying the required endothermic heat of reaction without utilizing conventional methods of heat exchange or heat-transfer which have been ascertained to be unduly expensive and impractical.

A further object of my invention is the provision of a method for securing a satisfactory degree of conversion both with respect to quantity and quality of conversion products without resorting to inordinately large equipment or proportions of reactants and catalysts at any stage of the process.

An additional object of the invention is the provision of an improved method for preconditioning the catalyst, particularly catalysts of the type exhibiting a tendency to become reduced during the conversion stage and oxidized during the regeneration stage of the process whereby the disadvantages arising from such reactions are largely obviated.

Various other objects, features, and advantages of my invention will be apparent from a consideration of the following detailed description thereof given in connection with the appended drawing, wherein I have illustrated diagrammatically a suitable arrangement of apparatus and flow plan for the practice of the invention.

Referring to the exemplary process-flow illustrated, the charging stock introduced through line 1 is suitably a naphtha fraction having a large proportion of straight chain hydrocarbons of low anti-knock quality, that is, aliphatic hydrocarbons having from 6 to 14 carbon atoms in the carbon chain. The naphtha is charged by pump 2 through a heating coil 3 in furnace 4 wherein it is vaporized and heated to a temperature suitable for the subsequent conversion stage. Hydrocarbon vapors are withdrawn from the furnace through transfer line 5 and mixed with recycle gas containing a large proportion of hydrogen introduced through recycle line 6, and the

vapor mixture then passed into the bottom of conversion chamber 7.

A supply of a suitable active catalyst, for example molybdenum or molybdenum oxide supported on alumina, heated to an elevated temperature approximating the desired conversion temperature is maintained in collecting or surge drum 8. Hot catalyst is fed from drum 8 by an adjustable screw feeder 9 into the lower part of the reactor where it is picked up and suspended in the incoming vapor stream of naphtha and recycle gas and then passes upwardly through the conversion chamber. The quantity of catalyst introduced into the reaction zone is preferably maintained within definite limits pursuant to my invention as described in detail hereinafter. In passing upwardly through the reactor the straight chain hydrocarbons undergo a dehydrogenation and cyclization reaction and are largely converted to aromatic compounds, and altered hydrocarbon compounds generally, of enhanced anti-knock quality. I have ascertained that the extent and effectiveness of the conversion is largely dependent upon the concentration of catalyst in the reactor and this concentration is preferably maintained within definitely prescribed limits pursuant to my invention as hereinafter described.

The conversion products are withdrawn from the top of reactor 7 through line 10 and passed to a collecting or surge drum 11 wherein most of the suspended used catalyst settles out of the gas stream and collects in the bottom portion of the drum. The vapor stream is withdrawn from the upper part of drum 11 through line 12 and passed to a suitable gas-solid separator such as cyclone 13 wherein residual suspended catalyst is separated and returned to the drum through line 14. Vaporous conversion products are withdrawn from separator 13 and passed to the products' recovery system through line 17. If desired, a part of the used catalyst may be withdrawn from drum 11 through pipe 15 and recycled to reactor 7 by screw feeder 16 for reuse in the mixture undergoing conversion.

The vaporous reaction products withdrawn through line 17 are cooled to a temperature suitable for the separation of their condensable components from the hydrogen and normally gaseous hydrocarbons contained therein. This cooling may suitably be effected by passing the stream through heat exchanger 60 in heat exchange relation with the cooler recycle gas stream in line 6, and then further cooled in additional heat exchangers or cooler 61. From the final cooler 61, the stream is passed into a high pressure gas separator 63 by line 62. The desired liquid product is withdrawn from the bottom of gas separator 63 by line 65, and may thereafter be treated in accordance with conventional practice of stabilization and the like to produce a commercial gasoline of exceptionally high anti-knock characteristics, or in any other desired manner.

Gaseous constituents are withdrawn overhead from separator 63 through line 66, and are compressed to a pressure suitable for the separation and recovery of low boiling hydrocarbons, such as propane and butane, present therein by compressor 67 and then passed to absorbing tower 50. In tower 50 the gas is passed in counter-current flow with an absorbing oil, suitably the naphtha which is to be introduced through line 1, supplied at the top of the tower through line 68. Enriched absorbing oil is withdrawn from the bottom of the tower through line 69, and a

gaseous product including a large proportion of hydrogen and light hydrocarbon gases, such as methane and ethane, is withdrawn overhead through line 51.

Used catalyst is withdrawn from drum 11 by feeder 18 and introduced into a suitable means for purging it of hydrocarbon vapors entrained therewith, such as stripping tower 19. Tower 19 may be provided with suitable baffles or the like to facilitate intimate contact between the purging medium and the catalyst passing downwardly therethrough. A suitable purging medium such as steam is introduced at the base of tower 19 through line 20 and passes upwardly through the tower and exits from the top through line 21. The purging medium is passed from line 21 into a separator 22 wherein any entrained catalyst is separated. Catalyst thus separated is passed by line 23 into a collecting or surge drum 24. The purging medium is withdrawn from the top of separator 22 by line 25 and suitably may be treated to recover its hydrocarbon content, for example by introduction of the mixture into line 17.

Stripped catalyst is withdrawn from the bottom of stripping tower 19 by line 26 and passed to a stripped, used catalyst collecting or surge drum 24 by screw feeder 27. Catalyst is withdrawn from the bottom of drum 24 and introduced into regeneration chamber 29 by screw feeder 28. A suitable regeneration medium such as air, or a mixture of air and flue gas, is introduced into the bottom portion of regeneration chamber 29 through line 30 by pump 31. Used catalyst is picked up and suspended in the regeneration medium and passes upwardly through the regeneration chamber during which travel combustion of its carbonaceous content is effected and the catalyst thereby regenerated. The regeneration products are withdrawn from the top of the regeneration chamber through line 32 and all or a portion thereof passed by line 33 to a regenerated catalyst collecting or surge drum 34 wherein most of the suspended catalyst is settled out of the gas stream and collected in the bottom portion of the drum. Combustion or flue gases are withdrawn overhead through line 35 and pass to a suitable separating system such as cyclone 36 for separation of residual catalyst entrained in the gas, the catalyst thus separated being returned to drum 34 through line 38. The separated flue gas passes overhead from cyclone 36 and is withdrawn from the process through line 39.

To assist in the maintenance of the temperature of the regeneration zone within required limits, it has been found highly desirable under certain conditions to recycle cooled regenerated catalyst to this zone for absorbing a portion of the heat of regeneration thereby preventing the temperature in the zone from exceeding the maximum or allowable safe regeneration temperature. In such cases the regeneration products withdrawn through line 32 may be split, a suitable fraction thus being by-passed through line 70 by suitable adjustment of valves 71 and 72. The by-passed stream is passed through a suitable cooler or heat exchanger 73 wherein its temperature is reduced to a suitable degree. From cooler 73 the cooled stream is passed by line 74 to separator 75 wherein the suspended catalyst is separated from the flue gas which passes overhead through line 76. Catalyst is withdrawn from the bottom of separator 75 by line 77 and introduced into the cooled regenerated catalyst collecting or surge drum 78. From drum 78 the cooled catalyst

is introduced in the required amounts to regeneration chamber 29 by screw feeder 79.

The regenerated catalyst from drum 34 is preferably subjected to a purging treatment for removal of any oxygen-containing gases entrained or adsorbed thereon, and also in certain instances is preferably subjected to a preconditioning treatment with a reducing gas, preferably hydrogen.

The regenerated catalyst may suitably be purged by withdrawing it from drum 34 through line 40 and introducing it by screw feeder 41 to a stripping tower 42. A suitable purging medium such as flue gas is introduced into the bottom of stripping tower 42 through line 43 and passes upwardly through the tower in countercurrent flow with the catalyst introduced through feeder 41, thereby stripping or displacing from the catalyst any detrimental oxygen-containing gases. The stripping medium is withdrawn from the top of tower 42 through line 44 and passes to a suitable separator 45 for separation of entrained catalyst. Catalyst thus separated is passed from separator 45 into the stripped, regenerated catalyst collecting or surge drum 47 through line 46. Catalyst is fed from the bottom of stripping tower 42 into the bottom of drum 47 by line 48 and screw feeder 49.

The stripped used catalyst in certain instances is preferably subjected to a preconditioning treatment with hydrogen or other reducing gas. This is particularly the case with catalysts of the molybdenum oxide type which undergo a reduction during the conversion stage and oxidation during the regeneration stage of the process. This preconditioning treatment may be effected by suspending the regenerated catalyst in the recycle stream at a point in the recycle line at a sufficient distance from the reactor to permit the desired reduction by recycle gas to take place before the charging stock is mixed therewith. I have ascertained, however, that this preconditioning treatment may be most effectively practiced by performing it in a separate stage and utilizing a stream of recycle gas separate from that passed to the reactor, as shown. In accordance with this preferred procedure, a portion of the total recycle gas withdrawn from gas separator 50 by lines 51 and 52 is introduced into the bottom portion of a preconditioning tower 53. The gas thus introduced picks up and suspends stripped regenerated catalyst introduced from drum 47 through feeder 54 and the suspension passes upwardly through the tower 53 during which travel the desired preconditioning of the catalyst is effected. The suspension is withdrawn from the top of tower 53 by line 55 and passed to a suitable separator 56 wherein separation of the preconditioned catalyst is effected. The preconditioning gas is withdrawn from separator 56 through line 57 and is preferably thereafter used as fuel, or other purposes than recycling to the reactor 7 since it is normally too dilute in hydrogen for satisfactory utilization in the conversion step. Separated catalyst passes from separator 56 into surge drum 8 through line 58. The treatment to which the regenerated catalyst is subjected from the time of withdrawal through line 33 and its introduction into the reactor 7 by feeder 9 are so conducted that the catalyst when introduced into the reactor retains substantially all the heat imparted during the exothermic regeneration reaction, and is at a temperature approximating, and preferably somewhat above, the average desired conversion temperature.

The conditions maintained in conversion cham-

ber 7 in the practice of my invention are suitably illustrated by a consideration of a typical example.

In this example a low anti-knock East Texas heavy naphtha was charged through line 1 and recycle gas introduced from line 6 in the proportions of 2500 cu. ft. of recycle gas (measured at standard conditions) to one barrel of naphtha and regenerated catalyst of the molybdenum oxide-on-alumina-type charged to chamber 7 from drum 8 in the proportions, by weight, of catalyst to naphtha of 7:1. During the course of the reaction the following conditions obtained:

Temperature naphtha vapors leaving furnace 4	°F.	800
Temperature recycle gas in line 6	°F.	600
Temperature regenerated catalyst in drum 8	°F.	1150
Average inlet temperature of the naphtha, recycle gas and catalyst to chamber 7	°F.	1000
Outlet temperature of conversion products leaving chamber 7	°F.	900
Endothermic heat of reaction, B. t. u./lb. of naphtha		250
Pressure in chamber 7	lbs./sq. in. ga.	100
Average velocity of gas in chamber 7	ft./sec.	3
Concentration of catalyst in chamber 7 per cu. ft. of reaction space	lbs.	12
Specific heat of catalyst		0.26
Time of contact of hydrocarbon vapors with catalyst	sec.	40

In the practice of my invention under conditions such as employed in the above example, the regenerated catalyst is supplied to the chamber 7 at an elevated temperature somewhat above the desired average conversion temperature and in relatively large proportions adequate to supply sufficient heat during the conversion to maintain the reaction mixture at the desired average temperature level. To produce this result the catalyst-to-oil weight ratio may suitably be about 7:1 as shown in the example. In general, my invention contemplates the introduction of regenerated catalyst heated to the required elevated temperature in amounts expressed in terms of catalyst-to-oil ratio by weight of about 2 to 20 and preferably in the intermediate range of about 4 to 10. The desired ratio (R) may be determined for any given set of conditions by applying the following general formula—

$$R = \frac{(H_1 - H_2) + Wh}{W(T_1 - T_2)S}$$

wherein the symbols have the following meanings,

T_1 = average inlet temperature of the naphtha, recycle gas and catalyst to the reactor (in °F.)

T_2 = required outlet temperature from the reactor (in °F.)

H_1 = total heat of the naphtha + recycle gas at temp. T_1 (in B. t. u./hr.)

H_2 = total heat of the reaction products at temp. T_2 (in B. t. u./hr.)

W = naphtha charge in #/hr.

h = heat of reaction of the naphtha in B. t. u./# naphtha

S = specific heat of the catalyst

R = catalyst/naphtha feed weight ratio

Pursuant to my invention, a relatively high concentration of catalyst is maintained in the conversion zone 7 by maintaining the velocity of the gases passing through the reactor within relatively low limits, for instance a concentration of

12 lbs. per cu. ft. of reactor space and a velocity of 3 ft. per second as shown in the example above. If the catalyst particles are carried through the reactor at substantially the same speed as the gas particles, the concentration of the catalyst in the reactor will be directly dependent upon the feed ratio of catalyst-to-oil to said zone, and the catalyst concentration therefore may be increased under this condition by increasing the quantity of catalyst fed through the screw feeder. If the velocity of the gases exceeds a definite maximum value termed its "transport velocity," the catalyst is carried along at substantially the same speed as the gas particles. The transport velocity is determinable and may be ascertained by calculation and experiment for any given set of conditions including such factors as the viscosity of the gas, density and size of the particles, and the like. By maintaining the velocity of the gases in the reactor below the transport velocity, slippage of the catalyst in the gas stream is induced and the concentration of the catalyst to unit volume of oil vapor in the reactor is increased relative to the ratio of catalyst to unit volume of oil vapor fed to the reactor. The extent of this increase is dependent upon the velocity and increases with lower velocities and corresponding greater slippage. With sufficiently low gas velocities the concentration of the catalyst in the reactor reaches a point where it is independent of the catalyst-to-oil feed rate.

The concentration of the catalyst maintained within the reaction zone in accordance with my invention may be expressed in terms of the weight of catalyst per unit volume of reaction space. I have found that satisfactory results are produced with a relatively high concentration, for example a catalyst concentration of about 12 lbs. per cu. ft. of reactor as given in the example; however the concentration may be varied over a range from about 2 to 25 lbs. per cu. ft. of reactor space and preferably is maintained within the range of 6 to 16 lbs. These concentrations may be produced by maintaining the velocity of the hydrocarbon gases in the conversion chamber 7 below the transport velocity, suitably in the range of about 1 to 25, and preferably in the range of about 2 to 7 feet per second.

The present invention does not reside in the selection or use of any particular catalyst per se, and any catalyst capable of catalyzing the desired dehydro-aromatization reaction may be used. As examples of suitable catalysts for use in my process, I prefer the oxides of the metals of the left-hand column of group VI of the periodic table, particularly chromium, molybdenum and tungsten but I can also use other metallic oxides and other metallic compounds, particularly oxides of the metals of the left-hand columns of groups IV and V of the periodic table such as titanium, cerium, thorium and vanadium. Moreover, while these catalytic oxides can be used alone or on various supports including magnesia, I find it preferable to utilize them on alumina, particularly an Activated Alumina or on alumina gel, as a support and in general the catalytic oxide or other catalytic compound should be the minor constituent, usually from 1 to 40% by weight of the total catalyst including the support, although the optimum percentage varies, of course, with the catalyst used. It will also be apparent that mixed catalysts can be used, for instance a mixture of chromium oxide and molybdenum oxide alone or preferably on an alumina support, and in this case the active catalytic oxides should be

from about 10 to about 25% by weight of the total catalyst. Another catalyst which can be used is magnesium chromite either alone or on a suitable support, preferably alumina. In fact, any aromatization catalyst or material that promotes the dehydrogenation and cyclization of aliphatic hydrocarbons can be used. As those skilled in the art know, the efficiency of a given catalyst is largely determined by its method of preparation and consequently some catalysts give better results than others under a specific set of conditions. Such differences obviously can be overcome to some extent by suitable slight alterations in the operating conditions within the limits taught in the present specification. However, of these catalysts I prefer to use a catalyst comprising molybdenum oxide supported on alumina, particularly Activated Alumina, and the molybdenum oxide should most advantageously constitute from about 2 to about 10% by weight of the total catalyst since this has been found to include a very sharp optimum range which gives the best results. Larger amounts of molybdenum oxide can, however, be used. The catalyst is preferably employed in finely divided or powdered condition, for example, most of the catalyst should be of sufficient fineness to pass a 200 mesh screen. My preferred catalyst may be made in accordance with the following procedure:

Dissolve 165 kilograms of ammonium para molybdate in sufficient distilled water to yield 1500 liters of solution. Place 1500 kilograms of granular Activated Alumina in an apparatus which can be evacuated and add the molybdate solution. Agitate the mixture and then apply a vacuum pump and reduce the pressure to 30-40 millimeters of mercury. Then allow the pressure to rise to atmospheric. Lower the pressure a second time to the same level and then allow it to rise again to atmospheric pressure. Repeat this procedure a third time and then drain the remaining liquid from the impregnated alumina. Air-dry the latter on screens or other suitable containers using layers of about one inch in depth. At the end of this time place the dried material in a furnace in a suitable container and heat it at a temperature of 1200° F. for one hour. Cool to atmospheric temperature, pulverize and store in closed containers until ready for use.

The pressure maintained in the conversion reactor 7 is preferably maintained at 100 lbs. gauge but satisfactory results are attainable with pressures ranging from about 50 lbs. to 450 lbs.

The amount of added hydrogen or recycle gas introduced through line 6 in the practice of my invention may be expressed as the number of mols of hydrogen per mol of charge calculated on the basis of the mean molecular weight of charge which for the East Texas heavy naphtha used in the illustrated example was in the vicinity of 120. The practice of my process will usually require the addition of hydrogen in amount ranging from between 0.5 to 9 and preferably about 2 to 3 mols of hydrogen per mol of hydrocarbons charged.

The present application discloses certain features previously disclosed in my co-pending application Serial No. 274,670 filed May 20, 1939, now U. S. Patent 2,253,486, and constitutes a continuation-in-part of said earlier-filed application.

I claim:

1. A process for the conversion of aliphatic hydrocarbons to aromatic hydrocarbons which comprises suspending a powdered dehydrogenating and cyclicizing catalyst of sufficient fineness that most of it passes a 200-mesh screen previously heated to a temperature approximating the con-

version temperature in vapors of the aliphatic hydrocarbons, passing the suspension upwardly through a conversion zone at a velocity within the range of about 1 to 3 feet per second under superatmospheric pressure and at elevated temperatures suitable to largely dehydrogenate and cyclicize the aliphatic hydrocarbons to aromatic compounds, and maintaining the proportions of heated catalyst and hydrocarbons introduced to the conversion zone within limits in terms of catalyst-to-hydrocarbons, by weight, of about 2-20 to 1 so as to supply the endothermic heat of reaction without supplying extraneous heat to the reaction mixture to any substantial extent.

2. A process for the conversion of aliphatic hydrocarbons to aromatic hydrocarbons which comprises suspending a powdered dehydrogenating and cyclicizing catalyst of sufficient fineness that most of it passes a 200-mesh screen previously heated to a temperature approximating the conversion temperature in vapors of the aliphatic hydrocarbons, passing the suspension upwardly through a conversion zone at a velocity within the range of about 1 to 3 feet per second under superatmospheric pressure and at elevated temperatures suitable largely to dehydrogenate and cyclicize the aliphatic hydrocarbons to aromatic compounds, and maintaining the proportions of heated catalyst and hydrocarbons introduced to the conversion zone within limits in terms of catalyst-to-hydrocarbons, by weight, of about 4-10 to 1 so as to supply the endothermic heat of reaction without supplying extraneous heat to the reaction mixture to any substantial extent.

3. A process for the conversion of aliphatic hydrocarbons to aromatic hydrocarbons which comprises suspending a powdered dehydrogenating and cyclicizing catalyst of sufficient fineness that most of it passes a 200-mesh screen and previously heated to a temperature approximating the conversion temperature, in vapors of the aliphatic hydrocarbons, passing the vapors upwardly through a conversion zone at a velocity within the range of about 1 to 3 feet per second under superatmospheric pressure and at elevated temperatures suitable to largely dehydrogenate and cyclicize the aliphatic hydrocarbons to aromatic compounds, and maintaining the proportions of heated catalyst and hydrocarbon reactants introduced to the conversion zone within such limits as to supply the endothermic heat of reaction without supplying extraneous heat to the reaction mixture to any substantial extent and within the range of about 2-20 parts of catalyst to 1 part of said reactants by weight, the said maintained velocity of the hydrocarbon vapors through said conversion zone being such as to produce a relatively great concentration of catalyst in said zone.

4. A process for the conversion of aliphatic hydrocarbons to aromatic hydrocarbons, which comprises suspending a powdered dehydrogenating and cyclicizing catalyst of sufficient fineness that most of it passes a 200-mesh screen and previously heated to a temperature approximating the conversion temperature in vapors of the aliphatic hydrocarbons, passing the vapors upwardly through a conversion zone within the range of about 1 to 3 feet per second in the presence of added hydrogen and under superatmospheric pressure and elevated temperatures suitable largely to dehydrogenate and cyclicize the aliphatic hydrocarbons to aromatic compounds and hydrogen, the proportions of heated catalyst and reactants introduced to the conversion zone

being maintained within such limits as to supply the endothermic heat of reaction without supplying extraneous heat to the reaction mixture to any substantial extent and within the range of 2 to 20 parts of catalyst to 1 part of reactants by weight, withdrawing the conversion products and separating the used catalyst and hydrogen produced, from the other components of the conversion mixture and each other, splitting the stream of hydrogen and utilizing one portion to precondition the used catalyst for reuse and the other portion as the added hydrogen present in the conversion stage.

5. A process for the conversion of aliphatic hydrocarbons to aromatic hydrocarbons which comprises heating and vaporizing said aliphatic hydrocarbons to a temperature suitable for conversion, suspending a powdered dehydrogenating and cyclicizing catalyst of sufficient fineness that most of it passes a 200 mesh screen and previously heated to a temperature approximating the conversion temperature in said vapors of the aliphatic hydrocarbons, passing the vapors upwardly through a conversion zone at a velocity within the range of about 1 to 3 feet per second under superatmospheric pressure and at elevated temperatures suitable to largely dehydrogenate and cyclicize the aliphatic hydrocarbons to aromatic compounds, and maintaining the proportions of heated catalyst and hydrocarbon reactants introduced to the conversion zone within such limits as to supply the endothermic heat of reaction without supplying extraneous heat to the reaction mixture to any substantial extent and within the range of about 2-20 parts of catalyst to 1 part of said reactants by weight, the said maintained velocity of the hydrocarbon vapors through said conversion zone being such as to produce a relatively great concentration of catalyst in said zone.

6. A process for the conversion of aliphatic hydrocarbons to aromatic compounds, which comprises suspending a powdered dehydrogenating and cyclicizing catalyst of sufficient fineness that most of it passes a 200 mesh screen and previously heated to a temperature approximating the conversion temperature in a mixture of added hydrogen and vapors of the aliphatic hydrocarbons, the feed weight ratio of the hydrocarbon reactants to the catalyst being maintained within such limits as to supply the endothermic heat of reaction without supplying extraneous heat to the reaction mixture to any substantial extent and within the range of about 2-20:1, passing the gases upwardly through a conversion zone at a velocity within the range of about 1 to 3 feet per second under superatmospheric pressure and at an elevated temperature suitable largely to dehydrogenate and cyclicize the aliphatic hydrocarbons in said conversion zone and within limits sufficient to produce a concentration of catalyst in said zone of between about 2 to 25 pounds per cubic foot of conversion space.

7. A continuous process for the conversion of a low octane naphtha to a highly aromatic high octane motor fuel, which comprises introducing a mixture of added hydrogen and vapors of said low octane naphtha and a finely divided dehydrogenating and cyclicizing catalyst of sufficient fineness that most of it passes a 200 mesh screen and previously heated to a temperature approximating the conversion temperature in a feed weight ratio of about 2 to 20 parts of catalyst to 1 part of naphtha into a catalytic conversion zone maintained under superatmospheric pres-

sure and at an elevated temperature, and passing the vaporous reactants upwardly through said conversion zone at a velocity maintained within the range of about 1 to 3 feet per second to provide a concentration of said finely divided catalyst in said conversion zone of not less than six pounds per cubic foot, the feed weight ratio of the hydrocarbon reactants to the hot catalyst being maintained within the above limits at a value effective to supply the endothermic heat of reaction without supplying extraneous heat to the reaction mixture to any substantial extent.

8. In a process continuous for dehydrogenating naphtha in the presence of a catalyst comprising an oxide of a metal included in the left-hand column of group VI of the periodic table to convert constituents of said naphtha to aromatic compounds, the improvement which comprises flowing the naphtha in the vapor phase at a relatively low velocity within the range of about 1 to 3 feet per second upwardly through a catalytic conversion zone maintained at an elevated temperature suitable for the desired conversion, continuously introducing said catalyst in powdered condition of sufficient fineness that most of it passes a 200 mesh screen and previously heated to a temperature approximating the conversion temperature into said conversion zone at a rate in proportion to the hydrocarbons introduced within the range of about 2 to 20 parts of catalyst to 1 part of naphtha by weight whereby the catalyst particles are circulated in the vapors undergoing conversion through the conversion zone and a relatively high concentration of cata-

lyst particles is maintained in said zone, the feed weight ratio of the hydrocarbon reactants to the hot catalyst being maintained within the above limits at a value effective to supply the endothermic heat of reaction without supplying extraneous heat to the reaction mixture to any substantial extent.

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