

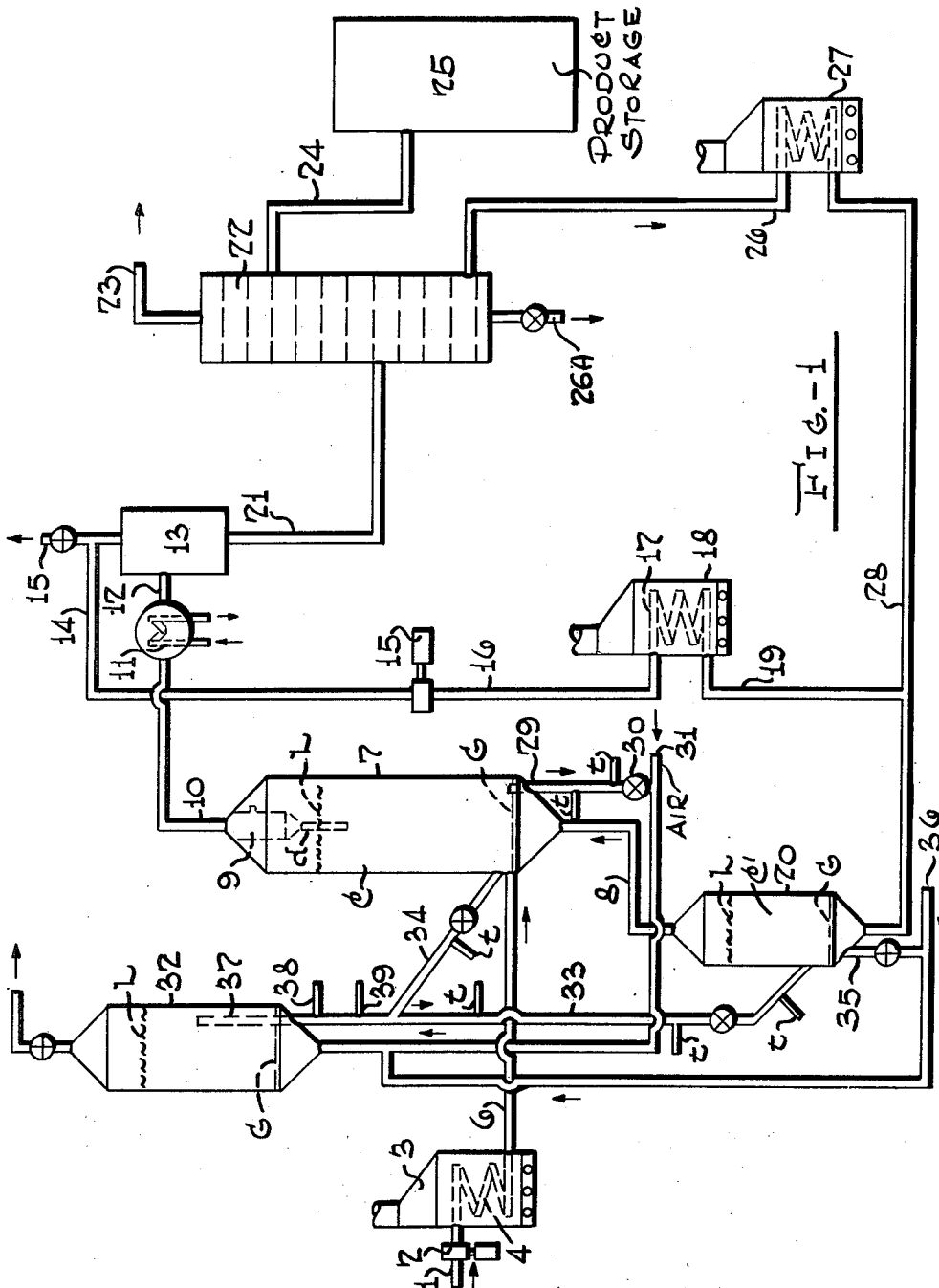
Dec. 21, 1954

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2 Sheets-Sheet 1



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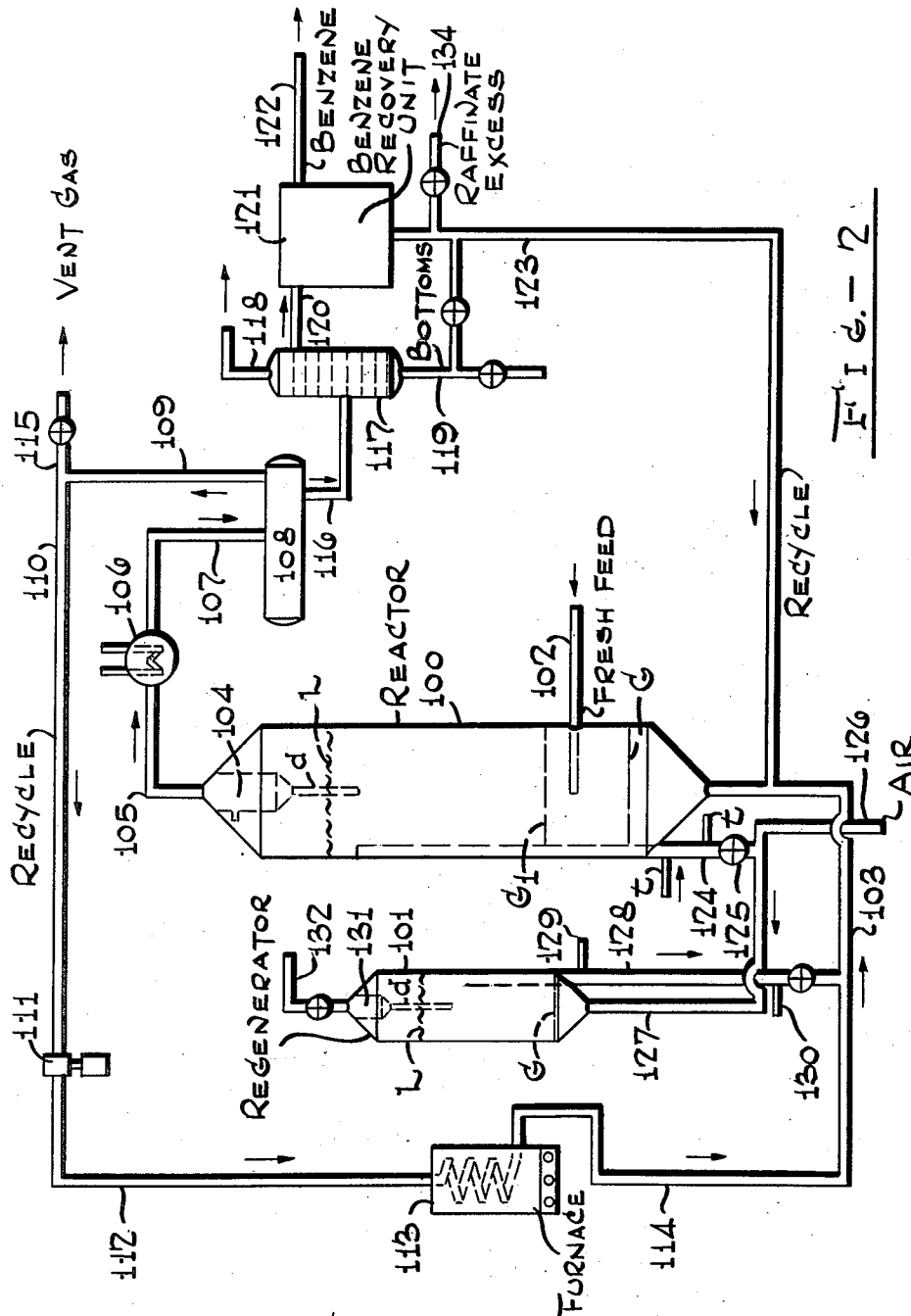
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REFORMING OF NAPHTHAS

Charles E. Hemminger, Westfield, and Wilfred O. Taff, Cranford, N. J., assignors to Standard Oil Development Company, a corporation of Delaware

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10 Claims. (Cl. 196—49)

The present invention relates to improvements in the reforming of virgin, straight run or cracked naphthas, and more particularly, it relates to the production of aromatic hydrocarbons.

Hydroforming is usually defined as an operation in which naphthene-containing naphtha is subjected to the influence of high temperatures and pressures in the presence of a solid catalytic material and added hydrogen for a period of time sufficient to convert naphthenes present in the naphtha to the corresponding aromatics by dehydrogenation. In addition to the dehydrogenation reactions indicated above, there is some isomerization of acyclic and cyclic paraffins which takes place. Although extraneous hydrogen may be added with the naphtha feed to the hydroforming zone, the process invariably operates so that there is no net consumption of hydrogen, and as a matter of fact, there is ordinarily a net production of hydrogen.

The present invention relates to a modification of the ordinary hydroforming process in that increased yields of the low boiling aromatics such as benzene and toluene are obtained. More specifically, the present invention involves increasing the yield of aromatics by recycling a portion of the product to the hydroforming zone where it contacts the hydroforming catalyst at a higher temperature than the fresh feed, and wherein heavy hydrocarbons containing 9 or 10 carbon atoms are treated to reduce the number of carbon atoms so as to form benzene, toluene and xylenes. In another modification of the present invention, the product of the hydroforming operation is solvent-extracted to recover benzene, and the raffinate containing primarily paraffins is recycled and subjected to the influence of the hydroforming catalyst at a higher temperature and, therefore, under more severe conditions than the fresh feed and a greater quantity of aromatics are thus produced from a given quantity of the original feed. The first mentioned modification would be especially applicable where the original feed was a wide boiling naphtha, say, one boiling from about 200–450° F. The product from this would contain C₉ and C₁₀ and higher aromatics. These C₉, C₁₀ and higher aromatics are present in the hydroformed product in the fraction boiling above 325° F. in 85–100% concentration, and the fraction containing them, therefore, possesses a high octane number of the order of 100 CFRR. However, these materials boil too high for commercial aviation and some grades of motor gasoline, and according to the first mentioned aspect of the present invention, this high boiling material is subjected to treatment with the hydroforming catalyst, but at a higher temperature as stated to split off or reduce the length of aliphatic hydrocarbon chains attached to the benzene nucleus and thus lower the boiling point. The removal of these side chains may be carried out to the extent that xylene, toluene and even benzene can be produced by subjecting this high boiling material to a high temperature treatment during a sufficiently long period of time.

With respect to the second aspect mentioned above, with a low-boiling feed stock such as a C₆ or C₇ cut in the broad range of about 150–250° F. the aromaticity of the hydroformed product is ordinarily not as high, and may be from about 60–85% aromatics or even lower depending on the feed stock and operating conditions. In this case, the hydroformed product may be subjected to solvent extraction to recover aromatics and the raffinate recycled but treated under more severe conditions than the fresh feed, to give an increased production of aromatics. By applying this multi-pass operation to a hexane feed cut, for example, where initial conversions of methylcyclopentane to benzene of about 75% on feed

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and of normal hexane to benzene of about 40% are obtained, ultimate benzene yields of the order of 90 and 75%, respectively, can be realized from these feed components.

The main object of the present invention is to produce aromatic hydrocarbons such as benzene, toluene, and the xylene isomers in increased yields from naphthas.

Another object of the present invention is to convert other constituents of a hydroformed naphtha into low boiling aromatics by a more severe second stage treatment, utilizing the high temperature level of the hot, freshly regenerated hydroforming catalyst in a circulating fluidized stream entering the primary reaction zone.

A specific object of the invention is to produce benzene in good yields and in a high state of purity from virgin naphtha by an improved fluid catalytic aromatization process.

Other and further objects and details of the invention will appear from the following more detailed description and claims.

In the accompanying drawings there is shown diagrammatically in Fig. I, an apparatus layout in which a preferred modification of the invention may be carried into effect; and in Fig. II, there is depicted diagrammatically, an apparatus layout in which another modification of the invention may be carried into effect.

Referring in detail to Fig. I, a naphtha fraction introduced into the system through line 1 is forced by pump 2 into the furnace 3, through a fired coil 4 in which the oil is preheated and vaporized, and thereafter withdrawn through line 6, and charged to fluid hydroforming reactor 7. The fluid reactor 7 contains a bed of finely divided fluidized catalyst C disposed as shown in reactor 7 between a foraminous member G and an upper dense phase level L. Simultaneously, there is discharged into reactor 7 at the bottom thereof, through line 8, a mixture of hydrogen, entrained catalyst particles and vaporized hydrocarbons. The latter, obtained from a subsequent step of the process, may consist essentially of treated vapors derived from the higher boiling aromatics in the product which have 9, 10 or more carbon atoms per molecule, that is to say, from benzene derivatives having from about 3 to 6 carbon atoms in a side chain or chains. This mixture of hydrogen, catalyst and aromatic hydrocarbons passes upwardly in reactor 7 through the foraminous member G and into the bed of catalyst C. Under conditions more fully set forth hereinafter, the desired conversion takes place, and the crude product issues from the dense bed and passes into a disengaging space disposed between L and the top of the reactor. In this disengaging space the concentration of catalyst in gasiform material decreases sharply upwardly. Before the vapors are withdrawn from the reactor, they are forced through one or more "cyclone" separators 9 wherein catalyst still entrained in the vapors is separated and returned to the dense bed via one or more dip pipes d.

The effluent from the reactor passes through line 10, thence through a cooler 11, and thereafter via line 12 into a separator 13. From separator 13, a gas rich in hydrogen is withdrawn overhead through line 14, passed through a compressor 15 and thence via line 16 into a fired coil 17 disposed in the furnace 18. This preheated hydrogen containing gas is withdrawn from furnace 18 through line 19 and passed into vessel 20 containing a bed of hot fluidized catalyst C'. The reaction which occurs in this vessel 20 may be essentially one of hydrocracking, in which the higher boiling naphtha aromatics containing 9 or more carbon atoms per molecule are dealkylated to form benzene, toluene, etc.

Referring again to separator 13, the hydrogen depleted crude product is withdrawn therefrom through line 21 and charged to a fractionating column 22. From fractionator 22, propane, butane and other light hydrocarbons are withdrawn through line 23 for suitable use, outside the system shown. Through line 24, the reformed naphtha product is withdrawn and delivered to a storage drum 25. A heavy fraction containing the C₉ and higher aromatics in the naphtha boiling range is withdrawn from fractionator 22 through line 26. While this will ordinarily be a bottoms cut, it may be desirable at times to withdraw through line 26A a tar bottoms above a vapor tempera-

ture of about 450°–500° F. The desired C₉ and higher aromatics stream in line 26 is charged to a furnace 27 where it is vaporized and thence passed via line 28 into the reactor 20 together with the hot hydrogen-containing gas from line 19. The dealkylated aromatic product is then withdrawn through line 8 along with entrained catalyst, and charged to vessel 7.

Periodically, it is necessary to regenerate the catalyst and toward this end the catalyst in reactor 7 is withdrawn through a standpipe 29 carrying the usual gas taps *t*, at a rate controlled by a valve 30, and charged into an air stream flowing in line 31 to form a suspension therein. This suspension is carried into regenerator 32 where it is maintained in the form of a dense fluidized bed extending from a grid or other foraminous member G' to an upper dense phase level L'. The catalyst, under conditions more fully set forth hereinafter, undergoes regeneration to remove carbonaceous and other deposits and then is withdrawn through a standpipe 33 carrying the usual gas taps *t*. Regenerated catalyst is charged in part to reactor 20 and the main portion passes via branch standpipe 34 into reactor 7. The catalyst in reactor 20 will, of course, also require regeneration and for this purpose it may be withdrawn through a standpipe 35 and charged into an air stream in line 36 and carried in suspension into regenerator 32.

An upper portion of line 33, namely, 37, may serve as a stripping section in which the catalyst is stripped of regeneration gases. For this purpose a suitable stripping medium as steam or nitrogen, is introduced through tap 38. The section of the line between taps 38 and 39, which may be of considerable length, may also serve as a treating section in which the catalyst is contacted with a hydrogen rich gas such as the recycle gas in line 14, introduced through tap 39, so as to reduce the regenerated catalyst to a desired degree. The exact procedure of this catalyst reduction is not part of this invention. The method, time of reduction and disposal of the water of reduction formed will vary with the catalyst and the exact mode of operation.

It is to be noted that reactor 7, reactor 20 and regenerator 32 are each provided with a foraminous member G which may be a grid or a screen and which serves the purpose of effecting good distribution of gasiform material fed into the bottom of the respective vessels. Also, in each case the reference L refers to the upper dense phase level of the fluidized bed of catalyst therein disposed.

Referring in detail to Fig. 2, 100 represents a reactor containing a fluidized bed of catalyst extending from the foraminous member G to the upper dense phase level L and 101 represents a regenerator to which there is fed fouled catalyst from reactor 100, which catalyst is treated with air in the said regenerator for the purpose of restoring the catalyst activity. In operation, fresh feed is charged to reactor 100 after suitable preheating in equipment (not shown) via line 102. In reactor 100, G₁ represents a second foraminous member such as a screen or a grid or a perforate baffle. It is to be noted that the fresh feed enters the reactor 100 at a point above the grid G but below G₁. Hydrogen-containing gas in line 103 is fed into the bottom of reactor 100 and passes upwardly through grids G and G₁ to form within the reactor, a dense fluidized bed of catalyst having an upper dense phase level at L. Under conditions more fully explained hereinafter, the desired reaction occurs and the product emerges from the dense phase and passes into a disengaging space positioned between L and the top of reactor 100. In this disengaging space, catalyst settles out of the gasiform material and descends into the said dense fluidized bed of catalyst. The gasiform material about to emerge from the reactor is forced through one or more "cyclone" separators 104 wherein catalyst still entrained in the vapors is separated and returned to the dense phase via one or more dip pipes *d*. The crude gasiform material issues from the reactor through line 105, is cooled in cooler 106 and thence passes via line 107 into a separator 108. From separator 108, a gas rich in hydrogen is withdrawn overhead through line 109, passed via line 110 through a compressor 111, and thence passed via line 112 to a re-heat furnace 113 and finally passes via line 114 into line 103 from which, as previously pointed out, it is charged into the bottom of reactor 100.

Referring again to separator 108, a portion of the overhead material withdrawn through line 109, may be vented from the system through line 115. This vent line pre-

vents hydrogen and normally gaseous hydrocarbon build-up in the system as well as providing an escapement for sulfur compounds which may be present in, or produced from sulfur bodies present in the original oil.

The heavier higher boiling portion of the material charged to separator 108 which consists of C₅+ hydrocarbons, is withdrawn through line 116 and charged to a fractionator 117. From fractionator 117, light ends are recovered overhead via line 118, while heavy bottoms are withdrawn through line 119. An intermediate boiling fraction containing benzene or toluene is withdrawn from fractionator 117 through line 120 and charged to an aromatic recovery unit 121. Either or both benzene and toluene may be produced, benzene recovery being shown for purposes of illustration. In its preferred form, the benzene recovery unit will comprise a solvent extraction system in which the extract is rich in benzene and the raffinate comprises mostly acyclic hydrocarbons. The benzene extract is withdrawn via line 122 and subjected to the usual finishing steps such as distillation, to recover a commercially pure benzene. The raffinate, on the other hand, is withdrawn from the benzene recovery unit via line 123 and recycled to line 103 where it mixes with the recycle gas and catalyst, and is carried with the latter into the bottom portion of reactor 100, thus completing the cycle of operations except for the regeneration of catalysts which will be described presently.

The bottoms fraction 119 from fractionator 117 may be discarded from the system, or it may be combined in line 123 with the raffinate from benzene recovery unit 121 and recycled to the hot catalyst zone at the bottom of reactor 100, in a manner analogous to the treatment of heavy ends in the process described in Figure 1. This recycle stream may also be preheated by a suitable heater or heat exchanging means, not shown, to further increase the severity of the second stage treatment.

The catalyst in regenerator 101 is in the form of a dense fluidized mass extending from G to L as previously noted. Catalyst to be regenerated may be withdrawn from reactor 100 through a standpipe 124 at a rate controlled by a valve 125. As usual, this standpipe is provided with a plurality of gas taps *t* through which small amounts of fluidizing gas or steam as a stripping gas may be introduced to maintain the column of catalyst in a mobile condition. Air is introduced into the regeneration system through line 126, picking up catalyst discharged from line 124 to form a suspension which is carried via line 127 into the bottom of regenerator 101. Under conditions more fully explained hereinafter, catalyst undergoes regeneration for the purpose of removing carbonaceous and other deposits therefrom and thereafter, the regenerated catalyst is withdrawn through aerated standpipe 128 and charged into line 103 to be returned to reactor 100 for further use therein. As in the case of Fig. 1 suitable catalyst stripping and treating gases may be introduced into line 128 by taps 129 and 130, respectively, to condition the regenerated catalyst before contacting the feed in line 103. The regeneration fumes formed from the generator 101 pass into the customary disengaging space located between L and the top of the regenerator. Fumes about to issue from the regenerator are forced through one or more "cyclone" separators 131 for the purpose of removing catalyst entrained therein, which removed catalysts is returned to the dense phase via one or more dip pipes *d*. The hot fumes finally issue from the regenerator through line 132. These may be utilized in the system to preheat the oil feed to form steam utilizable in the present system or otherwise in the refinery, or utilized in any known manner to recover its sensible heat content.

Referring again to the raffinate in line 123, a portion of this material is withdrawn from the system as a purge stream, either continuously or intermittently, through line 134.

In the modification illustrated diagrammatically by Fig. 2, it is preferred that the naphtha feed for benzene production comprise a cut boiling substantially between C₆ and 180° F. The catalyst in reactor 100 may be platinum, molybdena or chromia on alumina. The cyclohexane present in such a cut or fraction is substantially completely converted to benzene, under reaction conditions as hereinafter set forth. However, the conversion of the methylcyclopentane and the normal hexane present is not complete. Consequently, to obtain higher yields of benzene from the limited amount of this narrow cut feed available in ordinary oil refining and from the

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feed prepared by expensive distillation or other separation techniques, the present invention contemplates removing the benzene from the product by recycling the major portion of the raffinates obtained from the benzene extraction process, back to the hydro-forming zone. In this multi-pass operation, the unconverted methylcyclopentane and the normal hexane are at least partly converted to benzene. Hence, instead of obtaining a conversion of methylcyclohexane of about 75% and of normal hexane of only about 40% in a single pass operation, the present improved process contemplates ultimate conversion in the order of 90% of the methylcyclopentane originally fed, and 75% ultimate conversion of the normal hexane originally fed.

Exactly the same technique may be applied to other feed boiling ranges such as a C₇-230° F. cut for the production of toluene, or to broader cuts for the production of mixed aromatics. In any case, in connection with the apparatus depicted in Fig. 2, the raffinate is fed to the lower part of reactor 100 which is at the highest temperature. There it contacts the hot recycle gas and the hot catalyst from the regenerator before the recycle gas or the catalyst contacts the fresh feed which is introduced at a cooler point downstream. A modification of the present invention similar to Fig. 1 involves the use of two fluid reactors, one for treating the fresh feed and the second for treating the recycled raffinate, which second reactor will operate at a substantially higher temperature than the first-mentioned reactor as in Fig. 1. The present invention also contemplates the use of moving bed reactors, and operating with a platinum-alumina catalyst at two temperature levels involving continuous operation under conditions where catalyst regeneration is not required.

While the invention as disclosed in either Fig. 1 or Fig. 2 may be applied primarily to the manufacture of benzene, the same operations may be carried out for production of toluene or xylenes by choosing a feed of a suitable boiling range. The feed can also contain substantial portions of cracked stocks as from catalytic cracking or thermal cracking units.

Now, further to explain the present invention, the following specific examples are set forth dealing with detailed operating conditions and results secured.

EXAMPLE I

Conditions and results in design of Fig. 1

CONDITIONS IN REACTOR 7

Catalyst..... 10% molybdena on alumina.
Temperature..... 900° F.
Pressure..... 200 p. s. i. g.
Feed rate..... 0.3 lbs. oil/hr./lb. catalyst.

Catalyst to oil ratio in reactor 7=1.

CONDITIONS IN REACTOR 20

Catalyst..... 10% molybdena on alumina.
Temperature..... 1200° F.
Pressure..... 200 p. s. i. g.
Feed rate..... 1.0-3.0 lbs. oil/hr./lb. catalyst, depending on feed character.

Catalyst to oil ratio in reactor 20=5-10.

CONDITIONS IN REGENERATOR 32

Temperature..... 1200° F.
Pressure..... 200 p. s. i.
Contact time..... 10 minutes.

INSPECTION

	Feed	Product
Aromatics (C ₆ and C ₇) wt. percent.....	3	55
Paraffins.....	50	15
C ₈ -C ₁₀ aromatics.....	10	25
Naphthenes.....	37	5
Octane Number, CFRR.....	50	98
Yield, Vol. percent.....	100	75

EXAMPLE II

Conditions and results in design of Fig. 2

CONDITIONS IN REACTOR 100

Catalyst..... 10% molybdena on alumina promoted by zinc.
Temperature..... 950° F.
Pressure..... 200 p. s. i. g.
Catalyst to oil ratio..... 1.

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CONDITIONS IN REGENERATOR 101

Temperature..... 1200° F.
Pressure..... 200 p. s. i. g.
Contact time..... 10 minutes.

INSPECTION

	Feed	Product
Aromatics.....	5.2	55
Paraffins.....	68.3	40
Naphthenes.....	26.5	5
Benzene Yield on Fresh Feed.....	5.1	35

It is to be understood, of course, that the foregoing conditions are merely for illustration and are not to be construed as placing any limitation on the invention, for the temperature, pressure and other conditions as set forth numerically above may be varied within reasonable limits to effect the objects of this invention. Desirable temperature limits, for example, may include the range from about 850-950° F. for the main hydroforming reaction and about 1100-1200° F. for the catalyst regeneration and second reaction zones, which operate at essentially the same temperature. In any case the temperature of the various streams entering the second reaction zone is controlled so as to keep the temperature level in this zone at least 150° F. higher than the temperature in the main hydroforming reactor. By this means the desired reaction of increased severity can be carried out in this second reaction zone and the total product therefrom, including gas, hydrocarbon vapors and hot catalyst conveyed directly to the first reaction zone to supply the necessary heat of reaction for hydroforming the fresh feed stream.

With respect to the catalyst used in the main hydroforming and secondary treating zones, hereinbefore described, it is pointed out that this may comprise the VI Group metal oxides such as molybdenum or chromium carried on a suitable spacing agent such as alumina or an alumina-silica gel to which zinc may be added as a promoter. The hydroforming catalyst may also, however, include platinum catalysts or palladium catalysts carried on a suitable support such as alumina, or alumina carrying a small amount of silica, which catalyst may also contain a small amount of hydrogen fluoride. Operating pressures within the range from about 10 to 200 p. s. i. g. and temperatures within the broad limits described above may be chosen based on the activity of the particular catalyst and the characteristics of the particular feed stock employed.

Numerous modifications of the present invention will be apparent to those who are familiar with this particular art without departing from the spirit thereof.

What is claimed is:

1. In a process for the conversion of a naphtha hydrocarbon feed to a product rich in C₆-C₈ aromatics involving the treatment of hydrocarbon vapors with a fluidized hydroforming catalyst and a recycled hydrogen-containing gas under hydroforming conditions of temperature and pressure resulting in the gradual formation of a contaminating carbonaceous deposit on the surface of the catalyst, the improvement which comprises removing from the hydroforming reaction zone a product vapor stream and a stream of spent fluidized catalyst, isolating from said product vapor stream a stream of hydrogen-containing recycle gas, a liquid product stream rich in C₆-C₈ aromatics and a second liquid product stream containing hydrocarbons convertible to C₆-C₈ aromatics under more severe hydroforming conditions of temperature and catalyst to oil ratio, introducing said spent catalyst into a fluid bed maintained in a separate regeneration zone and regenerating the catalyst therein to remove at least a part of said carbonaceous deposit therefrom by an oxidative combustion process whereby the temperature of said catalyst is raised to a level at least 150° F. higher than the temperature of said first-mentioned hydroforming reaction, separately heating said recycle gas stream to at least the same high temperature level, injecting said hot regenerated catalyst and said hot recycle gas into a second hydroforming zone together with said second liquid product stream whereby at least a portion of the hydrocarbons therein are converted to C₆-C₈ aromatics at said higher temperatures utilizing the same catalyst as in the first reaction zone, and conveying the total reaction mixture from said second

hydroforming zone directly into said first hydroforming zone together with fresh hydrocarbon feed vapors to supply the catalyst, recycle gas and heat necessary to maintain the hydroforming reaction therein.

2. The process according to claim 1 in which the ratio of hot catalyst and hot recycle gas to oil in said second high temperature hydroforming zone is at least 5-10 times higher than in said first-mentioned hydroforming zone.

3. The method according to claim 1 in which the hydroformed product from said first reaction zone is separated by fractional distillation into a product stream rich in said low-boiling aromatics and a heavier bottoms fraction rich in higher-boiling aromatics, said second reaction zone operates at a temperature at least 150° F. higher than said hydroforming reaction zone, and said higher-boiling aromatics are hydrodealkylated in said second reaction zone to produce additional low-boiling aromatics.

4. The method according to claim 3 in which said hydroforming reaction takes place at a temperature of from about 850° to 950° F. and said hydro-dealkylation takes place at a temperature of from about 1100° to 1200° F.

5. The method of producing aromatics from a hydrocarbon oil fraction containing naphthenes which comprises vaporizing said oil, contacting said vapors at an elevated temperature and pressure and in the presence of added hydrogen with a moving stream of a fluidized hydroforming catalyst maintained as a catalyst bed in a first reaction zone, forming therein a product including a hydrogen-containing gas and low-boiling aromatics having not more than 7 carbon atoms, separating from said product a normally liquid hydrocarbon stream comprising non-aromatic constituents convertible to said low boiling aromatics by a severe hydroforming treatment, heating a stream of said non-aromatic constituents, a stream of said catalyst and a stream of said hydrogen-containing gas to a temperature higher than that required in said first reaction zone and passing the thus heated materials into a second reaction zone at a temperature at least 150° F. higher than the temperature of the first reaction zone, carrying out in said second reaction zone an endothermic hydroforming reaction utilizing the same catalyst as in the first reaction zone resulting in the formation of additional low-boiling aromatics and conveying the thus cooled catalyst and total reaction product from said second reaction zone directly into said first reaction zone together with fresh hydrocarbon feed vapors so as to supply the catalyst, hydrogen-containing gas and heat required so as to maintain the desired relatively cooler operating temperature in said first reaction zone, and recovering a separate stream rich in said low-boiling aromatics from the product leaving said first reaction zone.

6. The method according to claim 5 in which said hydrocarbon oil fraction has a boiling range from about C₆ to 180° F. and said low-boiling aromatic product consists essentially of benzene.

7. The method according to claim 5 in which said hydrocarbon oil fraction has a boiling range within the limits of from about 180° F. to 250° F. and said low-boiling aromatic product consists essentially of toluene.

8. The method of producing benzene from a vaporized C₆-180° F. naphtha hydrocarbon fraction which comprises passing said naphtha vapors into a hydroforming reaction zone at a temperature of from about 850-950° F. and a superatmospheric pressure of from about 10 to 200 p. s. i. g. in the presence of a bed of fluidized hydroforming catalyst comprising a Group VI metal oxide supported on an alumina-containing catalyst carrier, recovering from said hydroforming reaction a product vapor stream substantially enriched in hydrogen and benzene and at the same time causing a relatively small amount of a carbonaceous deposit to be laid down on the catalyst, separating said product vapor stream to recover therefrom a hydrogen-rich gas, a benzene product stream and a recycle hydrocarbon stream comprising non-aromatic constituents of said product vapor, continuously withdrawing from said hydroforming zone an aliquot portion of the catalyst therein and subjecting said withdrawn catalyst in a fluid bed in a separate regeneration zone to an oxidative combustion treatment to effect the at least partial removal of the carbonaceous deposit therefrom and heat the regenerated catalyst to

a temperature of between about 1100° and 1200° F., separately heating at least a portion of said hydrogen-rich gas and recycle hydrocarbon streams and introducing them together with a stream of said hot regenerated catalyst into a second reaction zone maintained thereby at a temperature of between about 1100° and 1200° F., converting a portion of said recycle hydrocarbon stream therein into additional product benzene, and conveying the total contents of said second reaction zone including hydrogen, hydrocarbons and hot catalyst continuously and directly into said first reaction zone together with said naphtha feed vapors to supply the necessary heat of reaction for said first-named hydroforming reaction.

9. The method of producing low-boiling aromatics from a vaporized naphtha hydrocarbon fraction which comprises passing said naphtha vapors into a hydroforming reaction zone at elevated temperature and pressure in the presence of a bed of fluidized hydroforming catalyst maintained therein, recovering from said hydroforming reaction a product vapor stream substantially enriched in hydrogen and low-boiling aromatics and at the same time causing a relatively small amount of a carbonaceous deposit to be laid down on the catalyst, separating said product vapor stream to recover therefrom a hydrogen-rich gas, a purified low-boiling aromatic hydrocarbon stream and a recycle hydrocarbon stream comprising other normally liquid constituents of said product vapor, continuously withdrawing from said hydroforming zone an aliquot portion of the catalyst therein and subjecting said withdrawn catalyst in a separate regeneration zone to an exothermic chemical treatment to effect the at least partial removal of the carbonaceous deposit therefrom and heat the regenerated catalyst to a temperature at least 150° F. above the temperature in said first-named hydroforming zone, separately heating at least a portion of said hydrogen-rich gas and recycle hydrocarbon streams and introducing them together with a stream of said hot regenerated catalyst into a second reaction zone maintained at essentially the same temperature as said regeneration zone, converting a portion of said recycle hydrocarbon stream therein into additional low-boiling aromatics, and conveying the total contents of said second reaction zone including hydrogen, hydrocarbons and hot catalyst continuously and directly into said first reaction zone to replenish the bed of catalyst therein and supply a major part of the necessary heat of reaction for said first-named hydroforming reaction.

10. In the process of converting vaporized hydrocarbons in the naphtha boiling range into low-boiling aromatics by a hydroforming treatment at elevated temperature and pressure with a moving stream of particles of a finely divided hydroforming catalyst, the improvement which comprises separating the total hydroformed product stream into a recycle gas stream, a product aromatic stream and a recycle hydrocarbon stream, said recycle hydrocarbon stream including non-aromatic hydrocarbons convertible into low boiling aromatics by a second stage hydroforming reaction at the same pressure and at a temperature at least 150° F. higher than that of said first-named hydroforming treatment and including also higher-boiling aromatic hydrocarbons convertible into low-boiling aromatics by a concomitant hydro-dealkylation reaction in the presence of the same hydroforming catalyst at said pressure and said higher temperature, continuously withdrawing from the zone of said first-named hydroforming treatment a stream of catalyst particles cooled by said treatment, heating said catalyst to such a higher temperature and introducing said hot catalyst directly into a second treating zone together with preheated streams of said recycle hydrocarbon, continuously conveying a stream of hot catalyst and reaction products from said second treating zone directly into said first treating zone, together with fresh naphtha feed vapors, to supply the necessary heat of reaction for said hydroforming treatment, and thereby recovering an improved yield of low-boiling aromatics in said product aromatic stream.

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