July 23, 1968 J. ENG ET AL 3,394,199

HYDROCARBON CONVERSION PROCESS

Filed Feb. 20, 1961 2 Sheets-Sheet 1

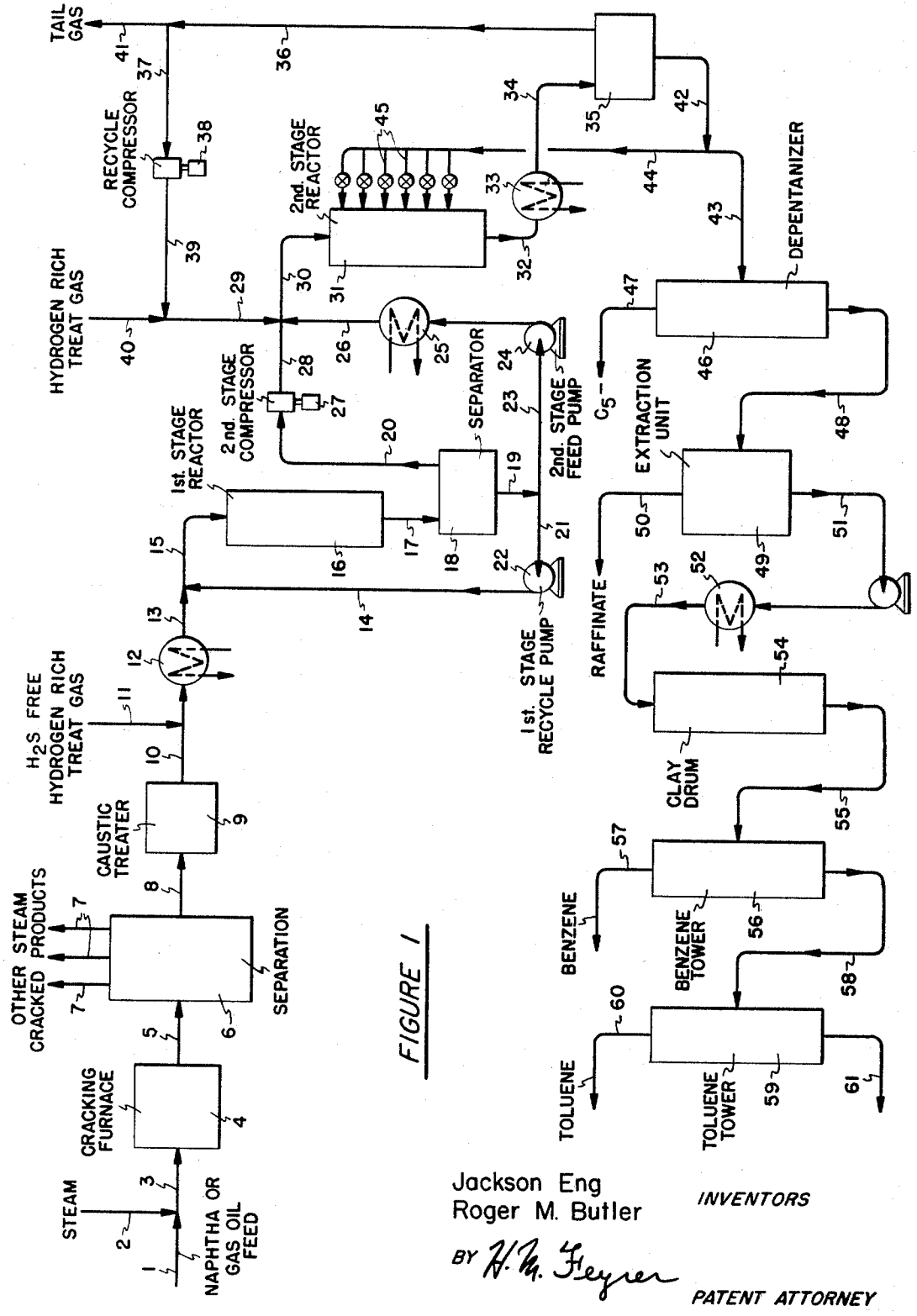


FIGURE 1

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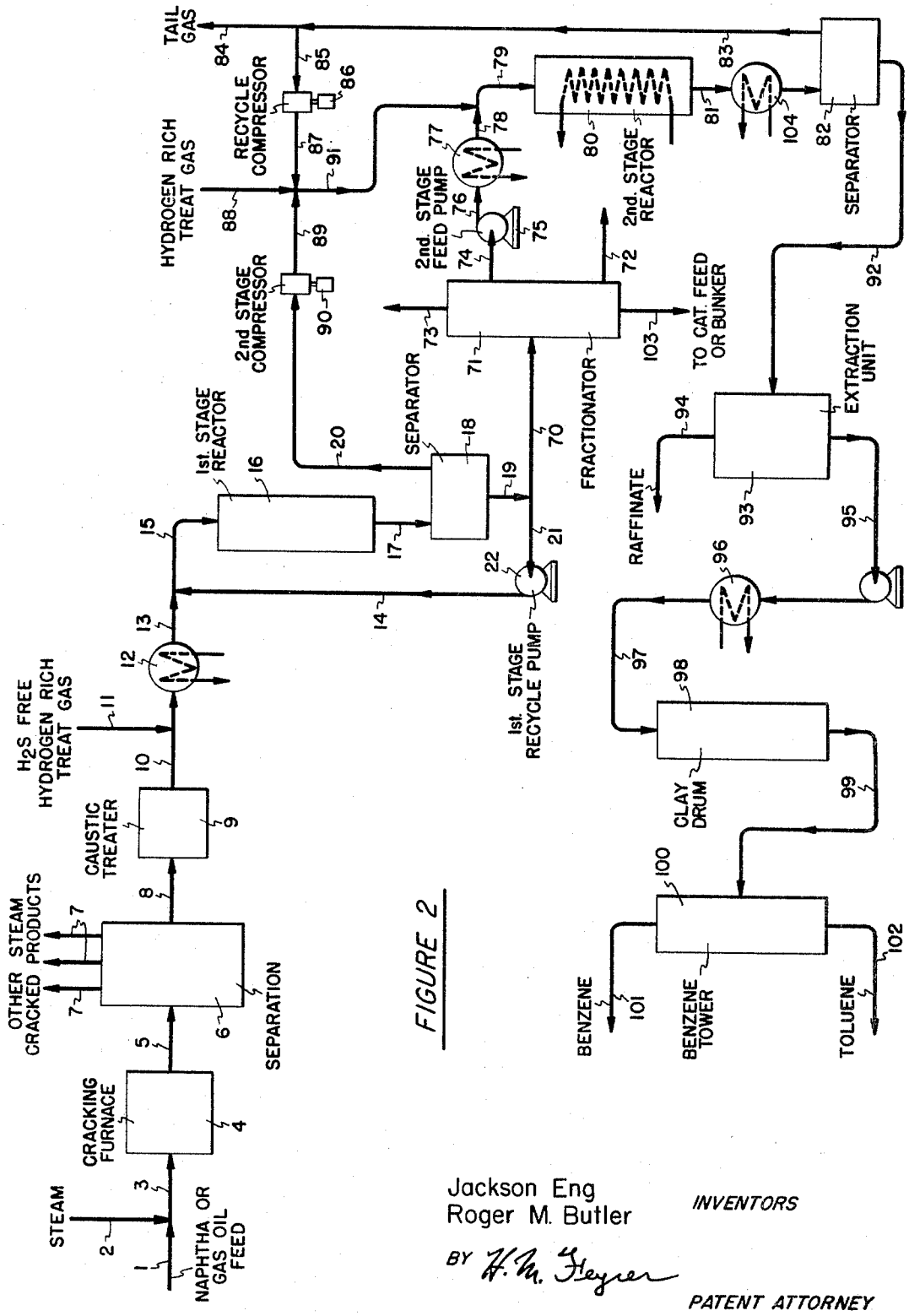


FIGURE 2

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3,394,199

HYDROCARBON CONVERSION PROCESS

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14 Claims. (Cl. 260—674)

The present invention pertains to an improved hydrocarbon conversion process and, more particularly, to an improved method for the production of benzene.

Steam cracking of hydrocarbon feedstocks is a well known and widely used process for the production of petrochemicals, particularly ethylene, propylene, and butadiene. A variety of feedstocks boiling up to and including the gas oil boiling range can be used, although usually a light gasoline or naphtha charge stock is used. The steam cracked product comprises, in addition to ethylene, propylene, and butadiene, a gasoline fraction and a residue or fuel oil fraction. The gasoline fraction has a Research Octane Number of about 95 to 100 and is, therefore, useful as a motor gasoline blending component. The steam cracked naphthas consist chiefly of aromatics and olefins, and the 140–200° F. fraction is especialy rich in benzene (approximately 40–70 vol. percent), and, therefore, a potential feedstock to an aromatics extraction plant. Glycol-type solvents are conventionally used for the extraction of aromatics from hydrocarbon fractions. Unfortunately, glycol-type solvents do not give a sharp separation of aromatics from olefins and particularly diolefins. A further difficulty is that appreciable quantities of thiophene occur in some steam cracked naphthas. This is extracted with the benzene resulting in an impure product. It is the object of this invention to provide an improved method for preparing aromatic hydrocarbons and, in particular, benzene.

It is also an object of this invention to provide an improved method for extracting aromatics from steam cracked naphthas.

It is a further object of this invention to provide a method for treating steam cracked naphthas to render them more amenable to extraction with glycol-type solvents.

These and other objects will appear more clearly from the detailed specification and claims which follow.

It has now been found that aromatic hydrocarbons can be efficiently extracted from steam cracked naphthas by means of glycol-type solvents, provided that the naphtha fraction is freed of diolefin or other polyunsaturates, and substantially freed of monoolefins prior to the extraction treatment.

In accordance with the present invention, the diolefins and monoolefins in steam cracked naphthas are saturated in an efficient two-stage hydrogen treatment. In the first stage, the naphtha is hydrofined with a reduced nickel catalyst in order to largely eliminate the diolefins and thereby stabilize the naphtha stream. Because of the high activity of the nickel metal catalyst, operations can be carried out at about 300° F. At this low temperature, polymer and coke formation are avoided and, consequently long catalyst life is possible. After the first stage treatment, the stabilized naphtha is hydrofined with the

usual cobalt oxide-molybdenum oxide type catalyst. While this second stage treatment requires temperatures of about 500° F. or higher, there is no polymer and coke formation problem in substantially completely saturating the rest of the monoolefins in this stage because the unstable diolefins have been removed. The treated product from the second stage is completely satisfactory for charging to the solvent extraction unit.

Reference is made to the accompanying drawings, (FIGURES 1 and 2) which illustrate diagrammatic flowplans of two embodiments of the process of the present invention.

Referring to FIGURE 1, a naphtha or gas oil feed is introduced through line 1 and mixed with the steam supplied through line 2 and enters cracking furnace 4 via line 3. The quantity of steam added is controlled so as to make the concentration of steam within the furnace from 50 to 90 mole percent, preferably 80 mole percent. In the cracking furnace, the temperature is raised to 1300 to 1450° F., preferably 1400 to 1440. The pressure within the cracking furnace is normally in the range of 10 to 20 p.s.i.g. The residence time in the furnace at cracking temperature is 0.2 to 2 seconds. The cracked products are discharged from the furnace via line 5 to separation equipment 6. This equipment is of a conventional nature and involves fractionation and, in some cases, extraction. The steam cracked naphtha is withdrawn from the separator, via line 8. Other steam cracked products leave via lines 7. The steam cracked naphtha is in the boiling range of C_5 to 400, preferably C_5 to 300. The naphtha in line 8 goes to the caustic treater 9 in which it is given a mild treat to remove mercaptans. Typical conditions would consist of a 10% treat with 20° Bé. caustic. The mercaptan-free naphtha leaves the treater via line 10 and is mixed with a stream of hydrogen sulfide-free, hydrogen-rich, treat gas, which enters via line 11. This gas will normally consist of caustic washed reformer tail gas and will have a concentration of about 80 mole percent hydrogen. Normally the quantity of treat gas will be in the range of 750 to 2000 standard cu. ft./bbl. of fresh feed. The mixed naphtha and hydrogen proceeds through heat exchanger 12, then via lines 13 and 15 into the first stage reactor 16. Recycled product is mixed with the naphtha via line 14. The heat added in exchanger 12 is controlled so that the feed entering the reactor is in the range of 250–300° F. Normally, low temperatures will be required with fresh catalyst and the temperature may be raised to the upper limit as the catalyst deactivates. The catalyst in the first stage reactor 16 is a reduced nickel type catalyst.

This may consist of the familiar type of catalyst known as nickel supported on kieselguhr. Such catalysts normally contain about 10 to 50 weight percent nickel, preferably 40 to 50. The preparation of a suitable nickel-on-kieselguhr catalyst has been described by L. W. Covert, R. Connor and H. Adkins (J.A.C.S., 54, 1951 (1932)). Catalysts of this type may be used in granular, pelleted or extruded form. Cylindrical tablets one-eighth inch in diameter by one-eighth inch tall are satisfactory although many other sizes can also be used.

Other types of supported nickel catalyst which can be used consist of nickel on precipiated calcium carbonate

and nickel on alumina. The pressure in **16** is in the range of 150–300 p.s.i.g. and the liquid hourly space velocity is in the range of .5 to 4 v./v.-hr. based on fresh feed. The product from the reactor proceeds via line **17** into separator **18**. Here the gas phase leaves via line **20** and the liquid phase via line **19**. Normally, the separator will be at a higher temperature than the feed inlet because of the considerable exothermic heat of reaction This is controlled by recycling product via line **21** and pump **22** and line **14**. The quantity of this recycle will normally be in the volume of 3 to 10 times that of the fresh feed. It should be chosen such that the temperature rise in the first stage reactor is less than 100° F., preferably less than 50° F. The liquid product is pumped via line **23** and pump **24**, through heater **25** line **26** and line **30** into the second stage reactor **31**. Hydrogen-rich gas from the first stage separator **18** passes through line **20**, is compressed by compressor **27**, and passes through line **30**. Recycle hydrogen and fresh hydrogen-rich treat gas also mix with the feed via line **29**. The second stage reactor contains a sulfided cobalt molybdate-on-alumina catalyst. The second stage hydrotreat vessel is charged with a conventional metal oxide-type hydrogenation catalyst, preferably a cobalt molybdate catalyst or one containing cobalt oxide and molybdenum oxide dispersed upon an alumina-containing support or base, preferably activated or adsorptive alumina. In general, such catalysts are prepared by first forming adsorptive alumina particles or hydrous alumina which on heating becomes adsorptive, and then compositing molybdenum oxide and cobalt oxide therewith. The molybdenum oxide can, for example, be added as a slurry or it may be applied as a solution of ammonium molybdate. The cobalt oxide is conveniently added as a salt such as cobalt nitrate or acetate, salts which are readily decomposed to cobalt oxide and volatile materials. The cobalt oxide and molybdenum oxide may be provided in equimolar amounts or a molecular excess of one over the other may be used. Suitable catalysts contain from about 5 to about 25 wt. percent cobalt oxide and molybdenum oxide, with the ratio of the former to the latter in the range of from about 1/5 to about 5/1. The pressure in this reactor is in the range of 400 to 1000 p.s.i.g., preferably 500 to 800 p.s.i.g. The inlet temperature is controlled by means of heater **25** to be within the range of 550 to 600, preferably about 575. Again there is a large exothermic heat of reaction and in the process shown in FIGURE 1, this is controlled by the addition of recycled product at intermediate points within the reactor. This cold quench will be at a temperature of 50 to 150° F. and will be in the liquid phase. As it enters the reactor, considerable vaporization occurs thus absorbing the heat of the reaction. The outlet temperatures from the reactor are maintained within the range of 550 to 700° F. and preferably within the range of 575–675° F. The liquid hourly space velocity in the second stage is in the range of .5 to 4 v./v.-hr. based on fresh feed. The product leaves the reactor via line **32** and is cooled and partially condensed in cooler **33**. If the proceeds to separator **35** via line **34**. The liquid product leaves the separator via line **42** and is divided into recycle which proceeds through line **44** to inlets **45** at reactor **31** and product which goes through line **43**. The gas from the separator proceeds through line **36** and is partially recycled via line **37**, compressor **38** and line **39**. The remainder of the gas leaves the unit via line **41**. The total hydrogen entering the second stage reactor is controlled so that it is in the range of 1000 to 3000 standard cu. ft. of hydrogen/bbl. of total liquid feed. The concentration of hydrogen in the product line **32** should be a minimum of 40 mole percent. The doubly hydrofined product enters the depentanizer **46** via line **43**. Here the pentanes and lighter components are taken overhead via line **47** and the higher boiling fraction leaves via line **48** to the extraction unit **49**.

This extraction unit may employ any conventional solvent such as diethylene glycol-water mixtures, triethylene glycol-water mixtures, dipropylene glycol-water mixtures or combinations of these solvents, sulfur dioxide, the sulfolanes, etc. The design and operation of such units is well known in the art. The raffinate consisting essentially of saturated hydrocarbons leaves the extraction unit via line **50**. The aromatic extract is pumped via line **51** through heater **52**. Here it is raised to a temperature of the order of 400–600° F., preferably 500° F. at a pressure of 200 to 500 p.s.i.g. The heated aromatic extract passes via line **53** into clay treating drum **54**. This contains active clay particles such as, for example, that known as *attapulgite*. The quantity of clay in drum **54** is chosen so as to make the liquid hourly space velocity about .5 to 1 v./v.-hr. The purpose of the clay treating step is to remove any traces of olefins and diolefins which may be present. The clay treated aromatic extract passes via line **55** to the benzene tower **56** in which an essentially pure benzene product is removed via line **57**. The bottoms from tower **66** passes through line **58** to tower **59** where essentially pure toluene is removed overhead in line **60** and heavier aromatics in line **61**.

FIGURE 2 shows a variation of the process presented in FIGURE 1. In this scheme, a gasoline fractionator is incorporated between the first and second stage hydrofiners. The purpose of this fractionator is to remove 140- and 200/400 streams that are suitable for incorporation into motor gasolines without further treatment. The process is exactly the same up to and including separator **18**. The first stage hydrofiner product from the separator passes through line **70** into the fractionator **71**. Here a 140–200° F. boiling range benzene heart cut is prepared. This passes via line **74** into second stage feed pump **75**. The 140- overhead stream and the 200–400 stream from the fractionator may be used as motor gasoline blending components. The 400+ fraction from the fractionator is highly aromatic and is unsuitable for gasoline or heating oil. It may be used either as a naphthalene concentrate or incorporated in cat cracker feed or bunker fuel. The second stage feed passes from the feed pump **75** through line **76** through heater **77** and line **78** and is mixed with hydrogen-rich treat gas which enters via line **91**. The combined feed and gas enter the second stage reactor **80** via line **79** at a pressure of 400–1000 p.s.i.g., preferably 500–800 p.s.i.g. The inlet temperature is controlled by means of heater **77** to be within the range of 550 to 600° F., preferably about 575° F. The exothermic heat of reaction is controlled in the process shown in FIGURE 2 by means of indirect internal reactor cooling, instead of the cooled quench injection shown in FIGURE 1. This is merely an alternative method. The design of these coolers should be such that the reactor temperature does not rise above 700° F. and preferably not above 675. The reactor products leave via line **81** and are cooled and partially condensed in heat exchanger **104**. The liquid is separated from the gas in separator **82**. The gas leaves via line **83** and is partially recycled via line **87** and line **91**. Tail gas may, if it is desired, pass to a hydrogen purification plant. The liquid from the separator **82** passes via line **92** into the extraction unit **93**. This extraction unit may employ any of the conventional solvents described in the discussion of FIGURE 1; its purpose is to separate the feed into an essentially saturated raffinate which leaves via line **94** and a saturate-free benzene concentrate which leaves via line **95**. This is clay treated by means of heater **96** line **97** and clay drum **98** at conditions similar to those used in FIGURE 1. The clay treated product passes through line **99** into the distillation tower **100** and essentially pure nitration grade benzene is removed overhead through line **101** and a nearly pure toluene stream is removed through line **102**.

The following examples are illustrative of the present invention.

Example 1

A gas oil derived from Western Canadian Crude is fed to a steam cracker.

GAS OIL FEEDSTOCK INSPECTIONS

° API, ASTM distillation	29.6
I.B.P.	438
5%	530
10%	571
50%	646
90%	732
F.B.P.	751
Wt. percent sulfur	0.70

In the cracker it is mixed with 160 lb. of steam per barrel of oil and maintained at 1430° F. for a residence time of 0.8 second. 45 wt. percent of the feed is converted to C_3 hydrocarbons and lighter. A 22 wt. percent yield of naphtha based on feed is obtained. After caustic treatment this has the inspections shown in column 1 of Table 1.

This naphtha is hydrofined over a reduced nickel on kieselguhr catalyst in the form of 8–12 mesh granules. The conditions used are:

Reactor temperature	300° F.—maintained essentially constant by cooling.
Pressure	200 p.s.i.g.
Treat gas	1000 s.c.f./bbl. of reformer tail gas.
Space velocity	1 v./v./hr.
Hydrogen consumption	470 s.c.f./b.

The inspections on the first stage hydrofined product are shown in column 2 of Table 1.

This material is fed to the second stage hydrofiner where it is treated over a sulfided cobalt molybdate-on-alumina catalyst of 5/64" diameter extrudate. The conditions used are:

Reactor temperature	625° F.
Pressure	500 p.s.i.g.
Treat gas	2000 s.c.f./bbl. of reformer tail gas.
Space velocity	1 v./v./hr.
Hydrogen consumption	580 s.c.f./b.

The inspections of the second stage hydrofined product are shown in column 3 of Table 1. This material is now depentanized and approximately 15 volume percent of pentanes are removed.

The remaining material is extracted in a Udex plant using diethylene glycol-water solvent and the extract is clay treated and distilled into essentially pure benzene and a bottom product consisting of toluene and higher boiling aromatics.

TABLE 1

	Naphtha Feed to 1st Stage	Product from First Stage	Product from Second Stage	Benzene Product
LV% Yield on Naphtha Feed	100	100	100	11.7
API	32.5	33.0	33.5	28.6
Sulfur, p.p.m.	1,700	1,700	----	2
Bromine No.	130	82	6	<0.5
Diene No.	30	5	<1	<0.1
Exist Gum [1] mg./100 ml.	5	1	1	[2]
Accel. Gum [1]	>3,000	----	----	----
Benzene, LV%	12	----	----	>99.5
Thiophene, p.p.m.	----	----	----	<10
MON Clear	83.1	85.0	----	----
+3 cc. TEL	84.5	86.3	----	----
RON Clear	98.1	98.2	----	----
+3 cc. TEL	99.2	99.9	----	----
15/5 Dist. I.B.P.	87	----	----	----
5%	97	----	----	----
10%	112	----	----	----
50%	235	----	----	----
90%	417	----	----	----
95%	428	----	----	----
F.B.P.	475	----	----	----
Freezing Pt., ° C.	----	----	----	5.3
Acid Wash Color	----	----	----	1
Acidity	----	----	----	[1]
Copper Corrosion	----	----	----	[2]

[1] Inhibited with 10 lb./MB (Active) Dupont No. 5.
[2] Nil.
[3] Pass.

Example 2

The results obtainable by employing the scheme illustrated in FIGURE 2 using the same feedstock and cracking conditions described in Example 1 are summarized below in Table 2.

TABLE 2

	Naphtha Feed to 1st Stg.	Product from 1st Stg.	200/400° F. Naphtha from 1st Stg.	140/200° F. Fraction to 2nd Stg.	Product from 2nd Stg.	Benzene Product
LV% Yield on Naphtha Feed	100	100	50.2	22.1	22.6	12.3
API Gravity	----	----	33.1	43.3	44.0	28.6
Sulfur, p.p.m.	----	----	1,400	95	8	1
Bromine No.	----	----	42	49	3	<0.5
Diene No.	----	----	4.1	2.1	<0.5	<0.1
Exist Gum, mg./100 ml [1]	----	----	1.0	1.0	----	[2]
Accel. Gum, mg./100 ml [1]	[3]	[3]	4.8	----	----	----
15/5 Dist. I.B.P.	----	----	195	135	----	----
5%	----	----	----	----	----	----
10%	----	----	----	----	----	----
50%	----	----	310	180	----	----
90%	----	----	----	----	----	----
95%	----	----	----	----	----	----
F.B.P.	----	----	406	205	----	----
Benzene, LV%	----	----	----	56	56	>99.5
Thiophene, p.p.m.	----	----	----	[4] 750	3.5	5.9
Freezing Pt., ° C.	----	----	----	----	----	5.2
Acid Wash Color	----	----	----	----	----	1
Acidity	----	----	----	----	----	[2]
Copper Corrosion	----	----	----	----	----	[5]

[1] Inhibited with 10 lb./MB (octane)Dupont No. 5.
[2] Nil.
[3] See Table 1.
[4] Value may be high due to olefin interference.
[5] Pass.

The foregoing description contains a limited number of embodiments of the present invention. It will be understood, however, that numerous variations are possible without departing from the scope of this invention.

What is claimed is:

1. A process for the separation of aromatics from a feedstock mixture containing, in a major proportion, said aromatics and mono-olefins, and a minor proportion of gum-forming constituents, comprising the steps of:

(1) contacting said feedstock in a first hydrogenation stage with a supported nickel catalyst which, under operating conditions, contains a major proportion of nickel in elemental form, in the presence of a gas in which the reactant thereof consists essentially of hydrogen and at an elevated temperature, whereby at least a substantial proportion of the gum-forming constituents are destroyed;

(2) passing at least a portion of the product stream from said first stage to a second, olefin removing, hydrogenation stage and contacting said portion with a hydrogenation catalyst different from said supported nickel catalyst of said first stage in the presence of hydrogen and at an elevated temperature whereby at least a substantial proportion of the mono-olefins in said portion are converted into paraffins; and

(3) recovering the aromatics from the product stream of said second stage.

2. A process according to claim 1, wherein a separate hydrogen feed is maintained to each of said hydrogenation stages.

3. A process for the separation of aromatics from a feedstock consisting of a raw steam cracked gasoline, comprising the steps of:

(1) distilling said raw steam cracked gasoline and recovering a light naphtha fraction therefrom predominating in hydrocarbons, said hydrocarbons being predominantly aromatics and mono-olefins with a minor proportion of gum-forming constituents;

(2) passing said fraction to a first, hydrogenation stage and contacting said fraction with a supported nickel catalyst which, under operating conditions, contains a major proportion of nickel in elemental form, in the presence of a gas in which the reactant thereof consists essentially of hydrogen and at an elevated temperature whereby at least a substantial proportion of said gum-forming constituents are destroyed without substantial hydrogenation of the aromatic and mono-olefin constituents;

(3) passing at least a portion of the product stream from said first stage to a second, olefin removing, stage and contacting said portion with a hydrogenation catalyst different from said supported nickel catalyst of said first stage in the presence of hydrogen and at an elevated temperature to convert the said mono-olefins of said fraction to paraffins; and,

(4) thereafter recovering the aromatics from the product stream of said second stage.

4. A process for the separation of aromatics from a feedstock consisting of raw steam cracked gasoline containing predominantly aromatics and mono-olefins with a minor proportion of gum-forming constituents, comprising the steps of:

(1) contacting said feedstock with a supported nickel catalyst which, under operating conditions, contains a major proportion of nickel in elemental form, in the presence of a gas in which the reactant thereof consists essentially of hydrogen and at an elevated temperature, whereby at least a substantial proportion of the gum-forming constituents are destroyed;

(2) passing at least a portion of the product stream from said first stage to a second, olefin removing, hydrogenation stage and contacting said portion with a hydrogenation catalyst different from said supported nickel catalyst of said first stage in the presence of hydrogen and at an elevated temperature whereby at least a substantial proportion of the mono-olefins in said portion are converted into paraffins; and

(3) recovering the aromatics from the product stream of said second stage.

5. A process for the separation of aromatics from a feedstock mixture containing aromatics and mono-olefins in major proportions and gum-forming constituents in a minor proportion which comprises:

(1) contacting said feedstock in a first hydrogenation stage with a supported nickel catalyst which, under operating conditions, contains a major proportion of nickel in elemental form, in the presence of a gas in which the reactant thereof consists essentially of hydrogen and at a temperature in the range 250–400° F., whereby at least a substantial proportion of the gum-forming constituents are destroyed,

(2) passing the product stream from said first stage to a second olefin removing hydrogenation stage and contacting said product stream in the presence of hydrogen with a cobalt oxide-molybdenum oxide catalyst on a support, at a pressure in the range 400 to 1000 p.s.i.g., a temperature in the range 550–700° F., and at a space velocity of 0.5 to 4.0 v./v./hr., whereby at least a substantial proportion of the mono-olefins in said product stream are converted into paraffins, and

(3) recovering the aromatics from the product stream of said second stage.

6. A process for the preparation of high purity aromatics which comprises:

(1) steam cracking a hydrocarbon feedstock boiling up to and including the gas oil boiling range,

(2) separating a naphtha fraction containing mono-olefins and polyolefins from the steam cracked products,

(3) treating the naphtha fraction with hydrogen in contact with a catalyst comprising reduced nickel on a support at temperatures of from about 250–400° F. and pressures of about 150–300 p.s.i.g. to hydrogenate a major proportion of the polyolefins therein in a first hydrofining stage,

(4) treating the hydrofined naphtha product from the first hydrofining stage with hydrogen in contact with a metal oxide type hydrogenation catalyst at temperatures of about 550–700° F. and pressures of about 400–1000 p.s.i.g. in a second hydrofining stage to substantially saturate the mono-olefins therein,

(5) subjecting the hydrofined liquid product from the second hydrofining stage to solvent extraction.

7. The process as defined in claim 6 wherein the temperature in the first hydrofining stage is controlled by recycling hydrofined liquid product thereto.

8. The process as defined in claim 6 in which hydrofined liquid product from the second hydrofining stage is depentanized before extraction.

9. The process as defined in claim 6 in which the aromatics fraction extracted from the hydrofined liquid product from the second hydrofining stage is clay treated at 400–600° F. to remove any traces of olefins or diolefins therefrom.

10. Process according to claim 6 in which the solvent is sulfolane.

11. Process according to claim 6 in which the solvent is a glycol-type solvent.

12. A process for the preparation of high purity benzene and toluene which comprises:

(1) steam cracking a hydrocarbon feedstock boiling up to and including the gas oil boiling range,

(2) separating a naphtha fraction containing mono-olefins and polyolefins from the steam cracked products,

(3) treating the naphtha fraction with hydrogen in contact with a reduced nickel catalyst at temperatures of from about 250–400° F. and pressures of about 150–300 p.s.i.g. to hydrogenate a major proportion of the polyolefins therein in a first hydrofining stage,

(4) fractionating the hydrofined naphtha product from the first hydrofining stage into a light fraction boiling up to 140° F., a 140–200° F. aromatics-rich heart cut and a heavy 200–400° F. fraction,

(5) treating the 140–200° F. fraction with hydrogen in contact with a metal oxide type hydrogenation catalyst at temperatures of about 550–700° F. and pressures of from about 400–1000 p.s.i.g. in a second hydrofining stage to substantially saturate the mono-olefins therein,

(6) recovering a hydrofined liquid product having a maximum bromine number of about 5,

(7) subjecting said hydrofined liquid product to solvent extraction to extract pure benzene and toluene therefrom.

13. A process for the preparation of high purity benzene and toluene which comprises:

(1) steam cracking a hydrocarbon feedstock boiling in the gas oil boiling range at temperature of at least about 1300° F.,

(2) separating a naphtha fraction from the steam cracked products,

(3) caustic washing said naphtha fraction to remove mercaptans therefrom,

(4) treating the caustic washed naphtha fraction with hydrogen in contact with hydrogen in contact with a reduced nickel catalyst at temperatures of from about 250–400° F. and pressures of about 150–300 p.s.i.g. to hydrogenate a major proportion of the polyolefins therein in a first hydrofining stage,

(5) fractionating the hydrofined naphtha product from the first hydrofining stage into a light fraction boiling up to 140° F., a 140–200° F. aromatics-rich heart cut and a heavy 200–400° F. fraction,

(6) treating the 140–200° F. fraction with hydrogen in contact with a metal oxide type hydrogenation catalyst at temperatures of about 550–700° F. and pressures of from about 400–1000 p.s.i.g. in a second hydrofining stage to substantially saturate the mono-olefins therein,

(7) subjecting the hydrofined liquid product from the second hydrofining stage to solvent extraction with a glycol-type solvent to extract pure benzene and toluene therefrom.

14. Process according to claim **13** in which said metal oxide type catalyst is sulfided.

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