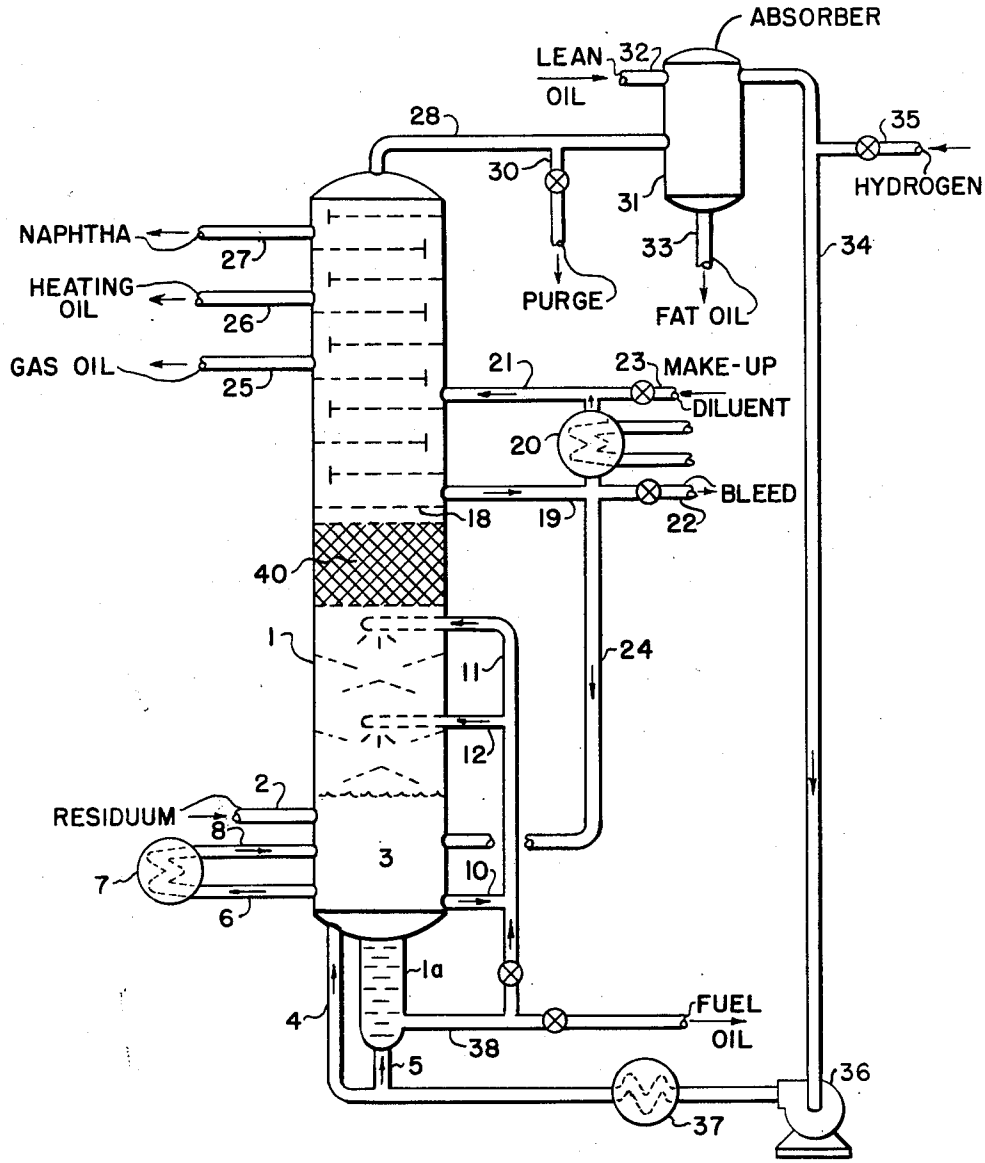
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HYDROGEN DONOR DILUENT CRACKING PROCESS

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HYDROGEN DONOR DILUENT CRACKING PROCESS

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This invention pertains to the converting of hydrocarbon oils by hydrogen donor diluent cracking (HDDC). More particularly it is concerned with the upgrading of petroleum oils, particularly heavy high boiling oils such as vacuum residua, by thermal cracking of the oil in the presence of a hydrogen donor diluent. By the process of this invention, low value hydrogen-deficient oils having coke forming propensities during thermal and catalytic cracking are converted to more volatile products, such as light distillates for use as motor fuels and gas oils suitable as feed stocks for catalytic cracking processes. A salient feature of this invention is that the cracking reaction, hydrogenation of the diluent, and separation of the conversion products occur in a single tower.

A process termed hydrogen donor diluent cracking (HDDC) has recently been proposed. In this process, a heavy, low value oil is upgraded by admixing it with a hydrogen donor diluent material, aromatic-naphthenic in nature, and cracking the mixture thermally as in a thermal cracking coil. The donor diluent is an aromatic-naphthenic material that has been especially prepared by partial hydrogenation from selected, normally surplusage, refinery streams such as catalytic cycle stocks and bottoms fractions, thermal tars therefrom, etc. The hydrogen donor diluent used has the ability to take up hydrogen in a hydrogenation zone and readily release it to a hydrogen-deficient oil in a thermal cracking zone. In this manner of hydrocracking oils, the oil being upgraded, usually a residual oil, is not contacted directly with the hydrogenation catalyst and does not, therefore, impair the activity of the catalyst by contamination and carbon deposition. The amount of concomitant light gases and coke produced by this process is relatively small, usually being about 5 to 10%. This technique of HDDC is more fully presented by co-pending application entitled, "Upgrading of Heavy Hydrocarbon Oils," by Langer, Serial No. 365,335, filed July 1, 1953, now abandoned.

In the previously described processes, the donor diluent and the conversion products, along with unconverted material, have been removed from the reactor as an equilibrium mixture of liquid and gases. The withdrawn materials were then separated to obtain the spent donor diluent, which was partially hydrogenated and recycled, and the unconverted refractory constituents of the feed, which were also recycled to obtain high conversions. Thus, in the past, maximum use has not been made of reactor capacity. The diluent has been used inefficiently because of its low concentration, and variations in product selectivities were difficult to obtain.

In previous HDDC processes wherein the conversion products and spent diluent are continuously removed from the cracking reactor, it is necessary to use a high diluent to residuum ratio in order to obtain high conversions, and also, it is necessary to recycle the unconverted constituents of the feed. The present invention proposes a novel method of operating the hydrogen donor diluent cracking process such that these disadvantages are overcome.

According to the present invention, a hydrogen donor diluent, described hereinafter, is admixed with a hydrogen-deficient oil and the resulting mixture is subjected to liquid phase thermal cracking in the lower portion of a reaction vessel. Vaporous conversion products are continuously removed from the reaction mixture and passed upwardly through the vessel. A free hydrogen-containing gas is bubbled or passed through the reaction mixture and then passed up through the tower with the vapors. The vapors and the hydrogen gas first pass through a catalytic hydrogenation zone and then are separated in the upper portion of the reaction vessel into various boiling range fractions. The diluent is selected such that it is the highest boiling range fraction removed as a vapor and thus it is the first fraction separated from the vapors. The regenerated donor diluent is separated from the vapors in the upper portion of the tower and refluxed through the hydrogenation zone to the reaction mixture.

By continuous withdrawal of the conversion products and spent diluent, the ratio of diluent to unconverted material in the liquid phase is greatly increased. Consequently, coke deposition during the cracking reaction is greatly minimized or eliminated, and much more efficient use is made of the diluent than has been made in previously proposed processes.

Because the hydrogenation of the spent diluent and the separation of the products occur in the same tower, certain heat economies are effected. By addition of the hydrogen-containing gas to the reaction mixture, which is not normally done in conventional HDDC processes, the problem of coking during the cracking of a heavy oil is particularly alleviated. Conventional HDDC is carried out by slow passage of the high temperature liquid reaction mixture through a vessel, thus giving coke an excellent chance to deposit on the vessel walls. In the present invention, the free hydrogen-containing gas is bubbled through the liquid, thus keeping it in a constant state of agitation. The presence of the hydrogen gas also prevents the formation of coke in an auxiliary manner, i. e., by direct hydrogenation.

The fractionation of the conversion products is carried out to an extent such that a gas having a high concentration of free hydrogen is removed overhead. By this means, the gas may be recycled with perhaps some concentration of the hydrogen in the gas as by absorption or adsorption of light hydrocarbons from the recycled stream and by addition of make-up hydrogen.

Thus, the present invention is concerned with a hydrogen diluent donor cracking process wherein heat soaking of the residuum-hydrogen donor diluent mixture, hydrogenation of spent donor, fractionation of products, and separation of the hydrogen gas for recycle are all carried out in the same tower. The process is carried out under relatively high pressures in the order of 100 to 600 p. s. i. to facilitate the various reactions and separations. The pressure may vary about ±10 p. s. i. over the tower under the reaction conditions.

A subsidiary feature of this invention is to utilize a hydrogenation gas, admitted to the base of the reactor tower, to strip a stream of high boiling bottoms removed from the tower to prevent the loss of diluent and lighter conversion products with the fuel oil bottoms.

In brief compass, the present invention proposes a hydrogen donor diluent cracking process which comprises maintaining in the bottom of a reaction tower under liquid phase hydrogen donor diluent cracking conditions a reaction mixture comprising a hydrogen-deficient charging stock and a hydrogen donor diluent, introducing a free hydrogen-containing gas into the bottom portion of the reaction tower, continuously withdrawing overhead from said reaction mixture reaction vapors comprising spent hydrogen donor diluent, hydrogen, and conversion products, passing the vapors under hydrogenating conditions upwardly through an intermediate portion of the reaction tower in the presence of a hydrogenation catalyst whereby the spent hydrogen donor diluent is regenerated, fractionating in the upper portion of the tower the hydrogenated vapors to obtain the hydrogen donor diluent and then products of intermediate boiling range, leaving free hydrogen-containing vapors, and returning the hydrogen donor diluent through the hydrogenation zone to the reaction zone.

Because materials not suitable as donor diluents will be cracked into the boiling range of the diluent, and because some of the donor diluent will be lost through the cracking reaction, a portion of the hydrogenated diluent is bled from the process as product and make-up diluent is refluxed through the hydrogenation zone to the reaction mixture.

In a preferred embodiment of the invention, the hydrogen-containing vapors removed overhead are subjected to a separation process to remove diluting light hydrocarbon gases, and make-up hydrogen gas is admitted to the vapors before they are recycled. Preferably the recycled vapors are heated to supply heat to the reaction zone.

It is an object of this invention to convert heavy hydrogen-deficient oils to more economically desired compounds. Another object is to propose a hydrogen donor diluent cracking process, and apparatus therefor, characterized by continuous vapor phase withdrawal of the conversion products.

These and other objects and advantages will become apparent as this description proceeds and the attached drawing, forming a part of this specification, is described in detail. The drawing schematically presents a preferred embodiment of this invention adapted to the conversion of petroleum residua. The drawing is presented for the purpose of illustration only and the invention is not to be limited thereby.

While the charging stocks for the present invention are preferably hydrogen deficient petroleum derived oils, such as crude oils, distillate and residual fractions therefrom, petroleum extracts, or mixtures thereof, this invention is capable of enjoying broader applications. Thus, coal tars, shale oils, tars, asphalts, etc., may also be processed by the present invention.

The sources of the aromatic-naphthenic oil that is partially hydrogenated to secure the hydrogen donor diluent have previously been described. In particular, it has been found that the thermal tars obtained by the thermal cracking of catalytic cycle stocks yield excellent donor diluents. It has also been found, however, that certain lube oil extracts, extracts of catalytic cycle stocks, heavy cycle stocks themselves, or bottoms from catalytic cracking can serve as a source of the donor diluent. In any case, the prime consideration for the selection of a donor diluent is that the diluent should be composed of predominant proportions of aromatic-naphthenic molecules or condensed ring structures having the ability to take up hydrogen in a hydrogenation zone and to release it in a thermal cracking zone. Such condensed ring structures are relatively refractory and will pass through the thermal cracking zone relatively unaltered. The condensed rings structure is susceptible to being recovered from the reaction mixture and being regenerated by partial hydrogenation. However, while a major proportion

of the donor diluent material can be continuously reused, there will normally be some loss of the donor diluent and this loss is made up from extraneous sources.

In the present invention, the boiling range of the diluent is particularly chosen such that the diluent will be, at the reaction or cracking temperature, partially vaporized in amounts sufficient to maintain the activity of the diluent when the vaporized portion is regenerated by partial hydrogenation and returned to the reaction mixture. Stated in another way, one of the principal features of the present invention is continuous vapor phase withdrawal of the conversion products such that the effective diluent to unconverted material ratio is substantially increased, but this invention also proposes that the withdrawal and regeneration of the spent diluent may advantageously be accomplished by vapor phase withdrawal of the spent diluent from the reaction mixture. Therefore, the diluent is selected so as to have a boiling range higher than the end boiling point of the desired vaporous conversion products. In a modification of the invention, however, the boiling range of the diluent is selected such that it is withdrawn for regeneration as a liquid.

In the following description of the attached drawing, the charging stock is a high boiling vacuum residuum, boiling preferably above 1050° F. The products desired include heating oils and lighter, i. e., products boiling below about 650° to 700° F. The boiling range of the diluent is consequently above about 700° F., e. g., 700° to 900° F. For this particular arrangement, the hydrogen donor diluent cracking temperature will be about 800° to 900° F. which will ensure a proper degree of vaporization of the diluent and products. With these particular materials, the degree of conversion obtained will be the range of 50 to 95%, 1050° F.+ conversion.

For convenience, the range of operating conditions pertinent to this invention are summarized in Table I.

With particular reference to the attached drawing, a charging stock, for example, a vacuum residua, is introduced into a reaction vessel or tower 1 by line 2. This charging stock may be suitably preheated to a temperature below incipient cracking temperatures. In the lower portion 3 of the tower, there is contained a liquid phase reaction mixture comprising the residuum and a hydrogen donor diluent in the ratio of about 1/20 to 1/1 vols. diluent/vol. residuum. The mixture is maintained at a suitable cracking temperature in the range of 750° to 1000° F. The donor diluent is selected to have a boiling range such that it will be partially vaporized at the cracking temperature and pressure. Thus the donor diluent will pass up through the tower along with the conversion products. In a modification of this invention, however, the diluent boiling range may be selected such that it remains unvaporized and is removed from the tower as a liquid.

A free hydrogen-containing gas is admitted to the base of the tower by lines 4 and/or 5. Preferably the major part of the hydrogen gas it admitted by line 5 to a stripping well 1-*a* at the base of the reactor. The gas so admitted will then strip from the bottoms bleed the material boiling in the diluent boiling range and lighter.

To maintain the reaction temperature, heat may be supplied to the reaction mixture, aside from heating the hydrogen gas and the residuum feed, by passing a portion of the reaction mixture by line 6 to an external heating means 7, e. g., a furnace, and back by line 8.

Vapors comprising conversion products, diluent and hydrogen gas are evolved from the reaction mixture and pass upwardly through the tower countercurrent to refluxed liquids which serve to remove heavy ends from the vapors. This reflux liquid comprises regenerated hydrogen donor diluent from the hydrogenation section and recycled liquid from the reaction mixture. In certain applications, it may be desired to introduce the fresh residual feed into the vapor phase above the reaction mixture

to aid in the scrubbing of the vapors. To provide for a high donor diluent/residuum ratio, materials may be withdrawn from the liquid phase by line **10** and introduced into the upper portions of the vapor phase by lines **11** and **12**. This recycled material may be suitably cooled to maintain the proper temperature gradient over the tower. Normally the amount of material recycled through lines **11** and **12** will amount to 0 to 1 vol./vol. of fresh feed.

The vapors rising from the reaction mixture after being scrubbed pass up through the hydrogenating catalyst bed **40** and the hydrogen of the vapors is introduced into the diluent. Some hydrogen gas is also introduced into the lighter fractions of the vapors to be removed as product. However, the amount of hydrogen so consumed will be relatively small as the products from the HDDC reaction are substantially saturated and will consume little, if any, additional hydrogen. If the hydrogenation catalyst is also a desulfurizing catalyst, some desulfurization, with consequent hydrogen consumption, of the product fractions may take place but this is a desirable reaction in any event.

The hydrogenation catalyst selected depends somewhat upon the type of donor and feed treated, the hydrogenation concentration, and the temperature of the hydrogenation zone. Preferably the hydrogenation zone comprises a fixed bed of particulate catalyst, e. g., 3/16" pills, and the catalyst used is a hydrogenation catalyst selected from the group consisting of cobalt, molybdenum, tungsten, nickel and their oxides and sulfides, mixtures thereof, platinum, etc. Advantageously, the catalyst is selected to be substantially sulfur resistant, although the sulfur content of vapors usually will not be so high as to be unduly detrimental.

The hydrogenated vapors issuing from the hydrogenation zone are then separated in the upper portion of the tower in a conventional high pressure fractionator arrangement with side stream draw-offs and side stream strippers (not shown). The initial fraction obtained from the vapors is the hydrogenated donor diluent. This fraction is refluxed downwardly through the hydrogenation zone to the reaction mixture. Additional hydrogenation of the donor diluent will take place in the liquid phase. A portion of the condensed donor diluent fraction is removed from draw-off plate **18** by line **19** and is passed through a cooler **20** and refluxed to the tower by line **21** to maintain the proper temperature level at this point.

As before explained, the concentration of the desired aromatic-napthenic constituent of the diluent fraction will be lessened by materials being cracked into the diluent boiling range and by loss of the aromatic-naphthenic compounds by cracking and/or condensation reactions. A portion of the donor diluent fraction pumped around is, therefore, withdrawn by line **22** as product, and make-up diluent comprising material previously described is added by line **23**. The bleed and make-up diluent will amount to about 10 to 25 vol. percent of fresh residuum feed.

In some applications, particularly where a low boiling diluent is used and high conversion is practiced, the hydrogenated diluent fraction rather than being refluxed interiorly in the tower, may be passed by line **24** directly to the liquid phase.

Normally, the space velocity in the hydrogenation section, including vaporized and liquid diluent will be 0.3 to 1.0 v./v./hr. Temperatures will range from about 650° to 850° F. under pressures of about 100 to 600 p. s. i. The hydrogen concentration of the vapors under these conditions is maintained at 50 to 75 vol. percent. At the hydrogenation zone, the composition of the hydrocarbons present in the zone would be approximately in the following range: the product vapors boiling below the initial boiling point of the diluent will compose approximately 20 to 50% of the hydrocarbons, the remainder being vaporized diluent passing up through the hydrogenation zone and refluxed liquid diluent passing down through the zone.

After the regenerated hydrogen donor diluent is separated from the vapors, the vapors are separated in the upper portion of the reaction tower into the various boiling range fractions. For example, if a 700° to 900° F. diluent is used then gas oils boiling in the range of 550° to 700° F. are withdrawn by line **25**. Heating oils boiling in the range of about 400° F. to 550° F. are removed by line **26** and naphthas boiling in the range of about C_5 to 400° F. are removed by line **27**.

The temperature of the top of the reaction tower is maintained relatively low, e. g., 90° to 150° F. vapor temperatures, so that the remaining uncondensed gases are composed largely of the unconsumed hydrogen. These uncondensed gases are removed by line **28** from the top of the tower. To maintain the effective hydrogen concentration, a portion of the gases may be purged as by line **30** and make-up hydrogen added. Thus 5 to 20% of the gas consisting of about 40 to 80% hydrogen may be purged and make-up hydrogen of a higher purity added. The make-up hydrogen used may come from any convenient source within the refinery. Thus hydrogen from naphtha hydroforming operations, from gasification reactions, from light gas reformation, etc. may be used.

Preferably, after purging out a portion of the vapors, the vapors are subjected to a separation process, e. g., absorption, adsorption, high temperature cracking, diffusion, low temperature fractionation, etc., to remove light hydrocarbon gases, thereby concentrating the remaining hydrogen. Thus line **28** transfers the vapors to an absorption system **31**, the operation of which is known by the art. Lean oil is admitted to the absorber by line **32** and scrubs the gases. Fat oil is removed from the absorber by line **33** and is regenerated in the customary manner. The effluent from the absorber is recycled by line **34** to the reactor after having make-up hydrogen added by line **35**. The recycled vapors are pressurized by blower system **36** and reheated by heating means including a furnace **37**, and passed, as previously described, by lines **4** and **5** into the reaction tower.

By this means of operation, i. e., low top tower temperature and absorption, the hydrogen added as make-up to the system is utilized more completely than in conventional HDDC operations.

It will be appreciated that the low quality of the feed charged to the reactor will cause a substantial build-up of uncoverted material such as metals, salts, extremely refractory hydrocarbons such as carboids, etc. in the liquid reaction mixture. For this reason, with most charging stocks, it is necessary to continuously or periodically bleed a portion of the liquid reaction mixture from the process. To accomplish this, a stripping zone **1–*a***, which may be suitably baffled to promote contacting and countercurrency, is affixed to the bottom portion of the reaction tower. A stripping gas comprising preferentially recycled hydrogen gas but including other gases such as steam is admitted to the stripping section **1–*a*** by line **5**. The temperature conditions of the stripping section are adjusted so that only materials boiling above the end boiling point of the diluent are removed from the tower by line **38**. The withdrawn material may be used as a fuel oil. Advantageously, the conversion conditions may be adjusted so that the amount of this fuel oil withdrawn will be sufficient to meet the fuel requirements of the process and, therefore, the process can be made balanced as to fuel requirements without the use of extraneous fuel sources. For material in the nature of vacuum residua, this bleed may amount up to about 75 vol. percent of the fresh feed, but is preferably 10–40 vol. percent of the fresh feed. A portion of the stripped bottoms in line **38** may be recycled by lines **11** and **12** to the tower to serve as reflux.

The range of pertinent operating conditions applicable to the above-described process are summarized in Table I. The table also presents a specific example of operating conditions.

Table I

	Range	Example
Pressure, p. s. i. g.	100 to 600	450.
Temperature, ° F. at —		
Cracking Zone	750 to 1000	850.
Hydrogenation zone	650 to 850	750.
Feed rate, v/v/hr.	0.5 to 10	2.0.
Diluent boiling range, ° F.	Any range between 450-1000	700-900.
Percent 700° F., conversion [1]	25 to 100	75.
Diluent/feed ratio	1/20 to 1/1	1/2.
Diluent make-up, vol./vol. fresh feed	Up to 1/4	1/10.
Treat gas rate [2], SCF/bbl. feed	2000 to 10,000	6000 (55% H_2).
Recycle gas, SCFB feed	0 to 10,000	4,700.
Gas purge rate, SCF/bbl. feed	500 to 5,000	750 (50% H_2).
Make-up gas rate, SCF/bbl. feed	500 to 5,000	1,300 (75% H_2).
Hydrogen consumption, SCF/bbl. feed	400 to 800	600.
Hydrogen concentration in hydrogenation zone, percent	50 to 75	55.
End point of vaporous products, ° F.	450 to 900	700.
Initial boiling point of stripped bottoms product, ° F.	550 to 1000	900.

[1] Based on fresh feed plus diluent make-up. 700° F. conversion is defined as: 100 vol. percent feed minus vol. percent products boiling above 700° F., based on feed, excluding coke.
[2] Comprises recycle gas plus make-up.

Table II presents an example of the products obtainable for the charging stock indicated when the process is operated in accordance with the example of Table I.

Table II

Charging stock: 12.9% West Texas residuum.	
Inspections:	
H/C atomic ratio	1.51
Gravity, °A. P. I.	7.5
Conradson carbon, wt. percent	21.4
Ash @ 800° C. wt. percent	0.18
Sulfur, wt. percent	3.32
Diluent make-up: 700° to 900° F. thermal tar.	
Inspections:	
H/C atomic ratio	1.11
Gravity, °A. P. I.	3.6
Conradson carbon, wt. percent	0.437
10% boiling point, °F.	690
Aniline point, °C.	91
Yields, percent on residuum:	
Coke, wt. percent	0.02
C_3-gas, wt. percent	2
C_4/430° F., vol. percent	18
430/700° F., vol. percent	53
700/900° F. (Diluent bleed), vol. percent	15
900° F. (Stripped bottoms), vol. percent	25

Various alternative modes of operation will be apparent to those skilled in the art. For example, while the process has been described using a diluent in the boiling range suitable for the product of predominantly heating oils and lighter products, other boiling ranges of diluents may be used. Thus, where gasoline boiling range hydrocarbons are desired as the predominant product, a diluent boiling in the range of, say, 500° to 650° F. can be used and all material boiling above 650° F. converted into material boiling below 500° F. except for the necessary bottoms bleed.

While internal recirculation and hydrogenation of the diluent has been described, it is within the concept of this invention to make liquid phase withdrawal of the diluent. Thus, as an example, if a 700° to 900° F. boiling diluent be used, an equilibrium mixture of unconverted charging stock and diluent can be withdrawn from the liquid phase at the bottom of the tower and a simple separation can be performed to separate spent diluent. This, for example, can be done by flashing at low pressure. The separated diluent fraction can then be introduced into the reaction tower above the hydrogenation zone as a liquid and will consequently pass down through the hydrogenation section, being regenerated, to the reaction mixture. A portion of the bottoms material separated from the diluent can be bled from the process as previously described and the remainder can be reheated and recycled to the reaction mixture.

If a low boiling diluent is used, e. g., 450° to 650° F. boiling range diluent, rather than to practice substantially complete conversion of materials boiling above 650° F., a substantial liquid phase bleed can be taken and can be fractionated to obtain a distillate fraction boiling above the boiling range of the diluent. Thus, with the above diluent boiling range, there may be separated from the liquid withdrawn from the reaction mixture a gas oil as a catalytic cracking stock boiling in the range of 650° to 1000° F., and the bottoms or ends remaining can then be recycled to the liquid phase for further treatment.

It can be seen that the process of this invention involves the treatment of the materials in three zones, i. e., a hydrogen donor diluent cracking zone, a diluent hydrogenation zone, and a product distillation zone. It will be apparent that in some instances these zones may be physically separate. For example, it may be desired to have the distillation zone separate from the combined cracking and hydrogenation zones. In this case, a suitable conduit arrangement can be used to convey vapors from the hydrogenation to the distillation zone, and to return the condensed donor diluent to the hydrogenation zone.

Having described the invention, what is sought to be protected by Letters Patent is succinctly set forth in the following claims.

What is claimed is:

1. A hydrogen donor diluent cracking process which comprises maintaining in the lower portion of a reaction zone a reaction mixture comprising a high boiling hydrogen-deficient charging stock and a hydrogen donor diluent under liquid phase hydrogen donor diluent cracking conditions, introducing a free hydrogen-containing gas into the reaction mixture, continuously withdrawing overhead from said reaction mixture reaction vapors boiling below the initial boiling point of said charging stock comprising spent hydrogen donor diluent, hydrogen and conversion products, passing said vapors under hydrogenating conditions upwardly through a hydrogenation zone through an intermediate portion of said reaction zone in the presence of a hydrogenation catalyst whereby said spent hydrogen donor diluent is regenerated, fractionating in a fractionation zone in the upper portion of said reaction zone the hydrogenated vapors to obtain first the hydrogen donor diluent, together with any reaction products in the boiling range of said donor diluent, and then products of intermediate boiling range, leaving free hydrogen-containing vapors, and returning the hydrogen donor diluent, together with said reaction products in said donor diluent boiling range to the reaction mixture.

2. The process of claim 1 wherein said hydrogen donor diluent is returned from said fractionation zone through said hydrogenation zone.

3. A hydrogen donor diluent cracking process which comprises maintaining in the lower portion of a reaction zone under liquid phase hydrogen donor diluent cracking conditions including a temperature in the range of 750° to 1000° F., a reaction mixture comprising a charging stock and a hydrogen donor diluent having a boiling range within the limits of 450° to 1000° F., introducing a high boiling charging stock into said lower portion at a rate within the range of 0.5 to 10 v./v./hr., introducing

a free hydrogen-containing gas into the bottom portion of said reaction zone, continuously withdrawing overhead from said reaction mixture reaction vapors boiling below the initial boiling point of said charging stock comprising spent hydrogen donor diluent and conversion products, passing said vapors under hydrogenating conditions upwardly through an intermediate portion of said reaction zone in the presence of a hydrogenating catalyst whereby said spent hydrogen donor diluent is regenerated, fractionating in the upper portion the hydrogenated vapors to obtain, first, said hydrogen donor diluent and reaction products boiling within the boiling range of the donor diluent and then products of intermediate boiling range leaving free hydrogen-containing vapors, and returning the hydrogen donor diluent and the reaction products boiling within the boiling range of the donor diluent so separated to said reaction mixture.

4. The process of claim 3 wherein said hydrogenating conditions include a pressure in the range of 100 to 600 p. s. i. g. and said hydrogenating catalyst includes a bed of particulate catalyst selected from the group consisting of cobalt, molybdenum, tungsten, nickel, their oxides and sulfides, mixtures thereof, and platinum.

5. The process of claim 3 comprising in addition thereto continuously withdrawing as product stripped bottoms from said reactions mixture having an average boiling range higher than that of said hydrogen donor diluent.

6. The process of claim 5 wherein said free hydrogen-containing gas is used to effect said stripping.

7. The process of claim 5 wherein a portion of said bottoms is recycled to said reaction zone below said intermediate portion and refluxed to said reaction mixture countercurrent to said vapors.

8. The process of claim 3 wherein said free hydrogen-containing vapors are recycled to said reaction mixture to form a major portion of said free hydrogen-containing gas.

9. The process of claim 3 wherein the concentration of hydrogen in said free hydrogen-containing vapors is increased by separating other gases from the vapors, and the concentrated vapors along with make-up hydrogen are recycled to said reaction mixture as said free hydrogen-containing gas.

10. The process of claim 3 wherein a portion of said reaction mixture is circulated to an external heating zone and back to maintain the reaction temperature.

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