

Feb. 14, 1961

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2,971,901

MULTICATALYST HYDROCONVERSION

Filed Oct. 17, 1958

3 Sheets-Sheet 1

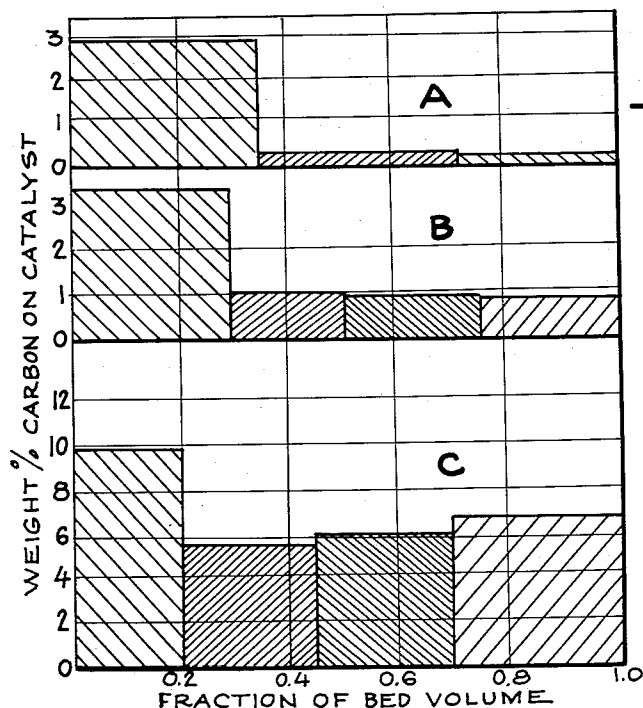


Fig. 1

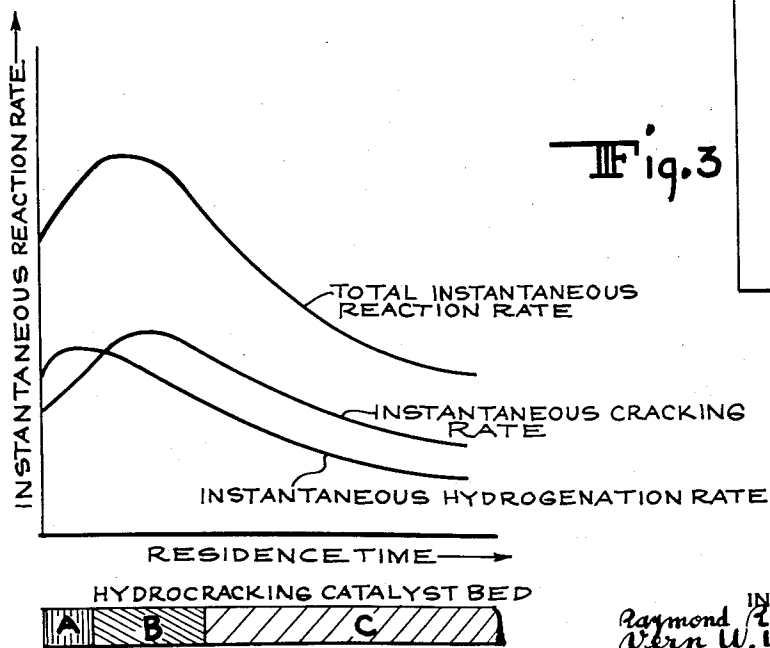


Fig. 2

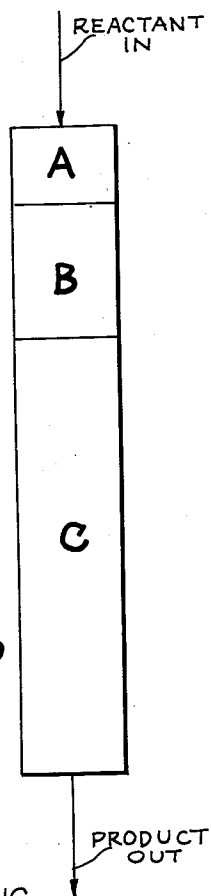


Fig. 3

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3 Sheets-Sheet 2

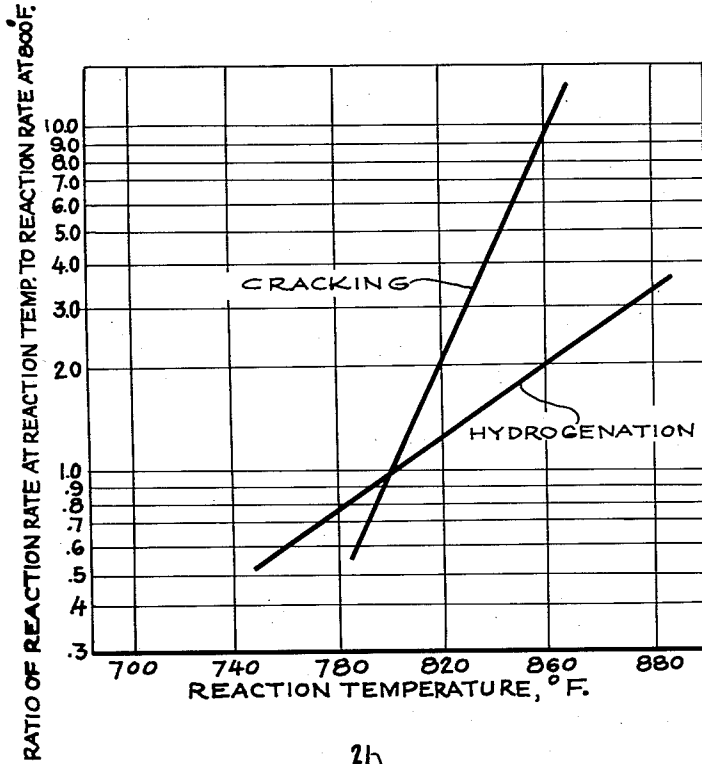


Fig. 4

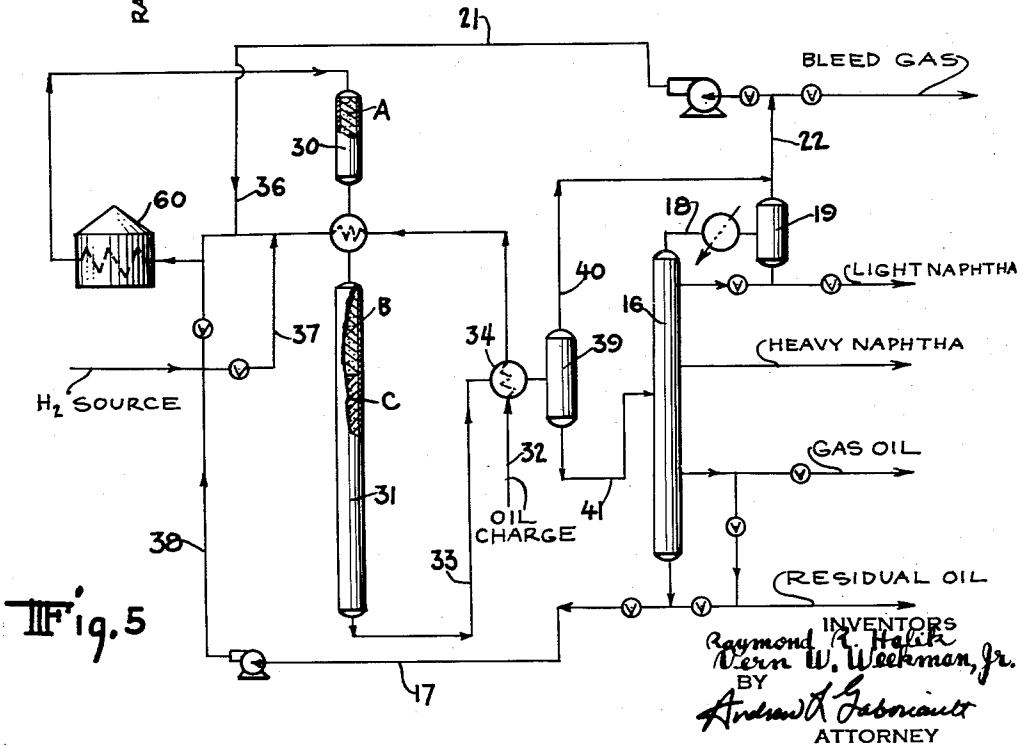


Fig. 5

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3 Sheets-Sheet 3

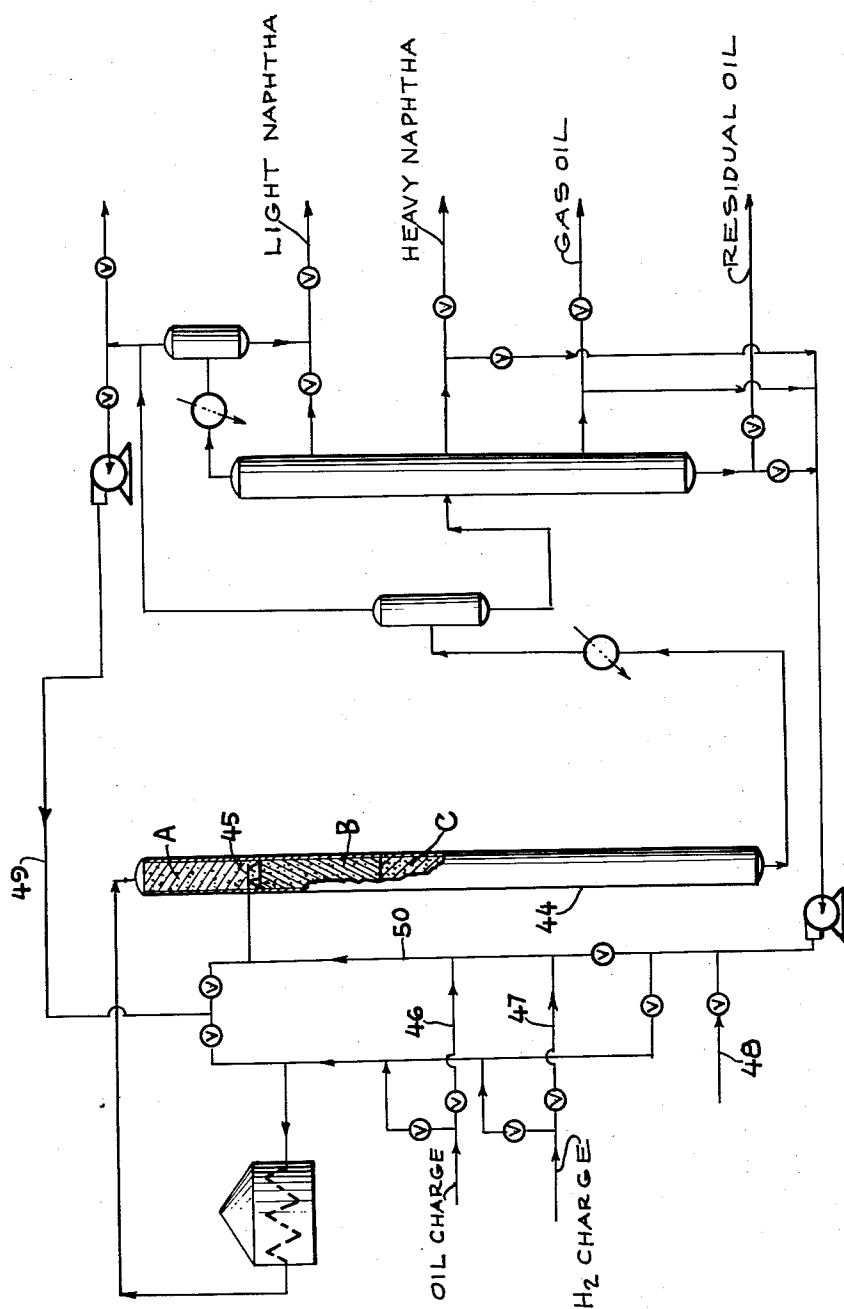


Fig. 19.6

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2,971,901

MULTICATALYST HYDROCONVERSION

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Filed Oct. 17, 1958, Ser. No. 767,924

6 Claims. (Cl. 208—59)

This invention deals with the catalytic conversion of hydrocarbons, in the presence of hydrogen, into more desirable hydrocarbon products, especially gasoline and fuel oil. More particularly, it relates to a technique for conducting such reactions so as to prolong the productive life of the catalyst between regenerations.

This invention will be best understood by referring to the attached drawings, of which:

Figure 1 is a graph showing the variation in carbon deposits through conventional hydrocracking reaction beds;

Figure 2 is a graph of the instantaneous reaction rates within a conventional hydrocracking reaction bed plotted against residence time of the reactant in the bed;

Figure 3 is a diagrammatic illustration of a hydrocracking reaction bed divided into three zones in accordance with this invention;

Figure 4 is a graph showing the variation in hydrogenation and cracking rates with temperature;

Figure 5 is a diagrammatic process flow sheet of one form of hydrocracking operation within the broad scope of this invention; and

Figure 6 is a diagrammatic process flow sheet illustrating another form of hydrocracking operation within the broad scope of this invention.

In conventional hydrocracking operations, a fixed bed of solid hydrocracking catalyst, made up of one or more components which promote hydrogenation and one or more components which promote cracking, is maintained within an enclosed reaction zone. The hydrocarbon charge is heated to about the desired reaction temperature and passed over the catalyst. The reaction is normally exothermic so that the reactor effluent will be at a higher temperature than the charge. The reaction deposits carbonaceous contaminant or coke on the catalyst which builds up over the course of time and eventually must be removed, usually by burning.

The prior art has indicated that in this type of hydrocracking operation the coke deposition on the catalyst in the reaction bed over the course of time is uneven. It has been demonstrated that the upper third, or sometimes as much as the upper half, of the reaction bed will bear an amount of carbonaceous contaminant substantially greater than the remainder of the bed. This phenomenon is illustrated in Figure 1. There, the carbonaceous deposit, as weight percent carbon on the catalyst, is plotted against the depth of the catalyst in the reaction bed, as the fraction of the bed depth from the inlet end thereof.

The graph marked "A" is based on data obtained from the isothermal bench scale operation of a hydrocracker employing a light gas oil charge and a platinum on silica-alumina catalyst at 1000 pounds per square inch pressure and a space velocity of 0.5 volume of charge (measured as 60° F. liquid) per volume of catalyst per hour.

The data for graph B were obtained in an isothermal bench scale hydrocracking operation at about 810° F.

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The charge was a heavy gas oil boiling above 650° F. and the catalyst was composed of the oxides of silicon, aluminum, molybdenum and cobalt. The reaction was conducted at 2000 pounds per square inch pressure and a space velocity of one volume of reactant per volume of catalyst per hour.

The data for graph C were obtained from an adiabatic pilot plant hydrocracking operation at 1500 pounds per square inch pressure. During this operation the temperature varied from 790° F. to 870° F. The charge stock was a gas oil boiling above 400° F. and the catalyst consisted of the oxides of silicon, aluminum, molybdenum and cobalt.

The graphs of Figure 1 make clear that the catalyst in the upper third of the reaction bed bears an amount of carbonaceous contaminant about three to five times that borne by catalyst in the remainder of the bed.

Obviously, in this situation, the initial portion of the reaction bed will become ineffective for practicing the desired hydrocracking process much more rapidly than the remainder of the bed. When the upper portion becomes ineffective, the portion next below operates as the inlet portion and the high coke laydown occurs there. Thus, this high coking area may proceed through the entire bed.

Under conventional practice the reaction temperature is increased to compensate for declining catalyst activity. However, this temperature cannot be increased indefinitely since the charge will undergo thermal conversion if its temperature is raised too high. Therefore, when the activity level of the catalyst is reduced by deposits of carbonaceous contaminant to some predetermined level, the entire unit must be shut down in order to regenerate the catalyst. Obviously the heavy coke laydown in the initial portion of the reaction bed accelerates the need for a shutdown.

The prior art has suggested that it would be desirable if the upper half or third of the reaction bed were contained in a separate vessel and that this heavily contaminated catalyst regenerated separately and more frequently than the catalyst in the remainder of the system, which becomes contaminated less rapidly. While this procedure probably would have certain economical advantages over the more conventional operation, it does not strike at the heart of this problem. It would be much more desirable to develop a technique by which the catalyst contaminated at a uniformly low level in the first place. Such an operation is the subject of this invention.

It is well-known that two basic reactions occur in a hydrocracking operation—cracking and hydrogenation. The catalysts customarily employed will contain one or more components which catalyze the hydrogenation reaction and another one or more components which catalyze the cracking reaction. Each reaction proceeds at a measurable rate and, in the usual practice, the rates of these two reactions are in balance so that the over-all reaction achieves conversion of the reactant to the desired products with a suitably low deposition of carbonaceous contaminant on the catalyst.

Copending application Serial Number 767,923, filed October 17, 1958, expounds the theory that the substantially higher coke deposits in the initial section of the hydrocracking bed result from the following sequence of reactions: Where the same catalyst is used throughout the length of the bed and the bed is kept under substantially the same reaction conditions throughout, there will be a tendency for sulfur and oxygen and olefins in the fresh charge to hydrogenate very rapidly upon first entering the reaction zone. At this first stage the hydrogenation reaction proceeds more rapidly relative to the cracking re-

action than will be the case in the main body of the reaction bed. Hydrogenation being highly exothermic, there is a substantial quantity of heat released in this initial stage. This heat then causes the cracking rate to accelerate, and in a second stage after the initial hydrogenation is completed the cracking rate will proceed more rapidly relative to the hydrogenation rate than is desirable. The accelerated cracking of the hydrocarbons will produce substantial quantities of unsaturates which will not be hydrogenated at the hydrogenation rate in this stage. These unsaturates, therefore, polymerize into carbonaceous contaminants or coke which contaminates the catalyst. After a period of this high rate of cracking, the temperature of the reactant will drop because of the endothermic nature of the cracking reaction and the cracking rate and hydrogenation rate will then be in balance and will remain in balance in the expected manner throughout the remainder of the bed.

This may be illustrated by reference to Figure 2, which plots the instantaneous reaction rate against residence time in the reaction bed. Of course, for conventional beds of uniform cross-sectional area, the residence time will be directly proportional to the distance within the catalyst bed away from the inlet end. The upper curve is a plot of the total instantaneous reaction rate. It goes through a peak in the initial stages of the catalyst bed.

The instantaneous hydrogenation rate, it is noted, is much higher than the instantaneous cracking rate in the early stages of the reaction. However, shortly thereafter, the cracking rate accelerates rapidly while the hydrogenation rate begins to decline. Finally, after a period, the cracking rate and hydrogenation rate are in balance, as denoted by the constant distance between the two curves. The bar graph at the bottom of Figure 2 illustrates the three separate zones within the reaction bed when the same catalyst and reaction conditions are employed throughout the bed. The first zone (A) is the zone of rapid hydrogenation. This is followed by a zone (B) in which the cracking reaction predominates and the final zone (C) is one in which cracking and hydrogenation reactions are in balance. This last zone is the one that operates in the manner intended by conventional designs. It will normally take up the major portion of the residence time, usually at least a half, and more frequently about two-thirds of the residence time.

Serial Number 767,923 broadly claims a process in which the reaction conditions are adjusted so that they are more favorable to hydrogenation and less favorable to cracking in the zone where rapid cracking occurs in the prior art. Specifically, Serial Number 767,923 claims a process wherein the catalyst used in that zone has a higher ratio of hydrogenation activity to cracking activity than the catalyst in the major portion of the bed. The present invention is an improvement on the process of Serial Number 767,923, enabling one to more certainly avoid this high coke laydown in the initial portions of the reaction bed.

It is a major object of this invention to lengthen the period between required regenerations in catalytic hydrocracking processes.

Another object of this invention is to provide a fixed bed, catalytic hydrocracking process in which the deposition of carbonaceous contaminant is uniformly low throughout the length of the bed.

Another object of this invention is to avoid excessive coke laydown on all portions of a catalytic hydrocracking reaction bed.

These and other objects of the invention will be apparent from the following discussion of the invention.

Broadly, in this invention, the hydrocracking reaction is conducted in three different zones connected in series. These zones are so sized that at least one-half of the total reaction time is spent in the last zone. A catalyst is used in the second zone which has a higher ratio of hydrogenation activity to cracking activity than said

ratio for the catalyst in the last zone. The effluent from the first zone is cooled as it leaves the first zone, thereby further favoring the hydrogenation reaction as compared to the cracking reaction in that zone.

Obviously, the process outlined above, by depressing the tendency toward cracking and increasing the tendency toward hydrogenation, works against the observed tendency in this zone toward overcracking and promotes rapid hydrogenation of any cracked products, thereby reducing the high coke make which occurs in this region in conventional operations.

The basic process to which this invention applies is illustrated in Figure 3, which is a diagrammatic representation of a bed of particulate solids divided into three zones or regions, an inlet zone A, an intermediate zone B, and an outlet zone C. Zone C is so sized that the reactant will consume at least one-half of the total reaction time in zone C. Zone C is filled with a standard hydrocracking catalyst having both hydrogenation and cracking activity. Zone B is, in general, substantially larger than zone A and is filled with a catalyst having a ratio of hydrogenation activity to cracking activity which is greater than the same ratio for the catalyst in zone C. Zone A may, if desired, employ the same catalyst as zone C. Preferably, however, it uses a catalyst having a lower ratio of hydrogenation activity to cracking activity. As indicated, outlet zone C should be so sized that at least one-half, and preferably at least two-thirds, of the total residence time in the bed is consumed there. Zone B should be greater in size than zone A, for example, about twice the size of zone A. Thus, in a typical reactor of uniform cross-sectional area, inlet zone A might extend for one-ninth of the bed height, intermediate zone B for two-ninths of the bed height and outlet zone C for two-thirds of the bed height.

In order that the above-described rapid hydrogenation will occur in zone A in conventional operations, at least one of the following components should be present in the charge stock in the quantity indicated:

40 Nitrogen	0.05% by weight.
Sulfur	0.2% by weight.
Oxygen	0.1% by weight.
Total unsaturates	20% by volume.

Of course, the first three named components will normally occur in chemical combination with hydrogen and carbon in petroleum base charge stocks.

The catalysts in the various zones may be evaluated by any test or tests suitable for measuring hydrogenation activity and cracking activity. For example, hydrogenation activity may be measured as follows:

A standard 400° F. to 700° F. gas oil may be passed over the catalysts under test at identical operating conditions, for example, 750° F. reactant inlet temperature, a space velocity of one volume of reactant (measured as 60° F. liquid) per volume of catalyst per hour, a reactor pressure of 1500 pounds per square inch gauge and a hydrogen circulation to the reactor of 5000 standard cubic feet of hydrogen per barrel of reactant. A 400° F. to 650° F. fraction would then be removed from each product and the aniline point (A.S.T.M. test 611-53T) and A.P.I. gravity (A.S.T.M. test D287-54) of the fraction determined. The diesel index may then be calculated for each fraction as the product of the aniline point and A.P.I. gravity divided by 100. The higher the diesel index, the greater the hydrogenation activity of the catalyst.

The cracking activity of catalysts may be determined by passing a straight run paraffinic gas oil, boiling between 400° F. and 700° F., over the catalysts under test. The same operating conditions as those employed in the hydrogenation test may be used except that a higher temperature of 830° F. would be employed. The volume of material in the product which boils below 400° F. is measured. The greater this volume, the greater the cracking activity.

It will be appreciated that when this application speaks of cracking activity reference is made to the cracking activity in the presence of hydrogen.

The improvement of the instant invention is based on an observation illustrated in Figure 4. Figure 4 demonstrates the effect of temperature on the hydrogenation and cracking reactions in a hydrocracking operation. Against temperature there is plotted the ratio of the reaction rate at any given reaction temperature to the reaction rate at 800° F. Figure 4 was determined from a plot of actual operating data. It is apparent that the cracking rate is much more sensitive to temperature than the hydrogenation rate. On Figure 4 the cracking rate may be cut in half by a temperature reduction of less than 20° F. while the hydrogenation rate is only halved with a 60° F. reduction.

Thus, by cooling the effluent from zone A, a lower temperature in zone B will be maintained which will materially assist in maintaining the conditions in zone B relatively more favorable to hydrogenation and less favorable to cracking than in zone C. While both hydrogenation rate and cracking rate decrease with lower temperatures, the cracking rate decreases relatively more.

Moreover, the use of cooling in this fashion furnishes a means by which the ratio of hydrogenation rate to cracking rate may be easily adjusted in zone B to precisely operate that zone so that excess coke laydown on the catalyst is avoided.

A specific application of the use of lower temperatures in zone B to reduce the tendency toward cracking and increase the tendency toward hydrogenation of the products that are cracked is illustrated in Figure 5. Zones A, B and C employ different catalysts as indicated above. In addition zone A is contained in a separate vessel 30 from the vessel 31 which houses zones B and C. Cold charge oil in line 32 is heat exchanged with hot product from passage 33 in heat exchanger 34. Heated charge oil is then heat exchanged with the effluent from vessel 30 in heat exchanger 35. The charge oil is blended with recycle hydrogen from line 36, any needed fresh hydrogen from line 37 and a recycle hydrocarbon stream from line 38, and passed through furnace 60, wherein the entire mixture is heated to the desired hydrocracking conversion temperature. The heated mixture is then fed to the upper end of zone A in vessel 30.

The heat exchange in 35 should reduce the zone A effluent in temperature to less than 790° F. However, this temperature should not be reduced below 740° F., since at lower temperatures the reaction rate is too low to be practical. The hydrocarbons then pass through zones B and C to complete the conversion.

Heat exchange in exchanger 34 reduces the temperature of the product stream to a temperature suitable for separation of gaseous material and the gaseous material is then separated in a conventional high pressure separator 39.

The gaseous material, mainly hydrogen, from separator 39 passes overhead through line 40. It is then recycled to the reaction system through lines 22, 21 and 36.

The liquid material is removed from the lower section of high pressure separator 39 and passed to a fractionator 16 by means of line 41. In fractionator 16 the liquid reactor effluent may be divided into the conventional products. Gaseous material taken overhead may be passed to a low pressure separator 19 via line 18, wherein naphtha is removed therefrom by condensation. The remaining gaseous material then passes overhead to join the gases from high pressure separator 39.

If desired, gas oil, residual oil, or both, may be recycled to the reaction system via passages 17 and 38.

Figure 6 illustrates another way in which the temperature in zone B may be reduced. Reactor 44 is equipped with zones A, B and C having different catalysts and a liquid distributor 45 which may be of any conventional design, such as one or more perforated pipes. To

this distributor 45 there is supplied a fluid which is cooler than the temperature of the reactants at that point in sufficient quantity to lower the temperature in zone B, as specified above, and thereby favor the hydrogenation reaction over the cracking reaction in that zone. The cooling fluid may include a part of the oil charge supplied through lines 46 and 50; it may include make-up hydrogen added through lines 47 and 50; it may include recycle gas supplied through lines 49 and 50; it may include any recycled fraction of the total liquid reactor effluent. An entirely separate hydrocarbon stream may be used as a cooling fluid supplied through lines 48 and 50. This stream might consist of naphthenic hydrocarbon oils so that the well-known hydrogen transfer effect is obtained in which hydrogen would be transferred from the naphthenic oil to the hydrocarbon reactant. This transfer would be to the heavier fragments or free radicals of the reactant before they polymerize to coke and would therefore reduce aging.

Of course, any combination of the various cooling streams mentioned might be used.

The remainder of the operation disclosed in Figure 6 is the same as that shown in Figure 5.

As indicated above, any hydrocracking catalyst having both a hydrogenation component and a cracking component may be used in the main bulk of the reaction bed, zone C. As nonlimiting examples, the catalyst may employ nickel, molybdenum, platinum, palladium, ruthenium, tungsten or cobalt or the oxides or sulfides or these materials deposited on a silica-alumina, silica zirconia or silica magnesia base. The catalyst in the other zones may have similar components to the catalyst in zone C, of course having a higher ratio of hydrogenation activity to cracking activity in zone B than that in zone C.

In zone A the catalyst may be the same as in zone C but preferably should have a lower ratio of hydrogenation activity to cracking activity. The catalyst in any of the zones, rather than being of one composition, may be a mechanical mixture or alternate layers of a hydrogenation catalyst and a cracking catalyst.

Particularly favorable catalysts for use in this invention, especially in zone C, are the catalysts described in Serial Number 418,166, filed March 23, 1954, now abandoned, and in Serial Number 825,016, filed July 6, 1959. The former application is directed to a catalyst of 0.05 to 20 percent by weight of a platinum group metal deposited on a specified base, for example, silica-alumina. The latter application is directed to a catalyst comprising broadly 15 to 40 percent by weight silica, 3 to 20 percent by weight molybdenum trioxide, 1 to 8 percent cobalt oxide and the remainder alumina.

As indicated above, outlet zone C should comprise at least one-half of the residence time of the reactants and usually about two-thirds of the residence time. Zones A and B will tend to require more residence time at higher space velocities. Thus, at space velocities of 0.1 to 1.0 liquid volume per volume of catalyst per hour, these two zones need only take up one-third of the bed; at higher space velocities, e.g., 1 to 2, they should be longer. Generally, zone B may have twice the residence time of zone A. Other factors which will have an effect on the required residence time in zones A and B are feed stock type, temperature, hydrogen to oil ratio and hydrogen consumption.

As previously pointed out, the zone A effluent should be reduced in temperature to at least 790° F. It should not, however, be cooled below 740° F. or the reaction will not proceed at an economical rate.

Also, as previously noted, this invention applies to the processing of charge stocks containing at least one of the following components in at least the indicated quantity:

Nitrogen	0.05% by weight.
Sulfur	0.2% by weight.
Oxygen	0.1% by weight.
Total unsaturates	20% by volume.

The general range of operating conditions over which this invention will function are:

Pressure	500-10,000 pounds per square inch gauge.
Temperature	700-1000° F.
Space velocity	0.1-10 liquid volumes of reactant per volume of catalyst per hour.
Charge stocks	Petroleum or like hydrocarbon fractions boiling above 400° F.

Of course, in the optimum operation of this invention the coke deposits in all zones will build up at the same low rate and to the same level while the unit is on stream. However, this optimum operation need not be achieved within the broad scope of this invention, since any reduction in the coke laydown on the catalyst initially encountered by reactants will be an improvement over the art.

It will be obvious that the cooling of the zone A effluent which this invention requires may be accomplished by a variety of conventional means in addition to those illustrated. For example, cooling coils employing an exteriorly supplied cooling fluid may be embedded in the reaction bed between zones A and B.

Example

In a process for the hydrocracking of hydrocarbons according to the present invention, the reaction bed might be of uniform cross-section with zone A 4 feet in length, zone B 8 feet and zone C 38 feet in length. Zone A could be filled with an equal mixture of silica-alumina cracking catalyst and a hydrocracking catalyst comprising 15 percent by weight silica, 2.5 percent by weight cobalt oxide, 8 percent by weight cobalt oxide and 74.5 percent by weight alumina. Zone B might employ the hydrocracking catalyst used in zone A crushed to 42-48 mesh Tyler. Zone C might employ either the hydrocracking catalyst used in zone A in $\frac{1}{4} \times \frac{1}{4}$ inch pellets or pelleted 0.5 percent by weight platinum on silica-alumina hydrocracking catalyst.

A petroleum derived charge stock boiling within the range 400 to 850° F. could be supplied to zone A at 760° F. The pressure in the reactor could be 2000 p.s.i.g. The zone A effluent might be 830° F. and would be cooled by indirect heat exchange with fresh charge to 790° F. prior to its entry to zone B.

This invention should be understood to include all of the changes and modifications to the examples of the invention, herein chosen for purposes of disclosure, which do not constitute departures from the spirit and scope of the invention.

We claim:

1. A process for the hydrocracking of hydrocarbon fractions boiling above 400° F. which contain at least one of the following components in at least the quantity indicated: nitrogen 0.05 percent by weight; sulfur 0.2 percent by weight; oxygen 0.1 percent by weight; unsaturates 20 percent by volume, which comprises: maintaining a bed of hydrocracking catalyst within an enclosed housing; maintaining an outlet region of said bed which extends from the end thereof from which products are removed for a distance amounting to at least two-thirds of the total length of the bed containing a standard hydrocracking catalyst; maintaining a middle region of said bed which next precedes said outlet region with a catalyst having a higher ratio of hydrogenation activity to cracking activity than the catalyst in said outlet region; maintaining an inlet region of said bed extending from the end of the bed to which the hydrocarbon fraction is supplied to said middle zone filled with a catalyst having a lower ratio of hydrogenation activity to cracking activity than the catalyst in said middle zone; maintaining said bed at a pressure within the range 500

to 10,000 pounds per square inch gauge; supplying the hydrocarbon fraction and hydrogen to the inlet end of said bed at a temperature within the range 700° F. to 1000° F. and passing said petroleum fraction and hydrogen through the bed to effect a conversion, at least in part, to lower boiling products and produce exothermic heat of reaction throughout said bed; removing said products from said bed and separating the desired lower boiling products from higher boiling recycle material; returning said higher boiling recycle material to said bed and injecting said material into said bed at the inlet to said middle region at a temperature and in a quantity that the effluent from the first region of the bed thereabove is cooled to a temperature below about 790° F.; and passing said recycle material through the middle and outlet regions of the bed, whereby uniform deposition of carbonaceous contaminants on all of the catalyst in said bed is promoted.

2. A process for the hydrocracking of a high boiling petroleum fraction containing at least one of the following components in at least the quantity indicated: nitrogen 0.05 percent by weight; sulfur 0.2 percent by weight; oxygen 0.1 percent by weight and unsaturates 20 percent by volume by passing said fraction through a bed of solid particulate material, which comprises: maintaining a bed of said particulate material within a confined housing and maintaining within said bed three separate zones with differing particulate solids, an inlet zone, an intermediate zone and an outlet zone; sizing said outlet zone so that the hydrocarbon reactant must consume at least one-half of the total residence time in said bed in said outlet zone and providing that the particulate solid in said outlet zone is a standard hydrocracking catalyst with activity for both cracking and hydrogenation; sizing said intermediate zone so that the reactant will consume substantially more residence time in said intermediate zone than in said inlet zone and providing that the particulate solid in said intermediate zone is a catalyst with a substantially greater ratio of hydrogenation activity to cracking activity than the catalyst in said outlet zone; providing a particulate solid in said inlet zone which has a ratio of hydrogenation activity to cracking activity substantially less than the ratio of said activities of the catalyst in the intermediate zone; passing the high boiling petroleum fraction under hydrocracking conditions into said inlet zone and through said inlet zone to hydrocrack said fraction and produce exothermic heat of reaction; cooling the effluent from said inlet zone to a temperature below about 790° F. but above about 740° F. and passing said effluent after cooling through said intermediate and outlet zones to complete the desired conversion with a uniformly low deposit of carbonaceous contaminant on all of the catalyst in said housing and produce exothermic heat of reaction in said intermediate and outlet zones so that the average reaction temperature in said outlet zone is higher than in said intermediate zone.

3. A process for the catalytic hydrocracking of a high boiling liquid hydrocarbon charge containing at least one of the following components in at least the quantity indicated: nitrogen 0.05 percent by weight, sulfur 0.2 percent by weight, oxygen 0.1 percent by weight, unsaturates 20 percent by volume, which comprises: maintaining three separate hydrocracking zones arranged in series, each filled with a catalyst, the catalyst in the last zone being greater in volume than the total volume of catalyst in the other two zones and having both hydrogenation and cracking activity, the catalyst in the second of said zones having a greater ratio of hydrogenation activity to cracking activity than the ratio of said activities of the catalyst in the last of said zones and the catalyst in the first of said zones having a ratio of hydrogenation activity to cracking activity less than the ratio of said activities in the second of said zones; passing the hydrocarbon charge through said three zones in series under hydrocracking reaction conditions which include a re-

action temperature in each zone above 700° F. and which produce exothermic heat of reaction in all of said zones so that the hydrocarbons increase in temperature as they flow through each of said zones and cooling the effluent from the first of said zones between said first and second zones to a temperature below 790° F. but not below 740° F., whereby the average reaction temperature in said second zone will be below the average reaction temperature in the last of said zones.

4. The process of claim 3 wherein said cooling is effected by indirect heat exchange between the hot effluent of the first zone and cooler hydrocarbon charge prior to the charge being supplied to the first zone.

5. The process of claim 3 wherein the cooling is effected by direct heat exchange between the hot effluent

of the first zone and a cooler fluid mixed with said effluent.

6. The process of claim 5 wherein the cooler fluid is a hydrogen transfer agent capable of transferring hydrogen to the charge as it undergoes conversion in the second and last zones.

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