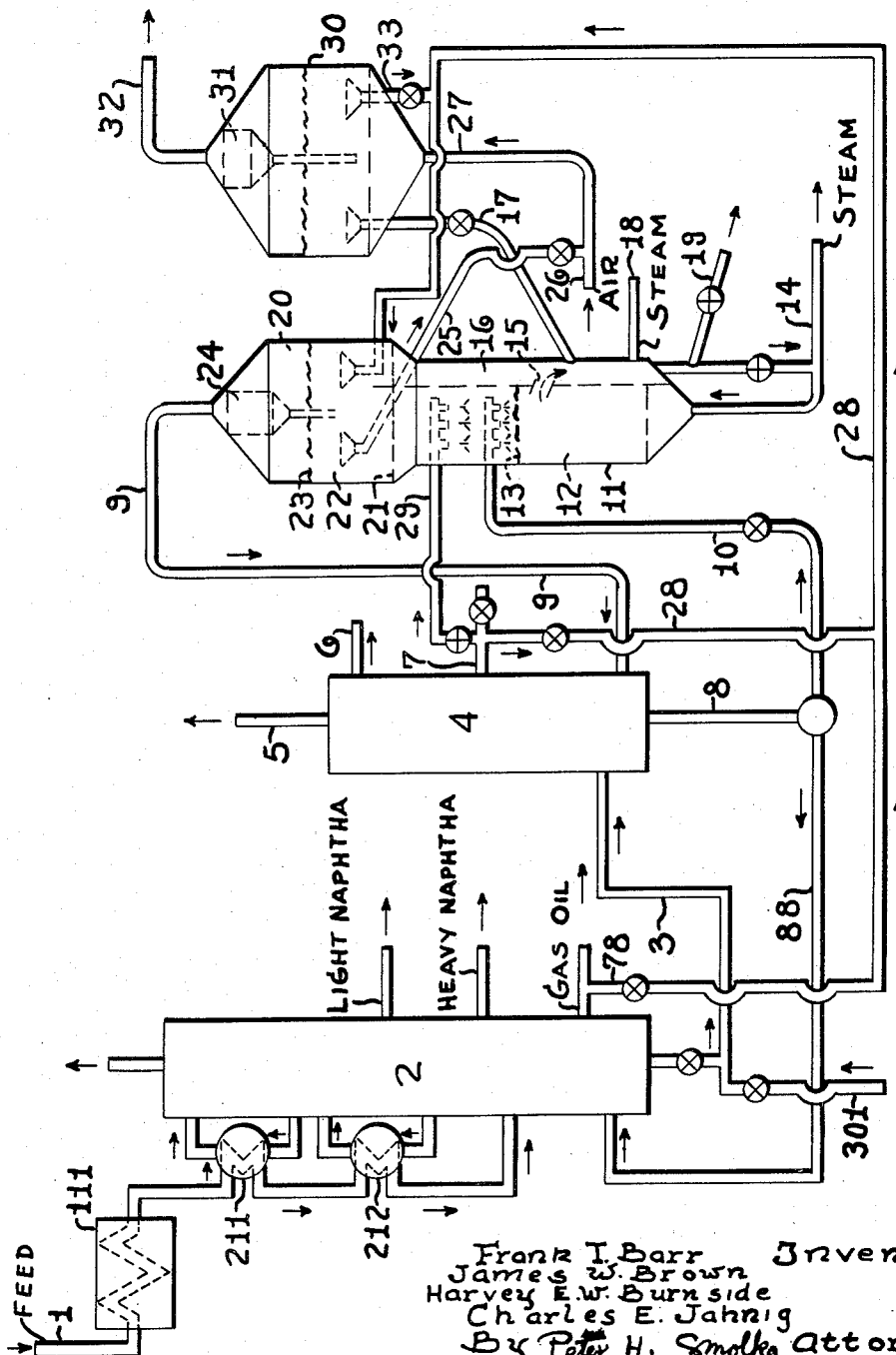


Dec. 4, 1956

F. T. BARR ET AL
INTEGRATED REFINING OF CRUDE OIL

Filed Aug. 5, 1952

2,773,017



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INTEGRATED REFINING OF CRUDE OIL

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Application August 5, 1952, Serial No. 302,646

6 Claims. (Cl. 196—52)

This invention relates to an integrated process for refining crude oil. In particular the invention relates to an improved refining process wherein crude petroleum is converted, in part catalytically, into large amounts of valuable distillate fractions such as high-octane gasoline and heating oil, with a minimum of residual or low-value products such as fuel oil or coke. Still more specifically the process of the invention permits refining crude oil into high quality products in greatly simplified equipment, making unusually effective use of the sensible heat of various streams for supplying heat to various stages of the process, while minimizing the cost of equipment. The main advantages of the process thus include reduction in the number of distillation towers, heaters and storage tanks, elimination of a vacuum tower, minimizing the heating and cooling of process streams by tubular or other indirect type heat exchangers, and substantial elimination of residual fuel oil production.

In existing refineries it has been customary to distill crude oil so as to separate it into various fractions such as virgin naphtha, gas oil and an atmospheric residuum. The separated gas oil is then catalytically cracked to form high-octane gasoline, virgin naphtha is separately upgraded by reforming where desired, and the long residuum is usually vacuum distilled to recover additional catalytic cracking feed stock. The resulting short or vacuum residuum, finally, may be used as a low-grade fuel, or further catalytic cracking feed stock may be obtained therefrom by coking. Each of these operations normally requires separate reaction zones, fractionating towers, heaters and coolers, as well as numerous storage tanks. Also, the vacuum distillation tower normally required in such refineries represents an expensive piece of equipment as regards both original investment and operation.

It is the principal object of the present invention to provide an improved, integrated refining process permitting maximum production of valuable products in simplified equipment and in a more economical manner. This and other objects and advantages of the invention will appear more clearly from the subsequent description and the drawing referred to therein.

The drawing diagrammatically illustrates an integrated refinery suitable for carrying out the invention.

Referring to the drawing, crude oil may be supplied to the process through line 1 which leads to atmospheric topping still 2. In still 2 the crude is fractionated to separate a bottoms fraction from the usual virgin distillate products such as gas, light and heavy naphtha, kerosene, heating oil, light gas oil, or various combinations thereof. The atmospheric bottoms or long residuum or topped crude may have an initial boiling point between about 500° and 750° F., e. g., 650° F., and is passed through line 3 to the bottom of fractionation tower 4 which serves to fractionate both said bottoms and the total cracked overhead products described hereafter. This tower may operate at atmospheric pressure, and avoids the need for a vacuum crude still, as will be described hereafter.

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The various fractions produced in tower 4 may include a gas stream 5 which consists largely of fixed hydrocarbon gases and hydrogen produced in the later described thermal and catalytic conversion zones; a largely catalytic gasoline stream 6; a gas oil stream 7 which contains some catalytic cycle oil as well as a more or less substantial amount of heavy virgin gas oil vaporized from the long residuum introduced through line 3; and a short residuum stream 8 which may have an initial cut point of between about 800 and 1100° F., e. g., 950° F. Essentially all or most of the heat required for vaporizing the lighter constituents of the long residuum 3 may be advantageously supplied by direct contacting with the catalytic cracking overhead products introduced into tower 4 through line 9. Additional heat may be added to long residuum stream 3 by indirect exchange against hot regenerated catalyst, coke, etc., or in a furnace, to obtain maximum vaporization of the long residuum in tower 4. The still temperature in tower 4 may advantageously be in the range of about 700 to 900° F.

Separate towers 2 and 4 will be used when it is desired to separate virgin naphtha, kerosene, heating oil, etc., free of cracked products. For example it may be advantageous to reform the virgin naphtha separately. Also, if the crude oil is fed directly into tower 4 relatively low-octane virgin naphtha is mixed with and reduces the value of the catalytically produced high-octane gasoline. However, such feeding of crude oil directly to tower 4 through alternate feed line 301 cheapens the process significantly and may therefore be preferred if virgin products need not be segregated, as when the crude feed contains only small amounts of naphtha.

The short residuum stream 8, which may amount to about 5 to 30 percent on crude, depending on the crude, is passed from tower 4 through line 10 to coker 11. Inert solids, such as powdered petroleum coke having a particle size in the range of about 40 to 500 or more microns, are maintained in coker or low-conversion vessel 11 by upflowing gases as a dense turbulent bed 12 having an upper lever 13 and an apparent density of about 10 to 50 lbs./cu. ft., with densities of about 0.01 to 5 lbs./cu. ft. in the more dilute phase above level 13. An inert gas such as steam or normally gaseous hydrocarbons may be introduced through line 14 to bring the total upward superficial velocity of the vapors in bed 12 to about 0.5–5 feet per second, so as to assure satisfactory fluidization. The temperature of the fluid bed 12 is maintained at about 800 to 1100° F., preferably 850 to 950° F. The optimum temperature must be high enough to effect the desired coking of the residuum but low enough to minimize formation of gasoline and lighter products by thermal cracking. The heat requirements of the coking zone are preferably supplied by direct or indirect heat exchange of the coke with hot regenerated catalyst, as described and separately claimed in copending applications Serial No. 227,169, filed on May 19, 1951, now Patent 2,734,850, and Serial No. 230,746, filed on June 9, 1951, now Patent 2,655,464, to which reference may be had for further details.

A part or all of the gas oil stream 7 may be passed from tower 4 directly to catalyst bed 23 through line 28. If desired, the oil may first be mixed with some hot catalyst withdrawn from regenerator 30 through standpipe 33. To distribute the oil feed and any added catalyst over the reactor cross-section, multiple injection points or a manifold may be used. The latter may be either above or below distributor plate 21. When catalyst is added below plate 21, coker feed from lines 10 and 29 should be introduced just above coke bed level 13, or at a lower point, so as to avoid knocking down of catalyst into the coking zone.

Gas oil from line 7 may also or alternatively be fed

through line 29 to a point below the catalyst bed, using either spray nozzles as illustrated or a bubble plate. The latter will prevent entrained coke dust from entering the catalyst bed. Collected dust is purged from the plate by overflowing a slurry stream which is fed to the coking zone. Part or all of gas oil stream 7 can be withdrawn from the system via line 77 to prevent an excessive build-up of refractory catalytic cycle stock. The light gas oil from crude still 2 may be either passed through line 78 and combined with the gas oil from product fractionator 4 for feeding to the cracking reactor 20, or disposed of as heating oil, diesel fuel or the like when economic conditions warrant.

For purposes of illustration, heating of the coke bed 12 by direct heat exchange is shown in the drawing. Coke from bed 12 thus overflows through orifice 15 into a mixing and elutriating section 16. In section 16 the heat content of the coke is raised to the desired extent by mixing with hot regenerated catalyst introduced through line 17 while an inert gas such as steam is introduced through line 18 into the bottom of elutriating section 16.

If desired, oil feed may also be added at this point. Vapors rising up serve to elutriate out the catalyst and carry it up to reactor 20. This is a particularly good point for feeding aromatic or refractory stocks such as cycle stock, clarified oil and the like. Virgin naphtha may be added at this point for reforming.

When such direct heat exchange between coke and catalyst is practiced, it is desirable to have one of the solids, preferably the coke, appreciably coarser or denser, or both coarser and denser, than the other solid. Thus, the coke particles may advantageously range in size between about 150 and 1000 microns while the catalyst particles may range between about 0 and 100 microns, a size differential of at least 50 microns being desirable between the two kinds of solids to facilitate separation. Under such conditions, by proper adjustment of the gas velocity in section 16 to a value equal to or slightly greater than the minimum fluidizing velocity of the coarser solids, but less than the velocity which will entrain them substantially, the coarser solids, such as coke, will gradually sink through the elutriating section 16 while picking up heat from the hot catalyst. At the same time the relatively fine solids, such as regenerated catalyst, will gradually rise through the elutriating section 16 while giving up heat, and will then be entrained into catalytic conversion vessel 20.

Reheated coke may then be withdrawn from section 16 and returned through line 14 to the coking zone 12 while net coke product may be recovered from line 19. In special cases where the heat transfer from the regenerated catalyst to the coking zone solids is insufficient to satisfy the heat requirements of the coking zone, a portion of the withdrawn coke may be passed through a separate heater or burner in a manner well known by itself. In some applications it may be undesirable to provide any heat for the coking step from the hot catalyst. All of the heat for coking may then be provided from a separate circulating fluid solids coke burner. Coking and cracking may be carried out in separate vessels. In any event the hot coke from the coke burner or heater may be returned to the coking zone to supply any or all of the heat required. Such a separate burner, in a form resembling regenerator 30 or a transfer line burner of the type described in copending application Serial No. 230,020, filed on June 5, 1951, now Patent 2,700,170, may also be necessary when the solids in the coker are not simply heat-carrying inerts, but are a non-expendable catalyst. Since some 10-20% coke (based on oil feed) is deposited on such a catalyst from the residuum even under conditions of low conversion, this coke must be substantially completely burned off between cycles so as to retain the desired catalytic effect in conversion zone 11. In any case, it is usually important to keep conversion in first

conversion zone 11 low, e. g. below 20 or at the most 30 or 35 volume percent of 430° E. P. gasoline (based on feed to the coker), as otherwise the resulting thermally cracked products will excessively dilute the more valuable products from catalytic zone 20. Or, if zone 11 is catalytic as well, low conversion is essential to avoid too rapid coking up of the catalyst.

Generally, however, it is preferred to use coke rather than catalyst in the low-conversion stage 11, since with coke there are no limitations on the amount of coke to be burned, and heat balance control is entirely at the command of the designer and operator. With catalyst in both stages such heat balance is more difficult to attain, and heat from the main catalyst regenerator is then at best put to such low grade uses as steam generation and the like.

Returning now to the description of the reaction in coker 11, the vapors liberated therein are passed upward to catalytic reactor 20 through supporting grid 21. In reactor 20 these rising vapors maintain a conventional powdered cracking catalyst as a dense turbulent bed 22 having an upper level 23. The catalyst may be clay or synthetic silica-alumina gel or the like as is well known, and may range in size between about 40 and 150 microns. The density and other physical characteristics of fluidized catalyst bed 22 are well-known and are essentially similar to those of fluidized coke bed 12 previously described. The temperature in reactor 20 is maintained at 850 to 1100° F., preferably at about 900 to 1000° F. depending on the particular feed stock and catalyst used, so as to obtain optimum catalytic cracking of the hydrocarbon vapors.

The cracked product vapors may be withdrawn from reactor 20 by way of a dust separating device such as cyclone 24 and passed through line 9 to the bottom of fractionation tower 4 where they supply the heat required for the desired partial vaporization of the topped crude stream 3 as previously described. Also the diluent effect of the relatively light product vapors in tower 4 aids vaporization of the topped crude, thereby increasing the proportion of distillate oil provided by tower 4 and suitable for catalytic cracking. Catalytic products such as high-octane gasoline are finally recovered from tower 4 as a sidestream 6 for example, or as an overhead product.

Spent catalyst, preferably after being stripped with steam or the like in a conventional manner, is withdrawn from reactor 20 through standpipe 25. After mixing with air or other oxygen-containing gas introduced at 26 the resulting suspension is passed by way of line 27 to regenerator 30. In regenerator 30 coke deposits are burned off the spent catalyst in a known manner while the catalyst is fluidized by the upflowing air. Flue gases may be removed overhead by way of cyclone 31 and exhaust line 32. Regenerated catalyst is finally returned to reactor 20 by way of previously mentioned standpipe 17 and elutriator 16, or through some equivalent direct or indirect heat exchange equipment adapted to transfer the surplus heat of the regenerated catalyst to the coking zone solids. Alternatively, some of the regenerated catalyst may also be returned directly to reactor 20 by way of previously mentioned standpipe 33 and cracking feed line 28.

Heat to the pipe still 2 is best supplied by recycling a part of the hot bottoms from fractionator 4 through lines 8 and 83. Feed preheat may also be supplied by exchange with sidestream pumparounds as indicated by heat exchangers 211 and 212. Indirect heat exchange between incoming feed and regenerated catalyst may likewise be resorted to. Also, hot coke from reactor 12 or line 19 may be added to the bottom of tower 2.

Heat to fractionator tower 4 is normally supplied in sufficient quantity by the cracked product vapors introduced into tower 4 from reactor 20 through line 9. Additional heat can be added if desired by adding hot coke to the bottom of tower 4.

Essentially all heat for the entire crude processing circuit is ultimately derived from catalyst regeneration, which, however, may be supplemented by some coke burning as desired, and as described above. At times it may also be advantageous to preheat the crude feed in a fired coil 111 which may conveniently burn some of the fuel gas produced in the process. It is generally desirable to preheat the crude feed to temperatures of about 650 to 800° F. before it is introduced into pipe still 2.

The foregoing description has been presented primarily in order to illustrate the invention. However, variations or modifications not specifically suggested herein may be devised by those skilled in the art without departing from the scope or spirit hereof. The invention is particularly pointed out and distinctly claimed in the appended claims.

We claim:

1. An integrated heavy oil refining process which comprises introducing an oil containing constituents vaporizable at atmospheric pressure into a fractional distillation zone operating at least at atmospheric pressure, removing and passing a gas oil fraction from said distillation zone to a fluid catalytic conversion zone, removing and passing a residual fraction boiling above about 800° F. from said distillation zone to a low-conversion zone wherein inert finely divided heat transfer solids are maintained as a fluid bed at a temperature in the range of 800 to 1100° F. and wherein said residual fraction is converted to coke and hydrocarbon vapors, directly passing the resulting hydrocarbon vapors without appreciable cooling to said fluid catalytic conversion zone wherein finely divided cracking catalyst is maintained as a fluid bed at a temperature in the range of 850 to 1100° F. and wherein said hydrocarbon vapors and gas oil fraction are converted into lower boiling catalytic conversion products, directly passing said lower boiling conversion products in vapor form from said catalytic conversion zone to said distillation zone, recovering a catalytically cracked gasoline fraction from an upper part of said distillation zone, removing coke-containing catalyst from said catalytic conversion zone, burning coke deposits therefrom in a regeneration zone in the presence of an oxygen-containing gas, returning catalyst so regenerated to said catalytic conversion zone, and supplying heat to said low-conversion zone by heat exchange between catalyst withdrawn from said regeneration zone and said heat transfer solids.

2. A process according to claim 1 including the additional step wherein said heat transfer solids containing deposited coke are withdrawn to a combustion zone, the coke deposit is burned with an oxygen-containing gas, and the resulting hot heat transfer solids are returned to the low-conversion zone.

3. An integrated process for refining crude petroleum oil which comprises introducing crude feed to a first fractional distillation zone, recovering and passing a bottoms fraction boiling above 500° F. from said first distillation zone to a second fractional distillation zone having a still temperature in the range of 600 to 900° F. and operating

at least at atmospheric pressure, removing and passing a residual fraction boiling above 800° F. from said second distillation zone to a fluid low-conversion zone wherein finely divided substantially inert heat-transfer solids are maintained as a dense fluid bed at a temperature in the range of 800 to 1100° F. and wherein the residual fraction is converted to coke and hydrocarbon vapors boiling predominantly above 400° F., passing the resulting hydrocarbon vapors without appreciable cooling and gas oil withdrawn from said second distillation zone to a fluid catalytic cracking zone wherein finely divided cracking catalyst is maintained as a dense fluid bed at a temperature in the range of 850 to 1100° F. and wherein said hydrocarbon vapors and gas oil are converted into lower boiling catalytic conversion products, passing said lower boiling conversion product without appreciable cooling from said catalytic cracking zone to said second distillation zone, removing a catalytically cracked gasoline fraction from an upper part of the said second distillation zone, removing carbon-containing catalyst from said catalytic cracking zone, burning coke deposits therefrom with an oxygen-containing gas in a regeneration zone at a temperature in the range of 1050 to 1250° F., returning regenerated catalyst from said regeneration zone to said catalytic cracking zone, and exchanging heat between regenerated catalyst and said finely divided heat transfer solids to supply heat to said fluid low-conversion zone.

4. A process according to claim 3 wherein a portion of said residual fraction from said second distillation zone is recycled to said first distillation zone, whereby heat is supplied for evaporating the virgin naphtha and lighter fractions from said crude feed.

5. A process according to claim 3 wherein a portion of the gas oil from the said second distillation zone is passed to the fluid low-conversion zone and another portion of the last-named gas oil is passed directly to said catalytic cracking zone.

6. The process of claim 3 wherein the exchange of heat between said regenerated catalyst and heat transfer solids is by direct contact between the solids in a heat exchange zone, with the solids being separated by elutriation, said regenerated catalyst passing upwardly and said heat transfer solids downwardly from said heat exchange zone.

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