

July 31, 1945.

E. V. MURPHREE ET AL

2,380,897

PROCESSING HYDROCARBON OILS

Filed Jan. 28, 1939

2 Sheets-Sheet 1

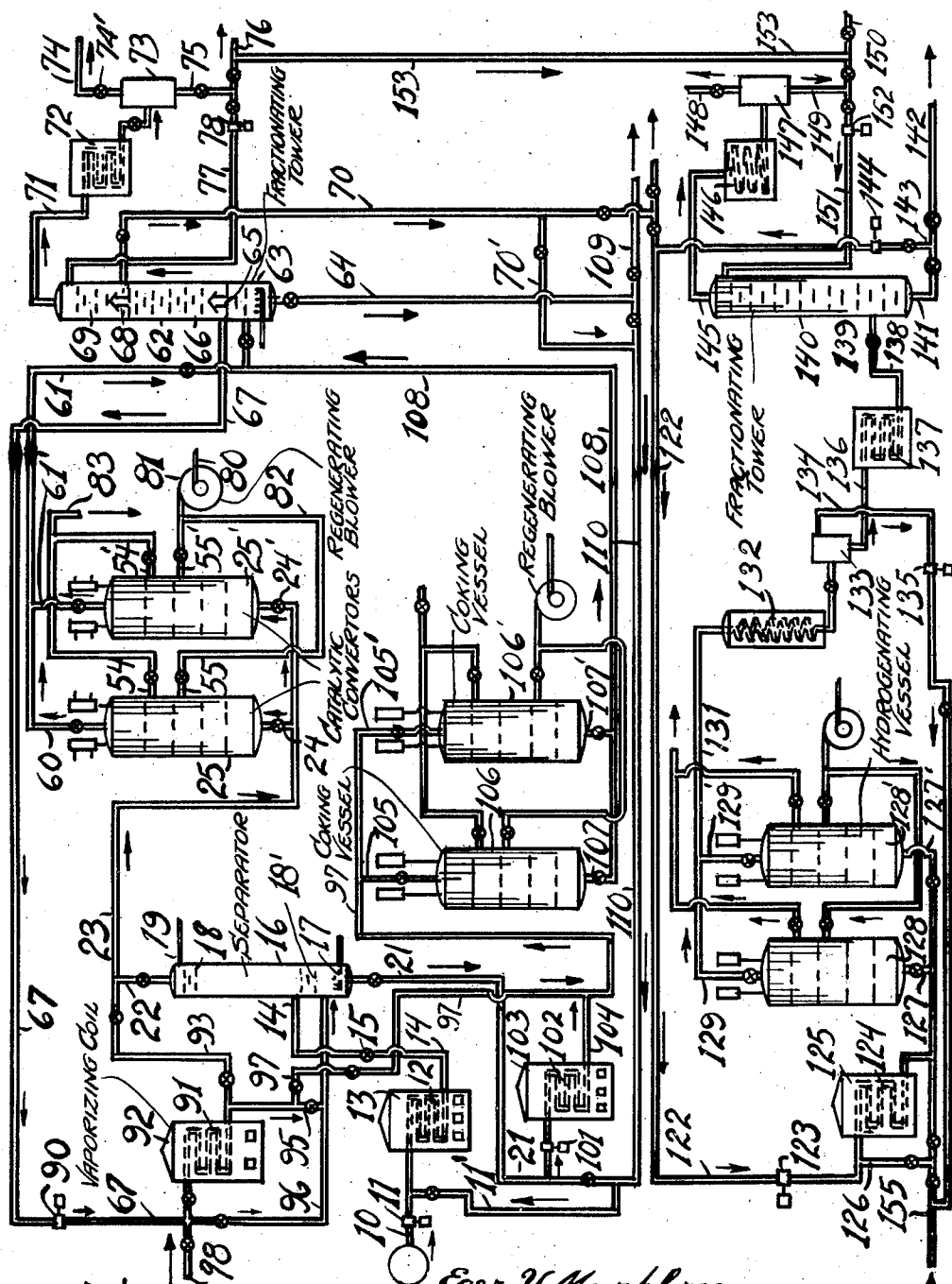


FIG.-1

Eger V. Murphree
 Alexis Voorhies, Jr.
 Leonard S. Bonnell Inventors
 Louis S. Tregre
 By *[Signature]* Attorney

July 31, 1945.

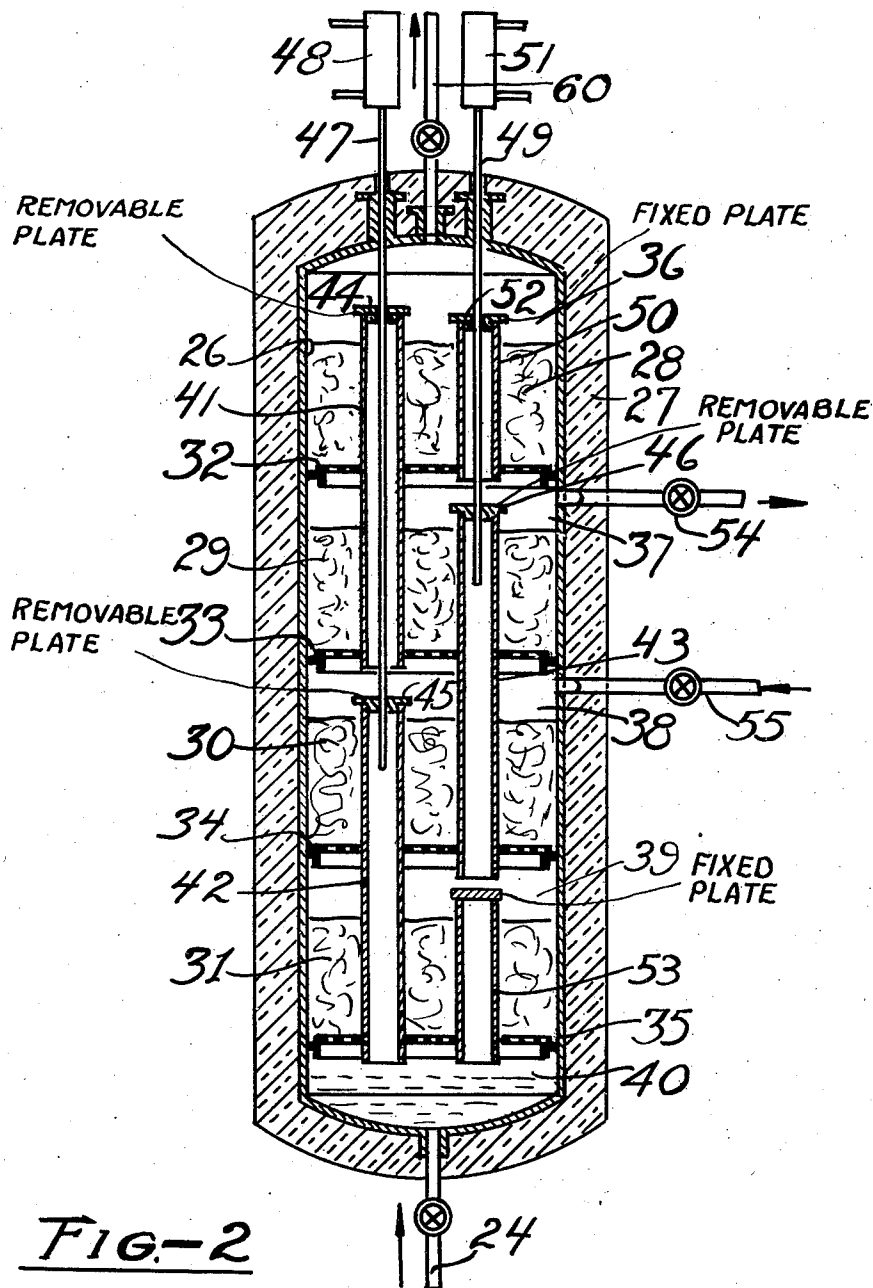
E. V. MURPHREE ET AL

2,380,897

PROCESSING HYDROCARBON OILS

Filed Jan. 28, 1939

2 Sheets-Sheet 2



Eger V. Murphree
Alexis Voorhies
Leonard S. Bonnell
Louis S. Tregre
Inventors
By *P. L. Young* Attorney

UNITED STATES PATENT OFFICE

2,380,897

PROCESSING HYDROCARBON OILS

Eger V. Murphree, Summit, N. J., Alexis Voorhies, Jr., Baton Rouge, La., Leonard S. Bonnell, Westfield, N. J., and Louis S. Tregre, Baton Rouge, La., assignors, by means assignments, to Standard Catalytic Company, a corporation of Delaware

Application January 28, 1939, Serial No. 253,248

3 Claims. (Cl. 196—49)

This invention relates to the processing of petroleum oil and pertains more particularly to a unitary process for conversion of crude residual oils into motor fuel of the gasoline type. The invention is more specifically directed to a combination process involving the catalytic conversion of selected fractions of the crude residual oil, the distillation and preconditioning of the oil prior to the catalytic conversion thereof and the further treatment of the catalytically converted products other than the desired distillate under conditions suitable for the production of a maximum yield of high anti-knock gasoline.

One of the ultimate objects of the invention is to provide an improved process for the conversion of residual petroleum oils into gasoline of high anti-knock quality.

A further general object of the invention is to provide a more economical process for producing gasoline from crude residual oils.

Other objects and advantages of the invention will become apparent from the more detailed description hereinafter in which reference will be made to accompanying drawings.

In the drawings, Fig. 1 is a simplified diagrammatic illustration of an apparatus suitable for carrying the invention into effect, and Fig. 2 is a cross sectional view of a catalytic converter adapted for use in the present invention.

Referring to the drawings, the reference character 10 designates a charge line through which crude residual stock may be introduced to the unit. This crude residual stock may be a topped crude from which the original virgin gasoline constituents have been removed, or it may be a reduced crude having kerosene and/or gas oil constituents removed, as well as the gasoline. The crude stock introduced through line 10 is charged by means of pump 11 to a heating coil 12 located in furnace 13. The oil during its passage through the heating coil is heated to a temperature sufficient to vaporize all crackable constituents contained therein.

According to one phase of the invention the oil passing through the heating coil 12 may be subjected to mild thermal cracking treatment of a nature capable of reducing the vaporizing temperature of the oil to a degree such that the resulting vapors upon passing through a catalytic conversion chamber hereinafter described will be at optimum temperature to produce a maximum yield of gasoline therefrom. By way of example, with certain highly active catalysts such as synthetic gels containing silica and alumina, it is desirable to carry out the catalytic conversion

under relatively low temperatures, such as, for example, from 700° F. to 800° F. to avoid over-cracking the vapors and thus form excessive gas and coke. In cases where such highly active catalysts are employed and the oil treated contains substantial quantities of constituents boiling above 700° F. to 800° F., it is desirable to subject such oil to a mild thermal cracking treatment to crack such high-boiling constituents into constituents which are vaporizable at a temperature below the temperature desired to be maintained in the catalytic converter.

When mild thermal cracking is carried out in the heating coil 12, the outlet temperature of the oil may be, for example of the order of 850° F., the oil being maintained within the coil for a period adequate to reduce the vaporizing temperature to the desired amount.

The oil after passing through the heating coil 12 is transferred through line 14 provided with a reducing valve 15 to a separator 16 in which vapors separate from unvaporized residue. The oil passing through the heating coil 12 may be subjected to pressure ranging from atmospheric to 700 lbs. or more per square inch.

The separator 16 is maintained under conditions which will vaporize all crackable constituents except a heavy semi-liquid residual product having from 1 to 2 A. P. I. gravity for example. Steam may be introduced into the bottom of the separator through line 17 to assist in vaporizing the oil.

Vapors liberated in the separator 16 after passing over baffles 18 positioned in the upper section of the separator for knocking back entrained solids and liquids are removed from the top thereof. Reflux oil may be introduced into the top of the separator through line 19.

The top temperature of the separator 16 should be controlled so that the sensible heat of the overhead products removed therefrom is sufficient to maintain the catalytic reaction chambers into which they are fed, as later described, under optimum temperature conditions to produce the maximum yield of gasoline with minimum coke and gas losses. The optimum temperature will depend upon the activity of the catalyst, the degree of conversion desired, and other factors, so that no specific temperature will be optimum for all conditions of operation.

In most cases the optimum temperature will vary between 750° F. and 900° F.

The degree of vaporization within the separator at a top temperature for the best catalytic cracking conditions may be controlled by regulat-

ing the pressure within the separator 16 and/or by varying the amount of steam introduced therein.

Vapors leaving the separator 16 pass through lines 22 and 23 to branch lines 24 and 24' leading to catalytic reaction chambers 25 and 25' respectively. These reactors may be of any desired type capable of effecting contact between the hydrocarbon vapors to be cracked and the catalyst to be employed for carrying out the conversion. Such catalysts may, for example, be an active or activated clay, but preferably a synthetic gel containing silica and alumina. When employing highly active catalysts such as specially prepared synthetic gels, it is desirable, when operating on heavy crude oils to maintain a relatively low temperature of an order of from 700-800° F. to avoid over-cracking of the oil, with the consequent formation of excessive gas and coke. While the invention is peculiarly adapted to the use of highly active catalysts, it will be understood that in its broader aspects it is not so restricted. As before pointed out, when highly active catalysts are employed in the treatment of heavy residual stocks, it is desirable to subject the residual stocks to mild thermal cracking treatment to reduce the vaporizing temperature of the oil to the desired catalytic cracking temperature.

As shown, the reaction chambers are the intermittent type wherein the catalyst is alternately subjected to cracking and regenerating treatment. It will be understood, however, that a continuous type reactor may also be employed, in which the catalyst either in powdered or molded form or in form of a slurry is introduced into the vapor stream either concurrently or countercurrently, and after contacting the vapors for the desired period is separated therefrom and regenerated in separate equipment.

One form of reactor capable of being employed in the invention is illustrated more in detail in Fig. 2. Both reactors 25 and 25' heretofore mentioned are of the same construction and consequently only one need be described in detail. For convenience, prime numbers will be employed for distinguishing the two reactors and the lines communicating therewith.

Referring to Fig. 2, the reactor may comprise an outer metal shell 26 provided with suitable insulation 27. Within the shell 26, a plurality of catalyst beds 28, 29, 30 and 31 are supported on grid plates 32, 33, 34 and 35 respectively, removably mounted on the outer shell 26. Between the individual catalyst beds and at the ends of the reactor are vapor spaces 36, 37, 38, 39 and 40.

Mounted within the reactor longitudinally thereof and supported by the grid plates are longitudinal vapor conduits 41, 42 and 43. Vapor conduit 41 provides communication between the upper vapor space 36 and the central vapor space 38. Vapor conduit 42 provides communication between the central vapor space 38 and the bottom vapor space 40. Vapor conduit 43 extends between the two intermediate catalyst beds and provides communication between the upper intermediate vapor space 37 and the lower intermediate vapor space 39. Vapor conduits 41, 42 and 43 are provided with valve caps 44, 45 and 46 respectively, which may be raised and lowered to open or close communication between the various vapor spaces previously described. Valve caps 44 and 45 for conduits 41 and 42 are operated by a single valve stem 47 extending in axial alignment with the conduits and projecting through a stuffing box in the top of the reaction chamber. This

valve stem may be operated by any suitable mechanism such as, for example, a hydraulic cylinder 48. Valve cap 46 for conduit 43 is operated by a similar valve stem 49 extending through a blank conduit 50 in the upper bed of catalyst 28. Valve stem 49 may be also operated by a suitable mechanism, such as a hydraulic cylinder 51. Blank conduit 50 is preferably of the same dimensions as the conduit 41 and has an enclosure plate 52 having a central opening through which the valve stem 49 projects. The lower bed of catalyst 31 is likewise provided with a blank conduit 53 of the same dimensions as conduit 42. By providing blank conduits 50 and 53, each catalyst bed contains the same volume of catalyst.

Leading from the upper intermediate vapor zone 37 and the central vapor zone 38 are valved pipes 54 and 55 respectively for the removal and introduction of vapors therefrom.

The internal ducts are preferably employed during regeneration. During the cracking cycle, valve caps 44, 45 and 46 are in closed position and hydrocarbon vapors entering through line 24 or 24' as the case may be, are caused to pass upward through the catalyst beds in series and then out through duct 50 or 50' as the case may be. During regeneration internal valve caps 44, 45 and 46 are preferably in raised position. The regenerating gases entering through valve line 55 are caused to pass through the individual catalyst beds in parallel and emerge through valve line 54. For example, the gases introduced through line 55 distribute themselves as follows. A portion of the regenerating gas passes downwardly through the lower intermediate bed 30 from whence it passes through vapor conduit 43 to vapor zone 37 and thence out through line 54. Another portion of the gases so introduced pass upwardly through the upper intermediate catalyst bed and emerge through line 54. A third portion of the regenerating gases passes upwardly from a central vapor zone 38 through vapor conduit 41 to the top vapor zone 36 thence downwardly through the upper catalyst bed and out through exit line 54. The remainder of the regenerating gases passes downwardly from the central vapor zone 38 to the bottom vapor zone 40 through conduit 42, then upwardly through the bottom catalyst bed, then through vapor conduit 43 to the outlet line 54. From the above, it will be noted that the regenerating gases pass in opposite directions through the adjacent beds.

The reaction chamber shown is provided with four catalyst beds. It will be understood, however, that any additional number of catalyst beds may be provided if desired. When more catalyst beds are employed, additional internal vapor conduits are provided connecting alternate vapor spaces as previously described.

Returning now to the treatment of the hydrocarbon vapors, for purposes of description, chamber 25 will be assumed to be on the cracking cycle while the chamber 25' is undergoing regeneration. The cracking conditions within the reaction chambers such as temperature, time of contact of oil vapors with the catalyst are preferably controlled to maintain an average of from 30% to 50% conversion into gasoline constituents during each passage of oil throughout the cracking cycle. The temperature and time of contact and other conditions necessary to obtain the desired conversion will depend on the activity of the catalyst which in turn will depend on the composition of the catalyst, the overall age of the catalyst in cases where the catalyst

gradually loses activity and the length of the cracking cycle. In the latter connection it should be understood that activity of the catalyst reduces as the cracking process continues due to more or less gradual accumulations of carbonaceous deposits which must be removed by regeneration periodically.

When employing freshly activated clay of the type known as "Super Filtrol" and when operating at an average cracking temperature of about 850° F. with a two hour cracking cycle the time of contact may be of the order of from 0.1 to 2 volumes of liquid feed per volume of catalyst to obtain about 40% conversion. The vapors after passing upwardly through the reaction chamber 25 and in series through the separate catalysts beds are withdrawn through line 60 from whence they pass through line 61 to the bottom section of a combined separator and fractionating tower 62. Entrained solids or liquids contained in the hydrocarbon vapors separate from the vapors in the bottom section 63 of the tower 62. The entrained products coupled with heavy condensate amounting, for example, from 1 to 10% of the total cracked products, collects in the bottom section 63 of the tower 62 and is withdrawn therefrom through line 64 and treated as herein-after described.

Uncondensed vapors from the bottom section of the fractionating tower 62 pass upwardly through a trap out tray 65 to the intermediate fractionating section 66 of the tower 62 wherein they are subjected to fractional condensation to separate a heavy gas oil fraction. This heavier gas oil fraction condensed in the intermediate section 66 collects in trap out tray 65 and is withdrawn therefrom through line 67 and treated as hereinafter described.

Vapors remaining uncondensed in the intermediate section 66 of the tower 62 pass upwardly through trap out tray 68 to the upper section 69 of the tower 62. The temperature of the top section of the tower 62 is controlled to condense constituents boiling above the desired motor fuel distillate. Condensate formed in the top section 69 collects in the trap out tray 68 and is withdrawn through line 70 and further treated as later described.

Vapors remaining uncondensed in the fractionating tower 62 pass overhead through line 71 to a condenser 72 wherein the desired distillate product is condensed. Products from the condenser 72 pass to a receiver 73 wherein the liquid distillate separates from residual gases. Residual gases so separated in the receiver 73 are removed therefrom through line 74 having a valve 74' for regulating the back pressure on the fractionator and catalytic converter. The residual gases withdrawn through line 74 may be passed to a suitable absorption unit not shown for removing gas constituents therefrom.

Distillate collected in the receiver 73 is withdrawn through line 75 and may be passed to a suitable storage tank (not shown) through line 76. A portion of the distillate so withdrawn may be returned through line 77 and pump 78 to the top of the fractionating tower 62 as reflux therefor.

The final distillate product withdrawn through line 76 may be subjected to any finishing treatment desired to produce the final market product. Such finishing treatments may, for example, comprise a stabilizing treatment, an acid treatment, clay treatment, hydrofining treat-

ment, doctor treatment, alkali treatment or any combination thereof.

When it becomes desirable to regenerate the catalyst in chamber 25 the valves in lines 24 and 60 are closed disconnecting reaction chamber 25 with the cracking circuit and valves in lines 24' and 61' are opened thus placing reaction chamber 25' on the cracking stream. Chamber 25' is then purged of oil vapors either by the introduction of steam or by placing the chamber under vacuum. During this purging operation the valve caps 44, 45 and 46 on internal ducts 41, 42 and 43 may be opened or closed.

After the reaction chamber 25 has been freed of residual oil vapors, regeneration is begun by introducing an oxidizing medium such as air or air diluted with inert gas such as steam or combustion gases. Referring to Figure 1 the regenerating gas is introduced through line 80 and forced by means of blower 81 through line 82 and conduit 55 to reaction chamber 25. During the regenerating process the internal valve caps 44, 45 and 46 are open and the regenerating gases pass in parallel through the individual catalyst beds as previously described. The spent regenerating gases are removed from the chamber through lines 54 and 83 and may be discarded. If desired, a portion of these spent gases after cooling may be recycled with the fresh oxidizing gas to assist in controlling the regeneration temperature in cases where the catalyst is adversely effected by high temperatures.

Returning to the fractionating tower 62, the heavy gas oil fraction withdrawn from trap out tray 65 is passed through line 67 and is forced by means of pump 90 to at vaporizing coil 91 located in furnace 92.

It has been found that cycle oil produced in a catalytic process of the nature before described has distinctly different characteristics than cycle oil produced from the conventional thermal cracking process. It is commonly known, for example, that cycle oil from a thermal cracking process is more refractory and more difficult to crack with the same conversion per pass than the original virgin gas oil charged to the unit. While this is generally true with respect to the cycle oil from the catalytic process, it has been found that the constituents which render said cycle oil more refractory or difficult to crack, are more highly concentrated in the lighter fractions of the gas oil than is the case with the thermal cracking process. This is particularly true when operating on paraffinic stocks or mixed paraffinic and naphthenic stocks, such as Mid-Continent crudes. When processing these stocks in a thermal cracking process, the A. P. I. gravity of the cycle gas oil decreases consistently as the boiling point increases. When processing the same stocks in a catalytic cracking process, however, the A. P. I. gravity remains substantially uniform with increasing boiling point over an extended range, thus showing that the lower boiling fractions of the cycle oil contain a higher percentage of refractory constituents than is the case with the gas oil from the thermal process. This makes it possible, by simple fractionation, to segregate from cycle gas oil resulting from catalytic cracking a higher boiling fraction which is more amenable to cracking treatment than that obtained by simple fractionation of gas oil vapors from a thermal cracking process. The amount of the heavier gas oil fraction segregated from the catalytic cycle oil in the trap-out tray 65 of the fractionating tower 62 for further catalytic cracking treatment

will depend upon the nature of the stock charged to the unit. When processing paraffin base oils, the heavier fraction collected in tray 65 may amount to as much as 70% of the total cycle gas oil fraction recovered from the process, whereas when operating on highly aromatic or naphthenic crudes, this fraction may not amount to more than 30% of the total gas oil produced in the process. The particular fraction for individual crudes to be collected in the trap-out tray 65 may be readily determined, however, by simple laboratory tests employed for determining paraffinicity or aromaticity of the oil. For example fractionation in the tower 62 may be controlled to remove from trap out tray 65 a fraction having an aniline point comparable to the aniline point of the oil passing to the catalytic cracking chambers 25 and 25'.

The heavier gas oil fraction during its passage through the vaporizing coil 91 is heated to the desired vaporizing and reaction temperature. The vapors from heating coil 91 may pass through line 93 merging with line 23 where they combine with vapors removed from the separator 16. A part or all of the vaporized oil from the heating coil 91 may be passed through line 95 to the bottom of the separator 16 to aid in stripping the oil therein or a part of the relatively heavy cold gas oil from line 67 may be by-passed around the vaporizing furnace 92 directly to the separator 16 as a quenching medium for the heated products from the heater oil coil 12. The above provision of lines 93, 95 and 96 permit a ready control of temperature within the separator 16 to obtain maximum vaporization of the products introduced therein. To assist in the vaporization, products from the heating coil 91 are preferably introduced into the separator at a point below the point of introduction of the heavy oil from the coil 12. If desired, baffle plates 181 may be provided within the separator between the points of entry of the heavy oil from the coil 12 and the vaporized oil from the coil 91 to assist in stripping the heavy oil of vaporizable constituents.

In some cases it may be desirable to pass a portion of the products from the vaporizing coil 91 through lines 95 and 97 to a coking unit hereinafter described to supply additional heat thereto.

Fresh gas oil may be introduced into the unit through line 98 and combined with the heavy oil fraction charged to the vaporizing coil 91. Such fresh gas oil may for example comprise virgin gas oil stripped from the crude residuum introduced through charge line 10. In certain of its broader aspects, the invention contemplates a process operating solely on gas oil. In such cases the charge line 10 and heating coil 12 may be omitted and all the gas oil charged through line 98 or as an alternative a part or all of the fresh gas oil charge may be introduced through charge line 10 and separately vaporized in heating coil 12. In the latter case, the fresh gas oil passing through coil 12 may be heated to a different temperature than the heavy recycle oil passed through the heating coil 91.

Returning to the heavy semi-solid residue withdrawn from the separator 16 through line 21. As previously stated, this product is a heavy semi-solid material totally unsuited for fuel oil purposes. According to the present invention, this heavy material from line 21 is charged by means of pump 101 to a coking coil 102 located in the furnace 103.

Products from the heating coil 102 pass through

line 104 to branch lines 105 and 105' leading to coking chambers 106 and 106'. These chambers may be in the form of digestion drums in which the products are permitted to digest until the residual material is converted completely to coke.

It is preferred, however, to provide the coking chambers 106 and 106' with suitable contact material upon which the coke collects and from which it may be periodically removed by oxidation. This contact material may be of catalytic nature, such as active or activated clays or synthetic gels of the nature previously described or it may be relatively inert contact material such as pumice, kaolin, fire brick, etc. As shown, these coking chambers are of the same construction as the catalytic converters 25 and 25' hereinbefore described, except that products from the coking coil are introduced into the top of the reactors rather than into the bottom. The products from the coking coil 102, upon being introduced into the top of the reactor 106 or 106' as the case may be, pass downwardly through the separate beds of contact mass in series. For descriptive purposes, it will be assumed that coking is being carried out in coke chamber 106 while chamber 106' is undergoing regeneration. Products from the coking process being carried out in reaction chamber 106 are removed from the bottom thereof through line 107 from whence they pass through line 108 to the fractionating tower 62 in which they combine with vapors from the catalytic converter 25 or 25'.

The products from the coking chamber will normally be in mixed liquid and vapor phase and may contain a considerable amount of entrained solids. These phases may be separated in the bottom of the combined separating and fractionating tower 62, as shown, or a separate separator may be provided in line 108 for separating vapors from liquid and only the vapors introduced into the fractionating tower. In the latter case the bottoms from such separator may be rejected from the system or combined with bottoms from the fractionating tower 62 and treated as later described.

Product separated in the bottom section 63 of the fractionating tower 62 and withdrawn therefrom through line 64 may be rejected from the process through line 109 or may be recycled through line 110 and pump 101 to the coking coil 102 where it is co-mingled with the residue from the separator 16 and coked therewith. As a further alternative a part or all of said residue may be passed through line 11' and combined with fresh oil in line 10.

Additional heat for accomplishing the coking in the coking chamber may be obtained by introducing a portion of the vapors from the vaporizing coil 91 through lines 95 and 97 as before described. Also in some cases it may be desirable to flux the heavy oil passing to the coking coil 102 with a lighter fraction to reduce coking difficulties in the coil. Such light fraction may for example comprise a portion of the light gas oil collected in trap-out tray 68 of the fractionating tower 62. To this end a part of such light gas oil withdrawn through line 70 may be passed through lines 70' and 110 respectively to the inlet side of coking coil 102.

When the contact mass contained in the coking chamber 106 becomes contaminated with coke so that it is desired to regenerate the same, the valves in lines 105 and 105' and valves in lines 107 and 107' are reversed, placing chamber 106' on the coking cycle and disconnecting chamber

106 from the coil. Chamber 106 may be then purged either by steaming or evacuation and regenerated as previously described with reference to chambers 25 and 25'. Chambers 106 and 106' are provided with suitable regenerating lines for effecting the regeneration similar to that described with reference to chambers 25 and 25'.

The light gas oil removed from the top section of the fractionating tower 62 through line 70 may be removed from the process. This product, however, is peculiarly suited for hydrogenation treatment for the production of high anti-knock aromatic gasoline. According to another phase of the invention, therefore, this product withdrawn through line 70 is passed through line 122 and pump 123 to a heating coil 124 located in furnace 125 where it is heated to a temperature suitable for carrying out the hydrogenation treatment for the production of aromatic gasoline. For example, the temperature may be of the order of 900° F. to 1200° F., the pressure may be 10 atmospheres or more, the amount of hydrogen employed may be of the order of 500-2500 cubic feet of hydrogen per barrel of oil treated. The heated products from the coil 124 after comingling with the desired amount of hydrogen introduced through line 155 are transferred through lines 127 and 127' to hydrogenating chambers 128 and 128'. If desired, a part of the hydrogen may be introduced through line 126 on the inlet side of the heating coil 124, to reduce coking within the coil.

The hydrogenating chambers 128 and 128' may be packed with suitable hydrogenating catalyst, such as, for example, the oxides or sulphides of group 6 of the periodic table or the chamber may contain a plurality of beds of catalyst. According to one of the features of the invention, these hydrogenating chambers are constructed in the same manner as the catalytic converters 25 and 25', heretofore described, so that the catalyst contained within the chamber may be periodically regenerated by the introduction of a regenerating medium such as an oxidizing gas or a sulphidizing gas to restore the activity of the catalyst.

For convenience, it will be assumed in the following description that chamber 128 is on the hydrogenating cycle whereas chamber 128' is undergoing regeneration. The combined vapors and hydrogen pass through the reactor 128 in contact with the hydrogenating catalyst maintained under temperature and pressure conditions suitable for bringing about the formation of low-boiling aromatics. The pressure within the reactor may range from 10-200 atmospheres or more.

Products after passing through the hydrogenating chamber 128 pass through lines 129 and 131 to a condenser 132 wherein oil constituents are condensed. Products from condenser 132 pass into hydrogen separator 133 wherein hydrogen is separated under pressure. The hydrogen so separated may be recycled through line 134 and compressor 135 to the hydrogenating unit. The oil is withdrawn from the separator through line 136 and passed through a heating coil 137 where it is heated to revaporizing temperature and thence through transfer line 138 having a reducing valve 139 to fractionating tower 140, containing suitable fractionating elements, such as bubble trays, for fractionally condensing undesired higher boiling constituents as reflux condensate. Products condensed in the fractionating tower 140 may be withdrawn from the bottom through line 141 and

may be removed from the process through line 142 or may be recycled to the heating coil for the hydrogenating chamber through line 143, pump 144 and line 122.

Vapors from the fractionating tower 140 pass overhead through line 145 to a condenser 146 wherein the desired distillate is condensed. Products from the condenser 146 pass to a receiver 147 wherein the liquid distillate separates from residual hydrocarbon gases. The residual gases are removed through line 148 and may be passed to a suitable absorption unit (not shown) for recovery of gasoline constituents therefrom.

The distillate formed in receiver 147 is withdrawn from the bottom thereof through line 149 and removed from the process through line 150. A part of this distillate may be recycled through line 151 and pump 152 to the top of the fractionating tower 140 as reflux therefor.

When operating the hydrogenating unit to produce aromatic gasoline, the resulting product may be blended directly with the catalytically cracked gasoline removed from the process through line 76. To this end, the blending line 153 connecting line 76 with line 150 is provided.

In lieu of subjecting the light gas oil collected in the top section of the tower 62 to destructive hydro-aromatizing conditions just described, this product may be subjected to a mild non-destructive hydrogenating treatment to form a high grade Diesel oil. When operating in this manner, the hydrogenating temperature may be of the order of from 500-800° F.; the volume of hydrogen consumed per volume of oil treated may be 500-1500 cubic feet of hydrogen per barrel of oil treated and the pressure maintained may be the same or somewhat higher than that for aromatic production.

Having thus described the preferred embodiment of the invention, it is understood that it embraces such other variations and modifications and such combinations and sub-combinations of steps as hereinafter claimed. It will be further understood that it is not the intention to unnecessarily restrict the invention or dedicate any novel features thereof.

We claim:

1. A method of processing residual oils which comprises distilling said residual oil to liberate vaporizable constituents and form a heavy fluid residue unsuitable for vapor phase cracking, passing the vapors so liberated in contact with a cracking catalyst in a cracking zone, maintaining said zone under conditions suitable for cracking said vapors into motor fuel constituents, maintaining said vapors in contact with said cracking catalyst for a period sufficient to effect the desired degree of conversion into motor fuel constituents, fractionally separating from said vapors a motor fuel fraction, a light condensate fraction boiling above the motor fuel boiling range, an intermediate condensate fraction and a heavy condensate fraction containing entrained unvaporized constituents, combining said last named heavy condensate fraction with said first named heavy fluid residue, subjecting the combined mixture to further cracking treatment to convert the same into a solid coke residue and vapors, vaporizing said intermediate condensate fraction, and combining the resulting vapors with vapors resulting from said first named distillation step prior to passing the same to the catalytic cracking zone.

2. In the process defined in claim 1, the further improvement which comprises passing said

light condensate fraction in admixture with hydrogen through an hydrogenating zone maintained under conditions capable of converting said fraction into motor fuel, and separating a motor fuel fraction from the hydrogenated products.

3. A method of processing crude residual oils containing constituents unvaporizable under temperature conditions optimum for carrying out catalytic cracking treatment, to form lower boiling hydrocarbons in the motor fuel boiling range, which comprises subjecting said residual oil to thermal cracking treatment controlled to reduce the vaporizing temperature of a substantial portion of said residual oil to a point such that the resulting vapors can be subjected to further cracking treatment in the presence of an active catalyst under conditions optimum for the production of a high ratio of motor fuel products to normally gaseous products, separating the products from said thermal cracking treatment into vapors and a heavy residue, passing said vapors in contact with an active solid catalyst maintained at a temperature optimum for production of a maximum ratio of motor fuel to normally gaseous products, maintaining said

5 vapors in contact with said catalyst for a period sufficient to effect the desired degree of cracking, fractionally separating from the resulting cracked products a motor fuel fraction, a light condensate fraction consisting principally of constituents boiling above the gasoline boiling range containing substantial quantities of constituents less amenable to catalytic cracking than the vapors from said first named thermal cracking treatment, and a heavier condensate fraction having cracking characteristics similar to the vapors from said first named thermal cracking treatment, passing the light condensate fraction in admixture with hydrogen through a hydrogenating zone maintained under hydrogenating conditions capable of converting the condensate fraction into motor fuel, separating a motor fuel from the resulting hydrogenated products and passing said heavier condensate in contact with said active solid catalyst to effect further conversion thereof.

EGER V. MURPHREE.
ALEXIS VOORHIES, JR.
LEONARD S. BONNELL.
LOUIS S. TREGRE.