May 15, 1956 N. J. G. ALOZERY ET AL 2,745,794

COMBINATION CRACKING PROCESS

Filed Jan. 21, 1953 4 Sheets-Sheet 1

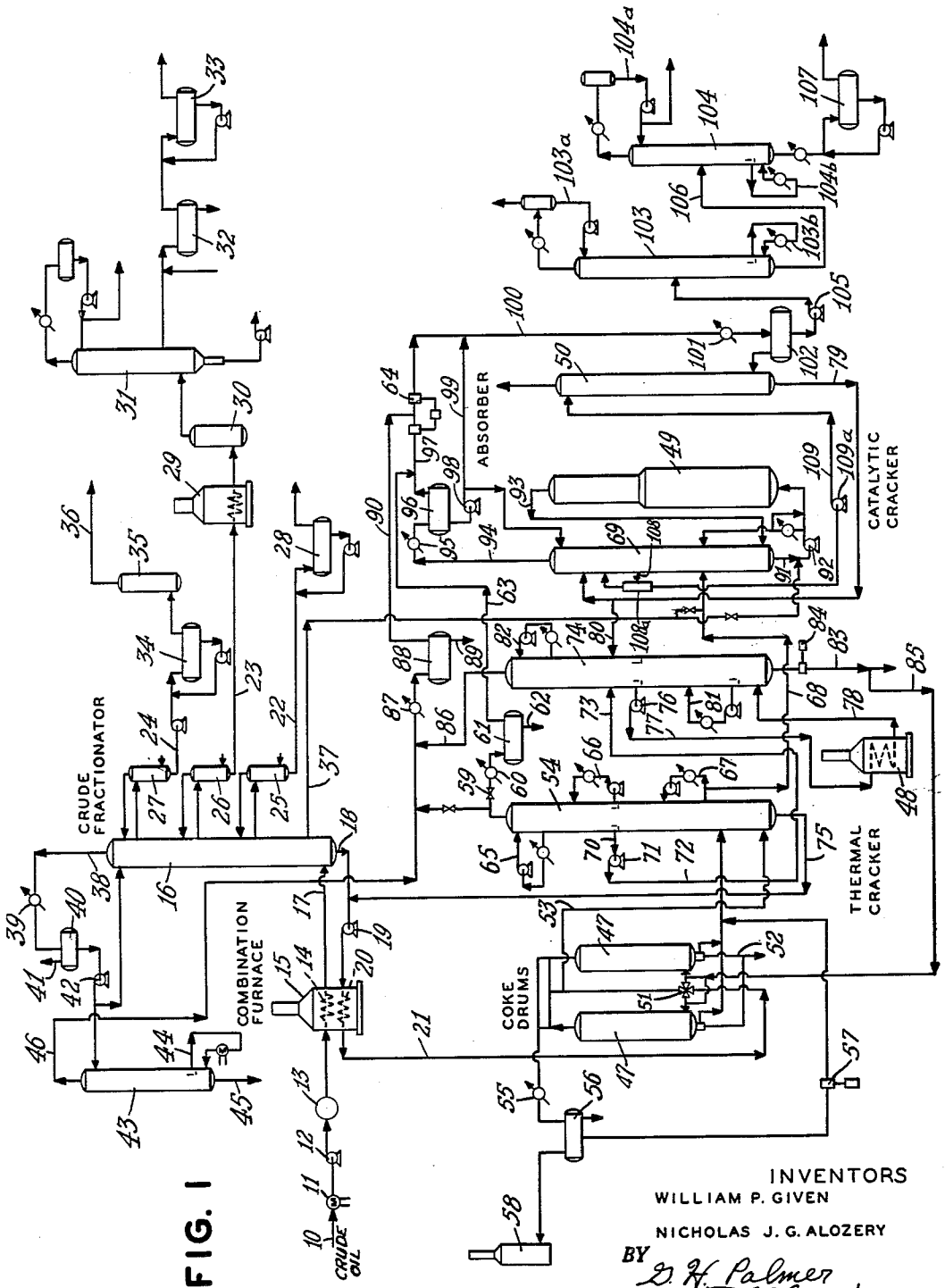


FIG. 1

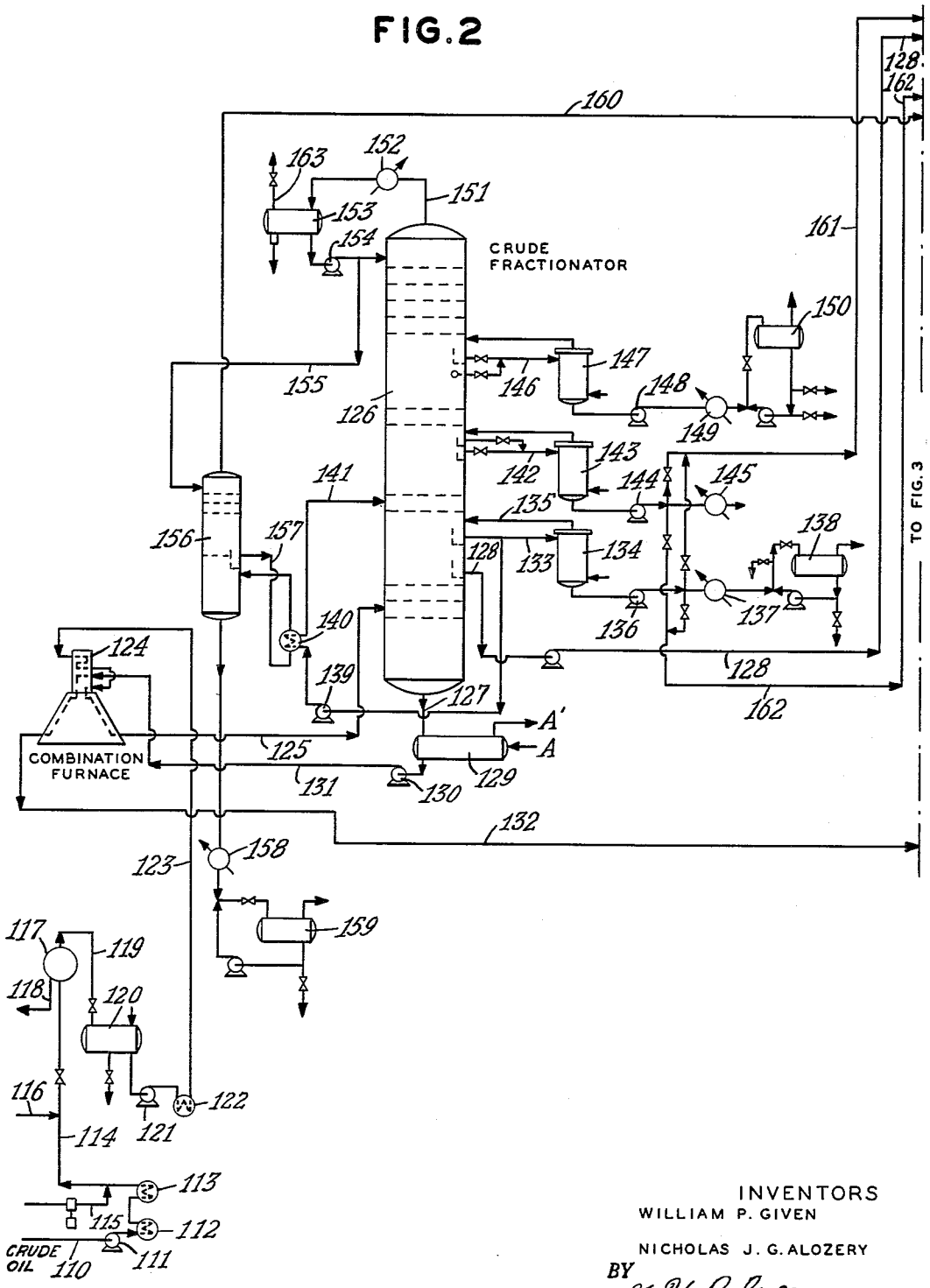
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COMBINATION CRACKING PROCESS

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FIG. 2



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COMBINATION CRACKING PROCESS

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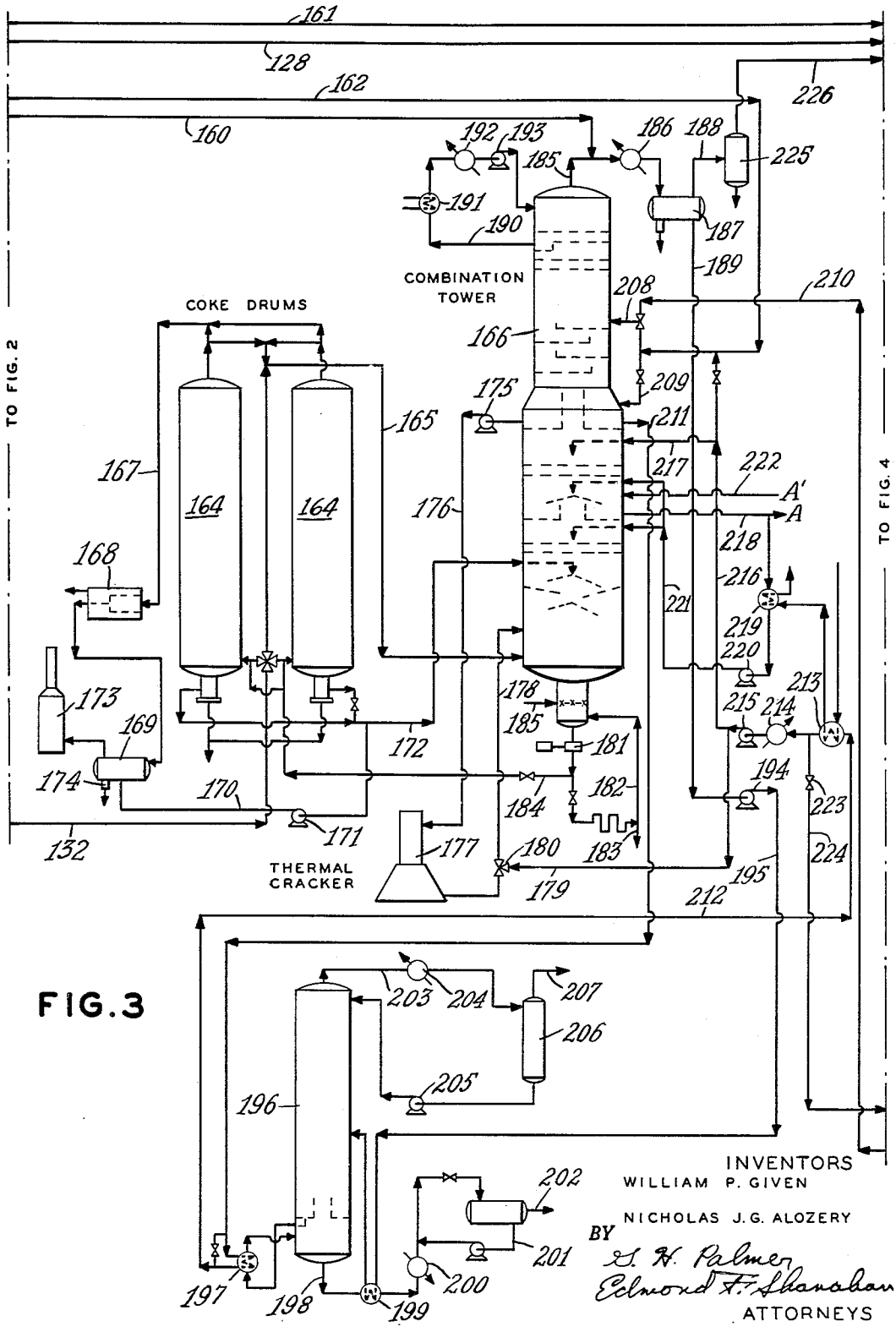


FIG. 3

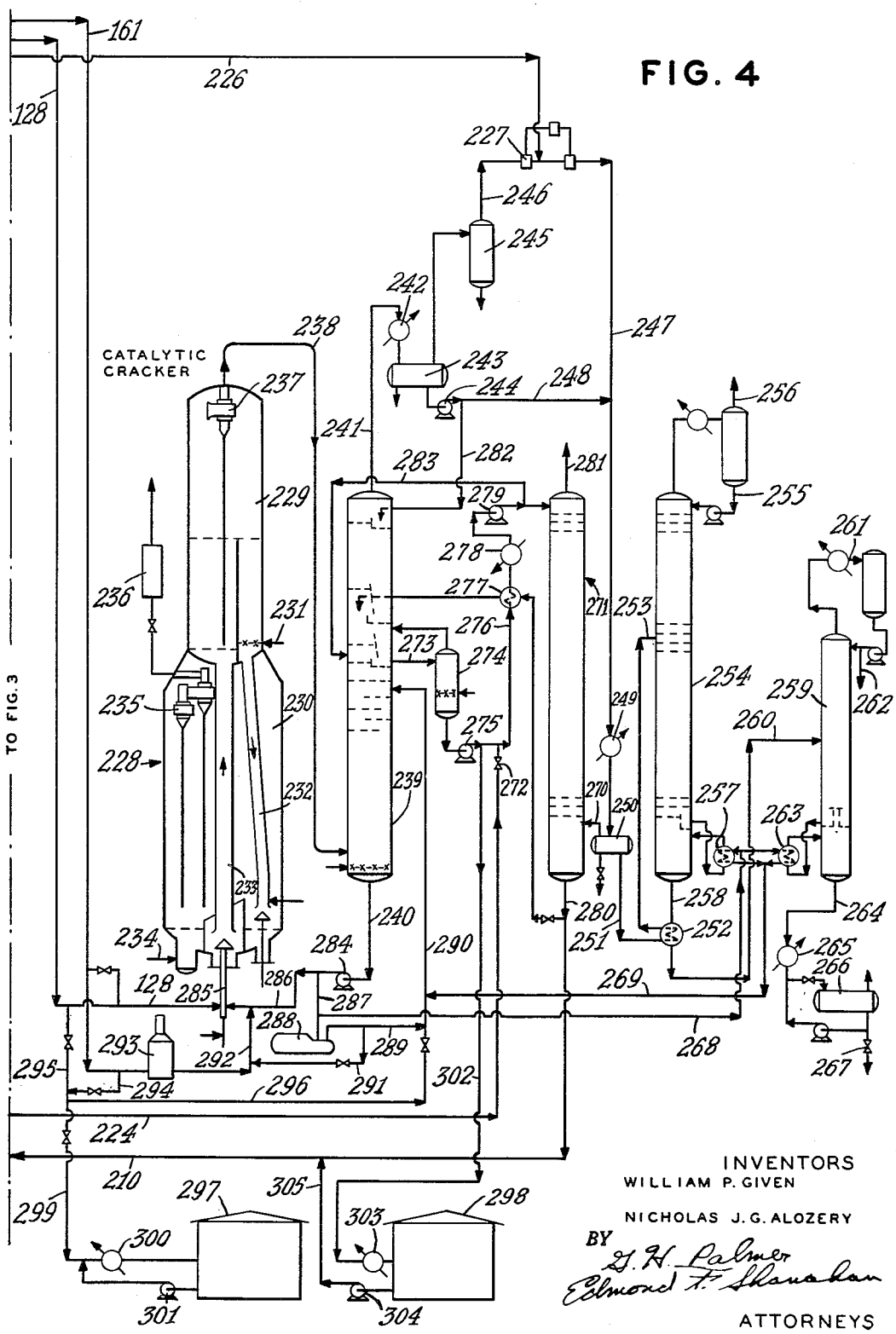
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COMBINATION CRACKING PROCESS

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FIG. 4



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2,745,794

COMBINATION CRACKING PROCESS

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Application January 21, 1953, Serial No. 332,392

11 Claims. (Cl. 196—49)

This invention is a petroleum refining process which begins with the fractional distillation of crude petroleum into reduced crude, heavy gas oil, and light straight run products, and then follows this fractionation with an integrated combination of catalytic cracking, thermal cracking, and coking to convert the reduced crude and heavy gas oil to cracked gasolines, coke, and a minimum of fuel oil or tar. The three methods of cracking are combined in the present invention into a single continuous process in which for the first time each of the cracking processes is continuously and automatically balanced with the other two by means of a common circulatory system which influences the operation of each component process in response to the variations in operation of the other components. The effluents of the three cracking processes are separated (in suitable fractional distillation zones) into rich gas, unstabilized gasoline, and other fractions. The rich gas from all three cracking steps is passed through a common absorber in which gaseous hydrocarbons are absorbed by a common absorber stream derived from at least one of the cracking effluents, the rich absorber oil is then combined with at least one of the cracking effluents. In each of the three cracking processes, it is customary to fractionate the effluent into a gasoline-containing distillate fraction and heavier fractions; but in the present combination process, it is a unique feature that the three effluent fractions are continuously in equilibrium with each other by virtue of gas oil streams which continuously flow between them without intermediate storage or wastage of heat or pressure head. Indeed, in one preferred species, the coker vapor effluent, the thermal cracking effluent, and a gas oil stream from the fractionation of catalytic cracking effluent all find their way into a single combination distillation tower. In a sub-species of the aforementioned preferred species, a gas oil fraction (mostly heavy coker gas oil) from the combination tower is mixed with reduced crude on its way from the crude fractionation to coking.

Although petroleum refining was at first merely a fractional distillation of the crude oil into a number of fractions, referred to as straight run products, modern demands for gasoline have resulted in the development of cracking processes for converting heavier petroleum components into gasoline. Each of the three cracking processes has its unique advantages and each complements the other two. Catalytic cracking is the blandest of the three; it is ideal for converting virgin gas oil to high yields of high octane gasoline. Thermal cracking is harsher and is not capable of doing with virgin feed what catalytic cracking does, but it has its best function in converting light cracked oils to gasoline. Thermal cracking produces lower octane but it is cheaper than catalytic cracking when employed with a feed which has been reduced by cracking to heavier hydrocarbons. Coking can be employed for a petroleum charge which is too heavy for either of the other two to digest efficiently. Coking is

adapted to the production of coke and coker hydrocarbon vapors; thermal and catalytic cracking are not satisfactory if coke production is high. Economic refining requires the employment of all three processes. In refineries of large capacity, for example 50,000 barrels per stream day or larger, satisfactory efficiency can be achieved with independent units; but in small refineries, provision for all these processes must be made with much lower capitalization and lower operating costs. Applicants' process is useful in a refinery of any size, but its advantages become particularly great in small refineries because it eliminates duplicate equipment and storage tanks; and it reduces operating costs because of savings in heat, pressure head, and maintenance costs.

Whatever the size of the refinery, however, the present process reduces the ground area occupied by uncombined units, the number of instrumentation centers, and the operating personnel.

Moreover, the component processes themselves are altered in character because the continuous flow of rich oil from the common absorber tower makes it possible to continuously and instantly shift light hydrocarbons from catalytic cracking effluent to thermal gasoline as desired. The composition of the heavy gas oil which is returned to the coke drums is not the same in the present combination process as in the coking process, because it contains some constituents derived from thermal cracking effluent, and even some derived from the catalytic cracking effluent by way of the absorber oil.

The behavior of the process is different by virtue of the absence of intermediate storage. Stored gas oil absorbs oxygen; when the oil is reheated for reintroduction into the process, some oxidation and deposition of clogging material takes place. Because of the continuing nature of the process, certain adjustments make themselves felt automatically. For example, the incoming crude passes in indirect heat exchange with reflux streams for the combination tower. When charge rate of the crude oil is increased or decreased, the various reflux streams are automatically cooled proportionately more or less. The absorber oil functions both as gas carrier and as relatively cool reflux; as it increases in flow to the combination fractional distillation tower, it simultaneously increases the gas content of that tower, increases the need for reflux, and itself provides the increased flow of reflux oil.

Other advantages of the invention may be understood after an explanation of the two specific embodiments which are shown by flow diagrams in the accompanying four sheets of drawings, in which:

Figure 1 is a flow sheet diagram of a combination refinery in which coker vapors and thermal cracking effluent are fractionated in separate fractional distillation towers; and

Figures 2, 3 and 4 represent a single flow sheet beginning on Figure 2 and extending on the right to Figure 3, and from the right side of Figure 3 to Figure 4. In this large flow sheet both coker vapors and thermal cracking effluent are introduced into a single combination fractional distillation tower.

In Figure 1, crude oil enters the refinery at **10**, is heated in heater **11** to a temperature of between about 100° F. and 300° F. and pumped by means of pump **12** to a pressure of preferably between 0 and 500 p. s. i. g. If the crude contains undesirable salts, such as magnesium chloride or calcium chloride, then it may be desalted in desalter **13**, which may be of any conventional type, for example, mixing with water caustic and electrostatically precipitating salts. The desalted crude is preferably preheated by heat exchange with reflux streams and then passes through crude coils **14** in combination furnace **15** and is heated to crude distillation temperatures of be-

tween 600° F. and 800° F. The crude enters crude fractionator 16 at the inlet 17 near the lower end. Reduced crude is withdrawn from the bottom of crude fractionator 16 through line 18, pumped by means of pump 19 through reduced crude coils 20 in combination furnace 15, whereby the reduced crude is heated to coking temperatures between 800° F .and 1000° F., and continuously fed to the coking step through line 21. Various light straight run fractions of successively lower boiling range may be withdrawn from crude fractionator 16 at successively higher elevations, for example, diesel oil through line 22, kerosene through line 23 and a petroleum solvent through line 24. Each of these fractions is withdrawn from fractionator 16 and steam stripped in steam strippers 25, 26 and 27, vapors being returned to the fractionator 16. The diesel oil is subsequently neutralized by washing with an aqueous alkali solution in caustic washer 28. The kerosene is converted to treated kerosene by being heated to 600° F. to 800° F. in furnace 29 at a pressure of about 50 p. s. i. g., contacted with bauxite in bauxite contacting drum 30, fractionated in rerun tower 31, and passed to sulphuric acid treatment 32, and caustic wash 33. In rerun tower 31, the kerosene is ridded of light vapors and heavy oils.

The solvent fraction may be pumped through a caustic washer 34 and a copper chloride contact tower 35, which serves to sweeten, i. e. convert mercaptan sulfur compounds to less odorous disulfides; a treated solvent is withdrawn at 36. Of course, the number of straight run products may be increased or reduced as economics dictate. For example, the diesel oil might be employed in part as catalytic cracking feed.

A gas oil fraction, the heaviest fraction next to the reduced crude withdrawn at 18, is withdrawn from the lower part of the crude fractionator 16 through line 37 for catalytic cracking.

Straight run gasoline is withdrawn as a light gaseous fraction from the upper end of the crude fractionator 16 at line 38. Of course, a heavy straight run gasoline might be withdrawn from crude fractionator 16 as a liquid. The light straight run gasoline is partially condensed by means of condenser 39, and separated in drum 40. Vapors are withdrawn through 41 and are combined with rich gas flowing to the absorber as will be explained hereinafter. The condensate is pumped by pump 42 into stripping tower 43 in which the light straight run gasoline is stripped by its own vapor reboiled by means of reboiling circuit 44. Light straight run gasoline, or light straight run naphtha, is withdrawn through line 45 and treated to caustic washing and copper chloride contacting (as shown in Figure 2) to produce a treated light straight run gasoline.

Rich gas leaves the upper end of stripper 43 through line 46 for subsequent combination with rich gas from coking, thermal cracking, and catalytic cracking.

In the lower part of Figure 1, the hot reduced crude (line 21) and the gas oil (line 37) from crude fractionator 16 are cracked in a continuous integrated system to produce coker gasoline (in coke drums 47), thermal gasoline (in thermal cracking furnace 48) and catalytic gasoline (in cracker 49); rich gas is obtained as a remainder from each of these cracking processes, and all the rich gases from the system, those derived from the three cracking processes and that derived by separation and stripping of the straight run gasoline are then passed through a single absorber 50.

The coke drums 47 are operated in alternate periods of coking and coke removal in the conventional manner. Hot reduced crude is diverted from line 21 into each of the drums 47 during alternate periods by means of the valve system indicated generally by the numeral 51. Coke is removed from these drums by conventional coke drills or hydraulic cutters, indicated diagrammatically by numeral 52. Coker hydrocarbon vapors are removed through line 53 and introduced into the lower end of a coker effluent fractionating tower 54. Coke drums 47 are also provided with a circuit for handling steam and hydrocarbon vapors produced when the coke is steamed out following a charging period. This circuit is comprised of a cooler 55, separator drum 56, and a pump 57, all of which serves the purpose of returning oil to the coker effluent fractionator 54. The blow-down stack 58 is also provided for the purpose of water scrubbing uncondensed vapors from drum 56. A light gaseous fraction is withdrawn from coker effluent tower 54. This gaseous fraction is comprised mostly of coker gasoline and it is necessary to separate it into a liquid distillate containing most of the coker gasoline, and a rich gas for diversion to absorber 50. Ordinarily, tower 54 may be operated at a pressure of 15 p. s. i. g. to 100 p. s. i. g. The coker hydrocarbon vapors may be passed through a valve 59, condenser 60, and into a drum 61, from which coker gasoline distillate may be withdrawn at 62. The rich gas from drum 61 leaves it by way of line 63 for compression in multistage compressor 64 up to the relatively high pressures of between 125 p. s. i. g. and 250 p. s. i. g., at which absorber 50 is operated.

Coker effluent fractionating tower 54 may be suitably refluxed at various elevations by withdrawing liquid, pumping it to a slightly higher pressure, cooling it, and returning it to the tower 54. Such combinations are indicated generally at 65, 66 and 67.

A part of the gas oil reflux withdrawn through line 67 is diverted to catalytic cracking by way of line 68 and catalytic cracking effluent fractionating tower 69. The most important intermediate stream of gas oil from coker effluent gas tower 54, however, is that withdrawn at 70, pumped by pump 71 through line 72 to an intermediate point 73 in thermal effluent fractionating tower 74. A heavy fraction is withdrawn from the bottom of coker effluent tower 54 through line 75 and recycled through pump 19, combination furnace 14 and line 21 to coke drums 47. However, instead of recycling all of this material, part or all may be diverted to catalytic cracker 49 or withdrawn from the system at this point as fuel oil product.

A gas oil stream is continuously withdrawn from thermal cracking effluent fractionating tower 74 and pumped by pump 76 through line 77 to thermal cracking furnace 48. The hot thermal cracking effluent returns to tower 74 from furnace 48 by way of line 78. Tower 74 may also be supplied with a rich gas oil from absorber 50 by way of line 79 and 80. Reflux circuits 81 and 82 are provided for the upper and lower regions of tower 74.

A tar fraction is withdrawn from the bottom of tower 74 by way of line 83 and pump 84. This tar may either be withdrawn as product or returned to coke drums 47 by way of line 85.

Thermally cracked gasoline is withdrawn as a vapor from the top of tower 74 by way of line 86 and joins the rich gas from the straight run gasoline in line 46. The vapor is partially condensed in condenser 87 and separated in drum 88 into thermal gasoline condensate, which is withdrawn at 89, and rich gas which is passed to an intermediate stage of multistage compressor 64 by way of line 90.

The catalytic cracker is supplied with fresh feed in the form of gas oil from crude fractionator 16 by way of line 37 which may conveniently join slurry oil withdrawn from the bottom of catalytic cracking effluent fractionator 69 at 91 and be pumped by pump 92 into catalytic cracker 49. Effluent from catalytic cracker 49 is introduced into the bottom of catalytic cracking effluent fractionator tower 69 via line 93. A vapor, comprised mostly of catalytic gasoline, is withdrawn overhead from tower 69 through line 94, partially condensed in condenser 95 and separated in drum 96 into rich gas and unstabilized catalytic gasoline. The former is passed to compressor 64 by way of line 97 and the latter is pumped by pump 98 and line 99 to be mixed with compressed rich gas in line

100, heated at heater 101, and separated into high pressure rich gas and unstabilized gasoline in high pressure separator drum 102. The catalytic gasoline is stabilized in depropanizer 103 and debutanizer 104, both of which are provided with conventional reflux circuits 103*a* and 104*a* at the top and reboiler circuits 103*b* and 104*b* at the bottom. The unstabilized gasoline moves from drum 102 by way of pump 105, and from tower 103 to 104 by way of line 106. It is caustically washed and withdrawn as a product at 107. Lean light gas oil for absorber 50 is obtained from catalytic cracking effluent fractionating tower 69 at 108, being steam stripped in steam stripper 108*a*, and transferred to absorber 50 by way of line 109 and pump 109*a*.

The combination refinery of Figs. 2, 3, and 4 employs a single combination tower (seen in Fig. 3) in place of coker effluent fractionating tower 54 (of Fig. 1) and thermal cracking effluent fractionating tower 74 of Fig. 1.

Crude petroleum enters the system at 110 and is pumped by means of pump 111 through a heat exchanger 112, in which the crude petroleum is raised to a temperature of about 125° F., by indirect heat exchange with exhaust steam; the crude oil is next passed through a heat exchanger 113 through which it is heated to a temperature of about 220° F., by indirect heat exchange with reflux oil withdrawn from the top of the combination tower, as will be described hereinafter. The heated stream of crude oil flows through line 114 picking up an aqueous alkali solution of sodium hydroxide at 115 and water at 116, and then enters desalting drum 117, in which settling takes place; salt solution is withdrawn from desalting settling drum 117 at 118, and dry desalted crude is withdrawn overhead through line 119, passed through surge drum 120, and pumped by pump 121 through heat exchanger 122, in which the crude is heated from a temperature of about 220° F., to a temperature of about 540° F. by indirect heat exchange with hot reflux streams from intermediate points of the combination tower as will be described hereinafter. The hot crude next flows via line 123 through crude heater lines in combination furnace 124, being heated to a temperature of about 725° F., and flows via line 125 to a flash zone in the lower portion of topping tower 126, also designated as the crude fractionator. The crude fractionator is typically about 95 feet in vertical extent, provided with about 33 trays, and controlled to provide a temperature gradient ranging from about 710° F. near the bottom to about 235° F. near the top. From the standpoint of steps in the combination process, the two most important streams withdrawn from crude fractionator 126 are the reduced crude stream withdrawn from the bottom by way of line 127 and the gas oil stream for catalytic cracking withdrawn a short distance above the flash zone through line 128. The reduced crude stream flows into a surge drum 129 in which it mixes with heavy gas oil from the combination tower, as will be described hereinafter. The mixture of reduced crude and heavy gas oil is pumped by means of pump 130 and line 131 through coking heater lines in combination furnace 124 and then by way of line 132 to the coke drums as will be described hereinafter. In addition to the reduced crude streams and the gas oil streams from crude fractionator 126, straight run products are withdrawn at successively higher elevations and light straight run gasoline is withdrawn as a vapor from the top.

Relatively heavy straight run sidestream withdrawn at 133 may be stripped in 134, vapors being returned via 135, and the stripped sidestream being pumped by way of pump 136 through cooler 137 to a caustic washer 138. This sidestream may be diesel oil, and may be subjected to subsequent conventional treatment, if desired. A part of the diesel oil side stream withdrawn at line 133 is pumped back to fractionator 126 by means of pump 139 being passed in heat exchange on the way through reboiler 140. The reintroduced stream serves as supplemental reflux in the lower section of the tower. Because of the withdrawal of several side streams further up, it would otherwise be necessary to reflux excessively in the upper part of the tower and greatly enlarge the upper part of the tower, in order to maintain adequate reflux in the lower section. Preferably, the refluxed stream enters crude fractionator 126 through line 141 at a point one or two trays above its point of withdrawal.

The next lighter sidestream is withdrawn a little ways farther up fractionator 126 at 142, stripped as before at 143, pumped and cooled at 144 and 145; this may be a kerosene stream suitable for finishing as a kerosene product.

Various other straight run streams may be withdrawn. In this example, the next higher stream, withdrawn at 146, stripped at 147, pumped by pump 148 through cooler 149 and caustic washer 150, may be heavy straight run gasoline suitable for conventional treating and blending.

Light straight run gasoline is withdrawn as vapor at the top of fractionator 126 through line 151, partially condensed in condenser 152, passed to reflux accumulator 153, and pumped by pump 154 in part back to fractionator 126 as reflux; the net production is diverted through line 155 to light straight run stripper 156 in which the light straight run gasoline is stripped by its own vapors, provided by heating a stream withdrawn through line 157, partially vaporized in previously mentioned reboiler 140. The stripped light straight run gasoline is withdrawn from the bottom of stripper 156, cooled at 158, caustic washed at 159, and is withdrawn from the combination circuit as light straight run gasoline product; or at least as light straight run gasoline suitable for such product finishing as may be desired.

Hydrocarbon vapors are withdrawn from the upper end of stripper 156 by way of line 160 at a temperature of about 143° F. and a pressure of about 60 lbs. per square inch. Usually these hydrocarbons will be butane and "lighter," i. e., of lighter boiling point; however, the gasoline fraction may be increased or reduced, depending upon the light straight run gasoline specifications, by varying the stripping temperature and/or pressure.

Extra piping circuits are provided so that the kerosene and/or diesel products may be reduced by diverting a part of either or both of these streams to catalytic cracking or to the combination tower for thermal cracking, thus maximizing gasoline production and minimizing kerosene and diesel production. Various reasons dictate such manipulation of the refinery, for example, product specifications, the nature of the crude, the state of storage capacity, etc. Line 161 is employed for diverting kerosene or diesel oil streams, or both to catalytic cracking. Line 162 is used for diverting kerosene or diesel oil streams or both to the combination tower.

Rich gas remaining uncondensed in reflux accumulator 153 is withdrawn through line 163 at relatively lower pressure, probably about 4 p. s. i. g., and like the stripped gas in line 160, eventually finds its way to a common compressor and absorber stream which will be discussed hereinafter.

Fig. 3 illustrates the coking and the thermal cracking sections of the combination refinery. The heated mixture of reduced crude and heavy coker gas oil, having been heated in combination furnace 124 to a temperature of about 930° F., is introduced during alternate periods of time into one or the other of paired coke drums indicated by the numeral 164. The hot oil is partially converted to coke in coke drums 164. During alternate periods of about 24 hours, solid coke is removed from the bottom of the coke drums 164. During charging periods, coker vapors are continuously withdrawn overhead through line 165 at a temperature of about 830° F. and at a pressure of about 80 p. s. i. g. and introduced into the lower end of a combination tower 166. Immediately after the charging period for each coke drum, the drum is purged with steam. The steam and purged vapors may be discharged into combination tower 166; but if desired, the

purging vapor mixture may be withdrawn through line **167**, cooler **168**, blow down accumulator drum **169**, whereby the condensed portion is separated and the hydrocarbons returned via line **170**, pump **171**, and line **172** to an intermediate point in the lower section of the combination tower **166**. Vapors from blowdown accumulator **169** are water scrubbed in blowdown stack **173**; a water draw-off **174** is provided on accumulator **169**.

The thermal cracking is carried out by continuously pumping, by means of pump **175**, a stream of light gas oil from an intermediate trap pan in combination tower **166** and passing it by way of line **176** to a thermal cracking furnace **177**, in which it is heated to a temperature of about 985° F., before return to a low point in combination tower **166** through line **178**. The pressure within combination tower **166** is typically 60 or 70 p. s. i. g., with 5 or 10 pounds pressure gradient from bottom to top. Pump **175** increases the pressure on the thermal cracking feed oil to a suitable thermal cracking pressure of about 1000 p. s. i. g. at the thermal cracking furnace inlet. The hot effluent leaves the thermal cracking furnace **177** at a pressure of about 720 p. s. i. g. In order to avoid coking in returning line **178**, relatively cool light gas oil, obtained in a manner to be described hereinafter, is introduced by line **179** at pressure reduction valve **180**, so that thermal cracking effluent is reduced in temperature to about 800° F. by direct heat exchange at the same time its pressure is reduced to about that in the lower part of combination tower **166**.

Tar is pumped from the bottom of combination tower **166** by means of pump **181**, cooled, and recirculated by a recirculation circuit **182**, the net tar production being withdrawn at **183**. However, if it is desired to further minimize tar production and maximize gasoline production, a substantial portion or all of the tar may be returned to coke drums **164** by way of recycle line **184**. The tar may be steam stripped just prior to withdrawal by means of steam stripper **185**.

A light gasoline fraction, comprised mostly of thermal and coker gasoline is withdrawn from the top of combination tower **166** at line **185**, partially condensed in condenser **186** and separated in gas separator **187** into a rich gas withdrawn at **188** and unfinished gasoline product at **189**. The composition of both the rich gas in **188** and the gasoline withdrawn from **189** is affected by commingling of rich gas diverted from the straight run gasoline to line **185** by way of line **160**. The combination of straight run gasoline rich gas with the vapor fraction from combination tower **166** makes unnecessary any separate provision for cooling or processing the rich gas from the light straight run gasoline.

Reflux for the top of combination tower **166** is provided by withdrawing liquid through line **190**, cooling in heat exchanger **191** by indirect heat exchange with the crude charge, for example heat exchanger **113**, seen in Fig. 2. The reflux is further cooled in coiler **192** and pumped back into combination tower **166** by pump **193**.

The mixture of thermal and coker gasoline (enriched to some degree by components of rich gas, may be light straight run gasoline) withdrawn from gas separator **187** by way of line **189** is pumped by pump **194** via line **195** to a conventional gasoline stabilizing tower **196**, whose reboiler **197** is heated by indirect heat exchange with a refluxing stream to be described hereinafter. Stabilized gasoline is withdrawn from stabilizing tower **196** at **198**, cooled in heat exchanger **199** by indirect heat exchange with the gasoline stream in line **195**, further cooled by cooler **200**, and caustic washed in caustic washing circuit **201**. Stabilized and washed gasoline is withdrawn at **202** as a product or for further treatment. Overhead vapors are withdrawn through line **203**, and partially condensed in condenser **204**, condensate being returned to stabilizing tower **196** as reflux by means of pump **205** and vapors being withdrawn from reflux accumulator **206** by way of line **207** for use as fuel gas.

At an intermediate point in combination tower **166**, third and fourth charge streams, supplementing coking effluent and thermal cracking effluent are introduced. One of these streams is surplus kerosene or diesel oil brought in through line **162** as previously described and introduced into tower **166** through line **208**, or in exceptional cases through line **209**. The fourth source of charge for combination tower **166** is a stream of rich oil derived from an absorber to be described hereinafter and containing components of rich gas from straight run gasoline, from coking and thermal cracking, and from a catalytic cracking step yet to be described. This rich gas is brought in via line **210**. The continuous flow of rich oil from the absorber to the upper part of combination tower **166** thus provides a common meeting place for products of every step in the combination process. It is a novel characteristic of the invention that at least part of the feed of combination tower **166** (or, if there are 2 towers, to the thermal carcking effluent fractionator) is a light gas oil, usually a rich oil of the nature already described, or lean oil derived from the effluent of catalytic cracking.

Combination tower **166** is refluxed at various elevations with light gas oil withdrawn at line **211**, cooled as previously described in reboiler **197**, returned by way of line **212** to heat exchanger **213** by indirect heat exchange with crude flowing through heat exchanger **122** in Fig. 2. The light gas oil reflux is further cooled in cooler **214** then pumped via pump **215** and lines **216** and **217** to various intermediate elevations in combination tower **166**.

Heavy gas oil is withdrawn from the lower section of combination tower **166** by way of line **218**. Part of this is cooled in heat exchanger **219**, preferably by indirect heat exchange by crude as in the case of heat exchanger **213** and then pumped by pump **220** back to combination tower **166** by way of line **221**. A second portion of the withdrawn heavy gas oil is diverted at A to surge drum **129** in Fig. 2, as previously described. Vapors from surge drum **129** return as indicated in A' by way of line **222** to combination tower **166**.

If desired, the combination tower may be employed as a source for part or all of the lean oil required in the absorption step. A suitable place for diverting gas oil to be used in this fashion is by way of valve **223** and line **224** immediately downstream from heat exchanger **213**.

Combination tower **166** performs the function of both coker gas fractionator **54** and thermal cracking effluent fractionator **74** in the embodiment of Fig. 1. There are certain important differences, however. Whereas in the embodiment of Fig. 1 heavy coker gas oil was withdrawn from an intermediate point in the coker gas fractionator **54** and diverted by way of line **68** to catalytic cracking effluent fractionator **69** whereby the heavy coker gas oil becomes part of the feed to the catalytic cracking step, in the combination tower version, the heavy gas oil stream withdrawn from combination tower **166** by way of line **218** finds its way back to the coke drums in the manner already described. Although the heavy gas oil withdrawn through line **218** contains substantially all of the heavy coker gas oil, it also contains a heavy component of the thermal cracking effluent. Other differences between the two embodiments illustrated follow from the fact that all the vapors from both the coker gas effluent and the thermal cracking effluent are mixed within the same fractional distillation zone and are withdrawn at a temperature somewhat higher than would be necessary to operate the coker gas effluent fractionator **54**. Combination tower **166** operates at a typical pressure of about 65 p. s. i. g., but the pressure may range from between 20 to 350 p. s. i. g. depending on economic and process conditions at the time of design. At pressures of 60 to 70 p. s. i. g., a typical temperature range within combination tower **166** would be from about 780° F. at the bottom to about 340° F. at the top. However, at higher pressures

these temperatures would be correspondingly higher; but whatever the pressure, the practical temperature ranges for the bottom and top respectively would be about 100° F. to 450° F. and 600° F. to 900° F. A satisfactory gradient of temperature is continuously maintained by the rate of reflux at the different points already described. The primary object is to so regulate conditions within combination tower **166** as to fractionate satisfactorily, in the order of decreasing boiling range, tar, a heavy gas oil for further coking and for refluxing, a light gas oil for thermal cracking and for refluxing and a vapor fraction comprised of gasoline and lighter ends.

Combination tower **166** provides a convenient means for shifting cracking load back and forth between coke drums and thermal cracker so as to operate both at maximum load efficiency. The temperatures at the top of combination tower **166** are determined by the specifications laid down for the gasoline to be withdrawn there. The temperatures at the bottom of the combination tower are established by tar specifications. But at the elevations of the side streams (light gas oil for thermal charge and heavy gas oil for recycle to coker) withdrawn through lines **218** and **175**, the operator may manipulate temperatures so as to increase or reduce the flow of either of the streams, this manipulation being accomplished by increasing or decreasing the rate of reflux just above these points of withdrawal.

The rich gas separated from the overhead gasoline fraction through line **188** is passed through a drum **225** to entrap any uncondensed liquid and then by way of line **226** to an intermediate stage of a multistage compressor **227** (Fig. 4).

The principal source of fresh feed for catalytic cracking is the gas oil stream withdrawn from crude fractionator **126** through line **128**. Most of the time this stream will flow directly into catalytic cracker **228**. Catalytic cracker **228** is comprised of a reactor section **229**, superimposed on a regenerator section **230**, each containing its own fluidized bed of catalyst, not shown. Catalyst is continuously circulating from reactor **229** to regenerator **230** by way of steam stripper **231** and standpipe **232**; regenerated catalyst and catalytic cracking feed enter reactor **229** through catalyst riser **233**. Air is supplied to regenerator **230** at **234** and flue gas escapes by way of cyclone **235** and stack **236**. Catalytic cracking effluent is withdrawn by way of cyclone separator **237** and line **238** to the bottom of a catalytic cracking effluent fractionator **239**, in which said effluent is fractionated into slurry oil withdrawn at the bottom of fractionator **239** through line **240**, and vaporous hydrocarbons comprised of catalytic gasoline and lighter components withdrawn at the top, through line **241**.

The vapor fraction of the catalytic cracking effluent is partially condensed by means of condenser **242**, condensate is separated from vapor in gas separator **243**, the condensate being pumped up to a pressure of about 135 p. s. i. g. by means of pump **244** and the remaining vapor being passed through drum **245** to eliminate liquid and then by way of line **246** to multistage compressor **227**, in which it is compressed up to a pressure of about 135 p. s. i. g. At an intermediate stage it is commingled with rich gas from line **226** (separated from the thermal and coking gasoline taken overhead from combination tower **166**). The compressed rich gas is withdrawn from compressor **227** by way of line **247**, combined with the pumped condensate from line **248**, passed through cooler **249** to a high pressure gas separator **250**.

The condensate in high pressure gas separator drum **250** is an unstabilized catalytic gasoline. This gasoline is continuously withdrawn through line **251**, heated in heat exchanger **252** (by indirect heat exchange with depropanized gasoline), and passed by way of line **253** to depropanizer tower **254**, and is provided at its upper end with a reflux circuit **255** and a reflux discharge **256** and at the lower end with a reboiler circuit **257**. Depropanized gasoline leaves the bottom of depropanizer tower **254** through line **258**, is cooled in heat exchanger **252**, and enters debutanizer tower **259** by way of line **260**. Debutanizer tower **259** is provided with a reflux circuit **261** at its upper end and a tap **262** for withdrawing net butane production, a reboiler **263** is employed for introducing heat in the lower region of debutanizer tower **259**. Debutanized gasoline is withdrawn from the bottom of debutanizer tower **259** through line **264**, cooled in cooler **265**, caustic washed in caustic washer **266** and withdrawn at **267** as catalytic gasoline product. Reboilers **257** and **263** may conveniently derive their heat by indirect heat exchange with a stream of slurry oil from the bottom of catalytic effluent fractionator **239**, the slurry oil being brought to the reboilers by way of line **268** and returned to the catalytic cracking effluent fractionating system by way of line **269**.

Compressed rich gas derived from all the gasoline streams produced by the combination unit is withdrawn from high pressure gas separator **250** through line **270** and introduced into the bottom of an absorber tower **271**, in which it is counter-currently contacted with a downflowing light gas oil in order to absorb light hydrocarbons into a gas oil. A light gas oil is generally derived as a side stream of the catalytic cracking effluent fractionator **239**; but all or a part of the light gas oil for the absorber tower **271** may be obtained from combination tower **166**, for example through line **224** by manipulation of the valve **272**. The preferred arrangement, however, is to withdraw a stream of light gas oil from catalytic cracking effluent fractionator **239** through line **273**, passing it through steam stripper **274**, and pumping it by means of pump **275** through line **276**, cooling it by means of heat exchanger **277** and cooler **278** before pumping it by means of pump **279** into the top of absorber tower **271**. The light gas oil entering the top of the absorber tower **271** is conventionally referred to as "lean oil," and as "rich oil" as it leaves the bottom of absorber tower **271** by way of line **280**, because it has absorbed a substantial percentage of propane and heavier hydrocarbon vapor from the rich gas as said gas ascends through the tower. Conversely, the rich gas entering at line **270** leaves the top of absorber **271** through line **281** as "lean gas" suitable for fuel gas.

Catalytic cracking effluent fractionator **239** may be suitably refluxed by returning some of the overhead condensate from line **248** by way of line **282**; reflux for the mid-point may be obtained by diverting part of the lean light gas oil entering the top of absorber **271** and returning it to catalytic cracking effluent fractionator **239** by way of line **283**. The lower region of catalytic effluent fractionator **239** may be refluxed by returning some of the slurry oil, preferably slightly cooled. The piping arrangement for slurry oil withdrawn through line **240** provides for pumping it by means of pump **284** directly to the charging valve **285** in the bottom of the catalytic cracker **228** by way of line **286**. At least part of the slurry oil, sometimes all of it, flows directly to the catalytic cracker. A second part is continuously being returned to catalytic cracking effluent fractionator **239** as reflux by way of line **287**, boiler **288** (in which the slurry oil is cooled and steam is generated), and then by lines **289** and **290** to tower **239**. As a temperature control feature, some of the slurry oil may be diverted through line **291** and line **292** to the catalytic cracking charge.

Excess kerosene or diesel oil diverted to catalytic cracking by way of line **261** may be heated in furnace **293** before continuing to catalytic charge via lines **292** and **286**, and charge valve **285**. As a control feature, unheated material from line **161** may be diverted to catalytic cracking feed by way of lines **294** and **295**.

Although all the heavy gas oil may be continuously introduced directly into catalytic cracking by way of line **128**, all or a part of this fresh feed may be diverted through line **295** to line **296**, and then by way of line

290, to the lower region of catalytic cracking effluent fractionator 239, so that the fresh feed finds its way to the catalytic cracker 228 in a mixture with the slurry oil withdrawn at line 240. This would be done when there seemed to be inadequate downflowing liquid in fractionator 239 to wash down catalyst.

In a unit of the combination type herein described many savings have been effected by integrating all operations, eliminating duplicate pieces of equipment, eliminating storage capacity, saving heat and pressure head and making each product according to a specification which is continuously affected in most cases by the specifications of other products. However, in order to make it possible to temporarily shut down parts of the refinery for occasional emergencies, inspections, or repairs, 2 storage tanks are provided in which may be stored enough in process oil to keep most of the refinery running during shutdown of one component part. These storage tanks are the catalytic feed flywheel tank 297 and the termal feed flywheel tank 298. Catalyst feed flywheel tank 297 is filled during normal operations by way of lines 295 and 299, and cooler 300. When it is necessary to continue operation of the catalytic cracking section during shutdown of other sections of the refinery, catalytic feed is pumped through tank 297 by means of pump 301 and by way of lines 299, 295 and 128 to catalytic cracking feed valve 285. Tank 298 is filled with feed suitable for combination tower 166 during normal operation by diverting to it a stream of gas oil through line 302 and cooler 303. During shutdown of the catalytic cracking section, for example, thermal feed may be pumped by means of pump 304 from tank 298 through line 305 to line 210 to combination tower 166.

It will be understood that if the market for coke is poor and the market for fuel oil is good, that it will be preferred to do less coking, or to eliminate coking almost entirely, by diverting the heated reduced crude from furnace 20 or furnance 124 around the coke drums (which would be cut of use when operating in this manner) directly to fractionating tower 54 or combination tower 166, respectively. In this manner, the furnace ordinarily serving as a coker furnace actually functions as a vis-breaking furnace. The combination unit then functions as a balanced combination of vis-breaking, thermal cracking, and catalytic cracking instead of the combination of coking, thermal cracking, and catalytic cracking. A significant range of control of the relative productions of coke and fuel oil is possible by varying the amount of heavy gas oil from the combination tower recycled to the coker furnace, a reduction in recycle rate increasing the relative fuel production, and vice versa.

Additional advantages of my invention are illustrated in the following example. The reactants and their proportions, and other specific ingredients and conditions, are presented as being typical and should not be construed to limit the invention unduly. A Redwater crude of 34.9° A. P. I. was charged to the topping tower for several weeks and fractionated as follows:

TABLE

Fraction	Vol. percent on crude charge	° API Gravity	A. S. T. M. Distillation Range	
			I. B. P.	E. P.
Gas to Recovery	2.8			
Virgin Naphtha	22.2	62.5	97	339
Solvent	3.9	46.8	294	390
Kerosene	6.8	41.4	364	458
Diesel Oil	5.0	36.5	428	565
Unstripped Virgin Heavy Gas Oil	35.3	29.2	387	
Reduced Crude	24.0	11.8	476	

Temperatures and pressures within the topping tower

Operating pressure at flash zone	p. s. i. g.	9.0
Temperatures:		
Base	°F.	685
Flash zone	°F.	710
Virgin heavy gas oil drawoff	°F.	600
Diesel oil drawoff	°F.	455
Kerosene drawoff	°F.	394
Solvent drawoff	°F.	350
Fractionator top	°F.	235

The heavy gas oil fraction was withdrawn at a temperature of about 600° F. and pumped to a pressure of about 75 p. s. i. g. and injected into a fluidized catalytic cracker wherein the heavy gas oil was cracked at a temperature of about 960° F. and a pressure of about 15 p. s. i. g., a catalyst to oil ratio (basis of total reactor charge) of about 4 to 1 and a liquid volume conversion of 66 per cent (basis of fresh feed). The effluent of the catalytic cracking zone and part of the absorber rich oil were charged to the catalytic cracking effluent fractionating tower. A certain amount of a slurry and heavy gas oil is recycled to the catalytic cracking. The combined charge of heavy virgin gas oil plus rich oil charged to the section comprising the catalytic reactor and the catalytic effluent fractionating tower was approximately 2900 B. P. S. D. and produced yields which were about as follows on a weight per cent basis of this charge:

	Weight percent
Catalytic cracked gas and gasoline	37.0
Lean oil to absorber	57.5
Catalytic cracking-coke	5.5

The pressure in the catalytic fractionator was approximately 9.0 p.s. i. g.

The temperatures within the catalytic fractionator were:

Tower base	°F.	507
Cycle oil drawoff	°F.	448
Fractionator top	°F.	264

The light gaseous fraction was cooled to condense about 60 weight per cent, and the uncondensed portion was compressed to a pressure of 55 p. s. i. g. in the first stage of the compressor system. The condensed portion was pumped to a pressure of about 160 p. s. i. g., recontacted with vapors from the second stage of the compressor and the total effluent was flashed at a pressure of about 137 p. s. i. g. and a temperature of about 100° F. The uncondensed vapors (or rich gas) were charged to the absorber. The condensed portion was pumped to a pressure of about 310 p. s. i. g. and charged to the stabilizer.

The reduced crude from the bottom of the topping tower was mixed with heavy coker gas oil from the combination tower and the mixture was passed through a furnace and thence to coke drums to produce coke and coker vapors. The coker vapors at a temperature of approximately 830° F., part of the rich oil from the absorber, and thermal cracking effluent were then charged to a combination tower to produce fuel oil and thermal cracked gas plus gasoline which was destined for subsequent steps.

The combined total charge to the coking and thermal cracking section consisting of reduced crude and part of the absorber rich oil which comprised the net production of catalytic cycle oil, was approximately 2100 B. P. S. D., and produced yields which were about as follows on a weight per cent basis of the charge:

	Weight percent
Gas plus gasoline	57
Fuel oil	26
Coke	17

The thermal feed fraction from said combination tower was continuously passed through a thermal cracking furnace and thermally cracked at temperatures of 966° F. and pressures of 720 p. s. i. g. and the effluent quenched

to a temperature of 800° F. by either light or heavy gas oil from combination tower, and introduced into said combination fractionating zone at an intermediate point having a temperature of about 780° F.

The gasoline fraction from the upper end of the combination fractionating zone was cooled to a temperature of 100° F. to condense out a 70 weight per cent fraction of unstabilized gasoline. The uncondensed portion was then combined with the partially compressed uncondensed portion of the catalytic cracking effluent and further compressed to a pressure of 137 p. s. i. g. The condensed liquid was pumped to a similar pressure and mixed with the compressed vapor at a temperature of about 150° F. The mixture is then cooled to a temperature of about 100° F. and the high pressure liquid so formed was withdrawn as unstabilized gasoline having an A. S. T. M. end point of 400° F. The remaining rich gas is charged to the absorber. The rich absorber oil is then charged to the combination cracking zone.

A 60 weight per cent of said rich oil from the absorber was charged to the catalytic cracking effluent fractionation step.

The stabilized gasoline was then charged to a debutanizer tower which operated at a pressure of 80 p. s. i. g. An overhead liquid butane product was removed for vapor pressure blending purposes with the thermal and straight run gasolines. The debutanizer bottoms product was catalytic gasoline and was cooled and caustic washed.

The following yields were obtained when charging approximately 1200 B. P. S. D. of unstabilized separation liquid to the stabilizer and debutanizer section:

Charge:

Unstabilized gasoline to stabilizer___ percent__ 100

Yields:

	Vol. percent
Stabilizer gas	8.8
Butane produced	8.2
Debutanized catalytic gasoline	83.0

Although this process has been described and exemplified in terms of its preferred modifications and the above example, it is understood that various changes may be made without departing from the spirit and the scope of the disclosure and the claims. For example, as shown in the drawings and description, there may be either a single combination tower or two effluent fractionating towers in its stead. Moreover, a vacuum still might have been employed as follows:

Reduced crude from the crude fractionator (**16** or **126**) would be processed in a vacuum still to yield an overhead gas oil stream and a heavy residuum bottoms. The overhead gas oil would be charged to the catalytic cracker. The residuum bottoms would be processed in either of the following manners dependent on the refinery condition.

(*a*) Residuum bottoms would be charged to a heater and heated to coking temperatures and the effluent vapors would discharge to the combination tower.

(*b*) Residuum bottoms would be charged to a heater and subjected to a vis-breaking operation in which the viscosity of the vacuum residuum would be reduced to that of a marketable fuel oil. The vis-breaker gasoline and gas oil would then enter the thermal stream.

Herewith is presented a typical overall weight balance of all products from this refinery based on crude charge:

	Weight per cent
Total crude charge, 5500 B. P. S. D.	
Redwater crude, 34.9° API	100
Yields:	
Total gas	10.5
Total gasoline	57.5
Diesel oil	11.0
Fuel oil	10.5
Coke-thermal	7.5
Coke-catalytic	3.0

It will be further understood that "finely divided catalyst" includes not only powdered catalyst but those of the pellet type used in other types of catalytic cracking, since the invention is not restricted to fluidized catalytic cracking, although the latter is the preferred form.

Each step of the process is characterized by a tolerable range. The crude petroleum may have been an A. P. I. gravity between about 10 and 50. The reduced crude may have an initial boiling point ranging between 350° F. and 850° F., the heavy gas oil fractions boiling ranges being between 300 and 700° F. initial, and 600 to 1000° F. final. Pressures and temperatures in the topping tower may vary from zero to 50 p. s. i. g. and temperatures of between 400 and 800° F. at the bottom.

The boiling points referred to in the foregoing specification and in the boiling points given hereafter, refer to the American Society for Testing Materials standard procedures.

The reduced crude fraction leaves the topping tower at a temperature of between 350 and 775° F. and is heated to coking temperatures of between 770° F. and 1000° F. to enter the coke drums, producing coker hydrocarbon vapors having a final boiling of between 600 and 1100° F. These vapors are transferred from the coke drums to the coking effluent fractionation zone without cooling below a temperature of about 700° F. or being allowed to fall beneath the pressure of about 15 p. s. i. g. The vapors are fractionated at a pressure of between 15 and 200 p. s. i. g. and the product fractions always include a thermal cracking light gas oil fraction having a boiling range between 300 and 500° F. initial and 550 and 750° F. final. This thermal cracking feed fraction is withdrawn from the fractional distillation zone at a temperature of 600° to 800° F. and a pressure of 20 to 200 p. s. i. g. Thermal cracking conditions may be in a range of between 900 and 1100° F., under a pressure of 50 to 1000 p. s. i. g. The thermal cracking effluent is quenched to a temperature of between 900 and 700° F. The two gasoline-containing effluent fractions from the catalytic cracking fractionation zone and from either the combination fractionating zone or the coking effluent fractionation zone and the thermal cracking effluent fractionation zone, are cooled to a temperature of between 80° and 150° F. to condense unstabilized gasoline fractions, the uncondensed portions being compressed to absorber pressure which may be between 50 and 350 p. s. i. g.; the remaining unstabilized gasoline will have an A. S. T. M. end point of between 350 to 450° F. The absorber oil may be a lean gas oil side stream from the catalytic cracking effluent fractionation zone having an A. S. T. M. boiling range of between 350 to 500° F. initial and 600 to 700° F. final boiling point.

We claim:

1. A continuous process for refining crude petroleum to produce a maximum of light straight run products and cracked gasoline, which includes the steps of: continuously introducing said crude petroleum into a crude petroleum distillation zone and fractionating said crude petroleum therein into a reduced crude fraction having a boiling range starting between 400° F. and 900° F., a heavy gas oil fraction having a boiling range of between 300° F. to 700° F. initial and 600° F. to 1100° F. final, and straight run fractions of lower boiling range; continuously withdrawing said heavy gas oil from said crude petroleum distillation zone at a pressure between zero and 50 p. s. i. g. and a temperature between 500° F. and 800° F. and contacting said heavy gas oil with finely divided solid catalyst at cracking temperatures and pressures without reducing the temperature of said heavy gas oil below 450° F. and catalytically cracking said oil; withdrawing an effluent from said catalytic cracking step at a temperature of between 850° F. and 1000° F. and a pressure of between 5 and 50 p. s. i. g., and introducing said effluent into a vertically extended catalytic cracking effluent fractionation zone operating at a temperature of

between 90 to 200° F. at the top and 600 to 900° F. at the bottom, and a pressure of between zero and 50 p. s. i. g., without reduction in temperature of said effluent below 600° F. before introduction; withdrawing a light gaseous fraction having a boiling range ending at from 250 to 450° F. from the upper end of said catalytic cracking effluent fractionation zone, cooling it to a temperature not higher than 120° F. to partially condense at least 40 per cent weight of said effluent, compressing the uncondensed portion to an intermediate pressure of at least 25 p. s. i. g., pumping the condensed portion to a pressure of at least 125 p. s. i. g.; withdrawing said reduced crude from said crude petroleum fractionation step, heating it to coking temperatures of between 770° F. and 1000° F., and coking it to produce coke and coker hydrocarbon vapors having a final boiling point between 600° F. and 1100° F.; flowing said coker hydrocarbon vapors to a vertically extended combination fractionation zone without lowering their temperature below 780° F. before introduction therein; fractionating said hydrocarbon vapors within said combination fractionating zone under a pressure of 15 to 350 p. s. i. g. to obtain several fractions, including a light gas oil fraction having a boiling range between 200° F. to 500° F. initial and 550° F. to 750° F. final; withdrawing said light gas oil fraction from said combination fractionating zone at a temperature of 500° F. to 800° F. and a pressure of 15 to 350 p. s. i. g., thermally cracking said thermal cracking feed fraction at final temperatures of between 900° F. and 1100° F.; introducing effluent from said thermal cracking into said combination fractionating zone; withdrawing from said combination fractionating zone a heavy gas oil having a boiling range between 450° F. and 650° F. initial and 700° F. and 1000° F. final, and recycling said heavy gas oil for further coking in said coking step; withdrawing a fraction composed primarily of gasoline from the upper end of said combination fractionating zone, cooling to a temperature less than 200° F. to partially condense said fraction into an unstabilized gasoline; combining the uncondensed portion of said fraction with said compressed uncondensed portion of said catalytic cracking effluent and compressing the mixture to a pressure of between 125 and 350 p. s. i. g.; combining said compressed mixture with said condensed catalytic cracking effluent at a pressure of between 125 and 350 p. s. i. g. and a temperature of between 80° F. and 200° F., to produce a high pressure rich gas and a distillate liquid composed primarily of unstabilized gasoline having a final boiling point of 350° F. to 450° F.; forming a stream of lean light gas oil at least partly by withdrawing a stream having a boiling range of between 300° F. to 500° F. initial and 600° F. to 750° F. final boiling point from at least one of said effluent fractionation zones, counter-currently contacting said high pressure rich gas with said stream of lean light gas oil to absorb heavier hydrocarbons from said rich gas and thereby produce a lean gas and a rich oil; and returning said rich oil to at least one of said effluent fractionation zones.

2. A continuous process for refining crude petroleum to produce a maximum of light straight run products and cracked gasoline, which includes the steps of: continuously introducing said crude petroleum into a crude petroleum distillation zone and fractionating said crude petroleum therein into a reduced crude fraction having an initial boiling point between 400° F. and 900° F., a heavy gas oil fraction having a boiling range of between 300° F. to 700° F. initial and 600° F. to 1100° F. final, and straight run fractions of lower boiling range; continuously withdrawing said heavy gas oil from said crude petroleum distillation zone at a pressure between 0 and 50 p. s. i. g. and a temperature between about 500° F. and and 800° F. and introducing said heavy gas oil into a fluidized mass of cracking catalyst without reducing the temperatures or pressure of said heavy gas oil below 450° F. and 10 p. s. i. a. respectively and catalytically cracking said oil; withdrawing an effluent from said catalytic cracking step at a temperature of between 850 and 1000° F. and a pressure of between 5 and 50 p. s. i. g., and introducing said effluent into a vertically extended catalytic cracking effluent fractionation zone operating at a temperature of between 90 to 200° F. at the top and 600° F. to 900° F. at the bottom, and a pressure of between zero and 50 p. s. i. g. without reduction in temperature or pressure of said effluent below 600° F. and zero p. s. i. g. respectively; withdrawing a light gaseous fraction having a boiling range ending at from 250° F. to 450° F. from the upper end of said catalytic cracking effluent fractionation zone, cooling to a temperature of less than about 200° F. to partially condense at least 40 per cent weight of said effluent, compressing the uncondensed portion to an intermediate pressure of at least 25 p. s. i. g., pumping the condensed portion to a pressure of at least 125 p. s. i. g.; withdrawing said rereduced crude from said crude petroleum fractionation step, heating it to coking temperatures of between 770° F. and 1000° F., and coking it to produce coke and coker hydrocarbon vapors having a final boiling point between 600° F. and 1100° F.; introducing said coker hydrocarbon vapors into a vertically extended combination fractionation zone without falling below a temperature of 780° F. and a pressure of 15 p. s. i. g.; fractionating said hydrocarbon vapors within said combination fractionating zone under a pressure of 15 to 250 p. s. i. g. to obtain several fractions, including a thermal cracking light gas oil feed fraction having a boiling range between 200° F. to 500° F. initial and 550° F. to 750° F. final; withdrawing said thermal cracking feed fraction from said combination fractionating zone at a temperature of 600° F. to 800° F. and a pressure of 20 to 250 p. s. i. g., thermally cracking said thermal cracking feed fraction at final temperatures of between 900° F. and 1100° F., quenching said effluent to a temperature of between 650° F. and 850° F. and introducing the liquefied effluent into said combination fractionating zone; withdrawing from said combination fractionating zone a heavy gas oil having a boiling range between 450° F. and 650° F. initial and 700° F. and 1000° F. final, and recycling said heavy gas oil for further coking in said coking step; withdrawing a fraction composed primarily of gasoline from the upper end of said combination fractionating zone, cooling to a temperature of 80° F. to 200° F. to partially condense said fraction into an unstabilized gasoline; combining the uncondensed portion of said fraction with said compressed uncondensed portion of said catalytic cracking effluent and compressing the mixture to a pressure of between 125 and 350 p. s. i. g.; combining said compressed mixture with said condensed catalytic cracking effluent at a pressure of between 125 and 350 p. s. i. g. and a temperature of between 80° F. and 200° F. to produce a high pressure rich gas and a high pressure distillate liquid composed primarily of unstabilized gasoline having a final boiling point of 250° F. to 450° F.; forming a stream of lean light gas oil by withdrawing a stream having a boiling range of between 300° F. to 500° F. initial and 600° F. to 750° F. final boiling point from at least one of said effluent fractionation zones, countercurrently contacting said high pressure rich gas with said stream of lean light gas oil to absorb heavier hydrocarbons from said rich gas and thereby produce a lean gas and a rich oil; and returning said rich oil to at least one of said effluent fractionation zones.

3. A continuous process for refining crude petroleum to produce a maximum of light straight run products and cracked gasoline, which includes the steps of: introducing said crude petroleum into a crude petroleum distillation zone and fractionating said crude petroleum therein into a reduced crude fraction having an initial boiling range starting between 400° F. and 900° F., a gas oil fraction having a boiling range of between 300° F. to 700° F. initial, and 600° F. to 1100° F. final, and straight

run fractions of lower boiling range; continuously withdrawing said gas oil from said crude petroleum distillation zone at a pressure between 0 and 50 p. s. i. g. and a temperature between 500° F. and 800° F. and contacting at least part of said heavy gas oil with finely divided solid catalyst at cracking temperatures and pressures without reducing the temperature of said heavy gas oil below 450° F., and catalytically cracking said oil; withdrawing an effluent from said catalytic cracking step at a temperature of between 850° F. and 1000° F., and introducing said effluent into a vertically extended catalytic cracking effluent fractionation zone operating at a temperature of between 90° F. and 200° F. at the top, and 600° F. to 900° F. at the bottom, and a pressure of between 0 and 50 p. s. i. g. without reduction in temperature or pressure of said effluent below 600° F. and 0 p. s. i. g., respectively, prior to said introduction; withdrawing a light gaseous fraction having a boiling range ending at from 250° F. to 450° F. from the upper end of said catalytic cracking effluent fractionation zone, cooling to partially condense said effluent, compressing the uncondensed portion and pumping the condensed portion to higher pressures; withdrawing said reduced crude from said crude petroleum fractionation step, heating it to temperatures of between 770° F. and 1000° F., and coking any desired portion thereof to produce coke and coker hydrocarbon vapors having a final boiling point between 600° F. and 1100° F., introducing the uncoked portion and coker to effluent hydrocarbon vapors into a vertically extended combination fractionation zone; fractionating hydrocarbon vapors within said combination fractionation zone under a pressure of 15 to 350 p. s. i. g. to obtain several fractions, including a light gas oil fraction having a boiling range between about 200° F. to 500° F. initial and about 550° F. to 750° F. final; withdrawing said light gas oil fraction from said combination fractionating zone, heating said light gas oil fraction to thermally crack it at final temperatures of between 900° F. and 1100° F.; introducing effluent from said thermal cracking into said combination fractionation zone; withdrawing from said combination fractionation zone a heavy gas oil having a boiling range between 450° and 650° F. initial and 700° F. and 1000° F. final, and recycling said heavy gas oil to at least one of said heating steps; withdrawing a fraction composed primarily of gasoline from the upper end of said combination fractionating zone, cooling it to partially condense said fraction; forming a stream of rich gas derived at least in part from said uncondensed portions taken from said catalytic cracking effluent fractionation zone and said combination fractionation zone; forming a stream of lean light gas oil at least partly by withdrawing a stream having a boiling range of between 300° F. to 500° F. initial and 600° F. to 750° F. final boiling point from at least one of said effluent fractionation zones; countercurrently contacting said high pressure rich gas with said stream of lean light gas oil to absorb heavier hydrocarbons from said rich gas and thereby produce a lean gas and a rich oil; and returning said rich oil to at least one of said effluent fractionation zones.

4. A continuous process for refining crude petroleum to produce a maximum of light straight run products and cracked gasoline, which includes the steps of: continuously introducing said crude petroleum into a crude petroleum distillation zone and fractionating said crude petroleum therein into a reduced crude fraction having an initial boiling range starting between 400° F. and 900° F., a gas oil fraction having a boiling range of between 300° F. to 700° F. initial, and 600° F. to 1100° F. final, and straight run fractions of lower boiling range; continuously withdrawing said gas oil from said crude petroleum distillation zone at a pressure between 0 and 50 p. s. i. g. and a temperature between 500° F. and 800° F. and contacting at least part of said heavy gas oil with finely divided solid catalyst at cracking temperatures and pressures without reducing the temperature of said heavy gas oil below 450° F., and catalytically cracking said oil; withdrawing an effluent from said catalytic cracking step at a temperature of between 850° F. and 1000° F., and introducing said effluent into a vertically extended catalytic cracking effluent fractionation zone operating at a temperature of between 90° F. and 200° F. at the top, and 600° F. to 900° F. at the bottom, and a pressure of between 0 and 50 p. s. i. g. without reduction in temperature or pressure of said effluent below 600° F. and 0 p. s. i. g. respectively, prior to said introduction; withdrawing a light gaseous fraction having a boiling range ending at from 250° F. to 450° F. from the upper end of said catalytic cracking effluent fractionation zone, cooling to partially condense said effluent, compressing the uncondensed portion and pumping the condensed portion to higher pressures; withdrawing said reduced crude from said crude petroleum fractionation step; heating it to temperatures in the coking and vis-breaking range; passing hydrocarbon fluids derived from said heated reduced crude to a second effluent fractionation zone; fractionating said hydrocarbon fluid within said second effluent fractionation zone to obtain several fractions; forming a thermal cracking gas oil feed fraction having a boiling range between 200° F. to 500° F. initial and 500° F. to 750° F. final, at least partly from a gas oil sidestream withdrawn from one of said two previously mentioned effluent fractionation zones; thermally cracking said thermal cracking feed fraction at final temperatures of between 900° F. and 1100° F., quenching said effluent, and introducing it into a thermal cracking effluent fractionation zone; withdrawing overhead vapor fractions from said second effluent fractionation zone and said thermal cracking effluent fractionation zone, partially condensing them, and combining at least part of the uncondensed vapors remaining, with said uncondensed vapors from said catalytic cracking effluent fractionation zone to make up at least part of a rich gas under pressure; forming a stream of lean light gas oil by withdrawing a stream having boiling range of between 300° F. to 500° F. initial and 600° F. to 750° F. final boiling point from at least one of said effluent fractionation zones, countercurrently contacting said high pressure rich gas with said stream of lean light gas oil to absorb heavy hydrocarbons from said rich gas and thereby produce a rich gas and a rich oil and returning said rich oil to at least one of said effluent fractionation zones.

5. A continuous process for refining crude petroleum to produce a maximum of light straight run products and cracked gasoline, which includes the steps of: continuously introducing said crude petroleum into a crude petroleum distillation zone and fractionating said crude petroleum therein into a reduced crude fraction having an initial boiling range starting between 400° F. and 900° F., a gas oil fraction having a boiling range of between 300° F. to 700° F. initial, and 600° F. to 1100° F. final, and straight run fractions of lower boiling range; continuously withdrawing said gas oil from said crude petroleum distillation zone at a pressure between 0 and 50 p. s. i. g. and a temperature between 500° F. and 800° F. and contacting at least part of said heavy gas oil with finely divided solid catalyst at cracking temperatures and pressures without reducing the temperature of said heavy gas oil below 450° F., and catalytically cracking said oil; withdrawing an effluent from said catalytic cracking step at a temperature of between 850° F. and 1000° F., and introducing said effluent into a vertically extended catalytic cracking effluent fractionation zone operating at a temperature of between 90° F. and 200° F. at the top, and 600° F. to 900° F. at the bottom, and a pressure of between 0 and 50 p. s. i. g. without reduction in temperature of pressure of said effluent below 600° F. and 0 p. s. i. g. respectively, prior to said introduction; withdrawing a light gaseous fraction having a boiling range ending at from 250° F. to 450° F. from the upper end of said catalytic cracking effluent fractionation zone,

cooling to partially condense said effluent, compressing the uncondensed portion and pumping the condensed portion to higher pressure; withdrawing said reduced crude from said crude petroleum fractionation step; heating it to temperatures of between 770° F. and 1000° F. and coking at least part to produce coke and coker hydrocarbon vapors having a final boiling point between 600° F. and 1100° F., introducing said coker hydrocarbon vapors into a coker effluent fractionation zone without reduction in temperature of said effluent below 780° F.; fractionating said coker effluent within said coker effluent fractionating zone to obtain several fractions; forming a thermal cracking gas oil feed fraction having a boiling range between 200° F. to 500° F. initial and 500° F. to 750° F. final, at least partly from a gas oil sidestream withdrawn from one of said two previously mentioned effluent fractionation zones; thermally cracking said thermal cracking feed fraction at final temperatures of between 900° F. and 1100° F., quenching said effluent, and introducing it into a thermal cracking effluent fractionation zone; withdrawing overhead vapor fractions from said coker effluent fractionation zone and said thermal effluent fractionation zone; partially condensing them, and combining at least part of the uncondensed vapors remaining with said uncondensed vapors from said catalytic cracking effluent fractionation zone to make up at least part of a rich gas under pressure; forming a stream of lean light gas oil from a gas oil sidestream withdrawn from said catalytic cracking effluent fractionation zone; countercurrently contacting said high pressure rich gas with said stream of lean light gas oil to absorb heavy hydrocarbons from said rich gas and thereby produce a lean gas and a rich oil; and returning at least part of said rich oil to said thermal cracking effluent fractionation zone.

6. A continuous process for refining crude petroleum to produce a maximum of light straight run products and cracked gasoline, which includes the steps of: continuously introducing said crude petroleum into a crude petroleum distillation zone and fractionating said crude petroleum therein into a reduced crude fraction having an initial boiling range starting between 400° F. and 900° F., a gas oil fraction having a boiling range of between 300° F. to 700° F. initial, and 600° F. to 1100° F. final, and straight run fractions of lower boiling range; continuously withdrawing said gas oil from said crude petroleum distillation zone at a pressure between 0 and 50 p. s. i. g. and a temperature between 500° F. and 800° F. and contacting at least part of said heavy gas oil with finely divided solid catalyst at cracking temperatures and pressures without reducing the temperature of said heavy gas oil below 450° F., and catalytically cracking said oil; withdrawing an effluent from said catalytic cracking step at a temperature of between 850° F. and 1000° F., and introducing said effluent into a vertically extended catalytic cracking effluent fractionation zone operating at a temperature of between 90° F. and 200° F. at the top, and 600° F. to 900° F. at the bottom, and a pressure of between 0 and 50 p. s. i. g. without reduction in temperature or pressure of said effluent below 600° F. and 0 p. s. i. g. respectively, prior to said introduction withdrawing a light gaseous fraction having a boiling range ending at from 250° F. to 450° F. from the upper end of said catalytic cracking effluent fractionation zone, cooling to partially condense said effluent, compressing the uncondensed portion and pumping the condensed portion to higher pressures; withdrawing said reduced crude from said crude petroleum fractionation step; heating it to temperatures of between 770° F. and 1000° F. and coking at least part to produce coke and coker hydrocarbon vapors having a final boiling point between 600° F. and 1100° F.; introducing said coker hydrocarbon vapors into a coker effluent fractionation zone without reduction in temperature of said effluent below 780° F.; fractionating said coker effluent within said coker effluent fractionating zone to obtain several fractions; forming a thermal cracking gas oil feed fraction having a boiling range between 200° F. to 500° F. initial and 500° F. to 750° F. final, at least partly from a gas oil sidestream withdrawn from said catalytic cracking effluent fractionation zone; thermally cracking said thermal cracking feed fraction at final temperatures of between 900° F. and 1100° F., quenching said effluent, and introducing it into a thermal cracking effluent fractionation zone; withdrawing overhead vapor fractions from said coker effluent fractionation zone and said thermal effluent fractionation zone, partially condensing them, and combining at least part of the uncondensed vapors remaining with said uncondensed vapors from said catalytic cracking effluent fractionation zone to make up at least part of a rich gas under pressure; forming a stream of lean light gas oil by withdrawing a stream having boiling range of between 300° F. to 500° F. initial and 600° F. to 750° F. final boiling point from at least one of said effluent fractionation zones, countercurrently contacting said high pressure rich gas with said stream of lean light gas oil to absorb heavy hydrocarbons from said rich gas and thereby produce a rich gas and a rich oil; and returning said rich oil to at least one of said effluent fractionation zones.

7. A continuous process for refining crude petroleum to produce a maximum of light straight run products and cracked gasoline, which includes the steps of: continuously introducing said crude petroleum into a crude petroleum distillation zone and fractionating said crude petroleum therein into a reduced crude fraction having an initial boiling range starting between 400° F. and 900° F., a gas oil fraction having a boiling range of between 300° F., to 700° F. initial, and 600° F. to 1100° F. final, and straight run fractions of lower boiling range; continuously withdrawing said gas oil from said crude petroleum distillation zone at a pressure between 0 and 50 p. s. i. g. and a temperature between 500° F. and 800° F. and contacting at least part of said heavy gas oil with finely divided solid catalyst at cracking temperatures and pressures without reducing the temperature of said heavy gas oil below 450° F., and catalytically cracking said oil; withdrawing an effluent from said catalyst cracking step at a temperature of between 850° F. and 1000° F., and introducting said effluent into a vertically extended cataytic cracking effluent fractionation zone operating at a temperature of between 90° F. and 200° F. at the top, and 600° F. to 900° F. at the bottom, and a pressure of between 0 and 50 p. s. i. g. without reduction in temperature or pressure of said effluent below 600° F. and 0 p. s. i. g. respectively, prior to said introduction; withdrawing a light gaseous fraction having a boiling range ending at from 250° F. to 450° F. from the upper end of said catalytic cracking effluent, compressing the uncondensed portion and pumping the condensed portion to higher pressures; withdrawing said reduced crude from said crude petroleum fractionation step: heating it to temperatures of between 770° F. and 1000° F. and coking at least part to produce coke and coker hydrocarbon vapors having a final boiling point between 600° F. and 1100° F.; introducing coker hydrocarbon vapors into a coker effluent fractionation zone without reduction in temperature of said effluent below 780° F.; fractionating said coker effluent within said coker effluent fractionating zone to obtain several fractions; forming a thermal cracking gas oil feed fraction having a boiling range between 200° F. to 500° F. initial and 500° F. to 750° F. final, at least partly from gas oil sidestreams withdrawn from said coker effluent fractionation zone and said catalytic cracking effluent fractionation zone; thermally cracking said thermal cracking feed fraction at final temperatures of between 900° F. and 1100° F., quenching said effluent, and introducing it into a thermal cracking effluent fractionation zone; withdrawing overhead vapor fractions from said coker effluent fractionation zone and said thermal effluent fractionation

zone, partially condensing them, and combining at least part of the uncondensed vapors remaining with said uncondensed vapors from said catayltic cracking effluent fractionation zone to make up at least part of a rich gas under pressure; forming a stream of lean light gas oil by withdrawing a stream having boiling range of between 300° F. to 500° F. initial and 600° F. to 750° F. final boiling point from at least one of said effluent fractionation zones, countercurrently contacting said high pressure rich gas with said stream of lean light gas oil to absorb heavy hydrocarbons from said rich gas and thereby produce a lean gas and a rich oil; and returning said rich oil to at least one of said effluent fractionation zones.

8. A continuous process for refining crude petroleum to produce a maximum of light straight run products and cracked gasoline, which includes the steps of: continuously introducing said crude petroleum into a crude petroleum distillation zone and fractionating said crude petroleum therein into a reduced crude fraction having an initial boiling range starting between 400° F. and 900° F., a gas oil fraction having a boiling range of between 300° F. to 700° F. initial, and 600° F. to 1100° F. final, and straight run fractions of lower boiling range; continuously withdrawing said gas oil from said crude petroleum distillation zone at a pressure between 0 and 50 p. s. i. g. and a temperature between 500° F. and 800° F. and contacting at least part of said heavy gas oil with finely divided solid catalyst at cracking temperatures and pressures without reducing the temperature of said heavy gas oil below 450° F., and catalytically cracking said oil; withdrawing an effluent from said catalytic cracking step at a temperature of between 850° F. and 1000° F., and introducing said effluent into a vertically extended catalytic cracking effluent fractionation zone operating at a temperature of between 90° F. and 200° F. at the top, and 600° F. to 900° F. at the bottom, and a pressure of between 0 and 50 p. s. i. g. without reduction in temperature or pressure of said effluent below 600° F. and 0 p. s. i. g. respectively, prior to said introduction; withdrawing a light gaseous fraction having a boiling range ending at from 250° F. to 450° F. from the upper end of said catalytic cracking effluent, compressing the uncondensed portion and pumping the condensed portion to higher pressures; withdrawing said reduced crude from said crude petroleum fractionation steps: heating it to temperatures of between 770° F. and 1000° F. and coking at least part to produce coke and coker hydrocarbon vapors having a final boiling point between 600° F. and 1100° F.; introducing coker hydrocarbon vapors into a coker effluent fractionation zone without reduction in temperature of said effluent below 780° F.; fractionating said coker effluent within said coker effluent fractionating zone under a pressure of less than 100 p. s. i. g. to obtain several fractions; forming a thermal cracking gas oil feed fraction having a boiling range between 200° F. to 500° F. initial and 500° F. to 750° F. final, at least partly from a gas oil sidestream withdrawn from one of said two previously mentioned effluent fractionation zones; thermally cracking said thermal cracking feed fraction at final temperatures of between 900° F. and 1100° F., quenching said effluent, and introducing it into a thermal cracking effluent fractionation zone maintained at a pressure higher than said coker effluent fractionation zone and in a range of about 50 to 350 p. s. i. g.; withdrawing overhead vapor fractions from said coker effluent fractionation zone and said thermal effluent fractionation zone, partially condensing them, and combining at least part of the uncondensed vapors remaining with said uncondensed vapors from said catalytic cracking effluent fractionation zone to make up at least part of a rich gas under pressure; forming a stream of lean light gas oil by withdrawing a stream having boiling range of between 300° F. to 500° F. initial and 600° F. to 750° F. final boiling point from at least one of said effluent fractionation zones, countercurrently contacting said high pressure rich gas with said stream of lean light gas oil to absorb heavy hydrocarbons from said rich gas and thereby produce a rich gas and a rich oil; and returning said rich oil to at least one of said effluent fractionation zones.

9. A continuous process for refining crude petroleum to produce a maximum of light straight run products and cracked gasoline, which includes the steps of: continuously introducing said crude petroleum into a crude petroleum distillation zone and fractionating said crude petroleum therein into a reduced crude fraction having an initial boiling range starting between 400° F. and 900° F., a gas oil fraction having a boiling range of between 300° F. to 700° F. initial, and 600° F. to 1100° F. final, and straight run fractions of lower boiling range; continuously withdrawing said gas oil from said crude petroleum distillation zone at a pressure between 0 and 50 p. s. i. g. and a temperature between 500° F. and 800° F. and contacting at least part of said heavy gas oil with finely divided solid catalyst at cracking temperatures and pressures without reducing the temperature of said heavy gas oil below 450° F., and catalytically cracking said oil; withdrawing an effluent from said catalytic cracking step at a temperature of between 850° F. and 1000° F., and introducing said effluent into a vertically extended catalytic cracking effluent fractionation zone operating at a temperature of between 90° F. and 200° F. at the top, and 600° F. to 900° F. at the bottom, and a pressure of between 0 and 50 p. s. i. g. without reduction in temperature or pressure of said effluent below 600° F. and 0 p. s. i. g. respectively, prior to said introduction; withdrawing a light gaseous fraction having a boiling range ending at from 250° F. to 450° F. from the upper end of said catalytic cracking effluent, compressing the uncondensed portion and pumping the condensed portion to higher pressures; withdrawing said reduced crude from said crude petroleum fractionation step; heating it to temperatures of between 770° F. and 1000° F. and coking at least part to produce coke and coker hydrocarbon vapors having a final boiling point between 600° F. and 1100° F.; introducing coker hydrocarbon vapors into a coker effluent fractionation zone without reduction in temperature of said effluent below 780° F.; fractionating said coker effluent within said coker effluent fractionating zone to obtain several fractions; forming a thermal cracking gas oil feed fraction at least partly from a gas oil sidestream withdrawn from one of said two previously mentioned effluent fractionation zones; introducing said thermal cracking feed fraction into a thermal cracking effluent fractionation zone; withdrawing a sidestream having a boiling range between about 200° F. to 500° F. initial and 500° F. to 750° F. final from said zone; thermally cracking said thermal cracking effluent fractionation zone sidestream at final temperatures of between 900° F. and 1100° F., quenching said effluent, and returning it to said thermal cracking effluent fractionation zone; withdrawing overhead vapor fractions from said coker effluent fractionation zone and said thermal effluent fractionation zone, partially condensing them, and combining at least part of the uncondensed vapors remaining with said uncondensed vapors from said catalytic cracking effluent fractionation zone to make up at least part of a rich gas under pressure; forming a stream of lean light gas oil by withdrawing a stream having boiling range of between 300° F. to 500° F. initial and 600° F. to 750° F. final boiling point from at least one of said effluent fractionation zones, countercurrently contacting said high pressure rich gas with said stream of lean light gas oil to absorb heavy hydrocarbons from said rich gas and thereby produce a rich gas and a rich oil; and returning said rich oil to at least one of said effluent fractionation zones.

10. A continuous process for refining crude petroleum to produce a maximum of light straight run products and cracked gasoline, which includes the steps of: continously

introducing said crude petroleum into a crude petroleum distillation zone and fractionating said crude petroleum therein into a reduced crude fraction having an initial boiling range starting between 400° F. and 900° F., a gas oil fraction having a boiling range of between 300° F. to 700° F. initial, and 600° F. to 1100° F. final, and straight run fractions of lower boiling range; continuously withdrawing said gas oil from said crude petroleum distillation zone at a pressure between 0 and 50 p. s. i. g. and a temperature between 500° F. and 800° F. and contacting at least part of said heavy gas oil with finely divided solid catalyst at cracking temperatures and pressures without reducing the temperature of said heavy gas oil below 450° F., and catalytically cracking said oil; withdrawing an effluent from said catalytic cracking step at a temperature of between 850° F. and 1000° F., and introducing said effluent into a vertically extended catalytic cracking effluent fractionation zone operating at a temperature of between 90° F. and 200° F. at the top, and 600° F. to 900° F. at the bottom, and a pressure of between 0 and 50 p. s. i. g. without reduction in temperature or pressure of said effluent below 600° F. and 0 p. s. i. g. respectively, prior to said introduction; withdrawing a light gaseous fraction having a boiling range ending at from 250° F. to 450° F. from the upper end of said catalytic cracking effluent fractionation zone, cooling to partially condense said effluent, compressing the uncondensed portion and pumping the condensed portion to higher pressures; withdrawing said reduced crude from said crude petroleum step; separating it into light and heavy vacuum fractions by vacuum distillation; diverting at least part of said light vacuum fractions to said catalytic cracking step; subjecting said heavy vacuum fraction to heating to temperatures in the coking and vis-breaking range; passing at least a fraction derived from said heated vacuum fraction to a vertically extended combination fractionation zone; fractionating vapors within said combination fractionating zone under a pressure of 15 to 350 p. s. i. g. to obtain several fractions, including a light gas oil fraction having a boiling range between 200° F. to 500° F. initial and 500° F. to 750° F. final; withdrawing said light gas oil fraction from said combination fractionating zone, thermally cracking said light gas oil fraction at final temperatures of between 900° F. and 1100° F.; introducing effluent from said thermal cracking into said combination fractionating zone; withdrawing from said combination fractionating zone a heavy gas oil having a boiling range between 450° F. and 650 F. initial and 700° F. and 1000° F. final, and recycling said heavy gas oil to at least one of said heating steps; withdrawing a fraction composed primarily of gasoline from the upper end of said combination fractionating zone, cooling it to partially condense said fraction; forming a stream of rich gas derived at least in part from said uncondensed portions taken from said catalytic cracking effluent fractionating zone and said combination fractionating zone; forming a stream of lean light gas oil at least partly by withdrawing a stream having a boiling range of between 300° F. to 500° F. initial and 600° F. to 750° F. final boiling point from at least one of said effluent fractionation zones; countercurrently contacting said high pressure rich gas with said stream of lean light gas oil to absorb heavier hydrocarbons from said rich gas and thereby produce a lean gas and a rich oil; and returning said rich oil to at least one of said effluent fractionation zones.

11. A continuous process for refining crude petroleum to produce a maximum of light straight run products and cracked gasoline, which includes the steps of: continuously introducing said crude petroleum into a crude petroleum distillation zone and fractionating said crude petroleum therein into a reduced crude fraction having an initial boiling range starting between 400° F. and 900° F., a gas oil fraction having a boiling range of between 300° F. to 700° F. initial, and 600° F. to 1100° F. final, and straight run fractions of lower boiling range; continuously withdrawing said gas oil from said crude petroleum distillation zone at a pressure between 0 and 50 p. s. i. g. and a temperature between 500° F. and 800° F. and contacting at least part of said heavy gas oil with finely divided solid catalyst at cracking temperatures and pressures without reducing the temperature of said heavy gas oil below 450° F., and catalytically cracking said oil; withdrawing an effluent from said catalytic cracking step at a temperature of between 850° F. and 1000° F., and introducing said effluent into a vertically extended catalytic cracking effluent fractionation zone operating at a temperature of between 90° F. and 200° F. at the top, and 600° F. to 900° F. at the bottom, and a pressure of between 0 and 50 p. s. i. g. without reduction in temperature or pressure of said effluent below 600° F. and 0 p. s. i. g. respectively, prior to said introduction; withdrawing a light gaseous fraction having a boiling range ending at from 250° F. to 450° F. from the upper end of said catalytic cracking effluent, compressing the uncondensed portion and pumping the condensed portion to higher pressures; withdrawing said reduced crude from said crude petroleum fractionation step; separating it into light and heavy vacuum fractions by vacuum distillation; diverting at least part of said light vacuum fractions to said catalytic cracking step; subjecting said heavy vacuum fraction to heating to temperatures in the coking and vis-breaking range; passing at least a fraction derived from said heated heavy vacuum fraction to a second effluent fraction zone; fractionating vapors within said second effluent fractionation zone to obtain several fractions; forming a thermal cracking gas oil feed fraction having a boiling range between 200° F. to 500° F. initial and 500° F. to 750° F. final, at least partly from a gas oil sidestream withdrawn from one of said two previously mentioned effluent fractionation zones; thermally cracking said thermal cracking feed fraction at final temperatures of between 900° F. and 1100° F., quenching said effluent, and introducing it into a thermal cracking effluent fractionation zone; withdrawing overhead vapor fractions from said second effluent fractionation zone and said thermal cracking fractionation zone, partially condensing them, and combining at least part of the uncondensed vapors remaining with said uncondensed vapors from said catalytic cracking effluent fractionation zone to make up at least part of a rich gas under pressure; forming a stream of lean light gas oil by withdrawing a stream having boiling range of between 300° F. to 500° F. initial and 600° F. to 750° F. final boiling point from at least one of said effluent fractionation zones, countercurrently contacting said high pressure rich gas with said stream of lean light gas oil to absorb heavy hydrocarbons from said rich gas and thereby produce a rich gas and a rich oil; and returning said rich oil to at least one of said effluent fractionation zones.

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