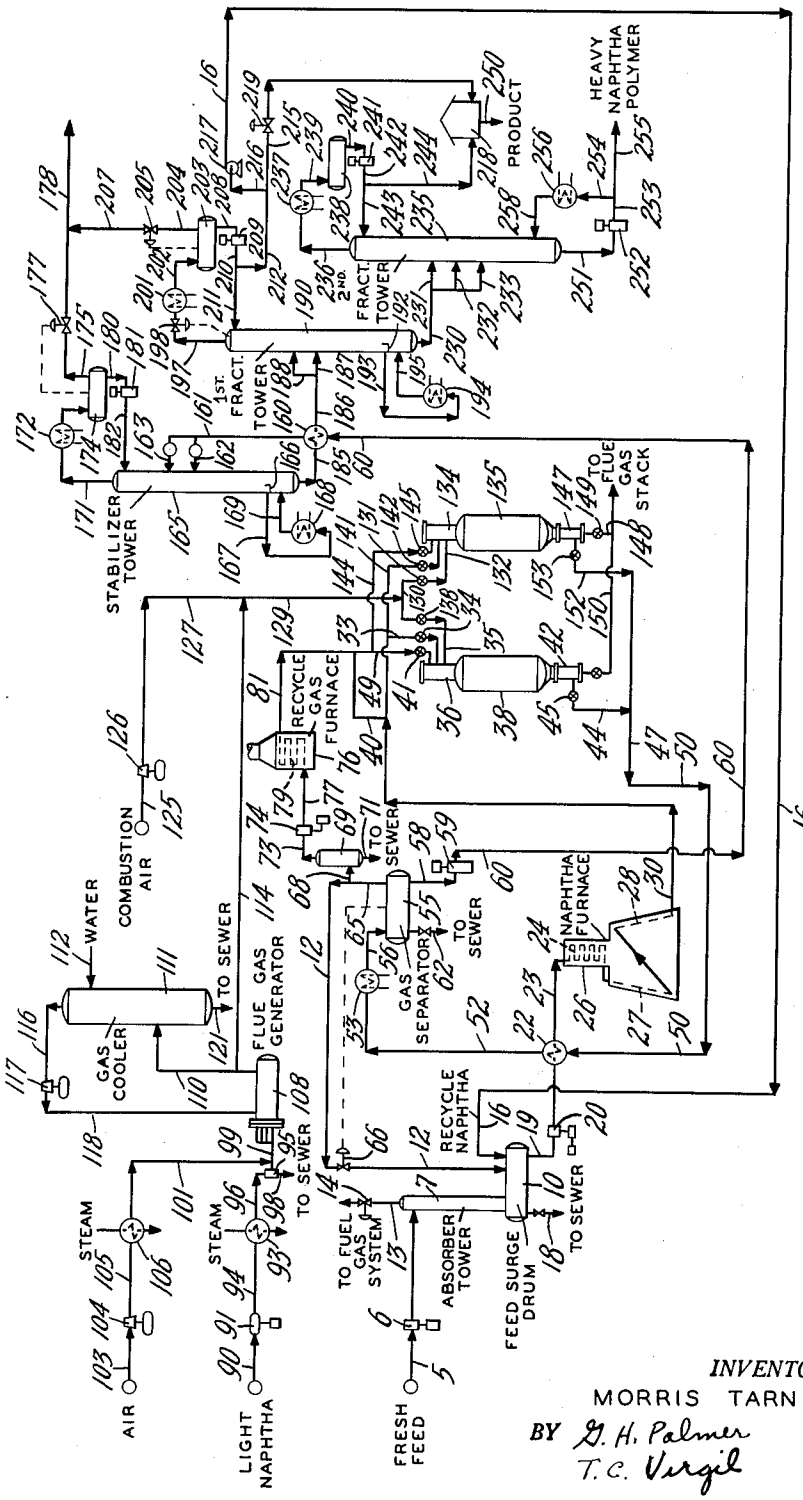
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REFORMING PROCESS

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REFORMING PROCESS

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This invention relates to an improved reforming process, and more particularly pertains to a hydroforming process for producing aviation gasoline of 100/130 grade, based on F–3 and F–4 ratings.

Usually a finished aviation gasoline is not produced from a hydroforming operation. The base stock which is necessary for the F–4 rating of the aviation gasoline is prepared by the hydroforming operation, and this stock is generally blended with other hydrocarbon materials in order to obtain a finished product which also has the desired F–3 rating. Normally, a once-through hydroforming operation produces also hydrocarbon material of 5 and 6 carbon atoms, but this material is not satisfactory for use in blending with the base stock to obtain the desired finished gasoline, because the F–3 rating thereof is not sufficiently high to effect this purpose.

The catalysts which are generally used in the hydroforming process possess isomerization properties, and so the lighter fractions produced in the operation can be repassed or recycled to improve the F–3 rating. The C_6 fraction may be processed to isomerize n-hexane, and improve the blending value thereof, and hence, effect an improvement in the quality of the finished gasoline. However, the quantity of n-hexane which remains unconverted after the recycle operation adversely influences the required specification of the gasoline.

To attain the desired finished gasoline from a hydroforming operation it is found advantageous in the present invention to recycle along with the C_6 fraction, hydrocarbon material of five carbon atoms. In the C_5 fraction there is present n-pentane, and this material is significantly higher in F–3 and F–4 ratings than n-hexane, and so, the amount which may not be converted to isopentane during the recycle step does not adversely influence the aviation gasoline specification as much as the n-hexane. Furthermore, the recycling of the C_5 fraction has the effect of lowering the concentration of n-hexane which might ultimately end in the finished gasoline, and hence render such material less effective in lowering the F–3 gasoline rating. The production of aviation gasoline having 100/130 rating from a hydroforming operation is quite unusual, and thus, this invention represents a major advance in the production of aviation gasoline.

By means of the present invention it is proposed preparing aviation gasoline by the method which comprises contacting naphtha with a reforming catalyst under suitable conditions in a reforming zone, separating the C_5–C_6 fraction which is present in the product thus produced, recycling part of the C_5–C_6 fraction to the reforming zone, separating an aviation base stock from the remainder of the product, and combining the base stock and the remainder of the C_5–C_6 fraction as the product of the process.

The naphtha which is processed by means of this invention is derived from any source, and it can be either straight run, cracked naphtha or mixtures of both. It is preferable, however, to employ a straight run naphtha, because the olefinic content is less and, therefore, better

aviation gasolines are produced. In general, the feed material contains a naphthene content of about 20 to about 60% by volume, and these naphthenes include those having 5 and 6 carbon atoms in the ring structure. Naphtha fractions containing large quantities of naphthenes are known to produce products which are exceptionally good as aviation base stocks. The percentage of naphthene compounds in the feed material is controlled by regulating the initial boiling point and the end point of the naphthene material. In this regard, it is customary to employ a feed material having an initial boiling point of about 80° to about 150° F. and an end point of about 280° to about 325° F. Such a material is suitable for the purposes of the present invention, however, in view that it is necessary to employ recycle stock comprised of light hydrocarbons having 5 and 6 carbon atoms, it is desirable to employ a feed material which contains these components. Accordingly, a preferred feed material is one which has an initial boiling point of about 90° to about 110° F. and an end point of about 290° to about 300° F. in order to have incorporated therein a significant number of hydrocarbons having about 5 to 6 carbon atoms in the molecule and little or none of those compounds having fewer carbon atoms. In this manner, the product produced from the reforming operation is certain to contain some hydrocarbons having 5 to 6 carbon atoms which can be used for blending with the aviation base stock, which is a highly concentrated fraction containing aromatic compounds.

This preferred feed material should be low in olefinic content, generally, about 0 to about 5 mol per cent, based on the fresh feed. The Watson characterization factor of the feed material can vary considerably, although those having a factor of about 11.0 to about 11.7 are preferred, because they usually contain unusually large quantities of naphthenic compounds. The sulfur content in the feed material may be important from the standpoint of affecting the yield and quality of the product material which is produced. For the production of aviation gasoline, it is usually desirable to employ a feed material having about 0.00 to about 0.20% by weight of sulfur, although stocks containing more sulfur than indicated can be used with less satisfactory results.

As is to be expected, hydrocarbons having 5 and 6 carbon atoms in the molecule are produced in the reforming operation by virtue of large molecules in the feed material being cracked into products of lower molecular weight. Under the conditions normally used for the reforming operation, there is an insignificant amount of cracking of hydrocarbon materials having 5 and 6 carbon atoms in the molecule, however, such compounds tend to undergo isomerization, dehydrogenation, etc. Any olefinic compounds which are produced by reason of cracking reactions are generally isomerized under conditions which usually occur in the reforming operation. Consequently, the yield of olefinic material is generally in the range of about 0 to about 5 mol per cent based on the product material. With the exception of changing the straight chain compounds having 5 and 6 carbon atoms to isomeric compounds and, in some cases, to aromatics through hydrocracking reactions, little of the material is converted to lower molecular weight hydrocarbons. Therefore, when employing a feed material having compounds of 5 and 6 carbon atoms, it is to be expected that the product will contain compounds of the same number of carbon atoms in greater quantity than in those operations in which the feed material was substantially free of such compounds.

The product derived from the reforming operation is subjected to a separation treatment in order to recover a fraction containing substantially all of the compounds having 5 and and 6 carbon atoms to the molecule. For

the purposes of this specification, such compounds are designated as the C_5–C_6 fraction. Recycling of the C_5–C_6 fraction is done in order to effect further isomerization of the straight chain hydrocarbons, particularly n-pentane and n-hexane. N-hexane has a low F–3 and F–4 rating; whereas its isomers are significantly better in this regard. In a once-through operation, the quantity of n-hexane which is found in the product is undesirably large and therefore it is difficult to obtain a satisfactory finished aviation gasoline. Hence, it is important to either lower the concentration of n-hexane in the product material, or to convert it to a more useful compound, i. e., isohexane. Recycling the C_5 fraction along with the C_6 fraction accomplishes a two-fold purpose, viz., (1) lowering the concentration of n-hexane in the final product, and (2) diluting the n-hexane with a more desirable material, i. e., n-pentane which can be converted to isopentane. Isopentane is especially desirable as an aviation gasoline component because the F–3 rating is higher than most of the isomerized C_6 hydrocarbons. Furthermore, n-pentane has a significantly higher F–3 rating than n-hexane. Therefore, the concentration of n-hexane is lowered by means of adding hydrocarbons which are higher in F–3 ratings, i. e., those having 5 carbon atoms, such as isopentane and n-pentane. Generally, in the reforming operation, the quantity of recycle employed is measured as the recycle ratio, which is defined as the ratio of the quantity of C_5–C_6 fraction which is recycled to the reforming zone to the quantity of fresh feed which is charged to the process, on a volumetric basis. Accordingly, the recycle ratio generally used is about 0.5 to about 3.0:1, preferably about 1.0 to about 2.0:1.

The aviation base stock is the portion of the liquid product which is highly concentrated in aromatic compounds containing at least seven carbon atoms. This base stock serves as the material which furnishes the required F–4 rating for the finished aviation gasoline. Usually, the base stock is separated from the total product as a material having an initial boiling point above that which would include a significant amount of the hydrocarbons containing six carbon atoms, for example, about 180° to about 225° F. The end point of the base stock is selected on the basis of eliminating substantially those aromatics having nine and more carbon atoms in the molecule. Such heavy aromatics have a boiling range which is too high according to specifications for satisfactory use as an aviation fuel. Normally, the C_9 aromatics have acceptable F–4 ratings and for this reason about 5–10% by weight can be included with the base stock. Consequently, the base stock has, in general, an end point of about 280° to about 325° F.

The catalyst employed for this process can be any suitable reforming catalyst which possesses the properties of dehydrogenation, hydrogenation, isomerization and cyclization. Such a type of catalyst is one which is normally used for the reforming operation and includes, for example, the oxide and/or sulfide of a group V or group VI metal of the periodic table as well as platinum and/or palladium either alone or supported on a carrier material. The carrier material used for this purpose is, for example, alumina in the form of a gel or the activated type, silica-alumina, silica, silica-magnesia, kieselguhr, pumice, fuller's earth, clays, etc. It is generally desirable to add a small amount of silica to the carrier material when employing alumina as a primary support in order to increase its heat stability. In this respect, about 0.1 to about 12%, preferably about 2 to about 8% of silica is employed for this purpose. Alumina is generally regarded as a favored support for the reforming reaction. The catalytic element or agent is employed in varying amounts for the reforming reaction. Generally, the catalytic agent is about 0.05 to about 25% by weight, based on the total catalyst. In the case of using a catalytic agent which is a sulfide and/or oxide of a metal of groups V and VI of the periodic table, it is preferred to employ about 1 to about 10% by weight based on the total catalyst and this agent. On the other hand, when platinum and palladium is employed as the catalytic agent, it is preferred to employ 0.1 to about 5% by weight thereof, based on the total catalyst. Specific examples of catalysts which are of use in the reforming operation are molybdena on alumina, platinum on alumina, chromia on alumina, tungsten oxide on alumina, etc. It is to be noted, however, that reforming operations generally employ either a molybdenum oxide on alumina catalyst, with or without being stabilized with silica, or platinum on alumina with or without silica being additionally present.

The reforming operation is generally conducted at an elevated temperature of about 850° to about 1050° F., preferably about 900° to about 975° F. At this elevated temperature, the pressure can be varied over a large range, namely, from about 25 to about 1000 p. s. i. g., although more usually the range is about 50 to about 500 p. s. i. g. The reforming operation is accomplished in the presence of hydrogen, either in the pure form or a diluted stream having light hydrocarbons present therein. The quantity of hydrogen added to the process will determine whether hydrogen is to be produced or consumed in the system. The higher hydrogen partial pressures tend to favor consumption of hydrogen, and therefore, in an operation involving a production of hydrogen, care should be taken to control the hydrogen partial pressure in a range where hydrogen is produced in quantities sufficient to sustain the process. The hydrogen partial pressure can be at least about 25 p. s. i. a. and it can be raised to the point where hydrogen is consumed in a hydroforming operation. More usually, in a reforming operation, the amount of hydrogen supplied is determined on the basis of the cubic feet of hydrogen (measured at 60° F. and 760 mm.) which are charged to the reforming zone per barrel of total oil feed. (The barrel being equal to 42 gallons.) Generally, for any type of reforming operation hydrogen can be supplied to the process at the rate of 500 to about 20,000 standard cubic feet per barrel. In a hydroforming operation, that is, where hydrogen is produced, it is preferred to employ a hydrogen rate of about 1500 to about 7500 SCFB. The hydrogen which is produced in the system is separated from the remainder of the reaction product, however, it usually does contain light hydrocarbons, that is, primarily those having about 1 to 3 carbon atoms in the molecule and some heavier compounds. This gas is recycled to the reforming zone, and it may contain at least 35% of hydrogen, although more usually it contains about 45 to about 80% by volume of hydrogen.

In a fixed and moving bed system, the severity of operation may be determined by the volumetric space velocity, which is measured as the volume of oil charged to the reforming zone, on an hourly basis, per volume of catalyst which is present therein. In the operation for producing aviation gasoline, the severity of the operation is generally high, and therefore, the volumetric space velocity is in the range of about 0.1 to about 4.5, more usually, about 0.2 to about 2.0, preferably about 0.3 to about 1.0. In a moving bed system, the severity of the operation may be determined by an additional factor, and that is, the catalyst to oil ratio, on a weight basis. In general, the catalyst to oil ratio for a moving bed system is about 0.05 to about 3.0:1, preferably about 0.2 to about 1.5:1, depending upon the catalyst used. The severity of the operation may be also measured by means of a weight space velocity factor which is measured as the weight of oil charged to the reaction zone, on an hourly basis, per weight of catalyst in the reaction zone. In general, the weight space velocity for a moving and fixed bed system generally runs about 0.10 to about 4.0, more usually, about 0.1 to about 2.0, preferably about 0.1 to about 1.0, depending upon the catalyst used.

For the purpose of this invention, the process can be operated as either a fixed or moving bed system. The fixed bed system can involve a fluid or non-fluid catalyst, and it can include at least two vessels, whereby one vessel is processing oil and the other vessel is undergoing regeneration in order to revivify the temporarily deactivated catalyst. In this manner, there is a continuous flow of processing materials, and hence, greater quantities of product are obtained. More usually, in commercial practice, four vessels are employed in the system in order to increase the capacity of operation. For a fixed bed system, the catalyst employed can be finely divided, granular, lump or pellets. In a moving bed system, the catalytic material can be granular or finely divided, however, the finely divided catalyst is utilized in order to operate by means of the fluid principle. In this regard, the finely divided catalyst has a particle size of about 0 to about 250 microns, more usually about 10 to about 100 microns. The processing materials are passed upwardly through a mass of finely divided catalyst at a superficial linear gas velocity of about 0.1 to about 50 feet per second, more usually about 0.5 to about 6 feet per second. For commercial operations, it is preferred to employ a superficial linear gas velocity of about 1 to about 2½ feet per second. At the latter velocities, a dense fluidized mass of catalyst is obtained which is optimum for contacting gas and solid particles. In a moving bed system, separate vessels are employed for the reaction and the regeneration, so that catalyst is being circulated from one processing zone to another in a continuous manner. The superficial linear gas velocities prevailing in all the processing zones fall within the range given above.

By reason of the reforming reaction, there is deposited on the catalyst carbonaceous material which causes temporary deactivation. In order to restore the catalyst activity, it is subjected to a treatment with an oxygen-containing gas, e. g., air, oxygen, diluted air containing about 1 to about 10% by volume of oxygen, etc. The carbonaceous material is removed from the catalyst through combustion at a temperature of about 600° to about 1250° F., more usually, about 950° to about 1150° F. Regenerated catalyst can be entirely free of carbonaceous material, or it can contain a small amount in the order of about 0.005 to about 0.1% by weight. In some instances, it is desirable to subject the catalyst to a stripping operation before regeneration. Stripping of the catalyst is effected by means of a stripping gas, such as nitrogen, steam, light hydrocarbons, hydrogen, recycle gas, etc. Stripping serves to remove from the catalyst any hydrocarbon material which is occluded and/or adsorbed thereby. Stripping is effected at essentially the same temperature which exists in the regeneration zone or it can be higher or lower than the regeneration temperature, depending upon the type of operation required.

In order to more fully understand the present invention, reference will be had to the accompanying drawing in which there is illustrated an operation by which aviation gasoline was produced having a 100/130 grade, based on F–3 and F–4 ratings, respectively.

In the figure, fresh naphtha feed is fed from a source **5** at a rate of 252 barrels per day (1 barrel is equal to 42 gallons), and by means of a pump **6** it is transported to the top section of an absorber tower **7**. The tower is maintained at a pressure of about 85 p. s. i. g. This tower serves for absorbing hydrocarbons from the gaseous product stream, which is introduced into the feed surge drum **10**, by means of a line **12**. The absorber tower **7** is superimposed on the feed surge drum, thus the surge drum serves as a receiver for the fresh feed flowing downwardly through the absorber tower, in countercurrent contact with the upflowing gaseous product stream, which is introduced through line **12**. About 10,600 cubic feet per hour of gaseous material (measured at 60° F. and 760 mm.) is discharged from the top of the absorber tower **7** through an overhead line **13**. The gas discharged from the top of the absorber tower is passed to the fuel gas system which is not shown in the drawing. There is installed in the overhead line **13**, a pressure control valve **14** for the purpose of maintaining the desired pressure in the absorber tower system. The fresh feed which is supplied through line **5** has the following characteristics:

° API gravity	64.2
IBP, ° F	108
5	154
10	158
20	179
30	192
40	203
50	212
60	221
70	234
80	244
90	263
E. P	299
Aromatics, vol. percent	16.0
Naphthenes, vol. percent	36.0

The fresh naphtha feed, laden with light hydrocarbon material, is mixed with a recycle naphtha, supplied at the rate of 365 barrels per day, through a line **16** leading to the top of the feed surge drum. This recycle naphtha has the following properties:

° API gravity	74.5
IBP, ° F	88
5	100
10	104
20	108
30	119
40	127
50	136
60	142
70	145
80	167
90	176
E. P	189

Any water which is present in the hydrocarbon material is withdrawn from the bottom of the surge drum **10** by means of a valved line **18**. The mixture of recycle naphtha and enriched fresh feed is withdrawn from the surge drum **10** through a bottom line **19**, and by means of the pump **20** installed therein, it is transported through a heat exchanger **22** at the rate of 617 barrels per day. Prior to entering the heat exchanger **22**, the temperature of the oil feed is 95° F. and as a result of being heated in the exchanger, the temperature is raised to 430° F. Thereafter, the total feed is passed through a line **23** which is connected to the convection coil **24** of a furnace **26**. After passing through coil **24**, the total feed is passed through coils **27** and **28** of the furnace in succession, and then, it leaves the furnace by means of line **30** at a temperature of 970° F.

The vaporized oil feed which leaves the furnace **26** passes first through line **30**, up to the junction with line **40** containing the heated recycle gas, following which the combined streams then enter line **33**, in which there is valve **34** in an open position, for entry of the combined feed material and recycle gas into the upper section **36** of the reaction vessel **38**. The reactor **38** is a cylindrical, vertical vessel in which there is supported pelleted molybdena-alumina catalyst containing 9% by weight of molybdena. This reactor vessel contains 7 tons of catalytic material for the reaction. The temperature in the reactor is such that the average is about 935° F., and the inlet pressure of the entering feed materials is about 277 p. s. i. g. The passage of reactant materials downwardly through the catalyst effects a pressure drop of about 7.1 p. s. i. g. Heated recycle gas containing about 69% by volume of hydrogen is supplied through a line **40** at the rate of about 79,400 cubic feet per hour. The

valve 41 in the recycle gas line 49 which branches from recycle gas line 40, is maintained in a closed position during this phase of the operation. The reaction product leaves the reactor vessel 38 and enters a section 42 connected to the bottom thereof wherein the temperature is about 897° F. The reaction product then passes into a line 44, containing valve 45 in the open position. This material then passes into a header 47. The mixture of recycle gas and reaction product passes into a line 50, and thence through heat exchanger 22, wherein the heat contained in the stream is indirectly transferred to the incoming feed material, which is fed to the furnace 26. The cooled recycle gas and reaction product having a resultant temperature of 310° F., passes into a line 52 before entering the cooler 53, wherein the temperature is further decreased, prior to passing to a gas separator 55 through a line 56. In the gas separator, the normal gaseous material under a pressure of 255 p. s. i. g. is separated from the liquid material. The liquid product is removed from the bottom of the gas separator through a line 58, and it is transported by means of pump 59 and line 60 to the product recovery system which will be discussed in more detail hereinafter. The temperature of the liquid product in the gas separator is 86° F. Any water which is present in the separator 55 is removed from the bottom thereof through a valved line 62.

The normally gaseous product material is removed from the top of the gas separator 55 through a line 65, and thence, it is divided so that a portion is passed through line 12 in which there is located a control valve 66 for the purpose of maintaining the pressure in the gas separator. The other portion of the normally gaseous product material flows through a line 68, which is connected with line 65, and thence flows into a knock-out drum 69, in which any entrained liquid material in the gas stream is removed from the bottom thereof through a line 71. The gaseous product material passes from the top of the knock-out drum by means of an overhead line 73, in which there is installed a compressor 74, which serves to compress the gaseous material to a pressure of about 300 p. s. i. g. and to transport it to recycle gas furnace 76 through a line 77. The temperature of the gaseous material at the inlet to the furnace 76 is 120° F. The gaseous material passes through coil 79 in the furnace, and then leaves the top thereof through a line 81. The outlet temperature of the gaseous material is 1100° F., which gaseous material is passed through the furnace at the rate of 84,700 cubic feet per hour.

The process is operated such that one vessel is being used to hydroform naphtha, while the other vessel simultaneously is undergoing regeneration to revivify the catalyst. In the description given above, vessel 38 is in the process cycle, consequently, the system is arranged to permit reactant materials to flow to this vessel, and prevent the flow of regeneration materials thereto. In view that an excessive quantity of heat is generated through the regeneration of the catalyst, it is necessary to employ a cooling means to remove this heat. In this illustration, the cooling medium employed is a flue gas which is prepared by burning a light naphtha (gaseous fuel can also be readily used), and then cooling the resultant flue gas prior to introducing the same into the vessel undergoing regeneration. In this regard, a light naphtha is supplied through a source 90, and then by means of a pump 91 it is passed into a steam heater 93 by means of line 94. The light naphtha is supplied at the rate of 35 barrels per day. In the heat exchanger 93, the steam indirectly vaporizes and heats the naphtha material to a temperature of 300° F. The heated naphtha vapor is then passed into a knock-out drum 95, by means of line 96, which leads from the heat exchanger 93. In the knock-out drum the separated entrained liquid material is removed therefrom through a bottom line 98; whereas the heated light naphtha vapor material passes into a line 99 to which there flows air at the rate of 75,000 cubic feet per hour, by means of a line 101. The air is supplied from a source 103, and then by means of a compressor 104 and line 105 it flows to a steam heater 106. In the steam heater, the temperature of the air is raised to 300° F., and thereafter it flows into line 101 previously mentioned. The combined streams of light naphtha vapor and air flow into a flue gas generator 108 wherein the light naphtha is burned to produce flue gas. The flue gas thus produced is discharged from the top of the flue gas generator by means of line 110, and then it passes into the bottom part of a gas cooler 111, wherein it flows upwardly in countercurrent contact with a downflowing stream of water, which is introduced at the top of the cooler tower by means of a line 112. A portion of the flue gas, which is produced in the generator 108, flows from line 110 into another line 114 which serves to transport this material to the vessel undergoing regeneration. The flue gas which is cooled in tower 111, flows from the top of the tower by means of an overhead line 116, and then it is transported by means of a compressor 117 and a line 118 to the flue gas generator 108 as recycle. The temperature of the recycled flue gas to the gas generator is 86° F. The water which is used in the cooling tower 111, for cooling the flue gas, is discharged from the bottom thereof by means of a valved line 121 and then discarded. The temperature of the flue gas used in the regeneration of the catalyst is controlled automatically by controlling the quantity of flue gas recirculated through the cooling tower 111.

The air used for the regeneration of catalyst is supplied from a source 125. This air is transported by means of a compressor 126 through a line 127, which connects to line 114 containing the flue gas, which serves as heat diluent. The combined streams of flue gas and air pass as a single stream through a conduit 129, and thence, it flows into a line 130 which divides in order to provide a flow of regeneration gas to either processing vessel depending upon the position of valves in the lines in question. In this example, line 130 contains a valve 131 which is maintained in the open position in order that the regeneration gas flows into the top section 134 of processing vessel 135. When processing vessel 38 is undergoing regeneration, valve 131 is closed, and the regeneration gas flowing through line 129 is passed through line 137 in which there is located a valve 138. Likewise, the combined feed material and recycle gas may be introduced into processing vessel 135 by means of line 141 in which there is installed a valve 142. Similarly, recycle gas for the purging or repressuring functions may be charged to processing vessel 135 through a line 144 which is connected to recycle gas line 49 and which contains a valve 145.

The average temperature of regeneration in vessel 135 is maintained at 915° F. The flue gas produced through the combustion of carbonaceous material on the catalyst first passes from processing vessel 135 into lower section 147 thereof. The flue gas then passes from the bottom of section 147 through a line 148 and a valve 149, and then into the vent line 150, which is connected to the flue gas stack (not shown). There is also connected to section 147 of processing vessel 135, a line 152 containing valve 153, for the purpose of transporting the reaction product from section 147 to the recovery system, when this processing vessel is being used on reaction cycle.

The liquid product which is discharged from the gas separator 55 flows through line 60 and then into a heat exchanger 160, wherein the temperature is raised from 86° F. to 207° F. The liquid product passes through line 60 at the rate of 125 barrels per day, and after being heated in exchanger 160, it passes through a line 161, before entering into either feed line 162 or 163 of the

stabilizer tower **165**. The top of the stabilizer tower is maintained at 145° F.; whereas the bottom temperature is 315° F. The pressure at the top of the stabilizer tower is approximately 178 p. s. i. g. The bottom of the tower is maintained at the desired temperature by withdrawing liquid from a trapout tray **166**, and passing the same through line **167** and into a heater **168**, before returning to the bottom of the tower through a line **169**. The overhead product passes from the top of the tower **165** through a line **171**, and then it passes through a condenser **172** prior to entering the accumulator **174**. Stabilizer gas is removed from the accumulator through an overhead line **175** at the rate of 1940 cubic feet per hour, and this line contains a control valve **177** by means of which the pressure in the stabilizer **165** is maintained. The stabilizer gas then passes into a line **178** which leads to a venting system, not shown. The liquid material in the accumulator **174** is withdrawn from the bottom thereof through line **180** and it is transported by means of a pump **181** and line **182** to the top of the stabilizer tower **165** as total reflux, at the rate of 440 barrels per day. The stabilized liquid product is withdrawn from the bottom of the stabilizer column **165** by means of a line **185**, and then it is passed through heat exchanger **160** wherein the heat contained by it is indirectly transferred to the incoming unstabilized liquid feed to the tower. From the heat exchanger **160**, the stabilized liquid product having a temperature of 165° F. passes at a rate of 529 barrels per day through line **186**, which later enters at either feed inlet **187** or **188** of the first fractionating tower **190**. In this tower, the temperature at the top is maintained at 216° F., and at the bottom, the temperature is 358° F. The pressure in this tower is maintained at 40 p. s. i. g. The bottom temperature is controlled by circulating liquid from the trapout tray **192** in the bottom of the tower, through a line **193** which leads to heater **194**, wherein the temperature is raised to the desired level, before it is returned to the tower by means of line **195**. The vapors at the top of the tower pass overhead through line **197**, in which there is located a control valve **198** for the purpose of maintaining the pressure in the system. The overhead vapors first pass into condenser **201**, then flow through line **202**, and finally pass into accumulator **203**. Normally gaseous materials are discharged from the accumulator by means of line **204**, in which there is situated valve **205** for controlling the pressure therein. This discharged gaseous material flows into line **207**, which leads to the venting system previously mentioned and not shown. The liquid product in accumulator **203** is withdrawn from the bottom thereof through a line **208**, and then it is pumped by means of pump **209** into line **210** before dividing into lines **211** and **212**. The liquid in line **211** is refluxed to the top of tower **190** at the rate of 220 barrels per day. The liquid in line **212** flows at the rate of 447 barrels per day, and this stream again divides so that a portion thereof, namely, 82 barrels per day pass through line **215** and the other portion, namely, 365 barrels per day pass through line **216**, whereby it is transported by means of pump **217** and line **16** and into feed surge drum **10**. The net production of light naphtha which flows through line **215** is then passed into a storage tank **218**. This line **215** contains a control valve **219** for the purpose of controlling the rate of liquid flow therethrough.

The liquid product is withdrawn from the bottom of tower **190** by means of line **230** at a temperature of 295° F. This liquid then passes into feed inlets **231**, **232** and **233** of the second fractionating tower **235**. The top temperature of the tower is maintained at 286° F.; whereas the bottom temperature is 380° F. The overhead vapors from tower **235** are first passed through a line **236**, then into a condenser **237**, and finally flow into an accumulator **238** by means of a line **239**. The pressure in the accumulator is maintained at 8.5 p. s. i. g., and the temperature of the liquid is 93° F. The liquid product is withdrawn from the bottom of accumulator **238** by means of a line **240**, and it is passed by means of a pump **241** into line **242** before dividing into two streams which flow through lines **243** and **244**. The liquid passing through line **243** is refluxed to the top of the tower at the rate of 1040 barrels per day. The net yield of liquid product passes through line **244** at the rate of 82 barrels per day, and it flows into storage tank **218** wherein it is mixed with the light naphtha, which was yielded from the overhead of the first fractionating tower **190**. The overhead product from tower **235** has the following properties:

°API gravity	45.5
IBP, °F	237
5	242
10	247
20	248
30	250
40	252
50	255
60	258
70	262
80	269
90	278
E. P.	293

The finished aviation gasoline product is withdrawn from the storage tank **218** by means of a bottom line **250**. A heavy naphtha polymer is withdrawn from the bottom of tower **235** through a line **251**, and then by means of pump **252** it is passed through a line **253** before dividing into lines **254** and **255**. The heavy polymer in line **254** is passed through a heater **256** wherein its temperature is raised to 216° F. and then it is returned to the bottom of the tower through a line **258**. The heavy polymer in line **255** represents the net yield of this material from the system, and it is passed to a fuel oil system not shown.

Having thus described my invention by reference to specific examples thereof, it should be understood that no undue limitations or restrictions are to be imposed by reason thereof, but that the scope of this invention is defined by the appended claims.

I claim:

1. A reforming process for producing a 100/130 grade aviation gasoline which comprises contacting a naphtha having about 20 to about 60% naphthenes with a catalyst selected from the group consisting of molybdenum oxide, chromium oxide, and platinum metal at a temperature of about 850 to 1050° F., a pressure of about 25 to about 1000 p. s. i. g., in the presence of hydrogen supplied at the rate of about 500 to about 20,000 SCFB, at a weight space velocity of about 0.10 to about 4.0, thus producing a reaction product including aromatics and acyclic hydrocarbons having 5 and 6 carbon atoms, separating a C_5–C_6 fraction from the reaction product, recycling a portion of the C_5–C_6 fraction at a recycle ratio of about 0.5 to about 3.0:1 for further contact with the catalyst, separating an aromatic fraction having a boiling point of about 180° to about 225° F. and an end point of about 280° to about 325° F. from the remainder of the reaction product, and combining the remainder of the C_5–C_6 fraction and the aromatic fraction as the product of the process.

2. A reforming process for producing a 100/130 grade aviation gasoline which comprises contacting a naphtha having about 20 to about 60% naphthenes with a molybdenum oxide catalyst, at a temperature of about 850° to about 1050° F., a pressure of about 25 to about 1000 p. s. i. g., in the presence of hydrogen supplied at the rate of about 500 to about 20,000 SCFB, at a weight space velocity of about 0.10 to about 4.0, thus producing a reaction product including aromatics and acyclic hydrocarbons having 5 and 6 carbon atoms, separating a C_5–C_6 fraction from the reaction product, recycling a portion of the C_5–C_6 fraction at a recycle ratio of about 0.5 to

about 3.0:1 for further contact with the molybdenum oxide catalyst, separating an aromatic fraction having an initial boiling point of about 180° to about 225° F. and an end point of about 280° to about 325° F. from the remainder of the reaction product, and combining the remainder of the C_5–C_6 fraction and the aromatic fraction as the product of the process.

3. A reforming process for producing a 100/130 grade aviation gasoline which comprises contacting a naphtha having about 20 to about 60% naphthenes with a molybdenum oxide catalyst in a reforming zone, at a temperature of about 900° to about 975° F., a pressure of about 50 to about 500 p. s. i. g., in the presence of hydrogen which is supplied to the reforming zone at the rate of about 1500 to about 7500 SCFB, at a weight space velocity of about 0.1 to about 1.0, thus producing a reaction product including aromatics and acyclic hydrocarbons having 5 and 6 carbon atoms, separating a C_5–C_6 fraction from the reaction product, recycling a portion of the C_5–C_6 fraction to the reforming zone at a recycle ratio of about 0.5 to about 3.0:1, separating an aromatic fraction having an initial boiling point of about 180° to 225° F. and an end point of about 280° to about 325° F. from the remainder of the reaction product, and combining the remainder of the C_5–C_6 fraction and the aromatic fraction as the product of the process.

4. A reforming process for producing a 100/130 grade aviation gasoline which comprises contacting a naphtha having about 20 to about 60% naphthenes with a fluidized mass of molybdenum oxide catalyst in the reforming zone,

at a temperature of about 900° to about 975° F., a pressure of about 50 to about 500 p. s. i. g., in the presence of hydrogen which is supplied to the reforming zone at the rate of about 1500 to about 7500 SCFB, at a weight space velocity of about 0.1 to about 1.0, thus producing a reaction product including aromatics and acyclic hydrocarbons having 5 and 6 carbon atoms, separating a C_5–C_6 fraction from the reaction product, recycling a portion of the C_5–C_6 fraction to the reforming zone at a recycle ratio of 0.5 to 3.0:1, separating an aromatic fraction having an initial boiling point of about 180° to about 225° F. and an end point of about 280° to about 325° F. from the remainder of the reaction product, and combining the remainder of the C_5–C_6 fraction and the aromatic fraction as the product of the process.

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