



PATENT SPECIFICATION

DRAWINGS ATTACHED

988.409

988.409



Date of Application and filing Complete Specification Dec. 19, 1962.

No. 47959/62.

Application made in United States of America (No. 170366) on Feb. 1, 1962.

Complete Specification Published April 7, 1965.

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Index at acceptance: —C5 EB2R

Int. Cl.: —C 10 g

COMPLETE SPECIFICATION

Improved Catalytic Reforming Process

5 We, ESSO RESEARCH AND ENGINEERING COMPANY, a Corporation organised and existing under the laws of the State of Delaware, United States of America, of Elizabeth, New Jersey, United States of America, do hereby declare the invention for which we pray that a patent may be granted to us, and the method by which it is to be performed, to be particularly described in and by the following statement:—

10 The present invention relates to an improved catalytic reforming process and, more particularly, to a process which permits the production of several grades of gasoline in a single operation catalytic reforming system.

15 Reforming or hydroforming of petroleum fractions in contact with catalysts particularly platinum-on-alumina catalysts is now well known and widely used in the petroleum industry for the production of high octane number gasolines. This process may be conducted continuously or non-regeneratively at temperatures of 750—1050° F. at relatively high pressures, i.e. above about 400 psig and correspondingly high hydrogen partial pressures. At lower pressures, higher yield octane relationships are achieved but with the formation of carbon deposits upon the catalyst which must be removed to maintain efficiency in the process. Regeneration may be carried out substantially continuously by separately regenerating the individual reactors in the system at frequent intervals as needed by the use of a "swing" reactor in a so-called cyclic process. Regeneration may also be carried out less frequently by shutting down the entire unit and regenerating all of the reactor vessels simultaneously in a so-called semiregenerative process.

40 In order to produce the different grades of gasoline marketed to satisfy the different octane requirements of various automobiles in use today resort is often made to so-called "blocked" operations in which a particu-

lar feed stock is hydroformed in a hydroformer unit under certain conditions to produce a relatively low octane product and the same, or a second, feedstock is hydroformed in the same unit under different conditions to produce a high octane number product or premium motor fuel. Blocked operations suffer from the disadvantage that extra tankage must be provided for the different feeds and products and it leads to additional operational requirements and problems. This type of operation has always been more expensive than a single-operation hydroformer.

It has now been found that several grades of gasoline can be prepared in a single operation hydroformer comprising several reactor vessels arranged in series with suitable means for reheating the reaction mixture prior to discharge thereof into the next reactor by withdrawing a portion of the total products after the first or one of the intermediate reactors. The withdrawn portion is condensed and separated from the normally gaseous components to provide a relatively low octane number gasoline which is suitable for use as regular grade gasoline or blend stock and an exceptionally high hydrogen content recycle gas. The reaction products remaining after such interreactor withdrawal, which may be only a relatively small percentage of the feed, is reheated with additional hydrogen gas and subjected to the conventional, more severe processing conditions in the remaining reactor or reactors of the series to prepare a high octane number or premium blend stock. The gaseous materials separated from the final high octane number product is of substantially lower hydrogen content than the gaseous material separated from the intermediate product and accordingly selective withdrawal of net gas make of lowest hydrogen content is achieved by withdrawal of said net gas make from the gaseous material separated from the final high octane number product.

[Price 4s. 6d.]



Another procedure for producing the two grades of gasoline with the reformer would be to bypass part of the feed around the reformer and blend this with part of the high octane reformat to produce the regular gasoline. Unblended reformat is then premium. As opposed to this alternate the process in accordance with the present invention will require a higher naphtha feed rate since feed by passing is not employed. However, total catalyst requirements will be considerably lower since a large portion of the feed is reformed to low severity which requires little catalyst, whereas the amount of feed to be reformed to high severity is sharply reduced. Low severity reforming is extremely selective to hydrogen and gasoline; therefore higher gasoline yields and higher hydrogen concentrations in the reforming system are achieved. This improvement in hydrogen concentration permits a reduction in operating pressure which further enhances gasoline yield. This increased feed rate results in an increase in the recycle gas compressor horsepower. Furthermore, the dual product withdrawal requires parallel heat exchange and a small incremental stabilizer. These added investments are overbalanced by the catalyst savings alone. However, additional investment savings are made by the elimination of blocked operation tankage normally required on both feed and products. Other advantages of the process are higher product credits for incremental salable products attained by improved yields.

The invention thus comprises a method for the production of at least two grades of motor gasoline in a single operation hydroformer which comprises charging a hydrocarbon feed stock boiling in the naphtha boiling range and hydrogen or a hydrogen containing gas to a hydroforming reactor system containing several reactor vessels in series, each of which is charged with an active platinum-alumina hydroforming catalyst, with means for reheating the reaction mixture between reactors, maintaining active hydroforming reaction conditions in each of the reactor vessels, maintaining the reactor pressures and temperatures within the range of 350—800 psig and 860—1050° F respectively, withdrawing a major proportion of the reaction mixture as discharged from at least one of the lead reactors of the series, condensing the withdrawn reaction mixture to recover an intermediate hydroformate of relatively low octane number and a high hydrogen content gas, passing the remainder of said reaction mixture through the remaining reactor vessels of the series and condensing reaction products discharged from the last reactor of the series to recover a high octane number hydroformate and a relatively low hydrogen-content gas.

The weight hourly space velocity (lbs. of naphtha feed per hour per pound of catalyst

in the reactor) is in the range of 4—12 in the first reaction zone, 2—6 in an intermediate reaction zone and 1—4 in the final reaction zone; this effects substantial catalyst savings.

In one form of the invention a major proportion of the hydrogen charged to the feed stream is passed to the first of the hydroforming reactors and the rest of the reaction mixture and the hydrogen is passed through the remaining reactor vessels of the series.

A portion of low hydrogen content gas may be selectively withdrawn as tail gas and remaining low hydrogen content gas combined with high hydrogen content gas to provide a high hydrogen content recycle gas and this gas mixture passed to the feed stream to the first of the hydroforming reactors.

The intermediate hydroformate and the high octane number hydroformate may be separately subjected to stabilization by distilling propane and lighter hydrocarbons therefrom, the overhead stream from both of these stabilizing steps combined, cooled to condense a portion of the vapors and the condensed liquid recycled as reflux to the respective stabilizing steps; uncondensed portions from the stabilizing steps may be passed to the bottom of the absorber-stripper, whereby excess butanes from the high severity reformat are transferred to the low severity reformat which may be deficient in butanes.

The hydrocarbon feed stock boiling in the naphtha boiling range and hydrogen treat gas may be charged to a reaction zone maintained under hydrofining reaction conditions to convert sulfur and nitrogen-containing impurities to hydrogen sulfide and ammonia respectively, excess treat gas hydrogen sulfide and ammonia from the hydrofined naphtha separated, the hydrofined naphtha passed through a reboiler absorber stripper counter-current to a gas containing light hydrocarbons to further free the hydrofined naphtha of gaseous impurities and undesired lighter hydrocarbons rejected while simultaneously absorbing the heavier hydrocarbons from the stripping gas and then passing the hydrofined naphtha from the absorber-stripper in admixture with a hydrogen-rich gas to the hydroforming reactor system.

Hydrocarbon feed stock boiling the naphtha boiling range may be fractionated into a light C_4 -160° F. fraction and heavy 160° F. to about 400° F. fraction, the latter heavy fraction and hydrogen charged to the hydroforming reactor system, and a portion of the light C_4 -160° F. fraction combined with low octane intermediate hydroformate to form a low octane product blend and with a portion of the light C_4 -160° F. fraction with the high octane number hydroformate to form a high octane product blend.

In the accompanying drawing, naphtha feed, which may be a virgin, a catalytically

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Stabilizing



cracked, or thermally cracked, or a Fischer Tropsch naphtha or a blend of two or more of these, is supplied via feed inlet line 10. The naphtha feed stock preferably will boil in the range of from butane to 430° F or it may be a narrower boiling cut from within this ~~boiled~~ range. Hydrogen or hydrogen rich gas is supplied through inlet line 11. The naphtha and the hydrogen-containing gas are heated separately or together in a preheater or furnace 12 and the preheated reactants are then supplied to a hydrofining reactor 14, if desired or necessary, in order to hydrogenate oxygen, sulphur and/or nitrogen compounds and saturate olefins. Ordinarily, it is preferred that the naphtha feed to the platinum hydroforming operation contain less than 40, preferably less than 10, ppm of sulfur and less than 10, preferably less than one, ppm of nitrogen. Hydrofining or treatment with excess hydrogen in contact with a suitable hydrogenation catalyst such as cobalt oxide-molybdenum oxide on alumina at temperatures of 500—700° F., pressures of 50 to 500 psig and feed rates of 1 to 10 V/V/Hr. can be used to remove these impurities to the above indicated levels.

The effluent from the hydrofining reactor is discharged via line 15, cooled at 16 and discharged into separator 17. The excess hydrogen treat gas, hydrogen sulfide and ammonia are taken overhead from separator 17 via line 18 and the hydrofined naphtha feed is withdrawn via line 19 and supplied to absorber-deethanizer column 20. The temperature and pressure conditions in column 20 are maintained at a suitable level to permit the ethane and lighter hydrocarbons to be vaporized for passage overhead through line 21 for use as fuel. Recovery of propane and butane is held at high levels by the absorption of these components in the hydrofined naphtha.

The liquid product withdrawn from the bottom of the column is heated in heater 22 and recycled to the bottom of column 20 in order to provide stripping of ethane and lighter components. Ordinarily the absorber-deethanizer column is operated at near hydrofiner pressure and with temperatures of 350—550° F. at the bottom and 100—160° F. at the top.

The deethanized liquid is withdrawn from the bottom of column 20, through line 23, and charged to debutanizer column 24. The lighter products, propane and butane, are taken overhead through line 25 condensed at 26 and a portion of the resultant condensate is pumped back to column 24 as reflux via line 28, while the remainder is passed via line 29 to C₃/C₄ splitter. The propane and butanes withdrawn via line 29 are treated for hydrogen sulfide and water removal if desired and necessary and fed to C₃/C₄ splitter column 30. Propane is taken overhead through line 31 condensed at 32 stored in vessel 33 and a portion of the condensate is pumped back to column 30 as reflux via line 34 while the remainder is passed via line 35 to C₃ LPG storage.

The liquid product withdrawn from the bottom of the column 30 is heated in heater 36 and recycled to the bottom of the column in order to provide the energy required for separation. The column may be operated at 180—300 psig with temperatures of 180—250° F. at the bottom and 100—150° F. at the top. The butane is withdrawn from the column bottoms via line 37 and a major portion is sent to C₄ LPG storage via line 38 and the remainder used for gasoline blending via line 39. By this processing scheme satisfactory quality C₃ and C₄ LPG can be produced from both the reformer and crude gases if desired.

At the debutanizer column 24, liquid from the bottoms is withdrawn and heated in heater 40 and recycled to the bottom of the column in order to provide the energy required for the separation. The bottoms product is a debutanized hydrogen-treated naphtha and is withdrawn via line 41 and fed, if desired, to naphtha splitter column 42. The lighter products are taken overhead through line 43 condensed at 44 and a portion of the resultant condensate is pumped back to column 42 as reflux via line 46 while the remainder is passed via line 47 to gasoline blending. The light fraction withdrawn via line 43 is desirably a C₅-160° F. boiling range cut which would be only slightly benefited by passage through the hydroforming section. If higher volatility gasolines are desired, or more C₄ and lighter material is desired or if excess reactor capacity is available, feed splitting may be omitted and the entire naphtha charged to the hydroforming section.

The heavier components, for example, a 160-340° F. boiling range hydroformer feed is withdrawn from splitter 42 heated in heater 48 and returned to the bottom of column 42 to provide the energy required for separation. Also, this heavier material is withdrawn via line 49, combined with hydrogen-rich recycle gas supplied through line 50 before or after preheater 51 and discharged into the lead reactor 53 of the hydroforming section through line 52.

Reforming or hydroforming of petroleum fractions in contact with platinum-on-alumina is now well known and widely used in the petroleum industry for the production of high octane number motor gasolines. When conducted at relatively high pressures and correspondingly high hydrogen partial pressures, this process is substantially continuous or nonregenerative. At lower pressures, higher yield-octane relationships are achieved but with carbon deposition upon the catalyst which must be removed to maintain efficiency

in the process. This, in turn, requires the provision of costly regeneration facilities. Regeneration may be carried out substantially continuously by separately regenerating the individual reactors at frequent intervals by the use of a "swing" reactor in a so-called cyclic process. Regeneration may also be carried out less frequently with the entire unit shut down and all reactors regenerated simultaneously in a so-called semiregenerative process. Basically, therefore, it becomes a question of economic balance between paying the extra cost for regeneration facilities and operation with resulting higher yields and the simpler, less costly operation afforded by a nonregenerative unit with its lower yield-octane relationship.

Employing a semiregenerative type of unit the operating conditions are pressures 350—800 psig, preferred 500 psig, with temperatures of 860—1050° F., preferred 900 at beginning of cycle and ending at 1000° F. at end of cycle with catalyst quantities expressed as weight hourly space velocities based on oil feed to each reactor (lb./Hr./lb. of catalyst) of 4—12 in first reactor, preferred 8, with 2—6 in the second reactor, preferred 4, and 1—4 in the third reactor, preferred 2. The catalyst quantities are of course a function of operational severity, catalyst activity, and activity maintenance. The temperature, pressure and space velocity are so adjusted in the several reactors that the low octane intermediate product separated after the first or second reactor of the series has a research O.N. (clear) of 60 to 70, preferably 70, and the high octane number hydroformate separated after the last reactor has an octane number of 90—100.

The reaction mixture is withdrawn from reactor 53 and a portion thereof, generally less than 50 per cent, is passed via line 54 to reheat furnace 55 into hydroforming reactor 56 for further hydroforming. The remainder is passed via line 57 to cooler or condenser 58 and thence to gas-liquid separator 59. In view of the fact that dehydrogenation is the principal reaction that occurs in the initial stages of the hydroforming operations, the gas produced in this section is particularly rich in hydrogen (about 85 to 98 mol percent). The gas separated at 59 is withdrawn via line 60, compressed in compressor 62 and recycled to the hydroforming section via line 50 to reactor 53 with a minor portion being passed via line 63 controlled by valve 72 to preheater 55 and reactor 56. This recycle gas is usually joined at the compressor suction with incremental recycle gas from separator 68 passing through line 70.

The amount of material withdrawn from reactor 53 via line 57 is controlled primarily by valve 61 with valve 69 as an aid in control. The liquid intermediate hydroformate product is withdrawn from separator 59 via

line 74 and passed to a stabilizer column 75. The stabilized low octane number intermediate hydroformate is passed through line 76 to product blending where it may be advantageously blended with the C₄-160° F. fraction withdrawn from the feed splitter 42 through line 47 and butane from line 39 and if desired a portion of the high octane reformate to produce a regular grade of motor gasoline. While the drawing shows withdrawal of low octane number intermediate hydroformate after only one reactor such withdrawal could be made after a second or later reactor in the hydroforming section which section can also, if desired, have more than the three reactors shown in the drawing. Also, if desired, intermediate hydroformate can be withdrawn simultaneously from more than one point in the hydroforming section, mixed together with other interreactor withdrawals or handled separately in the same way as the stream passed through line 57 as described above, particularly if a higher octane number regular gasoline is required or a number of different grades of gasoline or naphtha speciality products is to be produced.

As indicated above, a portion of the reaction products from the first portion of the hydroforming section is charged via line 54 to reheat furnace 55 and thence to hydroforming reactor 56. Also as indicated, additional recycle gas can be supplied via line 63 controlled by valve 72 to the reaction mixture passing through reactor 56. This operation is repeated as often as is necessary to hydroform the normally liquid hydrocarbons to the highest desired octane number. The reaction products are withdrawn from the final hydroforming reactor 65 via line 66, cooled in condenser 67 and passed into gas-liquid separator 68. Gaseous materials are taken overhead from separator 68 through valved line 69. The hydrogen content of this stream will be less than that in line 60. The net make gas or tail gas is removed from the hydroforming reactor system via line 71 and used as fuel or preferably as the treat gas supplied directly to the feed hydrofining step as through line 11. The remainder of the gas removed from separator 68 is passed through line 70 for addition to the high hydrogen content gas in line 60 for compression and recycling to the hydroforming reactors. In some cases where only a very minor portion of the feed is processed to the highest octane number all of the gas separated at 68 will be withdrawn as tail gas. In extreme cases part of the tail gas will have to come from a lower octane separator as the total gas separated at 68 will be less than the total net gas make on the unit.

The high octane number final hydroformate liquid is withdrawn from separator 68 through line 77. If desired, some of the C₄-

160° F. fraction removed from feed splitter 42 through line 47 may be blended with the high octane reformat before or after stabilization. If it is to be blended before some is passed through line 78 for blending with the high octane number hydroformate in line 77 whereupon the high octane number hydroformate or the blend is charged to stabilizer 79 for the removal of propane and lighter hydrocarbons and most, if not all, of the butane. The stabilized high octane number hydroformate is withdrawn via line 80 and passed to premium gasoline storage or blending. The light ends from stabilizer 79 and stabilizer 75 are combined and condensed at 81 and discharged into separator 82, the liquid material is pumped back to columns 75 and 79 as reflux and the gaseous product is withdrawn from the system via line 83 to absorber-demethanizer column 20 for recovery of propane and excess butane. By means of this common overhead equipment on the stabilizers it is possible to transfer butane from the high octane stream which has an excess to the lower octane stream which is frequently deficient in butane without having to add any investment to recover, separate, store, and blend the butane.

EXAMPLE

A 160/340° F. vapour temperature (VT) North African naphtha is reformed semiregeneratively to 70 and 97 RON Clear to produce components for a 98 RON (2.3 cc TEL) premium and a 88 RON (3 cc TEL) regular gasoline by blending with C₄/160° F. VT virgin naphtha. The ratio of premium to regular is 1:2-1/2.

In this application the unit is fed 1900 B/SD of naphtha with 1360 B/SD being reformed to 70 RON Clear and 540 B/SD reformed to 97 RON Clear.

The total 340° F. VT distillate from the

primary distillation unit is desulfurized in a hydrofiner containing cobalt molybdate catalyst. Hydrofining conditions are 350 psig, 550° F., 4.0 V/H/V space velocity with the total reformer net tail gas used on a once-through basis for hydrogen treat gas.

Following hydrotreating the naphtha is stabilized and split to provide a light virgin naphtha gasoline blend stock and the 160/340° F. VT reformer feed.

The total feed to the first reactor is preheated to 900–1000° F. and combined with 6000 SCF/B of recycle gas.

Following the first reactor the bulk of the reaction mixture is withdrawn and quenched to condense the 70 RON product. It is then stabilized and sent to gasoline blending. The gas separated from the 70 RON product is sent to the recycle gas compressor.

The remainder of the reaction mixture from the first reactor is joined by an incremental 1000 SCF/B (based on total feed to the first reactor—this amounts to 3500 SCF/B of high severity feed), preheated to 900–1000° F. and passed to the second reactor. Following the second reactor it is reheated to 900–1000° F. and passed through the third reactor. It is then quenched to condense the 97 RON product. This liquid stream is also stabilized and sent to gasoline blending.

The vapor from the 97 RON separator is split with an amount equivalent to the total net gas make being withdrawn to the naphtha hydrofiner and the remainder being combined with the 70 RON separator gas and sent to the recycle gas compressor. The recycle gas compressor discharge is split with the major portion being combined with the naphtha feed to the first reactor and the remainder combined with the portion of the first reactor product which is processed to 97 RON.

The feed naphtha inspections are:

Nominal Boiling Range - - - - -	160/340° FVT
API - - - - -	58
Watson K (Watson K is the cube root of the average boiling point in degrees Rankin divided by the specific gravity expressed as the ratio of the density to that of water at 60° F.) - - - - -	11.98
Composition, LV%	
Paraffins - - - - -	52.5
Naphthenes - - - - -	39.3
Aromatics - - - - -	8.2
ASTM Distillation	
10% - - - - -	220° F.
30% - - - - -	239
50% - - - - -	263
70% - - - - -	289
90% - - - - -	315



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The average cycle yields on a fresh feed basis are:

		70 RON	97 RON
5	C ₈ +LV% - -	93.0	73.7
	H ₂ Wt. % - -	0.75	1.2
	C ₁ Wt. % - -	0.4	1.9
	C ₂ Wt. % - -	0.4	3.8
	C ₃ Wt. % - -	1.4	6.6
	iC ₄ LV% - -	0.8	4.7
10	nC ₄ LV% - -	1.1	6.0
	iC ₅ LV% - -	1.5	4.2
	nC ₅ LV% - -	1.1	3.0

The catalyst is a 1/16" alumina extrudate impregnated with 0.3 wt. percent platinum. The space velocity (based on total feed) is 4.5 in the first reactor, 9.0 in the second, and 3.5 in the third reactor. The last reactor outlet pressure is 400 psig.

WHAT WE CLAIM IS:—

1. A method for the production of at least two grades of motor gasoline in a single operation hydroformer which comprises charging a hydrocarbon feed stock boiling in the naphtha boiling range and hydrogen to a hydroforming reactor system containing several reactor vessels in series, each of which is charged with an active platinum-alumina hydroforming catalyst, the weight hourly space velocity (lbs of naphtha feed per pound of catalyst in the reactor) being in the range of from 4—12 in the first reaction zone, from 2—6 in an intermediate reaction zone and from 1—4 in a final reaction zone, with means for reheating the reaction mixture between reactors, maintaining active hydroforming reaction conditions in each of the reactor vessels, maintaining the pressure in the range of from 350—800 p.s.i.g. and the temperature in the range of from 860—1050° F, withdrawing a major proportion of the reaction mixture as discharged from at least one of the lead reactors of the series, condensing the withdrawn reaction mixture to recover an intermediate hydroformate of relatively low octane number and a high hydrogen content gas, passing the remainder of said reaction mixture through the remaining reactor vessels of the series, condensing reaction products discharged from the last reactor of the series to recover a high octane number hydroformate and a relatively low hydrogen-content gas.
2. The method as claimed in claim 1 characterized by passing a major proportion of the hydrogen supplied to the feed stream to the first of the hydroforming reactors and passing the remainder of the said reaction mixture and the remainder of said hydrogen through the remaining reactor vessels of the series.
3. The method as claimed in claims 1 or claim 2 further characterized by selectively

withdrawing a portion of said low hydrogen content gas as tail gas and combining remaining low hydrogen content gas with said high hydrogen gas to provide a high hydrogen content recycle gas and passing said gas mixture to the feed stream to the first of the hydroforming reactors.

4. The method as claimed in any of claims 1—3 characterized by charging said hydrocarbon feed stock boiling in the naphtha boiling range and hydrogen treat gas to a reaction zone maintained under hydrofining reaction conditions to convert sulfur and nitrogen-containing impurities to hydrogen sulfide and ammonia respectively, separating excess treat gas, hydrogen sulfide and ammonia from the hydrofined naphtha, passing the hydrofined naphtha through a reboiler absorber stripper countercurrent to a gas containing light hydrocarbons to further free the hydrofined naphtha of gaseous impurities and reject undesired lighter hydrocarbons while simultaneously absorbing the heavier hydrocarbons from the stripping gas and then passing the hydrofined naphtha from the absorber-stripper in admixture with a hydrogen-rich gas to said hydroforming reactor system.

5. The method as claimed in claim 4 in which the intermediate hydroformate and the high octane number hydroformate are separately subjected to stabilization by distilling propane and lighter hydrocarbons therefrom, combining the overhead stream from both of these stabilizing steps, cooling to condense a portion of the vapors, recycling condensed liquid as reflux to the respective stabilizing steps and passing the uncondensed portions from said stabilizing steps to the bottom of the absorber-stripper, whereby excess butanes from the high severity reformat are transferred to the low severity reformat which is deficient in butanes.

6. The method as claimed in any of claims 1—5 which comprises fractionating a hydrocarbon feed stock boiling in the naphtha boiling range into a light C₄-160° F. fraction and a heavy 160° F. to about 400° F. fraction, charging the latter heavy fraction and hydrogen to said hydroforming reactor system, and combining a portion of said light C₄-160° F. fraction with said low octane intermediate hydroformate to form a low octane product blend and a portion of said light C₄-160° F. fraction with said high octane number hydroformate to form a high octane product blend.

7. The method as claimed in claim 1 and substantially as described herein.

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The average cycle yields on a fresh feed basis are:

		70 RON	97 RON
	C ₅ +LV% - -	93.0	73.7
5	H ₂ Wt. % - -	0.75	1.2
	C ₁ Wt. % - -	0.4	1.9
	C ₂ Wt. % - -	0.4	3.8
	C ₃ Wt. % - -	1.4	6.6
	iC ₄ LV% - -	0.8	4.7
10	nC ₄ LV% - -	1.1	6.0
	iC ₅ LV% - -	1.5	4.2
	nC ₅ LV% - -	1.1	3.0

The catalyst is a 1/16" alumina extrudate impregnated with 0.3 wt. percent platinum.

- 15 The space velocity (based on total feed) is 4.5 in the first reactor, 9.0 in the second, and 3.5 in the third reactor. The last reactor outlet pressure is 400 psig.

WHAT WE CLAIM IS:—

- 20 1. A method for the production of at least two grades of motor gasoline in a single operation hydroformer which comprises charging a hydrocarbon feed stock boiling in the naphtha boiling range and hydrogen to a hydroforming reactor system containing several reactor vessels in series, each of which is charged with an active platinum-alumina hydroforming catalyst, the weight hourly space velocity (lbs of naphtha feed per pound of catalyst in the reactor) being in the range of from 4—12 in the first reaction zone, from 2—6 in an intermediate reaction zone and from 1—4 in a final reaction zone, with means for reheating the reaction mixture between reactors, maintaining active hydroforming reaction conditions in each of the reactor vessels, maintaining the pressure in the range of from 350—800 p.s.i.g. and the temperature in the range of from 860—1050° F, withdrawing a major proportion of the reaction mixture as discharged from at least one of the lead reactors of the series, condensing the withdrawn reaction mixture to recover an intermediate hydroformate of relatively low octane number and a high hydrogen content gas, passing the remainder of said reaction mixture through the remaining reactor vessels of the series, condensing reaction products discharged from the last reactor of the series to recover a high octane number hydroformate and a relatively low hydrogen-content gas.

2. The method as claimed in claim 1 characterized by passing a major proportion of the hydrogen supplied to the feed stream to the first of the hydroforming reactors and passing the remainder of the said reaction mixture and the remainder of said hydrogen through the remaining reactor vessels of the series.

3. The method as claimed in claims 1 or claim 2 further characterized by selectively

withdrawing a portion of said low hydrogen content gas as tail gas and combining remaining low hydrogen content gas with said high hydrogen gas to provide a high hydrogen content recycle gas and passing said gas mixture to the feed stream to the first of the hydroforming reactors.

4. The method as claimed in any of claims 1—3 characterized by charging said hydrocarbon feed stock boiling in the naphtha boiling range and hydrogen treat gas to a reaction zone maintained under hydrofining reaction conditions to convert sulfur and nitrogen-containing impurities to hydrogen sulfide and ammonia respectively, separating excess treat gas, hydrogen sulfide and ammonia from the hydrofined naphtha, passing the hydrofined naphtha through a reboiler absorber stripper countercurrent to a gas containing light hydrocarbons to further free the hydrofined naphtha of gaseous impurities and reject undesired lighter hydrocarbons while simultaneously absorbing the heavier hydrocarbons from the stripping gas and then passing the hydrofined naphtha from the absorber-stripper in admixture with a hydrogen-rich gas to said hydroforming reactor system.

5. The method as claimed in claim 4 in which the intermediate hydroformate and the high octane number hydroformate are separately subjected to stabilization by distilling propane and lighter hydrocarbons therefrom, combining the overhead stream from both of these stabilising steps, cooling to condense a portion of the vapors, recycling condensed liquid as reflux to the respective stabilizing steps and passing the uncondensed portions from said stabilizing steps to the bottom of the absorber-stripper, whereby excess butanes from the high severity reformat are transferred to the low severity reformat which is deficient in butanes.

6. The method as claimed in any of claims 1—5 which comprises fractionating a hydrocarbon feed stock boiling in the naphtha boiling range into a light C₄-160° F. fraction and a heavy 160° F. to about 400° F. fraction, charging the latter heavy fraction and hydrogen to said hydroforming reactor system, and combining a portion of said light C₄-160° F. fraction with said low octane intermediate hydroformate to form a low octane product blend and a portion of said light C₄-160° F. fraction with said high octane number hydroformate to form a high octane product blend.

7. The method as claimed in claim 1 and substantially as described herein.

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