

INTEGRATED PROCESS FOR THE PRODUCTION OF
REFORMATE HAVING REDUCED BENZENE CONTENT

This invention relates to a process for the production of a more environmentally suitable gasoline by eliminating
5 a substantial portion of benzene in gasoline by pretreatment of raw naphtha. The process also provides a means to reduce endpoint (T_{90}) and increase driveability index in reformat-containing gasolines.

Gasolines manufactured to contain a higher concentra-
10 tion of aromatics such as benzene, toluene and xylenes (BTX) can adequately meet the octane requirements of the marketplace for a high octane fuel. Aromatics, particularly benzene, are commonly produced in refinery processes such as catalytic reforming which have been a
15 part of the conventional refinery complex for many years.

Recent concerns about volatility and toxicity of hydrocarbon fuel and the resultant environment damage has prompted legislation that limits the content and composition of aromatic hydrocarbons in such fuels. Some
20 of these limitations relate specifically to benzene which, due to its toxicity, will be substantially eliminated from the gasoline pool. Restrictions on the content of heavy aromatics will result from proposed end boiling point limits of gasoline fuels, referred to as T_{90} or (90 vol%
25 temperature). T_{90} limits curtail the presence of hydrocarbon components that boil above temperatures in a range of 177 to 221°C (350 to 430°F).

When hydrocarbons boiling in the gasoline boiling range are reformed in the presence of a hydrogenation-
30 dehydrogenation catalyst, a number of reactions take place which include dehydrogenation of naphthenes to form aromatics, dehydrocyclization of paraffins to form aromatics, isomerization reactions and hydrocracking reactions. The composition of the reformer effluent or
35 reformat is shifted toward higher octane value product. Catalytic reforming primarily increases the octane of motor

gasoline by aromatics formation but without increasing the yield of gasoline.

Reformates can be prepared by conventional techniques by contacting any suitable material such as a naphtha
5 charge material boiling in the range of C5 or C₆ up to 193°C (380°F) with hydrogen in contact with any conventional reforming catalyst.

Benzene content and T₉₀ are two important specifications for reformulated gasoline (RFG). It is
10 known that reformat is the major benzene source in the gasoline pool. One option to reduce benzene content in the reformat is to remove the C6 cut, which is rich in benzene and its precursors, e.g. methylcyclopentanes and hexanes, from a pretreated naphtha using a dehexanizer, in order to
15 divert these materials from the reformer. The C6 cut can be eventually combined with reformat in the gasoline pool. In order to meet the benzene specification in the gasoline pool, refineries need to convert benzene in this C6 cut, which can contain as much as 20 to 30% benzene.

It would be advantageous to provide a way of treating
20 naphtha reformer feedstock to produce gasoline of reduced benzene content and T₉₀ specifications wherein a C6 cut is diverted from the reformer and combined with reformat, without requiring further treatment of the C6 cut to meet
25 benzene specifications.

U.S. Pat. No. 4,927,521 to Chu discloses a process for pretreating naphtha prior to reforming, by contacting with
a zeolite catalyst, e.g., zeolite beta, containing at least one noble metal and at least one alkali metal, for the
30 purpose of producing higher yields of C4+ and C5+ gasolines.

The present invention relates to an integrated process for providing a gasoline boiling range reformat-containing product produced from naphtha. The process comprises:

35 1) pretreating a raw naphtha feedstream containing benzene by contacting with a hydrodesulfurization catalyst

thereby forming H₂S and thereafter cascading said feedstream over a noble metal-containing porous crystalline inorganic oxide catalyst comprising pores having openings of 12-member rings under conditions sufficient to effect benzene saturation, thereby providing a pretreated effluent of reduced benzene content;

2) dehexanizing the pretreated effluent to provide a C₆⁻ hydrocarbon-containing overhead stream and a bottoms stream containing C₇⁺ hydrocarbons;

3) reforming said bottoms stream to form a reformat stream; and

4) combining said reformat stream and said C₆⁻ hydrocarbon-containing overhead stream to provide a combined gasoline boiling range product.

The present invention can be described more particularly as an integrated process for providing a gasoline boiling range reformat-containing product produced from naphtha comprising:

1) pretreating a raw naphtha feedstream by contacting with a hydrodesulfurization catalyst and thereafter cascading said feedstream over a noble metal-containing zeolite catalyst comprising pores having openings of 12-member rings thereby providing a pretreated effluent;

2) stripping off C₄⁻ hydrocarbons, hydrogen sulfide and ammonia from the pretreated effluent thereby providing a stripped effluent;

3) dehexanizing the stripped effluent to provide a C₅ to C₆ hydrocarbon-containing overhead stream and a bottoms stream containing C₇⁺ hydrocarbons;

4) reforming said bottoms stream to form a reformat stream;

5) distilling said reformat stream to provide a C₁ to C₄ hydrocarbon-containing overhead and C₅⁺ reformat bottoms fraction; and

6) combining said C5+ reformat bottoms fraction and said C5 to C6 hydrocarbon-containing overhead stream to provide a combined gasoline boiling range product.

The present invention relates to a process wherein a raw naphtha feed is pretreated to saturate benzene and to convert back-end materials (C9+) into lighter naphtha in an existing naphtha pretreater used for hydrodesulfurization. The process employs a noble metal promoted porous inorganic oxide catalyst downstream of the hydrodesulfurization catalyst. The noble metal-containing catalyst saturates benzene and reduces end-point in the presence of hydrogen sulfide produced during hydrodesulfurization. This approach eliminates or reduces the need for a secondary reactor for benzene conversion of C6 cut from dehexanizer overhead.

Inasmuch as noble metal promoted catalysts are generally sensitive to hydrogen sulfide poisoning which strongly inhibits hydrogenation activity of the noble metal, the ability of the noble metal-containing catalyst to retain its hydrogenation activity while contacting the H₂S-containing effluent from the dehydrodesulfurization step is unexpected.

The Figure is a process flow diagram depicting a preferred multistage embodiment of the present invention wherein raw naphtha is pretreated in two stages prior to stripping, removal of a C5 and C6 cut, reforming of the C7+ dehexanizer bottoms and combining the unreformed C5 and C6 cut with C5+ reformat.

Feed

The raw naphtha feedstream can comprise a mixture of aromatic and paraffin hydrocarbons having boiling points about 1.5 to 5.0 or higher mole percent benzene. It can also contain various C7 to C10 aromatic hydrocarbons including toluene and C8 to C10 aromatic hydrocarbons. The feedstream can also contain C4 to C6 paraffinic hydrocarbons including butane, isopentane, isohexane and

n-hexane which are normally present at a concentration above 5.0 mole percent. C7 to C9 paraffinic hydrocarbons such as isoheptane and isooctane can also be present. The exact composition of the raw naphtha feedstream will depend on its source. It may be formed by blending all or a portion of the effluent of several different petroleum processing units. Two such effluents are the bottoms product of the stripper column used in FCC gas concentration units and stabilized reformates which contain C6 to C9 aromatic hydrocarbons.

The raw naphtha contains sulfur. Products of catalytic cracking usually contain sulfur impurities which normally require removal, usually by hydrotreating, in order to comply with the relevant product specifications. These specifications are expected to become more stringent in the future, possibly permitting no more than 300 ppmw sulfur in motor gasolines. As a practical matter, the sulfur content will exceed 50 ppmw and usually will be in excess of 100 ppmw and in most cases in excess of 500 ppmw. For the fractions which have 95 percent points over 193°C (380°F), the sulfur content may exceed 1,000 ppmw and may be as high as 4,000 or 5,000 ppmw or even higher, as shown below. The nitrogen content is not as characteristic of the feed as the sulfur content and is preferably not greater than 20 ppmw although higher nitrogen levels typically up to 50 ppmw may be found in certain higher boiling feeds with 95 percent points in excess of 193°C (380°F). The nitrogen level will, however, usually not be greater than 250 or 300 ppmw.

The raw naphtha feed to the process comprises a sulfur- and benzene-containing petroleum fraction which boils in the gasoline boiling range. Feeds of this type include light naphthas typically having a boiling range of C₆ to 166°C (330°F), full range naphthas typically having a boiling range of C₅ to 216°C (420°F), heavier naphtha fractions boiling in the range of 127°C to 211°C (260°F to

412°F), or heavy gasoline fractions boiling at, or at least within, the range of 166°C to 260°C (330°F to 500°F), preferably 166°C to 211°C (330°F to 412°F).

5 The raw naphtha may be obtained from straight run distillation or from a coker or FCC unit. Alternatively, pyrolysis gasoline may be used as well. However, diene-containing streams should be treated to reduce or remove sources of gumming, as necessary.

Process Configuration

10 Referring to the Figure, the raw naphtha feedstream 1 containing sulfur compounds, nitrogen compounds and benzene is passed to a pretreater 2 where it is treated in a first pretreating (hydrodesulfurization) zone 3 by hydrotreating the feed by effective contact of the feed with a
15 hydrotreating catalyst, which is suitably a conventional hydrotreating catalyst, such as a combination of a Group VI and a Group VIII metal on a suitable refractory support such as alumina, e.g., Co/Mo on alumina, under hydrotreating conditions, i.e., at elevated temperature and
20 somewhat elevated pressure in the presence of a hydrogen atmosphere. More specifically, such conditions include temperatures ranging from 204°C to 399°C (400°F to 750°F), 1-10 LHSV, 790 kPa to 5500 kPa (100-800 psig) total pressure, and a hydrogen circulation rate of 89 to 890 nl/l
25 (500-5000 scf/b).

Under these conditions, at least some of the sulfur is separated from the feed molecules and converted to hydrogen sulfide, to produce a hydrotreated product and hydrogen sulfide. One suitable family of catalysts which has been
30 widely used for this service is a combination of a Group VIII and a Group VI element, such as cobalt and molybdenum, on a suitable substrate, such as alumina. This hydrotreated product generally has a sulfur content of less than 5 ppm S, e.g., 1 ppm S.

The effluent is cascaded to a second pretreater zone 4 which contains a noble metal-containing porous inorganic oxide catalyst having pore openings of 12-member rings. Such porous inorganic oxides have pore windows framed by 12 tetrahedral members and include but are not limited to zeolites selected from the group consisting of zeolite beta, zeolite L, zeolite X, ZSM-12, ZSM-18, ZSM-20, mordenite and boggsite, zeolite beta being preferred. Faujasites such as Rare Earth Y (REY), Dealuminized Y (DAY), Ultrastable Y (USY), Rare Earth Containing Ultrastable Y (RE-USY), Si-Enriched Dealuminized Zeolite Y (LZ-210) (disclosed in U.S. Patents 4,711,864, 4,711,770 and 4,503,023) are also suited to use in the present invention.

In an alternative embodiment, the pretreater reactor zones may comprise separate reactors with an optional interstage separator therebetween.

The catalyst can contain from 0.1 to 1 wt %, preferably from 0.3 to 0.7 wt%, noble metal selected from the group consisting of platinum, palladium, iridium, rhodium and ruthenium. Platinum is preferred as well as combinations of platinum and palladium which are resistant to sulfur poisoning.

The noble metal component is preferably dispersed on the catalyst to provide a H/noble metal ratio of at least 0.8 as measured by hydrogen chemisorption, preferably at least 1.0 H/Pt metal ratio. The hydrogen chemisorption technique indicates the extent of noble metal agglomeration of a catalyst material. Details of the analytical technique may be found in Anderson, J.R., *Structure of Metallic Catalyst*, Chapter 6, p. 295, Academic Press (1975). In general, hydrogen chemisorbs selectively on the metal so that a volumetric measurement of hydrogen capacity counts the number of metal adsorption sites. Preferably the noble metal-containing catalyst has an alpha value higher than 100 and is unsteamed. The high acidity permits

operation at lower temperatures so as to minimize thermodynamic constraints on benzene saturation. Alpha value, or alpha number, is a measure of zeolite acidic functionality and is more fully described together with
5 details of its measurement in U.S. Patent No. 4,016,218, J. Catalysis, 6, pp. 278-287 (1966) and J. Catalysis, 61, pp. 390-396 (1980).

Process conditions in the second reaction zone depend on zeolite catalyst activity and extent of benzene
10 saturation and back-end bottoms required by product specifications. The total pressure and hydrogen partial pressure can be in the range of those used in conventional naphtha pretreating processes, e.g. 790-5500 kPa (100-800 psig), preferably 1135-4240 kPa (150-600 psig) total
15 pressure. Total pressure (or hydrogen partial pressure) can be higher if more benzene saturation is desired. More specifically such conditions for the second pretreating zone can be the same as those for the first pretreating zone as discussed above.

20 The hydrodesulfurization catalyst and the inorganic oxide catalyst of the second reaction zone can be loaded either in the same reactor or in separate reactors operating in a cascade mode without interstage operation.

The effluent 5 from the second pretreater zone 4 is
25 passed to a stripper 6 wherein ammonia, hydrogen sulfide and C1 to C4 hydrocarbons are stripped off as overhead 7. The stripper bottoms 8 are passed to a dehexanizer 9 wherein a C5 and C6 fraction 10 is taken off and a C7+ bottoms fraction 11 is passed to a reformer 12.

30 Reforming operating conditions include temperatures in the range of from 427°C (800°F) to 538°C (1000°F), preferably from 477°C up to 527°C (890°F up to 980°F) liquid hourly space velocity in the range of from 0.1 to 10, preferably from 0.5 to 5; a pressure in the range of
35 from about atmospheric up to 4900 kPa (700 psig) and higher, preferably from 700 kPa to 4200 kPa (100 psig to

600 psig) and a hydrogen-hydrocarbon ratio in the charge in the range from 0.5 to 20 and preferably from 1 to 10.

The reformer effluent 13 is passed to a distillation unit 14 wherein C1 to C4 hydrocarbons 15 are taken off.

5 The reformer effluent 16 is then combined with the C5 and C6 fraction 10 taken off the dehexanizer 9 to provide a combined gasoline boiling range product 17 which contains no greater than 1 wt% benzene, preferably no greater than 0.8 wt% benzene.

10 The combined gasoline boiling range product can exhibit enhanced end boiling point limits of gasoline fuels, referred to as T_{90} or (90 vol% temperature), reflecting the presence of fewer hydrocarbon components that boil above temperatures in a range of 177°C to 221°C
15 (350°F to 430°F). The combined gasoline boiling range product can have a T_{90} of no less than 166°C (330°F), preferably no less than 177°C (350°F). This indicates that the noble metal-containing catalyst in the second pretreating zone selectively converts C9+ bottoms into
20 isobutane and C5-C8 gasoline-range hydrocarbons.

The following example is provided to illustrate the invention.

Example

Two zeolite beta-containing catalysts containing noble
25 metal were evaluated under conditions compatible with conventional naphtha pretreating processes. The zeolite catalysts were evaluated in a hydrodesulfurization (HDS)/zeolite catalysts system using a commercial CoMo/Al₂O₃ catalyst as the desulfurization catalyst. The
30 experiments were conducted in a fixed-bed, down-flow, dual reactor pilot unit. The commercial HDS catalyst was loaded in the first reactor and the zeolite-containing catalyst downstream in a second reactor in a 1/2 volumetric HDS/zeolite catalyst ratio. The pilot unit was operated in
35 a cascade mode without interstage separation to remove zeolite catalyst poisoning ammonia and hydrogen sulfide

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from the first reactor effluent. The normal operating conditions were: 4.0 LHSV over HDS catalyst, 2.0 LHSV over zeolite catalyst, 712 nl/l (4000 scf/bbl) of hydrogen circulation rate, and 3900 kPa (550 psig) total pressure.

5 The HDS catalyst was kept at a constant 343°C (650°F) while the zeolite temperature was varied from 204°C-413°C (400°F-775°F) to obtain a wide range of conversion conditions. Table 1 lists properties of the naphtha used in the experiments.

10 The pretreated product contains ammonia and hydrogen sulfide derived from organic nitrogen- and sulfur-containing compounds in the feed to the pretreater. Such ammonia and hydrogen can be cascaded to the zeolite cracking zone. Because impurities in the feed such as
15 hydrogen sulfide and ammonia will deactivate the reforming catalyst, feed pretreating in the form of hydrotreating is usually employed to remove these materials.

Typically, feedstock and reforming products or reformat have the following analysis set out in Table 1
20 below:

TABLE 1

	API Gravity, °API	54.2
	Hydrogen, wt%	14.27
	Sulfur, ppmw	500
25	Nitrogen, ppmw	10
	Benzene, wt%	1.43
	Distillation (D2887), °C	(°F)
	IBP	-18 (-1)
	10%	57 (135)
30	50%	118 (244)
	90%	206 (402)
	EBP	279 (534)

The yields in the following Tables are based on raw naphtha feed.

CATALYSTS

Two catalysts were evaluated in the second pretreatment zone and their properties are set out in Table 2 below. Catalyst A was an unsteamed Pt/Beta/alumina catalyst of the present invention. Catalyst B was a steamed Mo/Beta/alumina catalyst used for comparative purposes.

TABLE 2

	<u>Catalyst A</u>	<u>Catalyst B</u>
10 Zeolite, wt%	65	65
Alumina, wt%	35	35
Platinum, wt%	0.5	--
Molybdenum, wt%	--	3.6
Alpha*	350	110
15 Surface Area*, m ² /g	459	422
n-C ₆ Sorption, cc/g	14.7	13.9
H/Pt Chemisorption	0.83	--

* Prior to metal addition

BENZENE SATURATION

20 The present invention serves to saturate benzene in the presence of hydrogen sulfide under naphtha pretreating conditions prior to dehexanizing. As shown in Table 3 below, the catalyst of the present invention achieved greater than 90% benzene saturation at temperatures above 288°C (550°F) while the Mo zeolite beta catalyst did not 25 saturate the benzene at all. In fact, the Mo zeolite beta catalyst produced more benzene.

TABLE 3
Benzene Saturation Activity

	<u>Feed</u>	<u>HDS</u>	<u>Catalyst-A</u>		<u>Catalyst-B</u>	
			288°C (550°F)	306°C (583°F)	316°C (600°F)	399°C (750°F)
5 Zeolite Temp.	-	-				
Benzene Yld, wt. %	1.4	1.5	0.1	0.01	1.6	1.7
Benzene Conv. %	-	-7	92	>99	-14	-21

Back-End Conversion

- 10 The noble metal-containing catalyst of the present invention was found to be very active for the conversion of 149°C (300°F) bottoms and for retaining products as C₃+ gasoline-range hydrocarbons. Both zeolite-beta catalysts A and B also gave high yields of i-butane which can be
- 15 alkylated with C₃-C₄ olefins to high-octane alkylate. The back-end conversion comparison for the two catalysts are shown in Table 4 below.

TABLE 4Back-End Conversion Activity

		<u>HDS only</u>	<u>Catalyst-A</u>		<u>Catalyst-B</u>	
			288°C <u>550°F</u>	306°C <u>583°F</u>	316°C <u>600°F</u>	399°C <u>750°F</u>
5	Zeolite Temp.	-				
	149°C ⁺ (300°F ⁺) Conv., %	4.6	32	91	20	43
	i-C ₄ Yield, wt.%	0.3	2.8	12.1	0.6	6.6
	C ₅ ⁺ Yield, wt.%	98.4	95.8	82.1	97.7	87.0

Driveability Index

T_{50} , T_{90} and driveability index (DI) of the pretreated naphtha by the HDS/zeolite-beta catalyst systems were reduced. The reformulated gasoline specification may limit T_{90} at 149°C (300°F). The driveability index is defined as follows:

$$\text{DI} = 1.5 \times T_{10} + 3.0 \times T_{50} + T_{90}$$

Some naphthas (or gasolines) have a DI of 1400 DI and lower DIs are desirable, e.g., 1300 or lower. As shown in the following table, the catalyst of the present invention is superior for the boiling point reduction and driveability index improvement:

TABLE 5
Driveability Index

		<u>HDS only</u>	<u>Catalyst-A</u>		<u>Catalyst-B</u>	
5	Zeolite Temp.	-	288°C	306°C	316°C	399°C
	149°C* (300°F*) Conv., %	4.6	<u>550°F</u>	<u>583°F</u>	<u>600°F</u>	<u>750°F</u>
<u>C₈+ Naphtha (based on D2887)</u>						
10	T ₁₀ , °C (°F)	63 (146)	56 (132)	29 (85)	59 (138)	48 (118)
	T ₃₀ , °C (°F)	123 (253)	117 (243)	93 (199)	118 (244)	112 (233)
	T ₉₀ , °C (°F)	206 (402)	178 (353)	135 (275)	194 (381)	178 (353)
	Driveability Index	1380	1280	1000	1320	1229

Integration With Reformer

To minimize benzene production in the conventional reforming complex, only C₇⁺ naphtha will be upgraded. Consequently, a dehexanizer before the reforming reactors to remove benzene and its precursors (methylcyclopentane and cyclohexane) is used. The impact of reforming C₇⁺ naphtha was examined using a kinetic model that simulates commercial reforming performance. The simulations were set at conditions to produce C₃⁺ reformates with an octane of 100 R+O. The C₃-C₆ cut produced from the HDS/Beta pretreating stage, which was removed by dehexanizer and not reformed, was then blended back into the C₃⁺ reformat as "combined C₃⁺ product." The following table illustrates the comparison between the two zeolite beta-containing catalysts.

TABLE 6

HDS/Zelite/Dehexanizer/Reforming Integration

<u>Pretreater Conditions</u>		<u>HDS only</u>	<u>Catalyst-A</u>		<u>Catalyst-B</u>	
5	Zeolite Temp.	-	288°C 550°F	306°C 583°F	316°C 600°F	399°C 750°F
	149°C* (300°F*) Conv., %	4.6	32	91	20	43
<u>Integrated Process Performance</u>						
	Total i-C ₄ , wt.%	1.2	3.8	12.8	1.5	7.4
10	H ₂ Product, nl/l (scf/b)	90 (505)	91 (510)	28 (160)	93 (520)	50 (280)
<u>Combined C₃* Product</u>						
	Octane, R+O	94.5	94.0	92.1	94.0	93.7
	Yield, vol.%	86.5	83.4	75.7	85.8	79.2
	Benzene, vol.%	1.9	0.8	0.6	2.1	2.0
15	T ₁₀ , °C (°F)	33 (92)	31 (88)	28 (82)	32 (90)	28 (82)
	T ₅₀ , °C (°F)	106 (223)	104 (219)	86 (186)	104 (219)	96 (205)
	T ₉₀ , °C (°F)	159 (319)	157 (315)	133 (272)	158 (317)	156 (313)
	Driveability Index	1126	1104	953	1109	1051

As shown in Table 6, the benefits found in the pretreating stage also cascade to the overall process integration. For example, benzene content, end-point, and driveability index of the combined C_3^+ product were reduced as compared to those for conventional HDS alone in the pretreater.

Incorporating a sulfur-tolerant noble metal promoted porous inorganic oxide, e.g., zeolite, catalyst into the bottom of a naphtha pretreating reactor can lower combined C_3^+ product (reformate and C_3 - C_6 dehexanizer overhead) end-point and benzene content for refiners with dehexanizers. Reformulated gasoline specifications limit benzene content to 1.0 % and T_{90} is much lower than that for conventional gasoline. Cascade HDS/zeolite followed by a dehexanizer can allow refineries to achieve benzene reduction and driveability index improvement by using a low-capital cost catalyst replacement approach. The present invention is especially attractive to refineries without C_3/C_6 isomerization units.

In a preferred embodiment the invention provides a process integrated into the reformer section of a refinery for the manufacture of high octane gasoline. The invention can improve the economics of meeting the benzene specification of the gasoline pool, preferably reducing the pool benzene content below 1% or 0.8 %.

IT IS CLAIMED:

1. An integrated process for providing a gasoline boiling range reformat-containing product produced from naphtha comprising:

- 5 1) pretreating a raw naphtha feedstream containing benzene by contacting with a hydrodesulfurization catalyst thereby forming H_2S and thereafter cascading said feedstream over a noble metal-containing porous crystalline inorganic oxide catalyst comprising pores having openings
10 of 12-member rings under conditions sufficient to effect benzene saturation, thereby providing a pretreated effluent of reduced benzene content;
- 2) dehexanizing the pretreated effluent to provide a C_6 - hydrocarbon-containing overhead stream and a bottoms
15 stream containing C_7+ hydrocarbons;
- 3) reforming said bottoms stream to form a reformat stream; and
- 4) combining said reformat stream and said C_6 - hydrocarbon-containing overhead stream to provide a
20 combined gasoline boiling range product.

2. The process of claim 1 wherein said noble metal-containing inorganic oxide catalyst comprises zeolite having pores with openings of 12-member rings selected from the group consisting of zeolite beta, zeolite L, zeolite X,
25 zeolite Y, Dealuminized Y, Ultrastable Y, Ultrahydrophobic Y, Si-Enriched Dealuminized Y (LZ-210), ZSM-12, ZSM-18, ZSM-20, mordenite and boggsite.

3. The process of claim 2 wherein said noble metal-containing zeolite comprising pores having openings of 12-member rings is zeolite beta.
30



4. The process of claim 1 wherein said catalyst contains from 0.1 to 1 wt % noble metal selected from the group consisting of platinum, palladium, iridium, rhodium and ruthenium.

5 5. The process of claim 2 wherein said zeolite is zeolite beta and said catalyst contains from 0.3 to 0.7 wt% platinum.

10 6. The process of claim 1 wherein said noble metal is dispersed on said catalyst to provide a H/noble metal ratio of at least 0.8 as measured by hydrogen chemisorption.

7. The process of claim 5 wherein said platinum is dispersed on said catalyst to provide a H/Pt metal ratio of at least 1.0 as measured by hydrogen chemisorption.

15 8. The process of claim 1 wherein said combined gasoline boiling range product contains no greater than 1 wt% benzene.

20 9. The process of claim 8 wherein said combined gasoline boiling range product has a T_{90} of no less than 177°C (350°F) and a Driveability Index of no greater than 1300.



10. An integrated process for providing a gasoline boiling range reformat-containing product produced from naphtha comprising:

5 1) pretreating a raw naphtha feedstream by contacting with a hydrodesulfurization catalyst and thereafter cascading said feedstream over a noble metal-containing zeolite catalyst comprising pores having openings of 12-member rings thereby providing a pretreated effluent;

10 2) stripping off C4⁻ hydrocarbons, hydrogen sulfide and ammonia from the pretreated effluent thereby providing a stripped effluent;

3) dehexanizing the stripped effluent to provide a C5 to C6 hydrocarbon-containing overhead stream and a bottoms stream containing C7+ hydrocarbons;

15 4) reforming said bottoms stream to form a reformat stream;

5) distilling said reformat stream to provide a C1 to C4 hydrocarbon-containing overhead and C5+ reformat bottoms fraction; and

20 6) combining said C5+ reformat bottoms fraction and said C5 to C6 hydrocarbon-containing overhead stream to provide a combined gasoline boiling range product.

