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DESCRIPTION

Field of the invention

[0001] The invention relates to petrochemistry, and more specifically, to methods and plants for the production of a base for the production of high-octane gasolines or high-octane gasoline components (concentrates of alkylaromatics) from raw hydrocarbon fractions, fractions of gaseous olefins and oxygenates, wherein starting components are mixed, feedstock is heated and supplied into a reactor, in which the reaction is carried out in the presence of a zeolite-containing catalyst, a gasoline base with improved antiknock rating or a concentrate alkylaromatics is isolated by the separation of the conversion product, while diverting simultaneously the reaction water and exhaust gases.

[0002] This description uses the following terms.

[0003] **Gasoline** is the main component of gasolines with improved antiknock rating for the production of commercial automotive and aviation gasolines with a low content of benzene, sulfur compounds and olefins, durene, and with no diene hydrocarbons in the composition, or a concentrate of alkylaromatics.

[0004] **Concentrate of alkylaromatics** (interchangeable with **concentrate alkylaromatics**) is aromatic hydrocarbon concentrate with the total content of alkyl-substituted aromatics (e.g. xylenes, ethylbenzene) exceeding the total content of non-alkyl-substituted aromatics (e.g. benzene, naphthalene).

[0005] **Feedstock hydrocarbon fractions** (interchangeable with **raw hydrocarbon fractions**) are various hydrocarbon fractions, such as 1) straight-run fractions; 2) fractions of hydrocarbons with higher benzene content, including, for example, a benzene fraction made of straight-run gasoline reformates; 3) fractions containing components prone to resin formation, such as dienes and aromatic olefins, including light fractions of pyrolysis gasolines and mixtures thereof with other fractions; 4) hydrocracking gasolines; 5) stable natural gasoline; 6) gas condensate distillate; and other similar products and mixtures thereof.

[0006] **Methanol and / or other oxygenates and / or olefin-containing feed-stock**, which means that a mixture of methanol and / or other oxygenates, including ethanol and water, and, in addition to or instead of it, a light olefin-containing feed-stock, such as various ethylene fractions, propane-propylene fractions, butane-butylene fractions, broad olefin containing fractions and mixtures thereof can be used. Thus, the feedstock can include in addition to the raw hydrocarbon fractions one or more of the following: methanol, ethanol, oxygenates, water, light olefins, light-olefinic fractions and combination and variations of these.

[0007] **Methanol and / or other oxygenates** are various oxygenate fractions that can include

one or more of the following: methanol, crude methanol, a mixture of alcohols, mixture of their ethers, water and combination and variations of these.

[0008] Zeolite-containing catalyst is one or a mixture of zeolite-containing catalysts used in processes of hydrocarbon and/or oxygenates and/or olefin conversion, such as 1) co-processing of oxygenates with hydrocarbons into predominantly gasoline-range hydrocarbons; 2) conversion of methanol into hydrocarbons; 4) co-processing of olefins with hydrocarbons; 5) olefin to gasoline processes; 6) catalytic deparaffinization; and other similar processes and combinations thereof.

[0009] Reaction water is the water chemically created during the production of gasoline from oxygenate-containing feedstock.

[0010] RON stands for the Research Octane Number. It is determined by the method described in GOST 8226-82: "The method is to compare the antiknock rating of test fuel with the antiknock rating of reference fuels expressed by the octane number" on a bench motor. RON characterizes the behavior of a gasoline engine in the modes of small and medium loads. To assess RON, there are also used indirect methods based, for example, on the analysis of chromatograms of gasoline or the analysis of the characteristics of the low-temperature gas-phase oxidation reaction simulating pre-ignition processes, or by the measurement of indirect gasoline physical quantities that correlate well with RON indicators, measured in accordance with GOST 8226-82. Measurement tools for indirect RON assessment are called Octane meters.

[0011] BRON stands for motor fuel component blending RON that is determined by calculation based on the analysis of RON of mixture of this component and straight-run gasoline taken in predetermined weight or volume ratios and analysis of straight-run gasoline RON (only weight BRONs are considered here). Analogous BRON can be calculated for other hydrocarbon fractions besides straight-run gasoline.

[0012] Reaction zone is an isolated space in a reactor, where the catalyst is located.

[0013] This description refers to the production of a major component of commercial gasoline. In some cases, depending on the feedstock, the conversion product may contain fractions having temperatures of over 210 °C in an amount of more than 2 vol. %, which does not meet the commercial gasoline requirements published in the European Norm EN 228:2008.

Prior art of the method

[0014] One of the aspects of the present invention relates to methods for producing motor gasolines having an improved antiknock rating of the raw hydrocarbon fractions, fractions of gaseous olefins and oxygenates, where the starting components are mixed, the feedstock is heated and supplied into a reactor, wherein the reaction is carried out in the presence of a

zeolite-containing catalyst, high-octane gasoline is isolated by separation of the conversion product, while diverting simultaneously the reaction water and exhaust gases.

[0015] The patent No. 2429910 of the Russian Federation for the invention, which was published in 2011, described a method for co-processing of low-octane hydrocarbon fractions and aliphatic alcohols and / or dimethyl ether.

[0016] This method is the closest to the present invention and is taken as the prototype for the method.

[0017] As an example of use of a reactor, the patent referred to the Patent No. 65045 of the Russian Federation for utility model No. 65045, which was published in 2007, which describes a plant for the production of synthetic gasoline from aliphatic alcohol, in particular, methanol, where a reactor with coaxial heat pipes is proposed. The heat pipes, due to coolant evaporating and condensing processes occurring in them, allow efficient redistribution of heat from heating zones adjacent to the surface of the heat pipes to heated zones also adjacent to the surface of the heat pipes. The distance between the surfaces of the heat pipes limiting the reactor space in the reactor described in the prior art should be no more than 150 mm. The ratio of the reaction zone volume to the reactor volume will be less than 0.3-0.5. Using another configuration of the heat pipes, for example as described in the Patent No. 2433863 of the Russian Federation, which was published in 2011, does not solve this problem either, because the reactor (reaction zone) volume increases as the square of its diameter and the required surface of the heat pipes necessary for the heat exchange with the reaction zone depends linearly on the volume of the reaction zone. Thus, the use of reactors with the heat pipes does not allow achieving any reduction in metal consumption of the reactor structure. This limits the creation of a large-scale reactor with heat pipes having a capacity of dozens and hundreds of cubic meters.

[0018] Furthermore, a part of the reactor space between the heat pipes, when it is used in the process described in the prototype, is used to heat the feedstock supplied into the reactor in order to provide the coolant condensation on the inner surface of the heat pipes in an amount necessary for removing heat from the inlet (front) catalyst bed, where exothermic reactions prevail, and where the heat removal is performed by evaporation of coolant in the inner space of the heat pipe. Thus, according to Examples 3-9 and 11-17 in the patent No. 2433863 of the Russian Federation, feedstock is supplied into the reactor at a temperature of up to 100-205 °C.

[0019] Thus, another disadvantage of this method is the limitation of production capacity of motor gasolines due to limiting the possibility of increasing the efficient and overall reactor volumes.

[0020] The third disadvantage of this method is the inability to control the temperature throughout the catalyst bed. Thus, in the lower catalyst bed, where the endothermic reactions prevail, the temperature will always be lower than the coolant condensation temperature on the

heat pipe walls. The temperature in the input (front) part of the catalyst bed, where the exothermic reactions prevail and where the heat removal is performed by evaporation of the coolant flowing over the surface of the heat pipes from above the condensation zone will always be higher than the temperature of coolant condensation / evaporation in the heat pipe. Thus the temperature at the top of the reaction zone will always be higher by 15-45 °C than the temperature at the bottom of the reactor. This leads to the fact that it is impossible to adjust the temperature at the bottom of the catalyst bed, without increasing the temperature at the top (front) part of the catalyst bed in the reaction zone limited by the heat pipe space. Thus, if during operation, the conversion of methanol is reduced down to 99.5 %, which will lead to an increase in the content of methanol in the reaction water, it is advisable to raise the temperature in the lower catalyst bed by 1-2 °C.

[0021] Another disadvantage of this method is the fact that upon increasing the methanol to hydrocarbon ratio, the amount of heat generated in the front catalyst bed increases too. This, in turn, limits the temperature of feedstock being supplied to the reactor as the feedstock of a lower temperature is supplied to the reactor and is superheated exactly in the reactor in the area limited by the surface of the heat pipes located upstream the front bed, which is necessary for creating a sufficient coolant condensation on the surface of the heat pipes to dissipate heat.

[0022] Another drawback of the described method is restrictions by the fractional composition of feedstock. Thus, according to the method described in the prototype, the feedstock is supplied into the reactor in the gaseous phase at a temperature of no higher than 210 °C, at the same time, the boiling point of n-octane will be 233 °C at 1 MPa, which is higher by 23 °C than indicated in the prototype.

[0023] Therefore, another drawback of this method is deterioration of the plant's capability of recycling feedstock having various fractional composition. Indeed, the normal boiling temperature of n-octane is 126 °C, and its boiling point is 233 °C at the pressure of 1 MPa, which does not allow using octane as feedstock at a pressure in the reaction zone of higher than 1 MPa.

[0024] Yet another drawback of the proposed method is that the high concentrations of light olefins and oxygenates in the feedstock, in particular, of methanol, dimethyl ether, ethylene lead to increased formation of heavy compounds contained in the conversion product liquid hydrocarbon fractions; some of these compounds have high melting points, the presence of which is undesirable in commercial gasolines and usually their concentration is no more than 2 % by weight.

[0025] Methods and plants for producing gasoline or a concentrate of aromatic hydrocarbons using two reactors connected in series, wherein the reactors each contain zeolite catalysts are disclosed in WO 2015/147700 A1 and US 4 835 329 A.

[0026] A process and plant for producing C₂-C₄ olefins from methanol and/or dimethyl ether is

disclosed in US 2009/124841 A1, wherein the methanol is preheated and reacted in a prereactor on a dehydrating catalyst, e.g. alumina, to obtain dimethyl ether, and the obtained methanol/dimethyl ether mixture is supplied to a reactor with a zeolite based catalyst which may consist of two or more series-connected reaction stages.

Summary of the method according to the invention

[0027] The present invention mainly aims to propose a method of producing gasoline from raw hydrocarbon fractions, fractions of gaseous olefins and oxygenates, as disclosed in appended claim 1, wherein the starting components flow is supplied to a unit for supplying the flow to be processed into a reactor, wherein the reaction is carried out in the presence of a zeolite-containing catalyst, gasoline is isolated by separation of the conversion product, while diverting simultaneously the reaction water and the exhaust gases, allowing improvement of the quality of the resulting gasoline by reducing the content of benzene in it, as well as refractory components, such as durene, as well as by the possibility of regulating the temperature throughout the catalyst bed without the use of complex heat exchange reactor equipment, as well as the possibility of increasing the useful volume of the reactor equipment, which is the technical problem specified.

[0028] The specified technical problem is solved by providing a method of producing gasoline from raw hydrocarbon fractions, fractions of gaseous olefins and oxygenates, wherein the feedstock component flow is supplied to a unit for supplying the flows to be processed into a reactor, wherein the reaction is carried out in the presence of a zeolite-containing catalyst, high-octane gasoline is isolated by separation of the conversion product, while diverting simultaneously the reaction water and the exhaust gases, wherein

1. (a) as a reactor, a reactor is used, which contains at least two reaction zones with a zeolite-containing catalyst,
2. (b) between which there are further arranged means for mixing the reaction product from the previous reaction zone and the supplied methanol and / or other oxygenates and olefin-containing feedstock, and
3. (c) by means of the unit for supplying flows the methanol and / or other oxygenates and olefin-containing feedstock are separated into at least two flows, wherein

a first flow of methanol and / or other oxygenates and olefin-containing feed-stock is directed into the space upstream the first reaction zone where it is mixed with the flow of raw hydrocarbon fractions and is supplied into the first reaction zone of the reactor, and

a second flow of methanol and / or other oxygenates and olefin-containing feedstock is mixed in the downstream reaction zones with the conversion product from the upstream reaction zone.

[0029] An example of such a reactor can be a vertical multiple-shell reactor or a horizontal reactor, the housing of which accommodates several series-connected catalyst baskets; an example of such solutions using a horizontal reactor is shown in Figure 2. An example of such a reactor can be single or multibed reactor or reactor system, such as arrangement of reactors in series, arrangement of reactors in parallel and combinations thereof. An example of such a reactor can be fixed bed reactor, fluidized bed reactor, reactors designed for continuous catalyst regeneration. An example of such a reactor can be isothermal, adiabatic or pseudo-isothermal reactor.

[0030] An example of the catalyst can be catalysts described hereinafter.

[0031] Due to these favorable characteristics, the following possibility appears:

- Increase in RON of gasolines produced using the method described;
- Reduction of the content of undesired compounds therein with high melting points and being prone to crystallization, such as, for example, durene, 2,6-dimethylnaphthalene;
- Reduction of the content of benzene therein, which is a toxic and carcinogenic substance having the maximum volatility (fugacity) among other aromatics;
- Increase in production of gasoline of higher quality and with an improved antiknock rating due to the possibility of increasing the reactor capacity.

[0032] In one of the embodiments, methanol and / or other oxygenates and / or olefin-containing feedstock are separated by means of said unit into at least two flows, the first flow is directed into the space upstream the first reaction zone where it is mixed with the flow of raw hydrocarbon fractions, and the next flows are mixed in the downstream reaction zones with the conversion product from the upstream reaction zone.

[0033] In another embodiment one or more of the reaction zones can be **not** supplied with flows of methanol and / or other oxygenates and / or olefin-containing feedstock as long as at least two reaction zones are supplied with the flows of methanol and / or other oxygenates and / or olefin-containing feedstock.

[0034] In another embodiment, methanol and / or other oxygenates and / or olefin-containing feedstock are separated by means of said unit into at least one flow, which is directed to the downstream reaction zone, where said stream is mixed with the conversion product from the upstream reaction zone, wherein the first reaction zone is pre-supplied with a pre-mixed flow of methanol and / or other oxygenates and / or olefin-containing feedstock and the flow of raw hydrocarbon fractions. The flow of methanol and / or other oxygenates and / or olefin-containing feedstock is mixed with the raw hydrocarbon fractions in the space upstream the first reaction zone or more upstream in the direction of the hydrocarbon feedstock supply flow, and prior to supplying to the first reaction zone, the mixed flow of methanol and / or other oxygenates and / or olefin-containing feedstock and the flow of raw hydrocarbon fractions are heated, e.g., they may be preheated prior to mixing, during mixing or upon mixing, or prior to

or upon introduction into the reaction zone.

[0035] In another embodiment flows of raw hydrocarbon fractions and flows of methanol and / or other oxygenates and / or olefin-containing feedstock can be heated prior to mixing, during mixing or upon mixing, or prior to or upon introduction into the reaction zone.

[0036] In another embodiment one or more reaction zones can be supplied with at least part of the raw hydrocarbon fraction flow.

[0037] The flow is directed from the unit said in the above five embodiments into the reaction zone directly or through a mixing zone located between the reaction zones and having means for mixing.

[0038] Indeed, the distributed supply allows improving the quality of the produced base stock for the production of gasoline by increasing the concentration of components having a high BRON, such as meta- and para-xylenes, in the resulting product, and allows reducing the concentration of undesirable durene, methylnaphthalenes having high melting temperatures, significantly decreasing the concentration of normal C6+ paraffins having an extremely low antiknock ratings, as well as significantly reducing the benzene content (see comparative Table 3 for examples No. 2 and No. 3).

[0039] Furthermore, the use of a reactor with supply of methanol and / or other oxygenates and / or olefin-containing feedstock to each or at least two reaction zones allows increasing production due to the ability to increase the efficient and total reactor volume. Thus, there is the possibility to simultaneously control the parameters of resulting gasoline due to the possibility to control the supply of methanol and / or other oxygenates and / or olefin-containing feedstock to the first and / or downstream reaction zones and / or the temperature of the mixture supplied to the first reaction zone or downstream reaction zones

[0040] If methanol and / or other oxygenates and / or olefin-containing feedstock is supplied for the conversion by not a single flow, but divided into 2-5 flows and added to the converted mixture successively prior to supplying the converted mixture into the downstream reaction zone, and there is a mixing device installed upstream each downstream reaction zone, the function of which can also be carried out by a protective bed (precontact) poured over the catalyst bed in a catalytic basket, and the function of the mixing device can also be performed by a channel connecting the catalyst baskets to be placed in a horizontal reactor, this will allow using a vertical or horizontal reactor of simple design, with a high efficiency of the reactor volume.

[0041] In one of the embodiments, function of a mixing device is carried out by a protective bed of neutral or weakly-reacting material, for example quartz, in forms such as granules; spheres; fraction of crushed materials.

[0042] In one of the embodiments, function of a mixing device is carried out by a channel

connecting the reactor spaces where the reaction zones are located.

[0043] In one of the embodiments, function of a mixing device is carried out by specifically controlled speed and/or direction of supplied feedstock flow, e.g. by using tangential supply of methanol and / or other oxygenates and / or olefin-containing feedstock to the reaction zone.

[0044] In one of the embodiments, the flow rate of methanol and / or other oxygenates and / or olefin-containing feedstock to each reaction zone is controlled, and the temperature of the feedstock to be supplied to the reaction zone is controlled so that the maximum catalyst bed temperature in the reaction zone does not exceed 420 °C in the production of a base stock for the production of gasolines and 500 °C in the production of alkylaromatics concentrate as a high-octane gasoline component.

[0045] Additionally, when decreasing the supply of methanol and / or other oxygenates and / or olefin-containing feedstock to the first reaction zone, the temperature of raw mixture supplied into the first reaction zone can be increased, and vice versa.

[0046] Additionally, when decreasing the supply of methanol and / or other oxygenates and / or olefin-containing feedstock to the reaction zone, the temperature of raw mixture supplied into the reaction zone can be increased, and vice versa.

[0047] Indeed, the optimal solution is the option of supplying methanol and / or other oxygenates and / or olefin-containing feedstock into the reaction zones so that the heating temperature of the catalyst bed in the reaction zone does not exceed 420 °C in the production of the base stock for the production of high-octane gasoline and 500 °C in the production of alkylaromatics concentrate, whereas the temperature of the catalyst end bed should be respectively by 40-70 °C below said maximum temperature. Meanwhile, this raw mixture can be supplied to the first reaction zone at a reduced temperature of up to 310 °C, whereby most of the methanol and / or other oxygenates and / or olefin-containing feedstock can be supplied to the first reaction zone, which allows controlling the supply of methanol and / or oxygenates and / or olefin-containing feedstock to the process by increasing the temperature of the feedstock mixture supplied into the first reaction zone while reducing the consumption of methanol and / or other oxygenates and / or olefins; and vice versa, decreasing the feedstock mixture temperature while increasing the flow rate of methanol and / or other oxygenates and / or olefins.

[0048] Additionally, if it is necessary to increase the flow rate of methanol and / or other oxygenates and / or olefins to the second and / or downstream shelves / reaction zones, methanol and / or other oxygenates and / or olefins may be supplied at a lower temperature to prevent overheating of the front bed resulting in rapid deactivation of the catalyst and higher gassing.

[0049] Additionally, if it is necessary to decrease the temperature of the second and/or downstream reaction zones, methanol and / or other oxygenates and / or olefins may be

supplied at a lower temperature to prevent overheating of the front part of the zone resulting in rapid deactivation of the catalyst and higher gassing.

[0050] In one of the embodiments, the flow rate of methanol and / or other oxygenates and/or olefin-containing feedstock to each reaction zone is regulated independently.

[0051] Due to this advantageous characteristic, there appears a possibility to mitigate uneven catalyst deactivation among reaction zones.

[0052] Example 1 of this patent provides information regarding adiabatic heating of the front catalyst bed during the conversion of hexane-heptane fraction mixture (83.5 wt. %), methanol (6.6 wt. %), water (7 wt. %) and ethylene fraction (2.9 wt. %). Adiabatic heating of the front catalyst bed in an isothermal reactor was about 60-65 °C, while, due to the conversion of n-hexane, n-heptane and naphthenes contained in the hydrocarbon phase, the catalyst temperature in the end catalyst zone was about 350-360 °C.

[0053] There is an embodiment, wherein raw hydrocarbon fractions or a mixture of raw hydrocarbon fractions and methanol and / or other oxygenates and / or olefin-containing feedstock are heated in two stages: at the first stage, raw hydrocarbon fractions or a mixture of raw hydrocarbon fractions and methanol and / or other oxygenates and / or olefin-containing feedstock are vaporized, heavier non-evaporated components are separated, and at the second stage the vaporized components are heated, e.g., preferably superheated.

[0054] Due to this advantageous characteristic, there appears a possibility to reduce the content of heavy high-boiling components in gasoline and reduce the coking rate of the catalyst, reduce potential coke deposition on the inner surface of pipes and hydrocarbon heating unit.

[0055] There is, moreover, an embodiment, wherein raw hydrocarbon fractions and methanol and / or other oxygenates and / or olefin-containing feedstock are supplied to the first reaction zone with joint flow. Due to this advantageous characteristic, there appears a possibility not to use any mixing device in the first reaction zone; also, the presence of water in the raw mixture inhibits radical oligomerization of dienes and aromatic olefins while heating the raw mixture, whereby they may be present in the raw hydrocarbon fractions.

[0056] There is also an embodiment, wherein raw hydrocarbon fractions and methanol and / or other oxygenates and / or olefin-containing feedstock are supplied to each reaction zone by separate flows, wherein they are pre-mixed, and a bed of granules of neutral material and / or a fraction of the crushed quartz placed upstream the frontal catalyst bed is used as a mixing device, or a line connecting the reactor spaces, where the reaction zones are located, is used as a mixing device. Due to this advantageous characteristic, there appears a possibility to use the protective catalyst bed or the line connecting the reaction zones or the line through which the mixture is supplied into the first reaction zone, as a mixing bed.

[0057] There is also an embodiment, wherein methanol and / or other oxygenates and / or olefin-containing feedstock are supplied by separate flows to the first or second and each downstream reaction zones, where methanol and / or other oxygenates are supplied in the gaseous phase at a temperature of no more than 380 °C, which prevents its decomposition upon heating. Due to this advantageous characteristic, there appears a possibility to reduce the loss of methanol for its conversion to carbon oxides during heating up to higher temperatures.

[0058] There is also an embodiment, wherein methanol and / or other oxygenates and / or olefin-containing feedstock are supplied by separate flows to the first and / or second **and** / or subsequent reaction zones, where methanol and / or other oxygenates are supplied in the gaseous phase at a temperature of no more than 380 °C, which prevents its decomposition upon heating. Due to this advantageous characteristic, there appears a possibility to reduce the loss of methanol for its conversion to carbon oxides during heating up to higher temperatures and possibility of finite control of temperature and/or chemical composition of the product of each reaction zone by controlling the amount of methanol and / or other oxygenates and / or olefin-containing feedstock coming to each reaction zone.

[0059] There is an embodiment, wherein a 1-2 % aqueous solution of an industrial demulsifier is added to a mixture of hydrocarbons and water supplied to the separation at a ratio of 1:50 - 1:200 to the volume of the reaction water, the aqueous solution of the demulsifier is mixed with the input flow in a laminar static mixer or directly in a hydrocarbons and reaction water condenser located in the separation unit. Due to this advantageous characteristic, there appears a possibility to separate efficiently the mixture of reaction water and liquid hydrocarbons (conversion product) in the separation unit, as well as to prevent entrainment of catalyst particles with the liquid products of conversion and catalyst dust deposition on the surface of internal devices of fractionation columns.

[0060] There is an embodiment, wherein isothermal reaction zones are used as the last one or two reaction zones of the reactor.

[0061] Due to this advantageous characteristic, there appears a possibility to use the proposed invention at a low flow rate of methanol and / or other oxygenates and / or olefin-containing feedstock per ton of convertible hydrocarbon feedstock. Indeed, at low flow rates of methanol less than 20 % by weight of the converted feedstock weight, to enhance RON of convertible mixtures, the proposed combined embodiment can be used, wherein the heat input to the last or to the penultimate and last reaction zones can be performed directly to the reaction zone. Moreover, it allows increasing the temperature in the last reaction zone in order to increase the methanol and / or other oxygenates and / or olefin-containing feedstock conversion as it decreases and with a decrease of catalyst activity during its use.

[0062] There is also an embodiment, wherein adiabatic reaction zones containing no heat exchange devices are used as the last one or two reaction zones of the reactor, at the same time, the feedstock flow to be supplied to the final and / or penultimate reaction zones is further

heated, e.g., preferably superheated no more than up to 500 °C.

[0063] Due to this advantageous characteristic, there appears a possibility to use the proposed invention at a low flow rate of methanol and / or other oxygenates and / or olefin-containing feedstock per ton of converted hydrocarbon feedstock. Indeed, at low flow rates of methanol less than 20 % by weight of the weight of converted feed-stock, to enhance RON of convertible mixtures the proposed combined embodiment can be used, wherein the heat input to the last or the last and the penultimate reaction zones can be carried out by heating, preferably preheating, the converted feed-stock to 500 °C before it is supplied to the respective reaction zones. This embodiment with the supply of heat to the last or the last and the penultimate reaction zones has also the advantage that with an increase in the useful volume of the reactor, the temperature can be adjusted in the last or the last and the penultimate reaction zones in the direction of the feedstock flow, regardless of the temperature in the first reaction zones (in the direction of the feedstock flow).

[0064] There is also an embodiment, wherein pseudo-isothermal reaction zones are used as the last one or two reaction zones of the reactor. In this embodiment, reaction zones contain heat-exchange devices which are unable to sustain constant temperature throughout all of the reaction zone but can partially mitigate temperature difference between front and end parts of the reaction zone without escalating the low of methanol and / or other oxygenates and / or olefin-containing feed-stock supplied to the reaction zone.

[0065] Due to this advantageous characteristic, there appears a possibility to use the proposed invention at a lower flow rate of methanol and / or other oxygenates and / or olefin-containing feedstock per ton of converted hydrocarbon feedstock.

[0066] There is another embodiment, wherein separated reaction water with an admixture of methanol and / or other oxygenates, is supplied to a distillation column, the distillation residue of which is water cleaned from residues of hydrocarbons with a reduced content of methanol and / or other oxygenates and the distillate of which is water with an increased content of oxygenates and hydrocarbons, while oxygenates, e.g. preferably methanol water and hydrocarbons produced during the distillation process are recycled to the reactor inlet to the beginning of the process in a mixture with methanol and / or other oxygenate and / or olefin-containing feedstock.

[0067] Due to this advantageous characteristic, there appears a possibility to increase the efficiency of the process for producing gasolines with improved antiknock rating due to more complete processing of all the components. The presence of water in the original raw mixture promotes the reduction in the coke deposition on the catalyst surface during its operation.

[0068] There is an embodiment, wherein a zeolite-containing catalyst is used, wherein a synthetic amorphous aluminosilicate or an inorganic oxide is used as a binder component, which is a mixture of alumina in an amount of 30-70 wt. % and silica in an amount of 70-30 wt. %. Due to this advantageous characteristic, there appears a possibility to increase the catalyst

useful life, as there is no destruction of the binder under the action of superheated steam produced in the process of converting feedstock and present in the convertible feedstock.

[0069] There is an embodiment, wherein the catalyst is used which was described in RU 2160161 for the purposes of producing liquid hydrocarbons of dimethyl ether. In this embodiment the use of said catalyst is not restricted to the products and feeds indicated in the initial patent. This catalyst is based on crystalline pentasil-type aluminosilicate having a molar ratio of $\text{SiO}_2/\text{Al}_2\text{O}_3 = 25 - 100$ characterized by the presence of residual amounts of sodium ions being equivalent to a content of 0.05 - 0.1 wt. % of sodium oxide, containing said crystalline aluminosilicate and binder, wherein it further contains cobalt oxide, oxides of rare earth elements and zinc oxide in the following ratio, wt.%:

Zinc oxide: 0.5 - 3.0

Oxides of rare earth elements: 0.1 - 5.0

Cobalt oxide: 0.05 - 2.5

Crystalline aluminosilicate: 63 - 69.8

Binder: the rest.

[0070] Due to this advantageous characteristic, there appears a possibility to increase the octane characteristics of the produced gasoline due to increased formation of isoparaffins and alkylaromatics, primarily meta-xylene and p-xylene with BRON of 117.5 and 116.4, respectively, with the observed decrease in the concentration of benzene in the produced gasoline up to 0.5-1.0 wt. %, with a reduction in the olefin content of down to 1.5-2 wt. %, with a reduction in the content of dienes and trienes to trace amounts, with a reduction in the content of sulfur-containing compounds by 5 to 20 times, with simultaneous formation of hydrogen sulfide exhausted with gases.

[0071] There is an embodiment, wherein the catalyst is used which was described in RU 2544017 for the purposes of aromatization of $\text{C}_3\text{-C}_4$ gases, low octane hydrocarbon fractions and aliphatic alcohols, and mixtures thereof. In this embodiment the use of said catalyst is not restricted to the products and feeds indicated in the initial patent. Catalyst is based on the pentasil group zeolites, which contains a mechanical mixture of two zeolites having a different silicate module: zeolite having a $\text{SiO}_2/\text{Al}_2\text{O}_3=20$ previously treated with an aqueous solution of alkali modified with oxides of rare earth elements (REE) in an amount of 0.5-2.0 wt. %, and zeolite having $\text{SiO}_2/\text{Al}_2\text{O}_3=82$ having a residual amount of sodium oxide of 0.04 wt. % taken in the ratio of 1.7 / 1 to 2.8 / 1, and the remainder being a binder in an amount of 20 to 30 wt. % of the catalyst weight.

[0072] Due to this advantageous characteristic, there appears a possibility to expand the

fractional composition of the converted hydrocarbon feedstock by including light hydrocarbon fractions.

[0073] There is an embodiment, wherein at least one reaction zone can contain zeolite-containing catalyst different from the catalysts in the other reaction zones. Due to this advantageous characteristic, there appears a possibility to expand the fractional composition of hydrocarbon feedstock and to selectively enhance conversion of undesirable gasoline components accumulating in downstream reaction zones.

[0074] There is another possible embodiment of the invention, where the temperature of the last reaction zone is increased during the catalyst operation each time by 1-2 °C upon an increase in allowable values of methanol and / or oxygenates concentration in the reaction water.

[0075] Due to this advantageous characteristic, there appears a possibility to increase the methanol and / or oxygenates conversion degree (to increase efficiency of the conversion process) and reduce the costs of recycling the reaction water.

[0076] There is an embodiment, wherein olefin-containing gas is used as olefins, selected from the group comprising ethylene-containing gas from plants for fractioning of exhaust gas from the catalytic cracking process, gas from the delayed coking process. Due to this advantageous characteristic, there appears a possibility to increase the efficiency of use of olefin-containing feedstock.

[0077] There is a further embodiment, wherein the reaction zones are located in multi-bed adiabatic reactor.

[0078] There is a further embodiment, wherein the reaction zones are located in in multi-bed isothermal reactor.

[0079] There is a further embodiment, wherein the reaction zones are located in in multi-bed pseudo-isothermal reactor.

[0080] There is a further embodiment, wherein the reaction zones are located in separate reactors, each one of which can be adiabatic, isothermal or pseudo-isothermal.

[0081] In another embodiment, the flow rate of methanol and / or other oxygenates and / or olefin-containing feedstock into the reactor is increased or decreased upon deviation of the parameters of the produced product from the set parameters.

[0082] In another embodiment, the flow rate of methanol and / or other oxygenates and / or olefin-containing feedstock into individual reaction zones is increased or decreased upon deviation of parameters of the flows coming from individual reaction zones from the set parameters.

[0083] All this allows increasing the total useful volume of reactors of the simplest design up to hundreds of cubic meters and simplify their design, and allows increasing the volume of produced gasoline and improve its antiknock rating.

[0084] The totality of essential features of the proposed invention is not known from the prior art for methods for the similar purpose, which allows making a conclusion that the invention as a method complies with the "novelty" criterion.

Prior art regarding a plant to be used for carrying out the method according to the present invention

[0085] A plant for the production of gasolines from raw hydrocarbon fractions, fractions of gaseous olefins and oxygenates, comprises a unit (1) for supplying methanol and / or other oxygenates and / or olefin-containing feedstock, a unit (2) for supplying raw hydrocarbon fractions, a unit for supplying flows to be treated into a reactor, and a reactor (6) having at least a reaction zone with a zeolite-containing catalyst and an outlet for subsequent separation of the conversion product.

[0086] A similar plant (a prototype) is described in the Russian patent for invention No. 2429910 (publ. 27.09.2011).

[0087] The disadvantage of the prototype is the lack of the ability to control the availability of heavy fractions in the hydrocarbon feedstock.

[0088] Another disadvantage of the described device is restrictions on fractional composition of feedstock. Thus, the description of prototype indicates that the feed-stock is supplied into the reactor in the gaseous phase at a temperature of no higher than 210 °C, at the same time, the boiling point of n-octane reaches 233 °C at 1 MPa, which is by 23 °C higher than indicated in the prototype. Therefore, another drawback of the plant is the reduction of capabilities of the plant for processing feed-stock of various fractional composition.

[0089] Another drawback of the prototype invention is the limitation of the output of motor gasolines by limiting the possibility of increasing the volume of the reactor.

[0090] This is because the distance between the surfaces of the heat pipes limiting the volume of the reaction zone should not exceed 150 mm. The ratio of the reaction zone volume to the reactor volume will be less than 0.3-0.5. In addition, the reactor volume increases as the square of its diameter, and the required surface for heat pipes for heat removal increases in direct proportion to its diameter. This limits the development of a large-scale reactor fitted with heat pipes having a volume of dozens of cubic meters. Furthermore, as noted above, it is impossible to adjust the temperature of the end catalyst bed (in the last, boundary reaction zone) regardless of the temperature at the top reaction zone.

[0091] A further disadvantage of the prototype is also that upon an increase in the methanol to hydrocarbon ratio, the amount of heat generated in the front layer of the catalyst increases. This in turn limits the temperature of the temperature of heating the feedstock supplied into the reactor.

Plant for carrying out the method of the invention

[0092] For carrying out the method according to the present invention it is therefore proposed a plant for producing gasolines from hydrocarbon fractions, fractions of gaseous olefins and oxygenates comprising a unit (1) for supplying methanol and / or other oxygenates and / or olefin-containing feedstock, a unit (2) for supplying raw hydrocarbon fractions, a unit for supplying flows to be treated into a reactor, and a reactor having at least a reaction zone with a zeolite-containing catalyst and an outlet for subsequent separation of the conversion product, wherein this plant allows at least mitigating the above drawback, namely, providing the improvement in the quality of the resulting gasoline due to reducing its benzene content, as well as content of high melting point components, such as durene, as well as due to the possibility of regulating the temperature throughout the catalyst bed without using a complicated heat-exchange reactor equipment, and the ability to increase the useful volume of the reactor equipment, which is the specified technical problem.

[0093] In view of this, the reactor is additionally provided with at least one more reaction zone, and the unit for supplying flows into the reactor is adapted to supply separate and / or pre-mixed flows of methanol and / or other oxygenates and / or olefin-containing feedstock and the flow of raw hydrocarbon fractions into the first reaction zone, and a flow of methanol and / or other oxygenates and / or olefin-containing feedstock into the second and / or subsequent reaction zone, wherein between the reaction zones, there is also located means for mixing the reaction products from the upstream reaction zone and supplied methanol and / or other oxygenates and / or olefin-containing feedstock.

[0094] Due to these advantageous characteristics, there appears a possibility to improve the quality of produced gasoline due to the reduction in the content of benzene and high melting point components such as durene in produced gasoline. Additionally, it becomes possible to adjust the temperature throughout the catalyst bed without using complicated heat-exchange reactor equipment.

[0095] Furthermore, it is possible to increase production of gasolines due to the possibility of increasing the volume of the reactor. Indeed, the use of a reactor with at least two reaction zones with methanol and / or other oxygenates and / or olefin-containing feedstock being supplied into each or at least two reaction zone allows increasing the production capacity due to the possibility of increasing the total volume of the reactor. Here, the reaction zones can be arranged in series in the vertical or horizontal direction.

[0096] In addition, the unit for supplying flows may contain a unit (5) for heating methanol and / or other oxygenates and / or olefin-containing feedstock, which has at least two outputs, the first of which is connected to the first input of the first reaction zone (61) directly or through the mixing zone, the second input of which is connected to the output of a unit (2) for supplying raw hydrocarbon fractions, wherein through the second output of the unit (5) there is supplied a flow of methanol and / or other oxygenates and / or olefin-containing feedstock to the second reaction zone. Here, the unit for supplying flows contains an additional heating unit (4), the input of which is connected to the output of the unit (2) for supplying raw hydrocarbon fractions directly, and the output is connected to the second input into the first reaction zone (61).

[0097] In addition, the unit for supplying flows may contain a unit (5) for heating methanol and / or other oxygenates and / or olefin-containing feedstock, which has at least two outputs, the first of which is connected to the first input of the first reaction zone (61) directly or through the mixing zone, the second input of which is connected to the output of a unit (2) for supplying raw hydrocarbon fractions, wherein through the second output of the unit (5) there is supplied a flow of methanol and / or other oxygenates and / or olefin-containing feedstock to the **third** reaction zone. Here, the unit for supplying flows contains an additional heating unit (4), the input of which is connected to the output of the unit (2) for supplying raw hydrocarbon fractions directly, and the output is connected to the second input into the first reaction zone (61).

[0098] In addition, the unit for supplying flows may contain a unit (5) for heating methanol and / or other oxygenates and / or olefin-containing feedstock having at least one output connected to the input into the second and / or subsequent reaction zone directly or through the mixing zone comprising means for mixing located between the reaction zones, and a unit (3) for mixing raw hydrocarbon fractions and methanol and / or other oxygenates and / or olefin-containing feedstock, whose output is connected to the input to the first reaction zone, directly or through an additional heating unit (4).

[0099] In addition, the unit for supplying flows may contain a unit (5) for heating methanol and / or other oxygenates and / or olefin-containing feedstock having at least one output connected to the input into the **third** reaction zone directly or through the mixing zone comprising means for mixing located between the reaction zones, and a unit (3) for mixing raw hydrocarbon fractions and methanol and / or other oxygenates and / or olefin-containing feedstock, which has at least two outputs which are connected to the inputs of the **first and second** reaction zones, directly or through an additional heating unit (4).

[0100] Means for mixing hydrocarbon fractions and oxygenates can be located upstream the first reaction zone.

[0101] Moreover, the plant may further contain a unit (10) for determining parameters of produced gasoline connected to a unit (11) controlling the flow rate of methanol and / or other oxygenates and / or olefin-containing feedstock, with which is additionally provided a unit (1) for supplying methanol and / or other oxygenates and / or olefin-containing feedstock.

[0102] Moreover, the plant may further contain a unit (10) for determining parameters of flow produced after individual reaction zones connected to a unit (11) controlling the flow rate of methanol and / or other oxygenates and / or olefin-containing feed-stock, with which is additionally provided a unit (1) for supplying methanol and / or other oxygenates and / or olefin-containing feedstock.

[0103] In addition, the unit (4) for heating raw hydrocarbon fraction may further include a unit for evaporating raw hydrocarbon fractions or mixtures of raw hydrocarbon fractions and methanol and / or other oxygenates and / or olefin-containing feed-stock and a unit for heating, e.g., preferably superheating, the formed vapors, wherein between them there is a separator for removal of heavy hydrocarbon fractions containing mainly hydrocarbon components with a molecular mass of no more than 180, with boiling point temperatures of up to 250 °C under normal conditions.

[0104] Due to this advantageous characteristic, in an embodiment using a superheater, there appears a possibility to reduce the content of heavy high-boiling components in gasoline and reduce the coking rate of the catalyst, reduce potential coke deposition on the inner surfaces of the raw hydrocarbon heater, in an embodiment superheater.

[0105] Moreover, a mixing device for mixing the reaction products of the upstream reaction zone and supplied methanol and / or other oxygenates and / or olefin-containing feedstock may be made as a bed of granules of a neutral material and / or a fraction of crushed quartz placed upstream the front catalyst bed.

[0106] Due to this advantageous characteristic, there appears a possibility to use the protective bed of granules of a neutral material and / or a fraction of crushed quartz placed upstream the front catalyst bed as a device for mixing the reaction products from the upstream reaction zone and supplied methanol and / or other oxygenates and / or olefin-containing feedstock.

[0107] Moreover, a line connecting the reaction zones (reactor spaces in which the reaction zones are located) may be used as a mixing device for mixing the reaction products of the upstream reaction zone and supplied methanol and / or other oxygenates feedstock and / or olefin-containing feedstock.

[0108] Due to this advantageous characteristic, there appears a possibility to use the line connecting the reaction zones as a device for mixing the reaction products of the upstream reaction zone and supplied methanol and / or other oxygenates and / or olefin-containing feedstock.

[0109] In addition, function of a mixing device may be carried out by a protective bed of neutral or weakly-reacting material, for example quartz, in forms such as granules; spheres; fraction of crushed materials.

[0110] In addition, function of a mixing device may be carried out by a channel connecting the reactor spaces where the reaction zones are located.

[0111] In addition, function of a mixing device may be carried out by specifically controlled speed and/or direction of supplied feedstock flow, e.g. by using tangential supply of methanol and / or other oxygenates and / or olefin-containing feedstock to the reaction zone.

[0112] Moreover, some or all the reaction zones may be made as separate adiabatic reactors, which are thermally insulated reaction vessels filled with catalyst and connected in series.

[0113] Moreover, the reaction zones may be located in multi-bed adiabatic reactor.

[0114] Moreover, the reaction zones may be located in in multi-bed isothermal reactor.

[0115] Moreover, the reaction zones may be located in in multi-bed pseudo-isothermal reactor.

[0116] Moreover, upstream the last one or two reaction zones, there may be located raw mixture heaters, e.g., preferably superheaters.

[0117] Due to this advantageous characteristic, there appears a possibility to use the proposed invention at a low flow rate of methanol per metric ton of convertible hydrocarbon feedstock. Indeed, at low flow rates of methanol less than 20 % of weight of the converted feedstock, to enhance RON of converted mixtures the proposed combined embodiment can be used, wherein a reaction mixture heater, e.g., preferably superheater, heating the reaction mixture up to temperatures of about 500 °C is located upstream the last or the penultimate and the last reaction zones. Additionally, there appears a possibility to raise the temperature in the reaction zones as the catalyst is deactivated to improve the efficiency of conversion of hydrocarbon feedstock and oxygenates.

[0118] Moreover, the plant may further contain a conversion product separation unit for removal of the reaction water with an admixture of methanol and / or other oxygenates, whose output is connected to the input of a rectification column, the distillation residue of which is water cleaned from residues of hydrocarbons with a reduced content of oxygenates, and the distillate of which are hydrocarbons and water with an increased content of oxygenates, wherein the distillation column output with the distillate is connected to the input of the unit for supplying methanol and / or other oxygenates and / or olefin-containing feedstock.

[0119] Due to this advantageous characteristic, there appears a possibility to increase the efficiency of gasolines production process due to a more complete processing of all components.

Brief description of the drawings

[0120] Other features and advantages of the invention clearly emerge from the description that is given below for illustration and is non-limiting, with reference to the accompanying drawings, where:

- Figure 1 schematically shows a general view of a plant for producing gasoline from raw hydrocarbon fractions according to the method of the present invention, wherein the unit for supplying flows contains a unit (5) for heating methanol and / or other oxygenates and / or olefin-containing feedstock having at least two outputs (an embodiment, wherein said unit has at least one outlet, is not shown in the drawings);
- Figure 2 shows a horizontal multiple-shell reactor;
- Figure 3 schematically depicts the steps of a process for producing gasolines from raw hydrocarbon fractions according to the invention.

[0121] According to Figure 1, a plant for production of a base stock for production of gasoline from raw hydrocarbon fractions, fractions of gaseous olefins and oxygenates includes a unit 1 for supplying methanol and / or other oxygenates and / or olefin-containing feedstock, and a unit 2 for supplying raw hydrocarbon fractions, whose outputs are connected to inputs of a unit 3 for mixing starting components, the output of which is connected to the input of a unit 4 for heating the feedstock (raw hydrocarbon fractions or mixtures thereof with methanol and / or other oxygenate and / or olefin-containing feedstock), whose output is connected to the inlet of a reactor 6, in which the reaction is carried out in the presence of a zeolite-containing catalyst, whose output is connected to the input of a unit 7 for separating the conversion product having the output 71 for removal of the reaction water and the outlet 72 for flue gases exhaust.

[0122] The reactor 6 is made as a multiple-shell reactor that includes reaction zones on each shelf / catalyst basket or as a horizontal reactor, the housing of which accommodates several series-connected catalyst baskets. Either every reaction zone, or part of the reaction zones may be made as individual thermally insulated vessels.

[0123] The unit 1 for supplying methanol and / or other oxygenates and / or olefin-containing feedstock is connected to the unit 5 for heating the unit for supplying flows. The unit 5 may have at least one or at least two outputs.

[0124] In case of at least one outlet (not shown in the drawings), said output is connected to the input to the second reaction zone 62 either directly or through the mixing zone having means for mixing located between the reaction zones, wherein the unit for supplying also contains a unit 3 for mixing the starting components, whose output is connected to the inlet to the first reaction zone 61 either directly or through the heating unit 4. In the proposed embodiment, the downstream (second) output of the unit 5 may also be connected to the inlet to the first reaction zone 61 (Fig. 1) directly or through the mixing zone.

[0125] In case of at least two outputs (Fig. 1), one of which is connected to the first input of the first reaction zone 61 either directly or through the mixing zone, the second input in which is

connected to the output of the unit 2 for supplying raw hydrocarbon fractions, wherein through the second output of the unit 5 the flow is supplied to the second reaction zone 62.

[0126] The separation unit 7 is connected to gasoline stabilization column 9, wherein the bottom product is stabilized gasoline with improved antiknock rating. Other products of the column are gasoline stabilization gases and reaction water trapped out of the column reflux tank.

[0127] Upstream the reaction zones, the plant further includes a mixing device installed for mixing the reaction products from the upstream reaction zone and supplied methanol and / or other oxygenates and / or olefin-containing feedstock. The mixing device is not shown in Figure 1.

[0128] In one of the embodiments, a mixer for mixing the reaction products from the upstream reaction zones and supplied methanol and / or other oxygenates and / or olefin-containing feedstock is made as a bed of granules of neutral material and / or a fraction of crushed quartz placed upstream the front catalyst bed, and a line connecting the reaction zones is used as a mixing device.

[0129] The plant further may include devices for heating the reaction mixture installed upstream the last or the penultimate and last reaction zones; the diagram of Fig. 1 does not show the heating devices.

[0130] The last one or two reaction zones of the reactor can be made as isothermal reactors.

[0131] The plant further includes a unit 10 for determining the parameters of the resulting gasoline, such as RON, connected to a unit 11 controlling the flow rate of methanol and / or other oxygenates and / or olefin-containing feedstock in the unit 1 for supplying methanol and / or other oxygenates and / or olefin-containing feedstock.

[0132] The plant further includes a reaction water distillation column 8, the bottom product of which is water cleaned of residues of hydrocarbons with a reduced content of methanol and / or other oxygenates, and the distillate of which is water with an increased content of methanol and / or other oxygenates and hydrocarbons.

[0133] The output of the conversion product separation unit for removal of the reaction water with an admixture of methanol can be connected to the input of the rectification column 8, wherein the output of the distillation column with distillate is connected to the input of the unit 1 for supplying methanol and / or other oxygenates and / or olefin-containing feedstock.

Embodiment of the invention

[0134] The gasoline base of raw hydrocarbon fractions, fractions of the gaseous olefins and

oxygenates is produced as follows (see Figures 1, 3).

[0135] Step A1. In the unit 3, starting components are mixed, which are supplied through the unit 1 for supplying methanol and / or other oxygenates and / or olefin-containing feedstock, and the unit 2 for supplying raw hydrocarbon fractions.

[0136] Step A2. The starting feedstock is heated in the unit 4 for heating the starting feedstock. During heating, if necessary, the heavy components are separated.

[0137] Step A3. The mixture is supplied from the unit 4 to the reactor 6, wherein the reaction is carried out in the presence of a zeolite-containing catalyst. The reactor contains at least two reaction zones, where the zeolite-containing catalyst is located. The mixture from step A2 is supplied into the first reaction zone 61. After heating in the unit 5, methanol and / or other oxygenates and / or olefin-containing feedstock are divided into several flows, the first flow is mixed with the converted hydrocarbon feedstock from the first reaction zone 61 in the mixing zone between the first and second reaction zones, and the subsequent flows are mixed in the downstream mixing zones (not shown in the drawings) with the conversion product from the upstream reaction zone.

[0138] A bed of granules of a neutral material and / or a fraction of crushed quartz placed upstream the catalyst bed placed on a shelf, or a line connecting the reaction zones are used as a mixing zone (a mixing device).

[0139] Methanol and / or other oxygenates are supplied in the gaseous phase at a temperature of no more than 380 °C preventing its decomposition upon heating.

[0140] The isothermal reactors are used as the last or the last and the penultimate reaction zones of the reactor. Alternatively, the adiabatic reactors are used as the last or last and the penultimate reaction zones of the reactor.

[0141] Zeolite-containing catalysts are used in the reaction zones, whose binder is a mixture of alumina in an amount of 30-70 wt. % and silica in an amount of 70-30 wt. %.

[0142] Alternatively, the catalyst is used for production of liquid hydrocarbons of dimethyl ether based on crystalline pentasil-type aluminosilicate having a molar ratio of $\text{SiO}_2/\text{Al}_2\text{O}_3 = 25 - 100$ characterized by the presence of residual amounts of sodium ions equivalent to a content of 0.05 - 0.1 % wt. of sodium oxide containing said crystalline aluminosilicate and binder, wherein it further contains cobalt oxide, oxides of rare earth elements and zinc oxide at the following component ratio, wt.%:

Zinc oxide: 0.5 - 3.0

Oxides of rare earth elements: 0.1 - 5.0

Cobalt oxide: 0.05 - 2.5

Crystalline aluminosilicate: 63 - 69.8

Binder: the rest.

Alternatively, the catalyst of light aliphatic hydrocarbon conversion is used characterized in that a mixture of pentasil group zeolites with a different silicate module is used as the HZSM-5 zeolite: 1) zeolite with $\text{SiO}_2/\text{Al}_2\text{O}_3=20$ previously treated with an aqueous solution of alkali, modified with oxides of rare earth elements (REE) in an amount of 0.5-2.0 wt. %, and 2) zeolite with $\text{SiO}_2/\text{Al}_2\text{O}_3=82$ with a residual amount of sodium oxide in an amount of 0.04 wt. % taken in a ratio of 1.7/1 to 2.8 / 1, and the remainder being a binder in an amount of 20 to 30 wt. % of the catalyst.

[0143] Step A4. The flow rate of methanol and / or other oxygenates and / or olefin-containing feedstock into the reaction zones is controlled (by the unit 1 and through picking methanol and / or other oxygenates and / or olefin-containing feedstock into the reaction zone from the unit 5), and the temperature of the mixture supplied to the first reaction zone (by the unit 4) is controlled so that the maximum temperature of the catalyst bed does not exceed 420 °C in the preparation of the base stock for the production of high-octane gasoline and 500 °C for the production of a concentrate alkylaromatics, and the temperature of the end catalyst bed in the reaction zone is respectively by 40-70 °C below the maximum temperature of the bed. If methanol is supplied in an amount of less than 20 to 25 wt. %, additional heat supply to the downstream reaction zones is required. To supply heat, hot stabilization gases from the fractionation unit 9 or heated circulating exhaust gases from the unit 7 can be used, or a heater can be installed on the line connecting 2 last zones, or heat is supplied using the heat pipes, or otherwise, or heat supply can be carried out through heating, e.g., preferably superheating the flow supplied into the last or the last and the penultimate reaction zones, but no more than up to 500 °C, or isothermal reaction zones with heat exchanging heat supply can be used as the last or the last and the penultimate reaction zones.

[0144] Step A5. The temperature of the last or the last and the penultimate reaction zones is decreased during catalyst operation each time by 1-2 °C as the allowable concentration of methanol in the reaction water increases. Figure 1 does not show a schematic drawing of additional heat supply to the last or the last and the penultimate reaction zones to control the temperature in this zone. To supply heat, heated stabilization gases from the fractionation unit 9 or circulating heated exhaust gases from the unit 7 can be used, or a heater can be installed on a line connecting the last 2 zones, or heat is supplied through the heat pipes or using other means.

[0145] Step A6. A 1-2 % aqueous solution of an industrial demulsifier is added to a mixture of hydrocarbons and water supplied to the separation at a ratio of 1:100 - 1:200 to the volume of the reaction water. Here, the aqueous solution of the demulsifier is mixed with the input flow in a laminar static mixer or directly in a condenser of hydrocarbons and reaction water located in the separation unit 6 upstream a three-phase separator.

[0146] Step A7. Gasoline is isolated by the separation of the conversion product while diverting simultaneously the reaction water and exhaust gases.

[0147] In the unit 7, the reaction mixture is cooled with partial condensation of C₅+ hydrocarbons and the aqueous portion, further, the three-phase separator located in the same location separates the mixture into the gaseous phase, hydrocarbon condensate and reaction water. The separated hydrocarbon condensate is supplied to a distillation unit 9, from where gasoline selected by stabilized vapor pressure and dried is picked up, and stabilization gases and reaction water with a high content of methanol are picked up from the reflux tank. The separated reaction water with an admixture of methanol and / or other oxygenates is supplied to the distillation unit 8, the distillation residue of which is water cleaned of residues of hydrocarbons with a reduced content of methanol and / or other oxygenates, and the distillate of which is water with a higher content of methanol and / or other oxygenates and hydrocarbons, wherein water with an increased content of oxygenates and hydrocarbons produced in the rectification process are recycled to the reactor inlet to the beginning of the process in a mixture with methanol and / or other oxygenate and / or olefin-containing feedstock.

[0148] Step A8. If the parameters of the produced gasoline with improved antiknock rating differ from the specified parameters that are measured by the unit 10 for determining the parameters of the resulting gasoline, the flow rate of methanol and / or other oxygenates and / or olefin-containing feedstock to the reactor is increased or decreased by the unit 11 controlling the flow rate of methanol and / or other oxygenates and / or olefin-containing feedstock.

[0149] The sequence of steps is exemplary and allows rearranging, adding or making some operations simultaneously without losing the ability of the production of a base stock for the production of gasolines from raw hydrocarbon fractions, fractions of gaseous olefins and oxygenates.

Industrial applicability

[0150] The proposed method for producing gasolines from raw hydrocarbon fractions can be embodied in practice, and upon embodiment, it provides an implementation of the declared application, which allows making a conclusion that the invention complies with the "industrial applicability" criterion.

[0151] In accordance with the proposed invention, tests and calculations for the production of gasolines from various hydrocarbon fractions, fractions of gaseous olefins and oxygenates have been conducted.

Example No. 1. Modeling the first reaction zone

[0152] A zeolite catalyst was used for the conversion of dimethyl ether after regeneration.

[0153] The process was carried out at atmospheric pressure in a reactor having a useful volume of about 1.5 liters and with thermal insulation made as an air gap between the external surface of the cylindrical reactor shell and an outer protection casing, and with compensation of heat losses using external heating of the protective casing. Previously, the catalyst bed having a volume of 1.2 liters (855 g) was heated to a temperature of 350 °C. A mixture of the following mass composition was supplied on the catalyst bed at a temperature of 335 °C:

- Methanol: 6.6 %
- Water: 7 %,
- Ethylene: 2.9 %, and
- Straight-run hydrocarbon fraction: 65-85 °C, 83.4 %, with the composition given in Table 3.

[0154] The mass feedstock flow rate was 2.6 kg / L of catalyst · h⁻¹. During steady state, the temperature of heating of the adiabatic catalyst bed in a narrow front bed (no more than one quarter of the charged catalyst) was about 60-65 °C (from 335 °C to 395-398 °C)), the end catalyst bed temperature was about 360 °C.

[0155] The mass yield of the C₅+ fraction from the first reaction zone was 97.5 % per the hydrocarbon fraction supplied (65-85 °C). RON for the C₅+ hydrocarbons fraction to be released from the conversion products was 67 units with RON of the initial fraction equal to 58 units. The methanol conversion was 99.9 %.

Example No. 2. Increase in RON of the product by increasing the methanol flow rate

[0156] It differs from the Example No.1 by that the temperature of the raw mixture was reduced down to 320 °C. A mixture of the following mass composition was supplied to the reactor inlet:

- Methanol: 10.7 %
- Water: 3.4 %
- Ethylene 2.6 %, and
- Fraction, 65-85 °C: 83.4 % the rest.

[0157] The mass feedstock flow rate was 2.6 kg / L of catalyst · h⁻¹. During steady state, the temperature of heating of the adiabatic catalyst bed in a narrow front bed (less than one quarter of the charged catalyst) was about 80-85 °C (from 320 °C to 402-406 °C)), the end

catalyst bed temperature was about 365 °C. The mass yield of the C₅+ fraction from the first reaction zone was 96.4 % per the hydrocarbon fraction supplied (65-85 °C). RON for the C₅+ hydrocarbons fraction to be released from the conversion products was 71 units with RON of the initial fraction equal to 58 units. The methanol conversion was 99.5 %.

Example No. 3

[0158] We used the same catalyst as that used in experiment 1. A flow reactor was used, wherein the catalyst was filled for the conversion of light hydrocarbons and oxygenates to aromatics concentrate in three beds in an amount of 18 ml of the catalyst each (a total amount of 40 g of catalyst). The catalyst beds were separated by quartz chip beds in an amount of 36 ml each forming the zones for mixing and heating the hydrocarbon feedstock. A narrow straight-run gasoline fraction at 65 - 85 °C having a density of 0.71 at a flow rate of 60 ml / hr and 80 wt. % methanol at a flow rate of 12 ml / hr were supplied to the first reaction zone.

[0159] Additionally, 80 wt. % methanol was supplied through the capillaries at a rate of 6 ml / hr to the second and third reaction zones through the mixing zones. The ratio of methanol / straight-run fraction was equal to 0.38.

Temperature in the reaction zones: 365 - 380 °C.

Reactor pressure: 0.3 - 0.35 MPa.

[0160] Material balance of experiment as of the 4th hour of the experiment is given in the Table below.

Table 1

Supplied	g / hr
to the first reaction zone	
Fraction 65-85 °C (RON 58)	42.7
Methanol	8.13
Water	2.03
To the second reaction zone additionally	
Methanol	4.065
Water	1.01
To the third reaction zone additionally	
Methanol	4.06
Water	1.015
Total supplied	62.98

Supplied	g / hr
Produced	
Liquid fraction of hydrocarbons (RON 87.5)	41.8
Reaction water	13.21
Hydrocarbon hydrogen-containing gas having a propane content of up to 39 wt. %	7.97
Conversion of methanol	99.7 %
C ₅ + hydrocarbon yield per gasoline supplied	0.975

Example No. 4 (comparative)

[0161] A flow reactor was loaded with 36 ml (26.7 g) of the catalyst, where a narrow straight-run gasoline fraction at 65 - 85 °C having a density of 0.711, at a flow rate of 36 ml / hr and 100 wt. % methanol at a flow rate of 12 ml / hr were supplied.

[0162] The ratio of methanol / straight fraction was equal to 0.371.

[0163] The temperature in the reaction zone was 365 - 380 °C.

[0164] The reactor pressure was 0.3 - 0.35 MPa.

[0165] The material balance as of the 4th hour of the experiment is shown in Table 2.

Table 2

Supplied	g / hr
Fraction at 65-85 °C (RON 58)	25.60
Methanol	9.50
Total supplied	35.10
Produced	
Liquid fraction of hydrocarbons (RON 86.3)	24.79
Reaction water	5.34
Hydrocarbon hydrogen-containing gas having a propane content of up to 42 wt. %	4.97
Conversion of methanol	100 %
C ₅ + hydrocarbon yield per gasoline supplied	0.968

[0166] Table 3 shows a comparison of the compositions in weight percentage of the initial narrow straight-run fraction C₆-C₇, and C₅+ hydrocarbon fraction from the catalysate hydrocarbon condensate. As seen from the comparison of examples when using a reactor with three reaction zones and with the distributed supply of methanol, the benzene and durene content in gasoline reduced, the content of alkylaromatics increased.

Table 3

Component description	Feedstock (fraction at 65-85 °C)	C ₅ + fraction from catalysate hydrocarbon condensate, Example 4 (comparative)	C ₅ + fraction from hydrocarbon condensate, Example 3 (reactor with 3 reaction zones)
C5-C6 olefins		0.20%	0.32%
Sum of C ₆ -C ₈ five-membered naphthenes	9.26%	8.18%	8.40%
Sum of C ₆ -C ₈ six membered naphthenes	5.97%	3.79%	2.82%
Isopentane	0.00%	3.91%	4.22%
Sum of C ₆ isoparaffins with two branches	0.49%	1.03%	1.16%
Sum of C ₇ isoparaffins with two branches	29.33 %	22.18 %	20.51 %
Sum of C ₆ isoparaffins with one side branch	9.34 %	11.26 %	11.53 %
Sum of C ₇ isoparaffins with one side branch	15.69 %	6.04 %	6.38 %
n-pentane		3.44 %	3.78 %
n-hexane	20.93 %	9.71 %	9.52 %
n-heptane	6.57 %	1.66 %	1.51 %
Benzene	1.53 %	1.12 %	0.76 %
Toluene	0.90 %	5.29 %	4.55 %
Xylenes and ethylbenzene		11.31 %	12.69 %
C ₉ aromatics		6.79 %	7.55 %
C ₁₀ aromatics		3.10 %	2.85 %

Component description	Feedstock (fraction at 65-85 °C)	C ₅ + fraction from catalysate hydrocarbon condensate, Example 4 (comparative)	C ₅ + fraction from hydrocarbon condensate, Example 3 (reactor with 3 reaction zones)
A ₁₁ ⁺		1.12 %	1.45 %
Total	100 %	100 %	100 %
Including aromatics	2.4 %	28.70 %	30.10 %
Including sum of tetramethylbenzenes		1.20 %	0.93 %
<u>Of these, durene (1,2,4,5-tetramethylbenzene)</u>		<u>0.46 %</u>	<u>0.30 %</u>

Example No. 5

[0167] This example demonstrates the possibility of using feedstock containing diene and triene hydrocarbons, as well as the ability to control the degree of conversion of feedstock by increasing the maximum process temperature.

[0168] A catalyst of conversion of light aliphatic hydrocarbons was used.

[0169] The feedstock was a mixture consisting of a mixture of SNG (stable natural gasoline) in an amount of 50 vol. % and a light fraction (initial boiling point: 120 °C) of catalytic cracking gasoline from Ufa Refinery. The total sulfur content in the mixture fraction was 0.008 wt. %. The content of diene and triene hydrocarbons was 0.3 wt. Into a flow reactor, 100 ml. (74 g) of fresh catalyst was loaded. During the experiment, to maintain RON at a level of about 92 units the volumetric feed-stock flow rate was reduced by 20 %, and the maximum temperature in the reaction zone was increased from 380 to 390 °C.

[0170] Table 4 shows the material balance of the experiment corresponding to the 203rd hour of the pilot plant operation. Diene and triene hydrocarbons were not detected in the produced gasoline. The total sulfur content was reduced to 0.001 wt. %. The methanol conversion during the experiment was no less than 99 % and was adjusted by increasing the temperature in the reaction zone and reducing the feedstock volumetric flow rate.

Table 4

Supplied	g / hr
Mixture of 50 % of SNG and light fraction catalytic cracking gasoline (initial boiling point: 120 °C).	50.47
Methanol	23.73

Supplied	g / hr
Total supplied	74.20
Produced	
Liquid fraction of hydrocarbons (RON 92)	44.90
Reaction water	13.39
Hydrocarbon hydrogen-containing gas having a propane content of up to 55 wt. %	15.9
Methanol conversion	100 %

Example No. 6 (calculated)

[0171] The content of methanol in the reaction water is 1.3 wt. %.

[0172] The reaction water is supplied into an evaporator with a vapor space where it is evaporated at a pressure close to atmospheric pressure. Vapors of the reaction water are supplied to a column still. The number of fractionating trays in the column is ten. Vapors from the top of the column are condensed and supplied partially as reflux to the column for reflux. The yield of distillate is 11 wt. % of the weight of the starting feedstock supplied to the column as vapor. The concentration of methanol in the distillate is 8-9 wt. %. This product is admixed to 100 wt. % methanol supplied for conversion together with the hydrocarbon feedstock. As a result, methanol at a concentration of 94 wt. % is supplied for conversion. The bottom product of the column is desalted and water cleaned of hydrocarbons, with the methanol content of no more than 0.2 wt. %.

[0173] Thus, as a result of measurement and calculations, it has been found that the separation of methanol and / or other oxygenates and / or olefin-containing feed-stock into multiple flows according to the method of the present invention with simultaneous utilization of a reactor comprising at least two reaction zones, allows increasing the useful reactor volume, simplify its design, and besides reduce gassing, reduce the consumption of methanol per unit of increment of RON of the base stock for the production of gasolines.

[0174] Accordingly, the present invention achieves the task set, i.e. increasing the quality of the produced high-octane gasoline due to the reduction of the content of benzene in gasoline, as well as the content of high melting point components such as durene, as well as due to the possibility of regulating the temperature throughout the catalyst bed without using complicated heat-exchange reactor equipment, as well as the possibility of increasing the useful volume of reactor equipment.

[0175] An additional advantage of the proposed solution is that:

- There is a low content of $C_1 - C_2$ hydrocarbons in exhaust gases, and since hydrogen is substantially insoluble in liquid hydrocarbons, this allows producing hydrogen-containing gas by using the C_3+ hydrocarbon absorption from exhaust gases with debutanized gasoline from the discussed plant.
- Instead of a factory gas fractioning plant, an absorption column can be used, into which gasoline is supplied that has been produced in the plant and cooled down to a temperature range of $-10\text{ }^{\circ}\text{C}$ to $+15\text{ }^{\circ}\text{C}$. Saturated gasoline is supplied from this column to the stabilization column (unit 9). This increases the yield of commercial propane-butane mixture (C_3-C_4) and allows producing hydrogen-containing gas with a low hydrocarbon content.
- As feedstock for the production of gasoline, instead of straight-run gasoline or in conjunction with it, there can be used side-cut fractions from various processes, including fractions containing benzene and so on, fractions containing diene hydrocarbons, including light catalytic cracking gasolines.
- Sulfur content in co-processed gasoline is 5-20 times lower than that in the original gasoline. This is because the catalyst runs on feedstock having a sulfur content in convertible gasoline of up to 500 ppm by weight. Sulfur is mainly trapped out as hydrogen sulfide along with exhaust gases and is partly sorbed on the catalyst and is removed during regeneration.

[0176] Gasolines produced in the processing by their basic characteristics (group composition, RON, sulfur content, etc.) comply completely with the fifth grade gasoline characteristics in terms of technical regulations under the Customs Union (CU TR 013/2011) "On requirements for automobile and aviation gasoline, diesel and marine fuel, jet fuel and heating oil".

[0177] The various embodiments of feedstocks, materials, processes, methods, equipment, and systems, set forth in this specification may be used may be used with each other in different and various combinations. Thus, for example, the configurations provided in the various embodiments of this specification may be used with each other; and the scope of protection afforded the present inventions should not be limited to a particular embodiment, configuration or arrangement that is set forth in a particular embodiment, example, or in an embodiment in a particular Figure.

REFERENCES CITED IN THE DESCRIPTION

Cited references

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Patentkrav

- 1.** Fremgangsmåde til fremstilling af benzin fra uforarbejdede carbonhydridfraktioner, fraktioner af gasformige olefiner og oxygenater, hvor råmaterialekomponentstrømmen leveres til enhed, som leverer strømmene, som skal
- 5 bearbejdes, til en reaktor, hvor reaktionen udføres i tilstedeværelse af en zeolitholdig katalysator, benzin med højt oktantal isoleres ved separering af omdannelsesproduktet, mens reaktionsvandet og spildgasserne samtidigt bortledes, **kendetegnet ved, at**
- 10 (a) som reaktor anvendes en reaktor, som indeholder mindst to reaktionszoner med en zeolitholdig katalysator,
- (b) mellem hvilke der er yderligere indrettet organ til blanding af reaktionsprodukterne fra den tidligere reaktionszone og det leverede methanol og/eller andre oxygenater og olefinholdigt råmateriale, og
- 15 (c) ved hjælp af enheden til levering af strømme separeres methanol og/eller andre oxygenater og olefinholdigt råmateriale i mindst to strømme, hvor
- en første strøm af methanol og/eller andre oxygenater og olefinholdigt råmateriale ledes til et rum opstrøms for den første reaktionszone, hvor den blandes med strømmen af uforarbejdede carbonhydridfraktioner og
- 20 leveres til den første reaktionszone af reaktoren, og
- en anden strøm af methanol og/eller andre oxygenater og olefinholdigt råmateriale blandes i nedstrømsreaktionszonerne med omdannelsesproduktet fra opstrømsreaktionszonen.
- 25 **2.** Fremgangsmåden ifølge krav 1 **kendetegnet ved, at** strømningsraten af methanol og/eller andre oxygenater og olefinholdigt råmateriale til hver reaktionszone styres, og temperaturen af råmaterialet, som skal leveres til reaktionszonen, styres, således at den maksimale katalysatorlejetemperatur i reaktionszonen ikke overstiger 420°C i fremstillingen af et grundstof til
- 30 fremstillingen af benzin, samtidig med reduktion af leveringen af methanol og/eller andre oxygenater og/eller olefinholdigt råmateriale til den første reaktionszone øges temperaturen af den uforarbejdede blanding, som leveres til den første reaktionszone, og vice versa.

3. Fremgangsmåden ifølge krav 1 **kendetegnet ved, at** methanol og/eller andre oxygenater og olefinholdigt råmateriale separeres ved hjælp af nævnte enhed til mindst en strøm, som ledes til nedstrømsreaktionszonen, hvor nævnte strøm blandes med omdannelsesproduktet fra opstrømsreaktionszonen, hvor den første reaktionszone **for-forsynes** med en forblandet strøm af methanol og/eller andre oxygenater og olefinholdigt råmateriale og strømmen af uforarbejdede carbonhydridfraktioner.
4. Fremgangsmåden ifølge krav 3 **kendetegnet ved, at** strømmen ledes fra nævnte enhed til reaktionszonen direkte eller gennem en blandingszone placeret mellem reaktionszonerne og med blandingsorgan.
5. Fremgangsmåden ifølge krav 3 **kendetegnet ved, at** strømmen af methanol og/eller andre oxygenater og olefinholdigt råmateriale blandes med de omdannelige carbonhydridfraktioner i rummet opstrøms for den første reaktionszone eller længere opstrøms i retningen af carbonhydridråmaterialeleveringsstrømmen.
6. Fremgangsmåden ifølge krav 3 **kendetegnet ved, at** forud for levering til den første reaktionszone forvarmes den blandede strøm af methanol og/eller andre oxygenater og olefinholdigt råmateriale og strømmen af uforarbejdede carbonhydridfraktioner.
7. Fremgangsmåden ifølge kravene 1 og 3 **kendetegnet ved, at** forud for separeringen af methanol og/eller andre oxygenater og olefinholdigt råmateriale i flere strømme, opvarmes en uforarbejdet carbonhydridfraktion eller en blanding af en uforarbejdet carbonhydridfraktion og methanol og/eller andre oxygenater og olefinholdigt råmateriale i to trin: i det første trin fordampes en uforarbejdet carbonhydridfraktion eller en blanding af en uforarbejdet carbonhydridfraktion og methanol og/eller andre oxygenater og olefinholdigt råmateriale, tungere ikke-fordampede komponenter separeres, og i det andet trin overophedes fordampede komponenter.
8. Fremgangsmåden ifølge krav 4 **kendetegnet ved, at** et leje af neutrale materialegranulater og/eller en fraktion af knust kvarts, som er placeret opstrøms

for det forreste katalysatorleje, eller en ledning, som forbinder reaktorrummene, hvor reaktionszonerne er placeret, anvendes som blandeorganer.

5 **9.** Fremgangsmåden ifølge kravene 1 og 3 **kendetegnet ved, at** methanol og/eller andre oxygenater leveres gennem nævnte enhed i den gasformige fase ved en temperatur på ikke mere end 380°C, hvorved nedbrydning deraf forhindres ved opvarmning.

10 **10.** Fremgangsmåden ifølge krav 1 **kendetegnet ved, at** en 1-2% vandig opløsning af et industrielt demulgeringsmiddel tilsættes til en blanding af carbonhydrider og vand leveret til separering i et forhold på 1:50 - 1:200 af volumenet af reaktionsvandet, den vandige opløsning af demulgeringsmidlet blandes med tilførselsstrømmen i en laminar statisk blander eller direkte i en carbonhydrid- og reaktionsvandskondensator placeret i separeringsenheden.

15

11. Fremgangsmåden ifølge krav 1 **kendetegnet ved, at** isotermeriske reaktionszoner anvendes som den sidste ene eller de sidste to reaktionszoner af reaktoren.

20 **12.** Fremgangsmåden ifølge krav 1 **kendetegnet ved, at** adiabatisk reaktionszoner, som ikke indeholder varmevekslerindretninger, anvendes som den sidste ene eller de sidste to reaktionszoner af reaktoren, samtidigt overophedes råmaterialestrømme, som skal leveres til de sidste og/eller næstsidste reaktionszoner, yderligere ikke mere end til 500°C.

25 **13.** Fremgangsmåden ifølge krav 1 **kendetegnet ved, at** separeret reaktionsvand med en blanding af methanol og/eller andre oxygenater leveres til en destillationssøjle, hvis destillationsrest er vand rensat for rester af carbonhydrider med et reduceret indhold af methanol, og hvis destillat er vand med et forøget indhold af oxygenater og carbonhydrider, mens methanolvand og
30 carbonhydrider fremstillet under destillationsprocessen genanvendes i reaktorindløbet i begyndelsen af processen i en blanding med methanol og/eller andre oxygenater og olefinholdigt råmateriale.

14. Fremgangsmåden ifølge krav 1 **kendetegnet ved, at** der anvendes en
35 zeolitholdig katalysator, hvor et syntetisk amorft aluminiumsilicat eller en

uorganisk oxid anvendes som en bindemiddelkomponent, hvilken er en blanding af aluminiumoxid i en mængde på 30-70 vægtprocent og siliciumdioxid i en mængde på 70-30 vægtprocent.

- 5 **15.** Fremgangsmåden ifølge krav 1 **kendetegnet ved, at** der anvendes en katalysator til fremstilling af flydende carbonhydrider af dimethylether baseret på krystallinsk aluminiumsilicat af pentasil-typen med et molforhold på $\text{SiO}_2/\text{Al}_2\text{O}_3 = 25 - 100$ **kendetegnet ved** tilstedeværelsen af restmængder af natriumioner, som svarer til et indhold på 0,05 – 0,1 vægtprocent natriumoxid, indeholdende
- 10 nævnte krystallinske aluminiumsilicat og bindemiddel, hvor den yderligere indeholder kobaltoxid, oxider af sjældne jordarter og zinkoxid i følgende forhold i vægtprocent:
- zinkoxid: 0,5 – 3,0
oxider af sjældne jordarter: 0,1 – 5,0
- 15 kobaltoxid: 0,05 – 2,5
krystallinsk aluminiumsilicat: 63 – 69,8
bindemiddel: resten.

- 16.** Fremgangsmåden ifølge krav 1 **kendetegnet ved, at** der anvendes en
- 20 katalysator af aromatiseringen af C_3 - C_4 -gasser, carbonhydridfraktioner med lavt oktantal og alifatiske alkoholer og blandinger deraf, baseret på zeolitter fra pentasilgruppen, som indeholder en mekanisk blanding af to zeolitter, som har et forskelligt silikatmodul:
- a. zeolit med en $\text{SiO}_2/\text{Al}_2\text{O}_3=20$ tidligere behandlet med en vandig
- 25 alkaliopløsning modificeret med oxider af sjældne jordarter (REE) i en mængde på 0,5-2,0 vægtprocent, og
- b. zeolit med $\text{SiO}_2/\text{Al}_2\text{O}_3=82$ med en restmængde af natriumoxid på 0,04 vægtprocent i forholdet på 1,7/1 til 2,8/1, og idet den resterende del er et bindemiddel i en mængde på 20 til 30 vægtprocent af katalysatorvægten.

30

- 17.** Fremgangsmåden ifølge krav 1 **kendetegnet ved, at** olefinholdig gas er valgt fra gruppen omfattende ethylenholdig gas fra anlæg til fraktionering af spildgas fra den katalytiske krakningsproces, gas fra den forsinkede koksdannelsesproces.

35

DRAWINGS

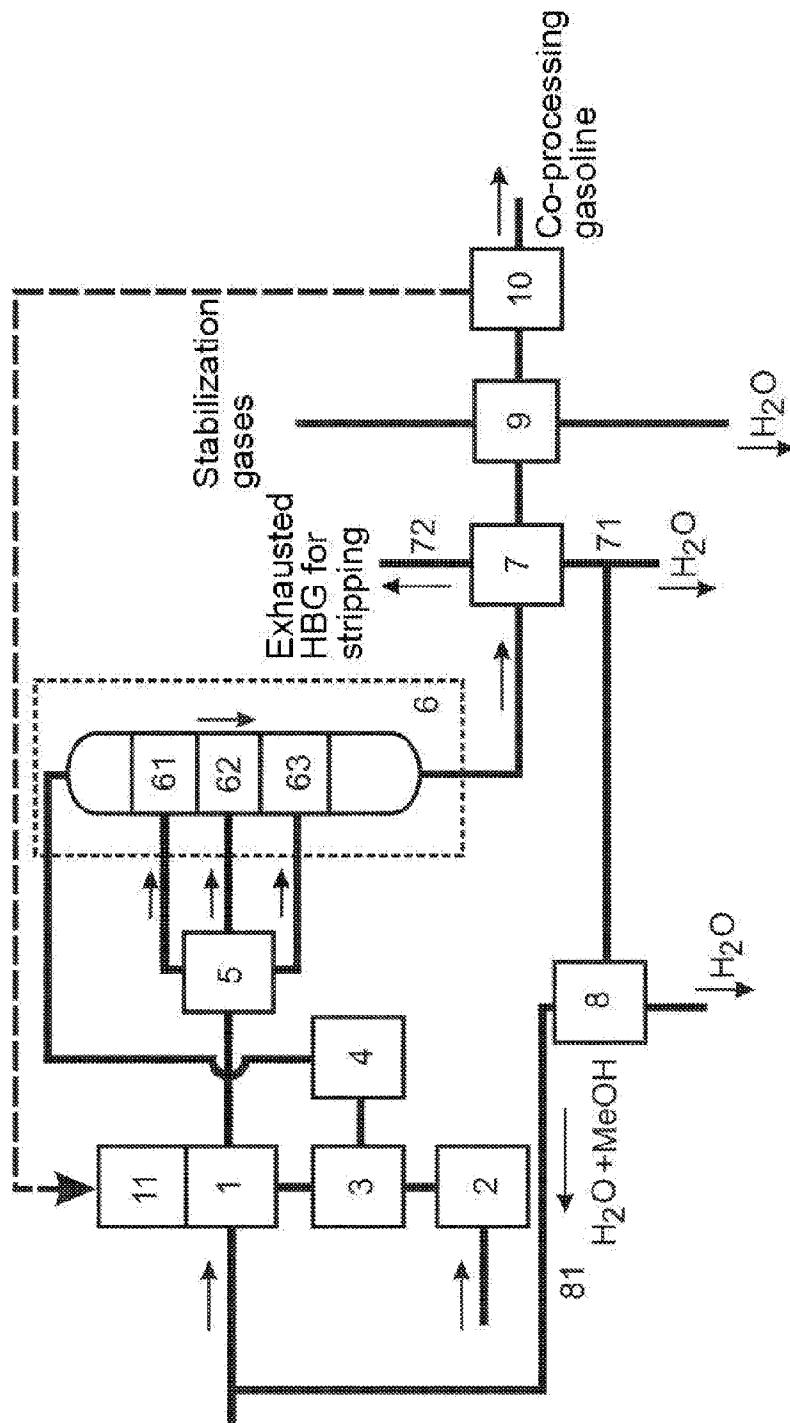


Fig. 1

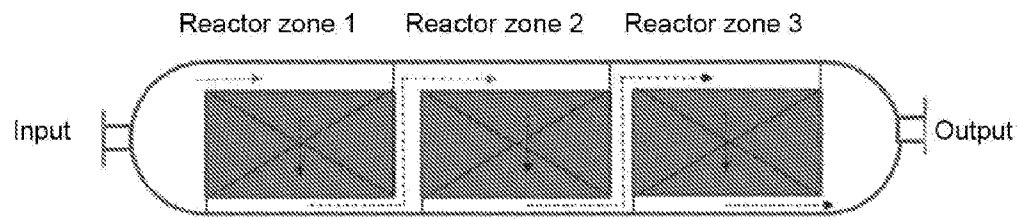


Fig. 2

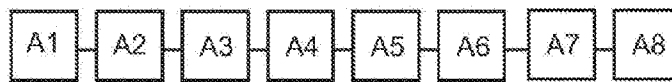


Fig. 3