

DEVICE AND METHOD FOR PREPARING AROMATIC HYDROCARBONS FROM NAPHTHA

TECHNICAL FIELD

The present application relates to a fluidized bed device and a method of using the device, belongs to the technical field of chemical industry, and particularly relates to a device and method for preparing aromatic hydrocarbons from naphtha.

BACKGROUND

Aromatic hydrocarbons (benzene, toluene and xylene, collectively referred to as BTX) are important organic chemical raw materials, wherein p-xylene (PX) is the most concerned product of the aromatic hydrocarbons. PX is mainly used for producing polyesters such as terephthalic acid (PTA), polyethylene terephthalate (PET), polybutylene terephthalate (PBT) and polytrimethylene terephthalate (PTT). In recent years, a yield and a consumption of the p-xylene in China have been increasing continuously. In 2021, a total import volume of the PX in China was about 13.65 million tons, and a foreign-trade dependence degree was about 38%.

Naphtha catalytic reforming technology is a main technical route for producing aromatic hydrocarbons. The composition of naphtha is very complex, and naphtha is not only a main raw material for catalytic reforming, but also a main raw material for ethylene preparation by cracking, and the composition plays an important role in economic benefits of a device. Generally speaking, a high potential content and a moderate distillation range of the raw material aromatic hydrocarbon are beneficial for catalytic reforming. However, high contents of linear-chain and branched-chain aliphatic hydrocarbons and low contents of naphthenic hydrocarbons and aromatic hydrocarbons are suitable for ethylene preparation by cracking. Usually, in order to make full use of naphtha resources and improve economic benefits, it is necessary to separate the linear-chain and branched-chain aliphatic hydrocarbons in the naphtha from the naphthenic hydrocarbons and the aromatic hydrocarbons first, the former are used as raw materials for ethylene production, and the latter are used as raw materials for a catalytic reforming device.

The distillation range of the fractions of naphtha is wide, it is difficult to efficiently separate the linear-chain and branched-chain aliphatic hydrocarbons from the naphthenic hydrocarbons and the aromatic hydrocarbons by a general separation method, in addition, it is difficult to convert the

linear-chain and branched-chain aliphatic hydrocarbons into the aromatic hydrocarbons by the catalytic reforming technology. The naphtha raw material used for catalytic reforming generally needs to be distilled to separate topped oil with a boiling point below 60°C, so as to increase a potential content of the raw material aromatic hydrocarbon for catalytic reforming. However, a fraction with a boiling point above 60°C still contains a large number of linear-chain and branched-chain aliphatic hydrocarbons difficult to be converted into the aromatic hydrocarbons. Therefore, converting the linear-chain and branched-chain aliphatic hydrocarbons into the aromatic hydrocarbons in a highly selective mode has always been a hot and difficult point in the development of a technology for preparing the aromatic hydrocarbons with the naphtha.

Due to the limitation of thermodynamic equilibrium, p-xylene in a xylene mixture produced by the naphtha catalytic reforming device only accounts for ~24%, and it is necessary to further increase a yield of the p-xylene through an isomerization-separation technology. Therefore, increasing the content of the p-xylene in the xylene mixture is an important way to reduce energy consumption for the production of the p-xylene.

SUMMARY

In one aspect of the present application, a device for preparing aromatic hydrocarbons from naphtha is provided, and the device can achieve the preparation of the aromatic hydrocarbons from naphtha with a low potential content of aromatic hydrocarbons, increase the content of p-xylene in mixed xylene, and reduce energy consumption for production.

The naphtha in the present application comprises components of C₄-C₁₂ linear-chain and branched-chain aliphatic hydrocarbons, naphthenic hydrocarbons and aromatic hydrocarbons.

The aromatic hydrocarbons in the present application refer to benzene, toluene and xylene, collectively called BTX.

The device for preparing the aromatic hydrocarbons from naphtha comprises a fluidized bed reactor and a riser reactor; wherein, an outlet of the riser reactor is connected to the fluidized bed reactor; and

the fluidized bed reactor is used for introducing a naphtha raw material to make contact with a catalyst from the riser reactor and react to generate a BTX-containing product gas stream and a spent catalyst, the product gas stream is subjected to gas-solid separation, the product gas stream after

separation is sent to a downstream working section, and unconverted naphtha after separation returns to the fluidized bed reactor as a raw material; and a part of low-carbon alkanes after separation return to the riser reactor as raw materials.

Preferably, the riser reactor is used for introducing a riser reactor raw material and a catalyst to react to generate aromatic hydrocarbons, and a stream containing the unreacted riser reactor raw material, aromatic hydrocarbons and the catalyst enters the fluidized bed reactor through the outlet of the riser reactor.

Preferably, the raw material comprises water vapor and the low-carbon alkane separated from the product gas stream.

Preferably, a water vapor content in the riser reactor raw material is 0wt% to 50wt%.

Preferably, an inlet of the riser reactor is connected with a fluidized bed regenerator, and the catalyst introduced into the riser reactor is a regenerated catalyst generated by the fluidized bed regenerator.

Preferably, the fluidized bed regenerator sequentially passes through a regenerator stripper and a regenerated slide valve to be connected to the inlet of the riser reactor through a pipeline.

Preferably, an inlet of the regenerator stripper extends into a regenerator shell of the fluidized bed regenerator and is located above a regenerator distributor.

Preferably, the fluidized bed reactor comprises a reactor shell, an area enclosed by the reactor shell is divided into a first gas-solid separation zone and a reaction zone from top to bottom, and a gas-solid separation device and a reactor gas collection chamber are provided in the first gas-solid separation zone; the reactor gas collection chamber is located on an inner top portion of the reactor shell, an inlet of the reactor gas collection chamber is communicated with a gas outlet of the reactor gas-solid separation device, and an outlet of the reactor gas collection chamber is communicated with a product gas conveying pipe; and a reactor distributor is provided on a lower portion of the reaction zone for introducing the naphtha raw material.

Preferably, the reactor gas-solid separation device is one or more groups of gas-solid cyclone separators, and each group of gas-solid cyclone separators comprises a first-stage gas-solid cyclone separator and a second-stage gas-solid cyclone separator.

Preferably, the device further comprises a fluidized bed regenerator connected with the fluidized bed reactor, the fluidized bed regenerator is used for introducing regeneration gas to convert the

spent catalyst into a regenerated catalyst.

Preferably, the fluidized bed reactor sequentially passes through a reactor stripper, a spent slide valve and a spent agent conveying pipe to be connected with the fluidized bed regenerator; wherein, an inlet of the reactor stripper extends into a fluidized bed reactor shell and is located below a catalyst outlet end of the reactor gas-solid separation device.

Preferably, the fluidized bed regenerator comprises a regenerator shell, and an area enclosed by the regenerator shell is divided into a second gas-solid separation zone and a regeneration zone from top to bottom; a regenerator gas-solid separation device and a regenerator gas collection chamber are arranged in the second gas-solid separation zone; the regenerator gas collection chamber is located on an inner top portion of the regenerator shell, and a flue gas conveying pipe is provided on the regenerator gas collection chamber; a gas outlet of the regenerator gas-solid separation device is communicated with the regenerator gas collection chamber; and a regenerator distributor is provided on an inner lower portion of the regeneration zone for introducing regeneration gas.

Preferably, the regenerator gas-solid separation device is one or more groups of gas-solid cyclone separators, and each group of gas-solid cyclone separators comprises a first-stage gas-solid cyclone separator and a second-stage gas-solid cyclone separator.

In another aspect of the present application, a method for preparing aromatic hydrocarbons from naphtha is provided, wherein the method comprises: preparing aromatic hydrocarbons by using the device for preparing aromatic hydrocarbons from naphtha above and a catalyst.

Preferably, the catalyst is a metal molecular sieve bifunctional catalyst.

Preferably, the metal molecular sieve bifunctional catalyst is a metal modified HZSM-5 zeolite molecular sieve;

The metal for the metal modification is selected from at least one of La, Zn, Ga, Fe, Mo and Cr; and

a method for the metal modification comprises: placing the HZSM-5 zeolite molecular sieve in a metal salt solution, soaking, drying and roasting to obtain the metal modified HZSM-5 zeolite molecular sieve.

Preferably, the method comprises: allowing the naphtha to enter the reaction zone of the fluidized bed reactor through the reactor distributor to make contact with the catalyst from the riser reactor to generate a product gas stream containing BTX, a low-carbon olefin, hydrogen, a

low-carbon alkane, combustible gas, a heavy aromatic hydrocarbon and unconverted naphtha, and coking and converting the catalyst into a spent catalyst at the same time; and

allowing the product gas stream to enter the reactor gas-solid separation device to remove the spent catalyst contained in the product gas stream, then allowing the product gas stream to enter the reactor gas collection chamber, and enter the downstream working section through the product gas conveying pipe.

Preferably, the unconverted naphtha after separation returns to the fluidized bed reactor as a raw material.

Preferably, a part of low-carbon alkanes after separation return to the riser reactor as raw materials.

Preferably, the BTX refers to benzene, toluene and xylene;

the low-carbon olefin refers to ethylene and propylene;

the low-carbon alkane refers to ethane and propane;

the combustible gas comprises methane, CO, and the like; and

the heavy aromatic hydrocarbons refer to aromatic hydrocarbons with a number of carbon atoms in a molecule greater than or equal to 9.

Preferably, the naphtha is selected from at least one of direct coal liquefaction naphtha, indirect coal liquefaction naphtha, straight-run naphtha and hydrocracking naphtha.

Preferably, the naphtha also contains unconverted naphtha separated from the product gas stream, and the unconverted naphtha comprises main components of C₄-C₁₂ linear-chain and branched-chain aliphatic hydrocarbons and a naphthenic hydrocarbon.

Preferably, a carbon content in the spent catalyst is 1.0wt% to 3.0wt%.

Preferably, process conditions of the reaction zone are: an apparent linear velocity of gas of 0.5 m/s to 2.0 m/s, a reaction temperature of 500°C to 650°C, a reaction pressure of 100 kPa to 500 kPa, and a bed density of 150 kg/m³ to 700 kg/m³.

Optionally, the apparent linear velocity of gas of the reaction zone is independently selected from any value or a value in a range between any two of 0.5 m/s, 0.6 m/s, 0.7 m/s, 0.8 m/s, 0.9 m/s, 1.0 m/s, 1.1 m/s, 1.2 m/s, 1.3 m/s, 1.4 m/s, 1.5 m/s, 1.6 m/s, 1.7 m/s, 1.8 m/s, 1.9 m/s and 2.0 m/s.

Optionally, the reaction temperature of the reaction zone is independently selected from any value or a value in a range between any two of 500°C, 510°C, 520°C, 530°C, 540°C, 550°C, 560°C,

570°C, 580°C, 590°C, 600°C, 610°C, 620°C, 630°C, 640°C and 650°C.

Optionally, the reaction pressure of the reaction zone is independently selected from any value or a value in a range between any two of 100 kPa, 125 kPa, 150 kPa, 175 kPa, 200 kPa, 225 kPa, 250 kPa, 275 kPa, 300 kPa, 325 kPa, 350 kPa, 375 kPa, 400 kPa, 425 kPa, 450 kPa, 475 kPa and 500 kPa.

Optionally, the bed density of the reaction zone is independently selected from any value or a value in a range between any two of 150 kg/m³, 200 kg/m³, 250 kg/m³, 300 kg/m³, 350 kg/m³, 400 kg/m³, 450 kg/m³, 500 kg/m³, 550 kg/m³, 600 kg/m³, 650 kg/m³ and 700 kg/m³.

Preferably, the method further comprises: introducing the riser reactor raw material and the catalyst into the riser reactor to react to generate aromatic hydrocarbons; and

allowing a stream containing the unreacted riser reactor raw material, aromatic hydrocarbons and the catalyst to enter the fluidized bed reactor from the outlet of the riser reactor.

Preferably, the catalyst is the regenerated catalyst from the fluidized bed regenerator.

Preferably, the regenerated catalyst sequentially passes through the regenerator stripper and the regenerated slide valve to enter the riser reactor.

Preferably, a carbon content in the regenerated catalyst is less than or equal to 0.5wt%.

Preferably, the riser reactor raw material comprises water vapor and the low-carbon alkane separated from the product gas stream.

Preferably, a water vapor content in the riser reactor raw material is 0wt% to 50wt%.

Preferably, process conditions of the riser reactor are: an apparent linear velocity of gas of 3.0 m/s to 10.0 m/s, a temperature of 580°C to 700°C, a pressure of 100 kPa to 500 kPa, and a bed density of 50 kg/m³ to 150 kg/m³.

Optionally, the apparent linear velocity of gas is independently selected from any value or a value in a range between any two of 3.0 m/s, 3.5 m/s, 4.0 m/s, 4.5 m/s, 5.0 m/s, 5.5 m/s, 6.0 m/s, 6.5 m/s, 7.0 m/s, 7.5 m/s, 8.0 m/s, 8.5 m/s, 9.0 m/s, 9.5 m/s and 10.0 m/s.

Optionally, the temperature is independently selected from any value or a value in a range between any two of 580°C, 590°C, 600°C, 610°C, 620°C, 630°C, 640°C, 650°C, 660°C, 670°C, 680°C, 690°C and 700°C.

Optionally, the pressure is independently selected from any value or a value in a range between any two of 100 kPa, 125 kPa, 150 kPa, 175 kPa, 200 kPa, 225 kPa, 250 kPa, 275 kPa, 300 kPa, 325

kPa, 350 kPa, 375 kPa, 400 kPa, 425 kPa, 450 kPa, 475 kPa and 500 kPa.

Optionally, the bed density is independently selected from any value or a value in a range between any two of 50 kg/m³, 60 kg/m³, 70 kg/m³, 80 kg/m³, 90 kg/m³, 100 kg/m³, 110 kg/m³, 120 kg/m³, 130 kg/m³, 140 kg/m³ and 150kg/m³.

5 Preferably, the method further comprises: allowing the spent catalyst to enter the reactor stripper from an opening end of an inlet pipe of the reactor stripper, and after the spent catalyst is stripped by the reactor stripper, allowing the spent catalyst to enter a downstream area through the spent slide valve and the spent agent conveying pipe.

Preferably, the downstream area is the fluidized bed regenerator.

10 Preferably, the regeneration gas is introduced into the regeneration zone of the fluidized bed regenerator through the regenerator distributor to make contact with the spent catalyst from the fluidized bed reactor, coke on the spent catalyst reacts with the regeneration gas to generate the flue gas, and the spent catalyst is converted into the regenerated catalyst at the same time.

15 Preferably, the spent catalyst sequentially passes through the reactor stripper, the spent slide valve and the spent agent conveying pipe to enter the fluidized bed regenerator, and makes contact and reacts with the regeneration gas to obtain the flue gas and the regenerated catalyst; and

the flue gas enters the regenerator gas-solid separation device to remove a regenerated catalyst contained in the flue gas, then enters the regenerator gas collection chamber, and enters the downstream working section through the flue gas conveying pipe.

20 Preferably, the regeneration gas is selected from at least one of oxygen, air and oxygen-enriched air.

Optionally, process conditions of the regeneration zone are: an apparent linear velocity of gas of 0.5 m/s to 2.0 m/s, a regeneration temperature of 600°C to 750°C, a regeneration pressure of 100 kPa to 500 kPa, and a bed density of 150 kg/m³ to 700 kg/m³.

25 Optionally, the apparent linear velocity of gas is independently selected from any value or a value in a range between any two of 0.5 m/s, 0.6 m/s, 0.7 m/s, 0.8 m/s, 0.9 m/s, 1.0 m/s, 1.1 m/s, 1.2 m/s, 1.3 m/s, 1.4 m/s, 1.5 m/s, 1.6 m/s, 1.7 m/s, 1.8 m/s, 1.9 m/s and 2.0 m/s.

30 Optionally, the regeneration temperature is independently selected from any value or a value in a range between any two of 600°C, 615°C, 630°C, 645°C, 660°C, 675°C, 690°C, 705°C, 720°C, 735°C and 750°C.

Optionally, the regeneration pressure is independently selected from any value or a value in a range between any two of 100 kPa, 125 kPa, 150 kPa, 175 kPa, 200 kPa, 225 kPa, 250 kPa, 275 kPa, 300 kPa, 325 kPa, 350 kPa, 375 kPa, 400 kPa, 425 kPa, 450 kPa, 475 kPa and 500 kPa.

5 Optionally, the bed density is independently selected from any value or a value in a range between any two of 150 kg/m³, 200 kg/m³, 250 kg/m³, 300 kg/m³, 350 kg/m³, 400 kg/m³, 450 kg/m³, 500 kg/m³, 550 kg/m³, 600 kg/m³, 650 kg/m³ and 700 kg/m³.

10 In the present application, a potential content of the aromatic hydrocarbons in the naphtha raw material is 0wt% to 80wt%, and a per-pass conversion rate of the naphtha is 70wt% to 95wt%. By using the device for preparing aromatic hydrocarbons from naphtha and the method for preparing aromatic hydrocarbons from naphtha based on the device, the finally obtained product is composed of: 60wt% to 75wt% BTX, 7wt% to 15wt% low-carbon olefin, 3wt% to 8wt% hydrogen, 2wt% to 7wt% low-carbon alkane, 4wt% to 6wt% combustible gas, 3wt% to 7wt% heavy aromatic hydrocarbon and 0.5wt% to 1wt% coke. The content of the p-xylene in the mixed xylene in the product is 50wt% to 65wt%.

15 The present application can have the beneficial effects as follows.

1) According to the present application, the linear-chain and branched-chain aliphatic hydrocarbons can be efficiently converted into aromatic hydrocarbons in a highly selective mode, the raw materials have a wide application range, and aromatic hydrocarbons may be prepared from the naphtha with a low potential content of aromatic hydrocarbons.

20 2) According to the present application, the aromatization of the low-carbon alkane is achieved by the riser reactor and the metal molecular sieve bifunctional catalyst, so that a yield of aromatic hydrocarbons of a technology for preparing aromatic hydrocarbons from naphtha is greatly increased.

25 3) In the aromatic hydrocarbon products produced by the present application, the content of the p-xylene in the xylene mixture is greater than 50wt%, which is much higher than a thermodynamic equilibrium content (about 24wt%), so that a yield of p-xylene can be effectively increased, and energy consumption for separation of the p-xylene can be greatly reduced.

30 4) The device for preparing aromatic hydrocarbons from naphtha of the present application comprises the fluidized bed reactor and the riser reactor. Because the low-carbon alkane is very stable and needs a high reaction temperature, in the device for preparing aromatic hydrocarbons from

naphtha in the present application, the high-temperature regenerated catalyst enters the riser reactor first to make contact with the low-carbon alkane, and the low-carbon alkane is subjected to an aromatization reaction under an action of the catalyst, so that the yield of aromatic hydrocarbons is increased; and then, the catalyst with reduced temperature is introduced into the fluidized bed reactor to make contact with the naphtha, which eliminates a local high-temperature zone in the fluidized bed reactor, thus effectively reducing the yield of low-carbon alkane and increasing the yield of aromatic hydrocarbon. According to the device for preparing aromatic hydrocarbons from naphtha in the present application, by connecting the high-temperature riser reactor and the relatively low-temperature fluidized bed reactor in series, the beneficial effects of reducing the yield of low-carbon alkane and increasing the yield of aromatic hydrocarbon are achieved.

BRIEF DESCRIPTION OF THE DRAWINGS

FIG. 1 is a schematic diagram of a device for preparing aromatic hydrocarbons from naphtha in one embodiment of the present application.

List of parts and reference symbols:

1 refers to fluidized bed reactor; 1-1 refers to reactor shell; 1-2 refers to reactor distributor; 1-3 refers to reactor gas-solid separation device; 1-4 refers to reactor gas collection chamber; 1-5 refers to product gas conveying pipe; 1-6 refers to reactor stripper; 1-7 refers to spent slide valve; and 1-8 refers to spent agent conveying pipe;

2 refers to fluidized bed regenerator; 2-1 refers to regenerator shell; 2-2 refers to regenerator distributor; 2-3 refers to regenerator gas-solid separation device; 2-4 refers to regenerator gas collection chamber; 2-5 refers to flue gas conveying pipe; 2-6 refers to regenerator stripper; and 2-7 refers to regenerated slide valve; and

3 refers to riser reactor.

DETAILED DESCRIPTION

The present application is described in detail hereinafter with reference to embodiments, but the present application is not limited to these embodiments.

The present application provides a device for preparing aromatic hydrocarbons from naphtha, which comprises a fluidized bed reactor and a riser reactor; wherein, an outlet of the riser reactor is

connected to the fluidized bed reactor; and

the fluidized bed reactor is used for introducing a naphtha raw material to make contact with a catalyst from the riser reactor and react to generate a BTX-containing product gas stream and a spent catalyst, the product gas stream is subjected to gas-solid separation, the product gas stream after separation is sent to a downstream working section, and unconverted naphtha after separation returns to the fluidized bed reactor as a raw material; and a part of low-carbon alkanes after separation return to the riser reactor as raw materials.

The BTX refers to benzene, toluene and xylene.

In one embodiment, the low-carbon olefin refers to ethylene and propylene.

The low-carbon alkane refers to ethane and propane.

The combustible gas comprises methane, CO, and the like.

The heavy aromatic hydrocarbon refers to an aromatic hydrocarbon with a number of carbon atoms in a molecule greater than or equal to 9.

In one embodiment, the naphtha is selected from at least one of direct coal liquefaction naphtha, indirect coal liquefaction naphtha, straight-run naphtha and hydrocracking naphtha.

In one embodiment, the naphtha also contains unconverted naphtha separated from the product gas stream, and the unconverted naphtha comprises main components of C₄-C₁₂ linear-chain and branched-chain aliphatic hydrocarbons and a naphthenic hydrocarbon.

In one embodiment, a riser reactor raw material comprises water vapor and low-carbon alkane separated from the product gas stream.

In one embodiment, a water vapor content in the riser reactor raw material is 0wt% to 50wt%.

In one embodiment, an inlet of the riser reactor is connected with a fluidized bed regenerator, and the catalyst introduced into the riser reactor is a regenerated catalyst generated by the fluidized bed regenerator.

In one embodiment, the fluidized bed regenerator sequentially passes through a regenerator stripper and a regenerated slide valve to be connected to the inlet of the riser reactor through a pipeline.

In one embodiment, an inlet of the regenerator stripper extends into a regenerator shell of the fluidized bed regenerator and is located above a regenerator distributor.

In one embodiment, the riser reactor is used for introducing the riser reactor raw material and a

catalyst to react to generate aromatic hydrocarbons, and a stream containing the unreacted riser reactor raw material, aromatic hydrocarbons and the catalyst enters the fluidized bed reactor through the outlet of the riser reactor.

In one embodiment, the fluidized bed reactor comprises a reactor shell, an area enclosed by the reactor shell is divided into a first gas-solid separation zone and a reaction zone from top to bottom, and a gas-solid separation device and a reactor gas collection chamber are provided in the first gas-solid separation zone; the reactor gas collection chamber is located on an inner top portion of the reactor shell, an inlet of the reactor gas collection chamber is communicated with a gas outlet of the reactor gas-solid separation device, and an outlet of the reactor gas collection chamber is communicated with a product gas conveying pipe; and a reactor distributor is provided on a lower portion of the reaction zone for introducing the naphtha raw material.

In one embodiment, the reactor gas-solid separation device is one or more groups of gas-solid cyclone separators, and each group of gas-solid cyclone separators comprises a first-stage gas-solid cyclone separator and a second-stage gas-solid cyclone separator.

In a preferred embodiment, the device further comprises a fluidized bed regenerator connected with the fluidized bed reactor, the fluidized bed regenerator is used for introducing regeneration gas to convert the spent catalyst into a regenerated catalyst. With reference to FIG. 1, the device comprises a fluidized bed reactor 1, a fluidized bed regenerator 2 and a riser reactor 3.

The fluidized bed reactor 1 comprises a reactor shell 1-1, a reactor distributor 1-2, a reactor gas-solid separation device 1-3, a reactor gas collection chamber 1-4, a product gas conveying pipe 1-5, a reactor stripper 1-6, a spent slide valve 1-7 and a spent agent conveying pipe 1-8.

The reactor shell 1-1 comprises a reactor upper shell and a reactor lower shell, a first gas-solid separation zone is enclosed by the reactor upper shell, and a reaction zone is enclosed by the reactor lower shell; and an outlet of the riser reactor 3 is provided in the reactor shell 1-1.

The reactor distributor 1-2 is provided on a lower portion of the reaction zone, and the reactor distributor 1-2 is used for introducing a naphtha raw material.

The reactor gas-solid separation device 1-3 and the reactor gas collection chamber 1-4 are also provided in the reactor shell 1-1; the reactor gas collection chamber 1-4 is located on an inner top portion of the reactor shell; a gas outlet of the reactor gas-solid separation device 1-3 is communicated with the reactor gas collection chamber 1-4; the reactor gas collection chamber 1-4 is

communicated with the product gas conveying pipe 1-5; and a catalyst outlet end of the reactor gas-solid separation device 1-3 is located above an opening end of an inlet pipe of the reactor stripper 1-6.

The reactor stripper 1-6 is provided below the reaction zone; an inlet of the reactor stripper 1-6 is located inside the reactor shell 1-1; an outlet of the reactor stripper 1-6 is located outside the reactor shell 1-1 and connected with the spent slide valve 1-7; and an opening end of the inlet of the reactor stripper 1-6 is located above the reactor distributor 1-2.

The spent slide valve 1-7 is provided below the reactor stripper 1-6; and an inlet of the spent slide valve 1-7 is connected to the outlet of the reactor stripper 1-6, an outlet of the spent slide valve 1-7 is connected to an inlet of the spent agent conveying pipe 1-8, and an outlet of the spent agent conveying pipe 1-8 is connected to a regenerator shell 2-1.

The spent slide valve 1-7 is used for controlling a circulating amount of a spent catalyst.

In a preferred embodiment, the reactor gas-solid separation device 1-3 is one or more groups of gas-solid cyclone separators, and each group of gas-solid cyclone separators comprises a first-stage gas-solid cyclone separator and a second-stage gas-solid cyclone separator.

The fluidized bed regenerator 2 comprises the regenerator shell 2-1, a regenerator distributor 2-2, a regenerator gas-solid separation device 2-3, a regenerator gas collection chamber 2-4, a flue gas conveying pipe 2-5, a regenerator stripper 2-6 and a regenerated slide valve 2-7.

The regenerator shell 2-1 comprises a regenerator upper shell and a regenerator lower shell, a second gas-solid separation zone is enclosed by the regenerator upper shell, and a regeneration zone is enclosed by the regenerator lower shell; and the outlet of the spent agent conveying pipe 1-8 is arranged in the regenerator shell 2-1.

The regenerator distributor 2-2 is provided on a lower portion of the regeneration zone, and the regenerator distributor 2-2 is used for introducing regeneration gas.

The regenerator gas-solid separation device 2-3 and the regenerator gas collection chamber 2-4 are also provided in the regenerator shell 2-1; the regenerator gas collection chamber 2-4 is located on an inner top portion of the regenerator shell 2-1; a gas outlet of the regenerator gas-solid separation device 2-3 is communicated with the regenerator gas collection chamber 2-4; the regenerator gas collection chamber 2-4 is communicated with the flue gas conveying pipe 2-5; and a catalyst outlet end of the regenerator gas-solid separation device 2-3 is located above an opening end

of an inlet pipe of the regenerator stripper 2-6.

The regenerator stripper 2-6 is provided below the regeneration zone; an inlet of the regenerator stripper 2-6 is located inside the regenerator shell 2-1; an outlet of the regenerator stripper 2-6 is located outside the regenerator shell 2-1 and connected with the regenerated slide valve 2-7; and an opening end of the inlet of the regenerator stripper 2-6 is located above the regenerator distributor 2-2.

The regenerated slide valve 2-7 is provided below the regenerator stripper 2-6; and an inlet of the regenerated slide valve 2-7 is connected to the outlet of the regenerator stripper 2-6.

The regenerated slide valve 2-7 is used for controlling a circulating amount of a regenerated catalyst.

In a preferred embodiment, the regenerator gas-solid separation device 2-3 is one or more groups of gas-solid cyclone separators, and each group of gas-solid cyclone separators comprises a first-stage gas-solid cyclone separator and a second-stage gas-solid cyclone separator.

An inlet of the riser reactor 3 is connected to the regenerated slide valve 2-7, and an outlet of the riser reactor 3 is connected to the reactor shell 1-1.

In order to achieve the aromatization of linear-chain and branched-chain aliphatic hydrocarbons and increase the yield of aromatic hydrocarbons and the content of p-xylene in mixed xylene, the present application provides a method for preparing aromatic hydrocarbons from naphtha, which comprises: preparing aromatic hydrocarbons by using the device for preparing aromatic hydrocarbons from naphtha above and a catalyst.

In one embodiment, the catalyst is a metal molecular sieve bifunctional catalyst.

In a preferred embodiment, the metal molecular sieve bifunctional catalyst is a metal modified HZSM-5 zeolite molecular sieve.

Metal for the metal modification is selected from at least one of La, Zn, Ga, Fe, Mo and Cr.

A method for the metal modification comprises: placing the HZSM-5 zeolite molecular sieve in a metal salt solution, soaking, drying and roasting to obtain the metal modified HZSM-5 zeolite molecular sieve. The metal modified HZSM-5 zeolite molecular sieve is used in Examples 1 to 5 below.

In a preferred embodiment, the method comprises the following steps.

a) The naphtha enters the reaction zone of the fluidized bed reactor 1 through the reactor distributor 1-2 to make contact with the catalyst from the riser reactor 3 to generate a product gas stream containing BTX, a low-carbon olefin, hydrogen, a low-carbon alkane, combustible gas, a heavy aromatic hydrocarbon and unconverted naphtha, and the catalyst is coked and converted into a spent catalyst at the same time. The product gas stream enters the reactor gas-solid separation device 1-3 to remove a spent catalyst contained in the product gas stream, then enters the reactor gas collection chamber 1-4, and enters the downstream working section through the product gas conveying pipe 1-5. The spent catalyst in the reaction zone enters the reactor stripper 1-6 from an opening end of an inlet pipe of the reactor stripper 1-6 for stripping, and after the spent catalyst is stripped, the spent catalyst enters the fluidized bed reactor 2 through the spent slide valve 1-7 and the spent agent conveying pipe 1-8.

b) The regeneration gas is introduced into the regeneration zone of the fluidized bed regenerator 2 through the regenerator distributor 2-2 to make contact with the spent catalyst, coke on the spent catalyst reacts with the regeneration gas to generate the flue gas, and the spent catalyst is converted into the regenerated catalyst at the same time. The flue gas enters the regenerator gas-solid separation device 2-3 to remove a regenerated catalyst contained in the flue gas, then enters the regenerator gas collection chamber 2-4, and enters the downstream working section through the flue gas conveying pipe 2-5. The regenerated catalyst sequentially passes through the regenerator stripper 2-6 and the regenerated slide valve 2-7 to enter the riser reactor 3.

c) The riser reactor raw material is introduced into the riser reactor 3, and makes contact and reacts with the regenerated catalyst from the fluidized bed regenerator 2, and the riser reactor raw material is converted into aromatic hydrocarbons under an action of the catalyst. Then, a stream containing the unreacted riser reactor raw material, aromatic hydrocarbons and the catalyst enters the fluidized bed reactor 1 from the outlet of the riser reactor 3.

The low-carbon olefin refers to ethylene and propylene.

The low-carbon alkane refers to ethane and propane.

The combustible gas comprises methane, CO, and the like.

The heavy aromatic hydrocarbon refers to an aromatic hydrocarbon with a number of carbon atoms in a molecule greater than or equal to 9.

In a preferred embodiment, the naphtha is selected from at least one of direct coal liquefaction

naphtha, indirect coal liquefaction naphtha, straight-run naphtha and hydrocracking naphtha.

In a preferred embodiment, the naphtha also contains unconverted naphtha separated from the product gas stream.

In a preferred embodiment, a carbon content in the spent catalyst is 1.0wt% to 3.0wt%.

5 In a preferred embodiment, process conditions of the reaction zone are: an apparent linear velocity of gas of 0.5 m/s to 2.0 m/s, a reaction temperature of 500°C to 650°C, a reaction pressure of 100 kPa to 500 kPa, and a bed density of 150 kg/m³ to 700 kg/m³.

Optionally, the apparent linear velocity of gas of the reaction zone is independently selected from any value or a value in a range between any two of 0.5 m/s, 0.6 m/s, 0.7 m/s, 0.8 m/s, 0.9 m/s, 10 1.0 m/s, 1.1 m/s, 1.2 m/s, 1.3 m/s, 1.4 m/s, 1.5 m/s, 1.6 m/s, 1.7 m/s, 1.8 m/s, 1.9 m/s and 2.0 m/s.

Optionally, the reaction temperature of the reaction zone is independently selected from any value or a value in a range between any two of 500°C, 510°C, 520°C, 530°C, 540°C, 550°C, 560°C, 570°C, 580°C, 590°C, 600°C, 610°C, 620°C, 630°C, 640°C and 650°C.

Optionally, the reaction pressure of the reaction zone is independently selected from any value 15 or a value in a range between any two of 100 kPa, 125 kPa, 150 kPa, 175 kPa, 200 kPa, 225 kPa, 250 kPa, 275 kPa, 300 kPa, 325 kPa, 350 kPa, 375 kPa, 400 kPa, 425 kPa, 450 kPa, 475 kPa and 500 kPa.

Optionally, the bed density of the reaction zone is independently selected from any value or a value in a range between any two of 150 kg/m³, 200 kg/m³, 250 kg/m³, 300 kg/m³, 350 kg/m³, 400 20 kg/m³, 450 kg/m³, 500 kg/m³, 550 kg/m³, 600 kg/m³, 650 kg/m³ and 700 kg/m³.

In a preferred embodiment, a carbon content in the regenerated catalyst is less than or equal to 0.5wt%.

In a preferred embodiment, the regeneration gas is selected from at least one of oxygen, air and oxygen-enriched air.

25 In a preferred embodiment, process conditions of the regeneration zone are: an apparent linear velocity of gas of 0.5 m/s to 2.0 m/s, a regeneration temperature of 600°C to 750°C, a regeneration pressure of 100 kPa to 500 kPa, and a bed density of 150 kg/m³ to 700 kg/m³.

Optionally, the apparent linear velocity of gas is independently selected from any value or a value in a range between any two of 0.5 m/s, 0.6 m/s, 0.7 m/s, 0.8 m/s, 0.9 m/s, 1.0 m/s, 1.1 m/s, 1.2 30 m/s, 1.3 m/s, 1.4 m/s, 1.5 m/s, 1.6 m/s, 1.7 m/s, 1.8 m/s, 1.9 m/s and 2.0 m/s.

Optionally, the regeneration temperature is independently selected from any value or a value in a range between any two of 600°C, 615°C, 630°C, 645°C, 660°C, 675°C, 690°C, 705°C, 720°C, 735°C and 750°C.

Optionally, the regeneration pressure is independently selected from any value or a value in a range between any two of 100 kPa, 125 kPa, 150 kPa, 175 kPa, 200 kPa, 225 kPa, 250 kPa, 275 kPa, 300 kPa, 325 kPa, 350 kPa, 375 kPa, 400 kPa, 425 kPa, 450 kPa, 475 kPa and 500 kPa.

Optionally, the bed density is independently selected from any value or a value in a range between any two of 150 kg/m³, 200 kg/m³, 250 kg/m³, 300 kg/m³, 350 kg/m³, 400 kg/m³, 450 kg/m³, 500 kg/m³, 550 kg/m³, 600 kg/m³, 650 kg/m³ and 700 kg/m³.

In a preferred embodiment, the riser reactor raw material comprises water vapor and low-carbon alkane separated from the product gas stream.

In a preferred embodiment, a water vapor content in the riser reactor raw material is 0wt% to 50wt%.

In a preferred embodiment, process conditions of the riser reactor are: an apparent linear velocity of gas of 3.0 m/s to 10.0 m/s, a temperature of 580°C to 700°C, a pressure of 100 kPa to 500 kPa, and a bed density of 50 kg/m³ to 150 kg/m³.

Optionally, the apparent linear velocity of gas is independently selected from any value or a value in a range between any two of 3.0 m/s, 3.5 m/s, 4.0 m/s, 4.5 m/s, 5.0 m/s, 5.5 m/s, 6.0 m/s, 6.5 m/s, 7.0 m/s, 7.5 m/s, 8.0 m/s, 8.5 m/s, 9.0 m/s, 9.5 m/s and 10.0 m/s.

Optionally, the temperature is independently selected from any value or a value in a range between any two of 580°C, 590°C, 600°C, 610°C, 620°C, 630°C, 640°C, 650°C, 660°C, 670°C, 680°C, 690°C and 700°C.

Optionally, the pressure is independently selected from any value or a value in a range between any two of 100 kPa, 125 kPa, 150 kPa, 175 kPa, 200 kPa, 225 kPa, 250 kPa, 275 kPa, 300 kPa, 325 kPa, 350 kPa, 375 kPa, 400 kPa, 425 kPa, 450 kPa, 475 kPa and 500 kPa.

Optionally, the bed density is independently selected from any value or a value in a range between any two of 50 kg/m³, 60 kg/m³, 70 kg/m³, 80 kg/m³, 90 kg/m³, 100 kg/m³, 110 kg/m³, 120 kg/m³, 130 kg/m³, 140 kg/m³ and 150 kg/m³.

In the embodiments of the present application, a potential content of aromatic hydrocarbons in the naphtha raw material is 0wt% to 80wt%, a per-pass conversion rate of naphtha is 70wt% to

95wt%, the unconverted naphtha is separated from the product gas and then returns to the fluidized bed reactor as a raw material, a part of low-carbon alkanes are separated from the product gas and then return to the riser reactor as raw materials, and the finally obtained product is composed of: 60wt% to 75wt% BTX, 7wt% to 15wt% low-carbon olefin, 3wt% to 8wt% hydrogen, 2wt% to 7wt% low-carbon alkane, 4wt% to 6wt% combustible gas, 3wt% to 7wt% heavy aromatic hydrocarbon, and 0.5wt% to 1wt% coke. The content of p-xylene in the mixed xylene in the product is 50wt% to 65wt%.

Example 1

A device shown in FIG. 1 is adopted in this embodiment.

In this embodiment, a naphtha raw material entering a fluidized bed reactor is direct coal liquefaction naphtha, and a potential content of aromatic hydrocarbons in the naphtha is 78wt%. The naphtha raw material entering the fluidized bed reactor further comprises unconverted naphtha separated from a product gas stream.

Process conditions of a reaction zone of the fluidized bed reactor are: an apparent linear velocity of gas of 0.5 m/s, a reaction temperature of 645°C, a reaction pressure of 100 kPa, and a bed density of 700 kg/m³.

Regeneration gas is air.

Process conditions of a regeneration zone of a fluidized bed regenerator are: an apparent linear velocity of gas of 0.5 m/s, a regeneration temperature of 745°C, a regeneration pressure of 100 kPa, and a bed density of 700 kg/m³.

A riser reactor raw material is a low-carbon alkane separated from the product gas stream.

Process conditions of a riser reactor are: an apparent linear velocity of gas of 3.0 m/s, a temperature of 690°C, a pressure of 100 kPa, and a bed density of 150 kg/m³.

A carbon content in a spent catalyst is 1.1wt%, and a carbon content in a regenerated catalyst is 0.1wt%.

The naphtha raw material entering the fluidized bed reactor has a per-pass conversion rate of 71wt%.

The product is composed of 74.5wt% BTX, 7wt% low-carbon olefin, 3wt% hydrogen, 2wt% low-carbon alkane, 6wt% combustible gas, 7wt% heavy aromatic hydrocarbon and 0.5wt% coke. A

content of p-xylene in the mixed xylene in the product is 51wt%.

Example 2

A device shown in FIG. 1 is adopted in this embodiment.

5 In this embodiment, a naphtha raw material entering a fluidized bed reactor is indirect coal liquefaction naphtha, and a potential content of aromatic hydrocarbons in the naphtha is 0.1wt%. The naphtha raw material entering the fluidized bed reactor further comprises unconverted naphtha separated from a product gas stream.

10 Process conditions of a reaction zone of the fluidized bed reactor are: an apparent linear velocity of gas of 2.0m/s, a reaction temperature of 510°C, a reaction pressure of 500 kPa, and a bed density of 150kg/m³.

Regeneration gas is oxygen.

15 Process conditions of a regeneration zone of a fluidized bed regenerator are: an apparent linear velocity of gas of 2.0m/s, a regeneration temperature of 610°C, a regeneration pressure of 500 kPa, and a bed density of 150kg/m³.

A riser reactor raw material comprises water vapor and a low-carbon alkane separated from the product gas stream, wherein a water vapor content in the riser reactor raw material is 50wt%.

Process conditions of a riser reactor are: an apparent linear velocity of gas of 10.0 m/s, a temperature of 580°C, a pressure of 500 kPa, and a bed density of 50 kg/m³.

20 A carbon content in a spent catalyst is 2.8wt%, and a carbon content in a regenerated catalyst is 0.3wt%.

The naphtha raw material entering the fluidized bed reactor has a per-pass conversion rate of 75wt%.

25 The product is composed of 66wt% BTX, 12wt% low-carbon olefin, 7wt% hydrogen, 3wt% low-carbon alkane, 5wt% combustible gas, 6wt% heavy aromatic hydrocarbon and 1.0wt% coke. A content of p-xylene in the mixed xylene in the product is 61wt%.

Example 3

A device shown in FIG. 1 is adopted in this embodiment.

30 In this embodiment, a naphtha raw material entering a fluidized bed reactor is indirect coal

liquefaction naphtha, and a potential content of aromatic hydrocarbons in the naphtha is 3wt%. The naphtha raw material entering the fluidized bed reactor further comprises unconverted naphtha separated from a product gas stream.

Process conditions of a reaction zone of the fluidized bed reactor are: an apparent linear velocity of gas of 1.2 m/s, a reaction temperature of 550°C, a reaction pressure of 120 kPa, and a bed density of 260 kg/m³.

Regeneration gas is oxygen-enriched air.

Technological conditions of a regeneration zone of a fluidized bed regenerator are: an apparent linear velocity of gas of 1.2 m/s, a regeneration temperature of 650°C, a regeneration pressure of 120 kPa, and a bed density of 260 kg/m³.

A riser reactor raw material comprises water vapor and a low-carbon alkane separated from the product gas stream, wherein a water vapor content in the riser reactor raw material is 25wt%.

Process conditions of a riser reactor are: an apparent linear velocity of gas of 7.0 m/s, a temperature of 630°C, a pressure of 120 kPa, and a bed density of 80 kg/m³.

A carbon content in a spent catalyst is 2.1wt%, and a carbon content in a regenerated catalyst is 0.2wt%.

The naphtha raw material entering the fluidized bed reactor has a per-pass conversion rate of 95wt%.

The product is composed of 61wt% BTX, 15wt% low-carbon olefin, 8wt% hydrogen, 7wt% low-carbon alkane, 5.2wt% combustible gas, 3wt% heavy aromatic hydrocarbon and 0.8wt% coke. A content of p-xylene in the mixed xylene in the product is 65wt%.

Example 4

A device shown in FIG. 1 is adopted in this embodiment.

In this embodiment, a naphtha raw material entering a fluidized bed reactor is straight-run naphtha, and a potential content of aromatic hydrocarbons in the naphtha is 46wt%. The naphtha raw material entering the fluidized bed reactor further comprises unconverted naphtha separated from a product gas stream.

Process conditions of a reaction zone of the fluidized bed reactor are: an apparent linear velocity of gas of 1.8 m/s, a reaction temperature of 600°C, a reaction pressure of 200 kPa, and a bed density

of 220 kg/m³.

Regeneration gas is air.

Process conditions of a regeneration zone of a fluidized bed regenerator are: an apparent linear velocity of gas of 1.8 m/s, a regeneration temperature of 700°C, a regeneration pressure of 200 kPa, and a bed density of 220 kg/m³.

A riser reactor raw material comprises water vapor and a low-carbon alkane separated from the product gas stream, wherein a water vapor content in the riser reactor raw material is 50wt%.

Process conditions of a riser reactor are: an apparent linear velocity of gas of 5.0 m/s, a temperature of 660°C, a pressure of 200 kPa, and a bed density of 110 kg/m³.

A carbon content in a spent catalyst is 1.5wt%, and a carbon content in a regenerated catalyst is 0.1wt%.

The naphtha raw material entering the fluidized bed reactor has a per-pass conversion rate of 86wt%.

The product is composed of 68wt% BTX, 10wt% low-carbon olefin, 6wt% hydrogen, 5wt% low-carbon alkane, 4wt% combustible gas, 6wt% heavy aromatic hydrocarbon and 1.0wt% coke. A content of p-xylene in the mixed xylene in the product is 63wt%.

Example 5

A device shown in FIG. 1 is adopted in this embodiment.

In this embodiment, a naphtha raw material entering a fluidized bed reactor is hydrocracking naphtha, and a potential content of aromatic hydrocarbons in the naphtha is 64wt%. The naphtha raw material entering the fluidized bed reactor further comprises unconverted naphtha separated from a product gas stream.

Process conditions of a reaction zone of the fluidized bed reactor are: an apparent linear velocity of gas of 1.0 m/s, a reaction temperature of 580°C, a reaction pressure of 150 kPa, and a bed density of 350 kg/m³.

Regeneration gas is air.

Process conditions of a regeneration zone of a fluidized bed regenerator are: an apparent linear velocity of gas of 1.0m/s, a regeneration temperature of 680°C, a regeneration pressure of 150 kPa, and a bed density of 350 kg/m³.

A riser reactor raw material comprises water vapor and a low-carbon alkane separated from the product gas stream, wherein a water vapor content in the riser reactor raw material is 40wt%.

Process conditions of a riser reactor are: an apparent linear velocity of gas of 7.0 m/s, a temperature of 650°C, a pressure of 150 kPa, and a bed density of 80 kg/m³.

5 A carbon content in a spent catalyst is 1.4wt%, and a carbon content in a regenerated catalyst is 0.5wt%.

The naphtha raw material entering the fluidized bed reactor has a per-pass conversion rate of 77wt%.

10 The product is composed of 71.3wt% BTX, 9wt% low-carbon olefin, 5wt% hydrogen, 2wt% low-carbon alkane, 6wt% combustible gas, 6wt% heavy aromatic hydrocarbon and 0.7wt% coke. A content of p-xylene in the mixed xylene in the product is 58wt%.

15 The above are only several embodiments of the present application, and are not intended to limit the present application in any form. Although the present application is disclosed in the preferred embodiments above, the preferred embodiments are not intended to limit the present application. The changes or modifications made by those skilled in the art by using the technical contents disclosed above without departing from the scope of the technical solution of the present application are equivalent to equivalent embodiments, and all fall within the scope of the technical solution of the present application.

Claims

1. A device for preparing aromatic hydrocarbons from naphtha, wherein the device comprises a fluidized bed reactor and a riser reactor; wherein, an outlet of the riser reactor is connected to the fluidized bed reactor; and

the fluidized bed reactor is used for introducing a naphtha raw material to make contact with a catalyst from the riser reactor and react to generate a BTX-containing product gas stream and a spent catalyst, the product gas stream is subjected to gas-solid separation, the product gas stream after separation is sent to a downstream working section, and unconverted naphtha after separation returns to the fluidized bed reactor as a raw material; and a part of low-carbon alkanes after separation return to the riser reactor as raw materials.

2. The device for preparing aromatic hydrocarbons from naphtha according to claim 1, wherein the riser reactor is used for introducing a riser reactor raw material and a catalyst to react to generate aromatic hydrocarbons, and a stream containing the unreacted riser reactor raw material, aromatic hydrocarbons and the catalyst enters the fluidized bed reactor through the outlet of the riser reactor.

3. The device for preparing aromatic hydrocarbons from naphtha according to claim 2, wherein the riser reactor raw material comprises water vapor and the low-carbon alkane separated from the product gas stream.

4. The device for preparing aromatic hydrocarbons from naphtha according to claim 2 or 3, wherein a water vapor content in the riser reactor raw material is 0wt% to 50wt%.

5. The device for preparing aromatic hydrocarbons from naphtha according to claim 1, wherein an inlet of the riser reactor is connected with a fluidized bed regenerator, and the catalyst introduced into the riser reactor is a regenerated catalyst generated by the fluidized bed regenerator.

6. The device for preparing aromatic hydrocarbons from naphtha according to claim 5, wherein the fluidized bed regenerator sequentially passes through a regenerator stripper and a regenerated

slide valve to be connected to the inlet of the riser reactor through a pipeline.

7. The device for preparing aromatic hydrocarbons from naphtha according to claim 6, wherein an inlet of the regenerator stripper extends into a regenerator shell of the fluidized bed regenerator and is located above a regenerator distributor.

8. The device for preparing aromatic hydrocarbons from naphtha according to claim 1, wherein the fluidized bed reactor comprises a reactor shell, an area enclosed by the reactor shell is divided into a first gas-solid separation zone and a reaction zone from top to bottom, and a gas-solid separation device and a reactor gas collection chamber are provided in the first gas-solid separation zone; the reactor gas collection chamber is located on an inner top portion of the reactor shell, an inlet of the reactor gas collection chamber is communicated with a gas outlet of the reactor gas-solid separation device, and an outlet of the reactor gas collection chamber is communicated with a product gas conveying pipe; and a reactor distributor is provided on a lower portion of the reaction zone for introducing the naphtha raw material.

9. The device for preparing aromatic hydrocarbons from naphtha according to claim 8, wherein the reactor gas-solid separation device is one or more groups of gas-solid cyclone separators, and each group of gas-solid cyclone separators comprises a first-stage gas-solid cyclone separator and a second-stage gas-solid cyclone separator.

10. The device for preparing aromatic hydrocarbons from naphtha according to claim 1, wherein the device further comprises a fluidized bed regenerator connected with the fluidized bed reactor, the fluidized bed regenerator is used for introducing regeneration gas to convert the spent catalyst into a regenerated catalyst.

11. The device for preparing aromatic hydrocarbons from naphtha according to claim 1, wherein the fluidized bed reactor sequentially passes through a reactor stripper, a spent slide valve and a spent agent conveying pipe to be connected with the fluidized bed regenerator; wherein, an inlet of the reactor stripper extends into a fluidized bed reactor shell and is located below a catalyst outlet end of

the reactor gas-solid separation device.

12. The device for preparing aromatic hydrocarbons from naphtha according to claim 1, wherein the fluidized bed regenerator comprises a regenerator shell, and an area enclosed by the regenerator shell is divided into a second gas-solid separation zone and a regeneration zone from top to bottom; a regenerator gas-solid separation device and a regenerator gas collection chamber are provided in the second gas-solid separation zone; the regenerator gas collection chamber is located on an inner top portion of the regenerator shell, and a flue gas conveying pipe is provided on the regenerator gas collection chamber; a gas outlet of the regenerator gas-solid separation device is communicated with the regenerator gas collection chamber; and a regenerator distributor is arranged on an inner lower portion of the regeneration zone for introducing regeneration gas.

13. The device for preparing aromatic hydrocarbons from naphtha according to claim 12, wherein the regenerator gas-solid separation device is one or more groups of gas-solid cyclone separators, and each group of gas-solid cyclone separators comprises a first-stage gas-solid cyclone separator and a second-stage gas-solid cyclone separator.

14. A method for preparing aromatic hydrocarbons from naphtha, wherein the method comprises: preparing aromatic hydrocarbons by using the device for preparing aromatic hydrocarbons from naphtha according to any one of claims 1 to 13 and a catalyst.

15. The method for preparing aromatic hydrocarbons from naphtha according to claim 14, wherein the catalyst is a metal molecular sieve bifunctional catalyst.

16. The method for preparing aromatic hydrocarbons from naphtha according to claim 15, wherein the metal molecular sieve bifunctional catalyst is a metal modified HZSM-5 zeolite molecular sieve;

metal for the metal modification is selected from at least one of La, Zn, Ga, Fe, Mo and Cr; and

a method for the metal modification comprises: placing the HZSM-5 zeolite molecular sieve in a metal salt solution, soaking, drying and roasting to obtain the metal modified HZSM-5 zeolite

molecular sieve.

17. The method for preparing aromatic hydrocarbons from naphtha according to claim 14, wherein the method comprises: allowing the naphtha to enter the reaction zone of the fluidized bed reactor through the reactor distributor to make contact with the catalyst from the riser reactor to generate a product gas stream containing BTX, a low-carbon olefin, hydrogen, a low-carbon alkane, combustible gas, a heavy aromatic hydrocarbon and unconverted naphtha, and allowing the catalyst to coke and converted into the spent catalyst at the same time; and

allowing the product gas stream to enter the reactor gas-solid separation device to remove a spent catalyst contained in the product gas stream, then enter the reactor gas collection chamber, and enter the downstream working section through the product gas conveying pipe.

18. The method for preparing aromatic hydrocarbons from naphtha according to claim 17, wherein the method further comprises: allowing the unconverted naphtha after separation to return to the fluidized bed reactor as a raw material.

19. The method for preparing aromatic hydrocarbons from naphtha according to claim 17, wherein the method further comprises: allowing a part of low-carbon alkanes after separation to return to the riser reactor as raw materials.

20. The method for preparing aromatic hydrocarbons from naphtha according to claim 17, wherein the low-carbon olefin refers to ethylene and propylene;

the low-carbon alkane refers to ethane and propane;

the combustible gas comprises methane and CO; and

the heavy aromatic hydrocarbon refers to an aromatic hydrocarbon with a number of carbon atoms in a molecule greater than or equal to 9.

21. The method for preparing aromatic hydrocarbons from naphtha according to claim 17, wherein the naphtha is selected from at least one of direct coal liquefaction naphtha, indirect coal liquefaction naphtha, straight-run naphtha and hydrocracking naphtha.

22. The method for preparing aromatic hydrocarbons from naphtha according to claim 21, wherein the naphtha also contains the unconverted naphtha separated from the product gas stream, and the unconverted naphtha comprises main components of C₄-C₁₂ linear-chain and branched-chain aliphatic hydrocarbons and a naphthenic hydrocarbon.

23. The method for preparing aromatic hydrocarbons from naphtha according to claim 17, wherein a carbon content in the spent catalyst is 1.0wt% to 3.0wt%.

24. The method for preparing aromatic hydrocarbons from naphtha according to claim 17, wherein process conditions of the reaction zone are: an apparent linear velocity of gas of 0.5 m/s to 2.0 m/s, a reaction temperature of 500°C to 650°C, a reaction pressure of 100 kPa to 500 kPa, and a bed density of 150 kg/m³ to 700 kg/m³.

25. The method for preparing aromatic hydrocarbons from naphtha according to claim 17, wherein the method further comprises: introducing the riser reactor raw material and the catalyst into the riser reactor to react to generate aromatic hydrocarbons; and

allowing a stream containing the unreacted riser reactor raw material, aromatic hydrocarbons and the catalyst to enter the fluidized bed reactor from the outlet of the riser reactor.

26. The method for preparing aromatic hydrocarbons from naphtha according to claim 25, wherein the catalyst is the regenerated catalyst from the fluidized bed regenerator.

27. The method for preparing aromatic hydrocarbons from naphtha according to claim 26, wherein the regenerated catalyst sequentially passes through the regenerator stripper and the regenerated slide valve to enter the riser reactor.

28. The method for preparing aromatic hydrocarbons from naphtha according to claim 26, wherein a carbon content in the regenerated catalyst is less than or equal to 0.5wt%.

29. The method for preparing aromatic hydrocarbons from naphtha according to claim 25, wherein the riser reactor raw material comprises water vapor and the low-carbon alkane separated from the product gas stream.

5 30. The method for preparing aromatic hydrocarbons from naphtha according to claim 25, wherein a water vapor content in the riser reactor raw material is 0wt% to 50wt%.

31. The method for preparing aromatic hydrocarbons from naphtha according to claim 25, wherein process conditions of the riser reactor are: an apparent linear velocity of gas of 3.0 m/s to 10.0 m/s, a temperature of 580°C to 700°C, a pressure of 100 kPa to 500 kPa, and a bed density of 50 kg/m³ to 150 kg/m³.
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32. The method for preparing aromatic hydrocarbons from naphtha according to claim 17, wherein the method further comprises: allowing the spent catalyst to enter the reactor stripper from an opening end of an inlet pipe of the reactor stripper, and after the spent catalyst is stripped by the reactor stripper, allowing the spent catalyst to enter a downstream area through the spent slide valve and the spent agent conveying pipe.
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33. The method for preparing aromatic hydrocarbons from naphtha according to claim 32, wherein the downstream area is the fluidized bed regenerator.
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34. The method for preparing aromatic hydrocarbons from naphtha according to claim 32, wherein the method further comprises: introducing the regeneration gas into the regeneration zone of the fluidized bed regenerator through the regenerator distributor to make contact with the spent catalyst from the fluidized bed reactor, allowing coke on the spent catalyst to react with the regeneration gas to generate the flue gas, and converting the spent catalyst into the regenerated catalyst at the same time.
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35. The method for preparing aromatic hydrocarbons from naphtha according to claim 34, wherein the spent catalyst sequentially passes through the reactor stripper, the spent slide valve and
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the spent agent conveying pipe to enter the fluidized bed regenerator, and makes contact and reacts with the regeneration gas to obtain the flue gas and the regenerated catalyst; and

the flue gas enters the regenerator gas-solid separation device to remove a regenerated catalyst contained in the flue gas, then enters the regenerator gas collection chamber, and enters the
5 downstream working section through the flue gas conveying pipe.

36. The method for preparing aromatic hydrocarbons from naphtha according to claim 34, wherein the regeneration gas is selected from at least one of oxygen, air and oxygen-enriched air.

10 37. The method for preparing aromatic hydrocarbons from naphtha according to claim 34, wherein process conditions of the regeneration zone are: an apparent linear velocity of gas of 0.5 m/s to 2.0 m/s, a regeneration temperature of 600°C to 750°C, a regeneration pressure of 100 kPa to 500 kPa, and a bed density of 150 kg/m³ to 700 kg/m³.