

FLUIDIZED BED DEVICE AND METHOD FOR PREPARING PARA-XYLENE AND CO-PRODUCING LIGHT OLEFINS FROM METHANOL AND/OR DIMETHYL ETHER AND BENZENE

TECHNICAL FIELD

The present invention relates to a device and a production method for producing para-xylene (PX) and co-producing light olefins, and is particularly suitable for a fluidized bed device and a production method for preparing the para-xylene and co-producing the light olefins by the alkylation of methanol and/or dimethyl ether and benzene, which belongs to the field of chemistry and chemical industry.

BACKGROUND

Para-xylene (PX) is one of the basic organic raw materials in the petrochemical industry, which has a wide range of applications in chemical fiber, synthetic resins, pesticides, pharmaceuticals and polymer materials. At present, the production of para-xylene mainly uses toluene, C₉ aromatics and mixed xylene as raw materials, and para-xylene is obtained by disproportionation, isomerization, adsorption separation or cryogenic separation. Since the para-xylene content in the product is controlled by thermodynamics, para-xylene only accounts for ~24% of the C₈ mixed aromatics, and the material circulation processing amount is large during the process, and the equipment is large and the operation cost is high. In particular, the three isomers of xylene have small differences in boiling points, and it is difficult to obtain high-purity para-xylene by conventional distillation techniques, and an expensive adsorption separation process must be employed. In recent years, many patents, domestic and abroad, have disclosed a new route for the production of para-xylene. The alkylation technology of methanol and/or dimethyl ether and benzene is a new way to produce para-xylene with high selectivity, which has been highly valued and paid great attention by the industry.

Light olefins, namely ethylene, propylene and butene, are two basic petrochemical feedstocks that are increasingly in demand. Ethylene and propylene are mainly produced from naphtha, depending on the petroleum route. In recent years, the non-petroleum route to produce ethylene and propylene has received more and more attention, especially the process route of the methanol conversion to light olefins (MTO), which is an important way to achieve petroleum substitution strategy, reduce and alleviate our demand and dependence for petroleum.

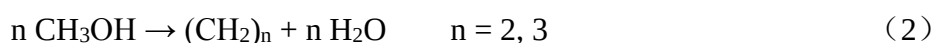
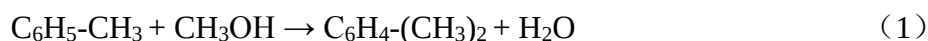
The preparation of para-xylene by alkylation of benzene and methanol and the preparation of toluene and xylene by alkylation of benzene and methanol are new ways to

increase the production of aromatics. The selective alkylation of benzene and methanol can produce highly selective para-xylene products, but the toluene required for this process is also an intermediate raw material for the production of para-xylene by an aromatics complex unit, which is in short supply in the market. While benzene is a by-product of the aromatics complex unit, it is estimated that an annual output of 800,000 tons of PX aromatics complex unit can produce about 300,000 tons of benzene. Therefore, the use of benzene as a raw material, alkylation with methanol to produce toluene and xylene will be an effective way to increase the production of aromatics.

A conventional toluene alkylation process involves mixing methanol and/or dimethyl ether and benzene and upstream of the reactor and then feeding the mixture together into the reactor. The reactor type includes a fixed bed and a fluidized bed. In order to increase the conversion rate of toluene, the phased injection of reactants has been employed in various fixed bed and fluidized bed processes.

The competition between the MTO reaction and the alkylation reaction is a major factor affecting the conversion rate of benzene, the yield of the para-xylene, and the selectivity of the light olefins. The process of simultaneously realizing two reactions in the same reactor is simple, but the conversion rate of benzene is low; the process of respectively realizing two reactions in different reactors is complicated, but the conversion rate of benzene and the yield of para-xylene are higher. Therefore, the process of the alkylation of methanol and/or dimethyl ether and benzene to prepare para-xylene and co-produce light olefins requires a major breakthrough in the process configuration and the reactor design, thereby coordinating and optimizing the competition between the alkylation reaction and the MTO reaction, and improving the conversion rate of benzene, the yield of para-xylene, and the selectivity of light olefins.

The above mentioned new routes for the preparation of para-xylene and light olefins are all acid-catalyzed reaction. Methanol-to-olefins reaction is inevitable in the process of preparing para-xylene by the alkylation of benzene and/or toluene-methanol and/or dimethyl ether and benzene based on the ZSM-5 molecular sieve catalyst. In the course of this reaction, the following reactions occur mainly:



Methanol is both a raw material for the alkylation reaction of benzene and/or

toluene-methanol and/or dimethyl ether and benzene, and a raw material for the MTO reaction, but the reaction rate of the MTO reaction is much higher than that of the alkylation reaction of benzene and/or toluene-methanol and/or dimethyl ether and benzene.

One of the characteristics of the MTO reaction is that the reaction rate is much higher than that of the alkylation reaction of methanol and/or dimethyl ether and benzene. Another important feature is that after the catalyst is carbonized, the conversion rate of the methanol decreases and the selectivity of the light olefins increases. Therefore, controlling the carbonation of the catalyst is an effective way to improve the selectivity of light olefins in the MTO reaction.

It can be seen from the above analysis that the technical field needs to coordinate and optimize the competition between the alkylation reaction and the MTO reaction from the two aspects of the catalyst design and the reactor design, so as to achieve synergistic effect and improve the conversion rate of benzene, the yield of para-xylene and the yield of light olefins.

SUMMARY OF THE INVENTION

According to an aspect of the present application, this is provided a turbulent fluidized bed reactor for preparing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene. The turbulent fluidized bed reactor resolves or improves the the competition problem between an MTO reaction and an alkylation reaction during the process of producing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene, thus achieving a synergistic effect between the MTO reaction and the alkylation reaction. By controlling the mass transfer and reaction, the competition between the alkylation reaction and the MTO reaction is coordinated and optimized to achieve a synergistic effect, thereby improving the conversion rate of benzene, the yield of para-xylene and the selectivity of light olefins.

As our experimental studies show, when benzene and/or toluene-methanol and/or dimethyl ether and benzene are co-fed and the content of the methanol in the raw material is low, the MTO reaction quickly consumes most of the methanol (alkylation reactant), inhibits the alkylation reaction of benzene and/or toluene-methanol and/or dimethyl ether and benzene, and the conversion rate of benzene and/or toluene is low. When the content of methanol in the raw material is excessively excessive, the difference in the diffusion speed between methanol and toluene and/or toluene in the molecular sieve pores makes the adsorption amount of benzene and/or toluene per unit time low, which is also unfavorable for the alkylation reaction of methanol and/or dimethyl ether and benzene. Therefore, optimizing the concentrations of

methanol and benzene and/or toluene in the reaction zone is an effective way to improve the conversion rate of benzene and/or toluene and the yield of para-xylene.

The turbulent fluidized bed reactor for preparing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene, comprises a first reactor feed distributor and a plurality of second reactor feed distributors, the first reactor feed distributor and the plurality of second reactor feed distributors are sequentially arranged along the gas flow direction in the reaction zone of the turbulent fluidized bed reactor.

Preferably, the number of the second reactor feed distributors is in a range from 2 to 10.

Preferably, the turbulent fluidized bed reactor comprises a first reactor gas-solid separator and a second reactor gas-solid separator, the first reactor gas-solid separator is placed in a dilute phase zone or outside a reactor shell, and the second reactor gas-solid separator is placed in the dilute phase zone or outside the reactor shell;

the first reactor gas-solid separator is provided with a regenerated catalyst inlet, a catalyst outlet of the first reactor gas-solid separator is placed at the bottom of a reaction zone, and a gas outlet of the first reactor gas-solid separator is placed in the dilute phase zone;

an inlet of the second reactor gas-solid separator is placed in the dilute phase zone, a catalyst outlet of the second reactor gas-solid separator is placed in the reaction zone, and a gas outlet of the second reactor gas-solid separator is connected to a product gas outlet of the turbulent fluidized bed reactor;

the reaction zone is located in a lower part of the turbulent fluidized bed reactor, and the dilute phase zone is located in an upper part of the turbulent fluidized bed reactor.

Preferably, the first reactor gas-solid separator and the second reactor gas-solid separator are cyclone separators.

Preferably, the turbulent fluidized bed reactor comprises a reactor heat extractor, and the reactor heat extractor is arranged inside or outside the shell of the turbulent fluidized bed reactor.

Preferably, the turbulent fluidized bed reactor comprises a reactor heat extractor, and the reactor heat extractor is arranged inside or outside the shell of the turbulent fluidized bed reactor.

Further preferably, the reactor heat extractor is arranged between the plurality of reactor feed distributors.

Preferably, the turbulent fluidized bed reactor comprises a reactor stripper, the reactor stripper passes through the reactor shell from the outside to the inside at the bottom of

the turbulent fluidized bed reactor and is opened in the reaction zone of the turbulent fluidized bed reactor, and a reactor stripping gas inlet and a spent catalyst outlet are arranged at the bottom of the reactor stripper.

Preferably, the turbulent fluidized bed reactor comprises a perforated plate located between the first reactor feed distributor and at least one of the second reactor feed distributors, the porosity of the perforated plate is less than or equal to 50%.

Preferably, the turbulent fluidized bed reactor comprises a perforated plate located between the first reactor feed distributor and the second reactor feed distributor closest to the first reactor feed distributor, the porosity of the perforated plate is in a range from 5% to 50%.

Preferably, the reactor stripper passes through the reactor shell from the outside to the inside at the bottom of the turbulent fluidized bed reactor and is opened in the reaction zone of the turbulent fluidized bed reactor, and a reactor stripping gas inlet and a spent catalyst outlet are arranged at the bottom of the reactor stripper;

the turbulent fluidized bed reactor comprises a perforated plate located between the first reactor feed distributor and at least one of the second reactor feed distributors, the porosity of the perforated plate is less than or equal to 50%;

the horizontal height of opening of the reactor stripper in the reactor shell is higher than that of the first reactor feed distributor and higher than that of the perforated plate.

In the present application, light olefins include at least one of ethylene, propylene and butene.

In the present application, methanol in the feedstock may be replaced in whole or in part by dimethyl ether, such as the feedstock containing dimethyl ether; the amount of methanol may be calculated by converting dimethyl ether into methanol having the same number of carbon atoms.

In the present application, “methanol and/or dimethyl ether” means that, including three cases: only methanol; or only dimethyl ether; or both methanol and dimethyl ether.

In the present application, “methanol and/or dimethyl ether and benzene” includes three cases: benzene and methanol; or benzene and dimethyl ether; or benzene and methanol and dimethyl ether.

Unless otherwise specified, the methanol in the present application may be replaced by all or part of dimethyl ether and the amount of methanol may be calculated by converting dimethyl ether into methanol having the same number of carbon atoms.

According to another aspect of the present application, this is provided a device for preparing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and

benzene. The device resolves or improves the the competition problem between an MTO reaction and an alkylation reaction during the process of producing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene, thus achieving a synergistic effect between the MTO reaction and the alkylation reaction. By controlling the mass transfer and reaction, the competition between the alkylation reaction and the MTO reaction is coordinated and optimized to achieve a synergistic effect, thereby improving the conversion rate of the benzene, the yield of the para-xylene and the selectivity of light olefins.

The device comprises at least one of the turbulent fluidized bed reactors according to any one of the mentioned above and a fluidized bed regenerator for regenerating a catalyst.

Preferably, the fluidized bed regenerator is a turbulent fluidized bed regenerator, and the fluidized bed regenerator comprises a regenerator shell, a regenerator gas-solid separator, a regenerator heat extractor and the regenerator stripper; the lower part of the fluidized bed regenerator is a regeneration zone, the upper part of the fluidized bed regenerator is a dilute phase zone of the regenerator, the regenerator feed distributor is placed at the bottom of the regeneration zone, the regenerator heat extractor is placed in the regeneration zone, and the regenerator gas-solid separator is placed in the dilute phase zone or outside the regenerator shell;

the inlet of the regenerator gas-solid separator is placed in the dilute phase zone of the regenerator, the catalyst outlet of the regenerator gas-solid separator is placed in the regeneration zone, and the regenerator stripper is opened at the bottom of the regenerator shell.

Preferably, the fluidized bed regenerator comprises a regenerator shell, a regenerator feed distributor, a regenerator gas-solid separator, a regenerator heat extractor, a flue gas outlet and a regenerator stripper;

the lower part of the fluidized bed regenerator is a regeneration zone, and the upper part of the fluidized bed regenerator is a dilute phase zone;

the regenerator feed distributor is placed at the bottom of the regeneration zone, the regenerator heat extractor is placed in the regeneration zone, the regenerator gas-solid separator is placed in the dilute phase zone or outside the regenerator shell, the inlet of the regenerator gas-solid separator is placed in the dilute phase zone, the catalyst outlet of the regenerator gas-solid separator is placed in the regeneration zone, the gas outlet of the regenerator gas-solid separator is connected to the flue gas outlet, and the regenerator stripper is opened at the bottom of the regenerator shell;

the spent catalyst outlet of the reactor stripper is connected to the inlet of an inclined spent catalyst pipe, a spent catalyst sliding valve is arranged in the inclined spent catalyst pipe, the outlet of the inclined spent catalyst pipe is connected to the inlet of a spent

catalyst lift pipe, the bottom of the spent catalyst lift pipe is provided with a spent catalyst lifting gas inlet, and the outlet of the spent catalyst lift pipe is connected to the dilute phase zone of the fluidized bed regenerator;

the bottom of the regenerator stripper is provided with a regenerator stripping gas inlet, the bottom of the regenerator stripper is connected to the inlet of an inclined regenerated catalyst pipe, a regenerated catalyst sliding valve is arranged in the inclined regenerated catalyst pipe, the outlet of the inclined regenerated catalyst pipe is connected to the inlet of a regenerated catalyst lift pipe, the bottom of the regenerated catalyst lift pipe is provided with a regenerated catalyst lifting gas inlet, the outlet of the regenerated catalyst lift pipe is connected to the regenerated catalyst inlet of the first reactor gas-solid separator, and the first reactor gas-solid separator is placed in the dilute phase zone or outside the reactor shell of the fluidized bed reactor.

According to still another aspect of the present application, this is provided a method for preparing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene. The method resolves or improves the the competition problem between an MTO reaction and an alkylation reaction during the process of producing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene, thus achieving a synergistic effect between the MTO reaction and the alkylation reaction. By controlling the mass transfer and reaction, the competition between the alkylation reaction and the MTO reaction is coordinated and optimized to achieve a synergistic effect, thereby improving the conversion rate of the benzene, the yield of the para-xylene and the selectivity of light olefins.

The method for preparing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene, at least one of the turbulent fluidized bed reactors according to any one of the mentioned above is used.

Preferably, a raw material A containing methanol and/or dimethyl ether and benzene is fed into the reaction zone of the turbulent fluidized bed reactor from the first reactor feed distributor and a raw material B containing methanol and/or dimethyl ether is respectively fed into the reaction zone of the turbulent fluidized bed reactor from a plurality of second reactor feed distributors respectively to be in contact with the catalyst, to form a material stream C containing para-xylene and light olefins products and a spent catalyst.

Preferably, the material stream C is separated to obtain para-xylene, light olefins, C₅₊ chain hydrocarbons, aromatic by-products and unconverted methanol, dimethyl ether and benzene;

wherein the unconverted methanol and dimethyl ether are fed into the reaction zone

of the turbulent fluidized bed reactor from a plurality of second reactor feed distributors, the aromatic by-products and the unconverted benzene are fed into the reaction zone of the turbulent fluidized bed reactor from the first reactor feed distributor to be in contact with a catalyst.

Preferably, the spent catalyst is regenerated by a fluidized bed regenerator and fed to the bottom of the reaction zone of the turbulent fluidized bed reactor.

Preferably, the method comprises the steps of:

(1) feeding a material stream A containing methanol and/or dimethyl ether and benzene into the reaction zone of the turbulent fluidized bed reactor from a first reactor feed distributor below the turbulent fluidized bed reactor to be in contact with the catalyst;

(2) feeding a material stream B containing methanol into the reaction zone of the turbulent fluidized bed reactor from 2 to 10 second reactor feed distributors to be in contact with the catalyst, to form a material stream C containing para-xylene and light olefins products and a spent catalyst; the 2 to 10 second reactor feed distributors are arranged in sequence above the first reactor feed distributor;

(3) separating the material stream C obtained from the step (2) to obtain a material stream C-1 containing unconverted methanol and dimethyl ether, a material stream C-2 containing unconverted benzene and aromatic by-products; the material stream C-1 is respectively fed into the reaction zone of the turbulent fluidized bed reactor from the 2 to 10 second reactor feed distributors to be in contact with the catalyst; the material stream C-2 is fed into the reaction zone of the turbulent fluidized bed reactor from the first reactor feed distributor to be in contact with the catalyst;

the aromatic by-products comprise toluene, o-xylene, m-xylene, ethylbenzene and C₉₊ aromatics;

(4) regenerating the spent catalyst obtained from the step (2) by the fluidized bed regenerator, the regenerated catalyst is fed to the first reactor gas-solid separator to removal the gas, and then the regenerated catalyst is fed to the bottom of the reaction zone in the fast fluidized bed reactor.

Preferably, in the mixture fed from the first reactor feed distributor into the turbulent fluidized bed reactor, the ratio of the molecular moles of aromatics to the carbon moles of methanol and/or dimethyl ether is greater than 0.5.

Further preferably, in the mixture fed from the first reactor feed distributor into the turbulent fluidized bed reactor, the ratio of the molecular moles of aromatics to the carbon moles of methanol and/or dimethyl ether is in a range from 0.5 to 5.

In the present application, the molecular mole refers to the number of moles of

molecules in the substance, and the carbon mole refers to the number of moles of carbon atoms in the substance.

Preferably, the molar ratio of all oxygen-containing compounds in the mixture fed from a plurality of second reactor feed distributors to the methanol in the mixture fed from the first reactor feed distributor into the turbulent fluidized bed reactor is greater than 1.

Further preferably, the molar ratio of all oxygen-containing compounds in the mixture fed from a plurality of second reactor feed distributors into the turbulent fluidized bed reactor to the methanol fed from the first reactor feed distributor is in a range from 1 to 20.

Further preferably, the spent catalyst passes through the reactor stripper, the inclined spent catalyst pipe, the spent catalyst sliding valve and the spent catalyst lift pipe into the dilute phase zone of the fluidized bed regenerator;

the regeneration medium enters the regeneration zone of the fluidized bed regenerator and reacts with the spent catalyst to removal coke, producing the flue gas containing CO and CO₂ and the regenerated catalyst, and the flue gas is discharged after dust removal by the regenerator gas-solid separator;

the regenerated catalyst passes through the regenerator stripper, the inclined regenerated catalyst pipe, the regenerated catalyst sliding valve and the regenerated catalyst lift pipe into the inlet of the first reactor gas-solid separator, and after the gas-solid separation, the regenerated catalyst enters the bottom of the reaction zone in the turbulent fluidized bed reactor;

the reactor stripping gas enters the reactor stripper via the reactor stripping gas inlet and contacts countercurrent with the spent catalyst, and then enters the turbulent fluidized bed reactor; the spent catalyst lifting gas enters the spent catalyst lift pipe via the spent catalyst lifting gas inlet and contacts cocurrent with the spent catalyst, and then enters the dilute phase zone of the fluidized bed regenerator;

the regenerator stripping gas enters the regenerator stripper via the regenerator stripping gas inlet and contacts countercurrent with the regenerated catalyst, and then enters the fluidized bed regenerator; the regenerated catalyst lifting gas enters the regenerated catalyst lift pipe via the regenerated catalyst lifting gas inlet and contacts concurrent with the regenerated catalyst, and then enters the inlet of the first reactor gas-solid separator, the first reactor gas-solid separator is placed in the dilute phase zone or outside the reactor shell of the fluidized bed reactor.

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

Further preferably, the regeneration medium is at least one of air, oxygen-poor air

or water vapor; and/or

the reactor stripping gas, the regenerator stripping gas, the spent catalyst lifting gas and the regenerated catalyst lifting gas are water vapor and/or nitrogen.

Preferably, the reaction conditions in the reaction zone of the turbulent fluidized bed reactor are: the apparent linear velocity of gas is in a range from 0.1m/s to 2.0m/s, the reaction temperature is in a range from 350°C to 600°C, the reaction pressure is in a range from 0.1MPa to 1.0MPa, and the bed density is in a range from 200kg/m³ to 1200kg/m³.

Preferably, the reaction conditions in the regeneration zone of the fluidized bed regenerator are: the apparent linear velocity of the gas is in a range from 0.1m/s to 2.0m/s, the regeneration temperature is in a range from 500°C to 750°C, the regeneration pressure is in a range from 0.1MPa to 1.0MPa, and the bed density is in a range from 200kg/m³ to 1200kg/m³.

In the present application, in the turbulent fluidized bed reactor, the catalyst in a fluidized state is in the dense phase zone of the lower part and the dilute phase zone of the upper part. The dense phase zone is the reaction zone of the turbulent fluidized bed reactor.

The present application provides a turbulent fluidized bed reactor for producing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene, the turbulent fluidized bed reactor comprises: a reactor shell 2, n reactor feed distributors (3-1~3-n), a reactor gas-solid separator 4, a reactor gas-solid separator 5, a reactor heat extractor (6), a product gas outlet 7 and a reactor stripper 8, wherein the lower part of the turbulent fluidized bed reactor 1 is a reaction zone, the upper part of the turbulent fluidized bed reactor 1 is a dilute phase zone, the n reactor feed distributors (3-1~3-n) are arranged from bottom to top in the reaction zone, the reactor heat extractor (6) is disposed in the reaction zone or outside the reactor shell 2, the reactor gas-solid separator 4 and the reactor gas-solid separator 5 are placed in the dilute phase zone or outside the reactor shell 2, the reactor gas-solid separator 4 is provided with the regenerated catalyst inlet, the catalyst outlet of the reactor gas-solid separator 4 is located at the bottom of the reaction zone, the gas outlet of the reactor gas-solid separator 4 is located in the dilute phase zone, the inlet of the reactor gas-solid separator 5 is located in the dilute phase zone, the catalyst outlet of the reactor gas-solid separator 5 is located in the reaction zone, the gas outlet of the reactor gas-solid separator 5 is connected to the product gas outlet 7, the reactor stripper 8 passes through the reactor shell from the outside to the inside at the bottom of the turbulent fluidized bed reactor and is opened in the reaction zone of the turbulent fluidized bed reactor 1, a reactor stripping gas inlet 9 are arranged at the bottom of the reactor stripper 8, and a spent catalyst outlet are arranged at the bottom of the reactor stripper.

In a preferred embodiment, the n reactor feed distributors(3-1~3-n) of the turbulent

fluidized bed reactor 1 are disposed in the reaction zone from bottom to top, $3 \leq n \leq 11$, and n is the total number of reactor feed distributors.

In a preferred embodiment, the horizontal height of opening of the reactor stripper 8 in the reactor shell 2 is higher than the height of the first reactor feed distributor, so as to avoid the direct entry of fresh catalyst into the reactor stripper.

In a preferred embodiment, the reactor gas-solid separator 4 and the reactor gas-solid separator 5 are cyclone separators.

The present application further provides a device for producing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene, the device comprising the turbulent fluidized bed reactor 1 described above and a fluidized bed regenerator 14 for regenerating a catalyst.

In a preferred embodiment, the fluidized bed regenerator 14 is a turbulent fluidized bed regenerator.

In a preferred embodiment, the fluidized bed regenerator 14 comprises a regenerator shell 15, a regenerator feed distributor 16, a regenerator gas-solid separator 17, a regenerator heat extractor 18, a flue gas outlet 19 and a regenerator stripper 20; wherein the lower part of the fluidized bed regenerator 14 is a regeneration zone, and the upper part of the fluidized bed regenerator 14 is a dilute phase zone; the regenerator feed distributor 16 is placed at the bottom of the regeneration zone, the regenerator heat extractor 18 is placed in the regeneration zone, the regenerator gas-solid separator 17 is placed in the dilute phase zone or outside the regenerator shell 15, the inlet of the regenerator gas-solid separator 17 is placed in the dilute phase zone, the catalyst outlet of the regenerator gas-solid separator 17 is placed in the regeneration zone, the gas outlet of the regenerator gas-solid separator 17 is connected to the flue gas outlet 19, and the inlet of the regenerator stripper 20 is connected to the bottom of the regenerator shell 15;

the spent catalyst outlet of the reactor stripper 8 is connected to the inlet of an inclined spent catalyst pipe 10, a spent catalyst sliding valve 11 is arranged in the inclined spent catalyst pipe 10, the outlet of the inclined spent catalyst pipe 10 is connected to the inlet of a spent catalyst lift pipe 12, the bottom of the spent catalyst lift pipe 12 is provided with a spent catalyst lifting gas inlet 13, and the outlet of the spent catalyst lift pipe 12 is connected to the dilute phase zone of the fluidized bed regenerator 14; and the bottom of the regenerator stripper 20 is provided with a regenerator stripping gas inlet 21, the bottom of the regenerator stripper 20 is connected to the inlet of an inclined regenerated catalyst pipe 22, a regenerated catalyst sliding valve 23 is arranged in the inclined regenerated catalyst pipe 22, the outlet of the inclined

regenerated catalyst pipe 22 is connected to the inlet of a regenerated catalyst lift pipe 24, the bottom of the regenerated catalyst lift pipe 24 is provided with a regenerated catalyst lifting gas inlet 25, the outlet of the regenerated catalyst lift pipe 24 is connected to the inlet of the reactor gas-solid separator 4.

Another aspect of the present application provides a method for producing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene, including:

sending a raw material containing methanol and/or dimethyl ether and benzene from the lowermost reactor feed distributor 3-1 of the turbulent fluidized bed reactor 1 into the reaction zone of the turbulent fluidized bed reactor 1, sending methanol from the reactor feed distributors 3-2 to 3-n in the turbulent fluidized bed reactor 1 into the reaction zone of the turbulent fluidized bed reactor 1, and contacting with a catalyst, to generate a material stream containing para-xylene and light olefins product and a spent catalyst containing carbon;

sending the material stream discharged from the turbulent fluidized bed reactor 1 containing para-xylene and light olefins product into a product separation system, obtaining para-xylene, ethylene, propylene, butene, C₅₊ chain hydrocarbons, aromatic hydrocarbon by-products and unconverted methanol, dimethyl ether and benzene after separation, in which aromatic by-products comprising toluene, o-xylene, m-xylene, ethylbenzene and C₉₊ aromatics, sending unconverted methanol and dimethyl ether from reactor feed distributor 3-2 to 3-n into the reaction zone of the turbulent fluidized bed reactor 1, sending the aromatic by-products and unconverted benzene from the reactor feed distributor 3-1 into the reaction zone of the turbulent fluidized bed reactor 1, and contacting with a catalyst to convert to product;

regenerating the spent catalyst by a fluidized bed regenerator 14, and after being gas-solid separated by a reactor gas-solid separator 4, the regenerated catalyst is fed to the bottom of the reaction zone in the turbulent fluidized bed reactor 1.

In a preferred embodiment, the method described in the present invention is carried out using the above-mentioned device for producing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene.

In a preferred embodiment, the spent catalyst passes through the reactor stripper 8, the inclined spent catalyst pipe 10, the spent catalyst sliding valve 11 and the spent catalyst lift pipe 12 into the dilute phase zone of the fluidized bed regenerator 14;

a regeneration medium enters the regeneration zone of the fluidized bed regenerator 14 from the regenerator feed distributor 16 and reacts with the spent catalyst to perform calcinations to produce the flue gas containing CO and CO₂ and the regenerated catalyst, and the

flue gas is discharged after dust removal by the regenerator gas-solid separator 17;

the regenerated catalyst passes through the regenerator stripper 20, the inclined regenerated catalyst pipe 22, the regenerated catalyst sliding valve 23 and the regenerated catalyst lift pipe 24 into the inlet of the reactor gas-solid separator 4, and after gas-solid
5 separation, the regenerated catalyst enters the bottom of the reaction zone in the turbulent fluidized bed reactor;

the reactor stripping gas enters the reactor stripper 8 via the reactor stripping gas inlet 9 and contacts countercurrent with the spent catalyst, and then enters the turbulent fluidized bed reactor 1; the spent catalyst lifting gas enters the spent catalyst lift pipe 12 via the spent
10 catalyst lifting gas inlet 13 and contacts concurrent with the spent catalyst, and then enters the dilute phase zone of the fluidized bed regenerator 14;

the regenerator stripping gas enters the regenerator stripper 20 via the regenerator stripping gas inlet 21 and contacts countercurrent with regenerated catalyst, and then enters the fluidized bed regenerator 14; the regenerated catalyst lifting gas enters the regenerated catalyst
15 lift pipe 24 via the regenerated catalyst lifting inlet 25 and contacts concurrent with the regenerated catalyst, and then enters the inlet of the reactor gas-solid separator 4.

In the method for producing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene as described herein, in the mixture entering from the lowest reactor feed distributor 3-1 of the turbulent fluidized bed reactor, the amount of
20 substance ratio of the aromatics to methanol is greater than 0.5, further preferably, greater than 1.

In the method for producing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene as described herein, the amount of substance ratio of the oxygen-containing compounds entering from the reactor feed distributors 3-2 to 3-n and
25 methanol entering from the reactor feed distributor 3-1 is greater than 1, more preferably greater than 5.

In a preferred embodiment, the catalyst comprises a HZSM-5 molecular sieve having both the function of alkylation of methanol and/or dimethyl ether with benzene, and aromatization of preparing olefins from methanol and methanol.

In a preferred embodiment, the catalyst comprises a HZSM-11 molecular sieve having both the function of alkylation of methanol and/or dimethyl ether with benzene, and aromatization of preparing olefins from methanol and methanol.

In a preferred embodiment, the carbon content of the regenerated catalyst is less than 0.5wt.%, and further preferably, the carbon content of the regenerated catalyst is less than

0.1 wt.%.

In a preferred embodiment, the reaction conditions in the reaction zone of the turbulent fluidized bed reactor are as follows: the apparent linear velocity of gas is in a range from 0.1 m/s to 2.0 m/s, the reaction temperature is in a range from 350°C to 600°C, the reaction pressure is in a range from 0.1MPa to 1.0MPa, and the bed density is in a range from 200kg/m³ to 1200kg/m³.

In a preferred embodiment, the reaction conditions in the regeneration zone of the fluidized bed regenerator are as follows: the apparent linear velocity of gas is in a range from 0.1 m/s to 2 m/s, the reaction temperature is in a range from 500°C to 750°C, the reaction pressure is in a range from 0.1MPa to 1.0MPa, and the bed density is in a range from 200kg/m³ to 1200kg/m³.

In a preferred embodiment, the regeneration medium is any one of air, oxygen-poor air or water vapor or a mixture thereof; the reactor stripping gas, the regenerator stripping gas, the spent catalyst lifting gas and the regenerated catalyst lifting gas are water vapor or nitrogen.

In the method for producing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene as described in the present application, the conversion rate of benzene is greater than 40%, the conversion rate of methanol is greater than 70%, and the selectivity of para-xylene is greater than 90%, the mass single-pass yield of para-xylene based on aromatics is greater than 32%, and the selectivity of light olefins (ethylene + propylene + butene) in chain hydrocarbons is greater than 70%, and good technical results have been achieved.

The main characteristics of the turbulent fluidized bed reactor in the present application are that the aromatic feedstock enter from the reactor feed distributor at the bottom-most, the oxygen-containing compound enters from n reactor feed distributors respectively, and the highly active regenerated catalyst directly enters the bottom of the reaction zone. The aromatic feedstock comprises fresh benzene, unconverted benzene and aromatic by-products, and the oxygen-containing compound comprises fresh methanol, unconverted methanol and dimethyl ether. First, in the lower part of the reaction zone, the catalyst has a high activity, which is advantageous to the alkylation reaction of benzene, the isomerization reaction of aromatic hydrocarbon by-products and methyl transfer reaction, etc.; second, using the method of multi-stage feeding of the oxygen-containing compound, only a small portion of the oxygen-containing compound is fed from the bottom of the reactor, the low concentration of the oxygen-containing compound in the bottom region and the high concentration of aromatics weaken the adsorption competition of the oxygen-containing compound with fast diffusion rate

in the molecular sieve pores to the aromatic hydrocarbons with slow diffusion rate, so as to ensure that most of the aromatics are adsorbed in the catalyst in the bottom region; third, most of the oxygen-containing compounds are fed from the middle part and the upper part, the conversion reaction of the oxygen-containing compound mainly occurs in the middle part and the upper part of the reaction zone, so as to avoid rapidly decreased activity of the high activity of the regenerated catalyst in the bottom zone due to the carbon formation during the MTO reaction; fourth, the amount of carbon in the catalyst is higher in the middle part and the upper part of the reaction zone, which is advantageous to improve the selectivity of light olefins in the MTO reaction; fifth, using the method of multi-stage feeding of the oxygen-containing compound which is advantageous to improve the selectivity of light olefins in the MTO reaction due to higher carbon content of the catalyst in the middle and upper part of the reaction zone, the concentration distribution of the oxygen-containing compound in the reaction zone is relatively uniform, providing sufficient alkylation reactants. After the aromatic hydrocarbons adsorbed in the catalyst are contacted with alkylation reactants, the alkylation reaction occurs quickly to improve the conversion rate of benzene and the yield of para-xylene.

From the above, the turbulent fluidized bed reactor of the present application may coordinate and optimize the competition between the alkylation reaction of methanol and/or dimethyl ether and benzene and the MTO reaction to achieve a synergistic effect, thereby improving the conversion rate of the benzene, the yield of para-xylene and the selectivity of light olefins.

The present application coordinates and optimizes the competition between the alkylation reaction and the MTO reaction by controlling the concentrations of methanol and/or dimethyl ether relative to benzene from the viewpoint of reactor design and process configuration, and improves the yield of para-xylene and the selectivity of light olefins to ensure that neither the situation of the inhibition of the alkylation reaction occurs due to the rapid consumption of most methanol and/or dimethyl ether by the MTO reaction, nor the situation against the alkylation reaction occurs due to far excess content of methanol and/or dimethyl ether, a large number of the MTO reaction occur, and lower adsorption amount of benzene in the catalyst per unit time.

The benefits brought out by the present application include:

(1) This is provided a fluidized bed reactor and device to achieve mass transfer control by distributing different raw materials stream in different regions in a co-feed system with a large difference in raw material reaction rates, so as to coordinate and optimize a co-feed system and improve the reaction yield.

(2) The method for producing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene, provided by the present application, has higher conversion rate of benzene and the selectivity of para-xylene, the conversion rate of benzene is greater than 40%, the selectivity of para-xylene is greater than 90%, the mass single-pass yield of para-xylene based on aromatics is greater than 32%, and good technical results have been achieved.

BRIEF DESCRIPTION OF THE DRAWINGS

FIG. 1 is a schematic diagram of a device for producing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene according to an embodiment of the present application.

The reference numerals in the figures are listed as follows:

1-turbulent fluidized bed reactor; 2-reactor shell; 3-reactor feed distributors (3-1~3-n); 4-reactor gas-solid separator; 5-reactor gas-solid separator; 6-reactor heat extractor; 7-product gas outlet; 8-reactor stripper; 9-reactor stripping gas inlet; 10-inclined spent catalyst pipe; 11-spent catalyst sliding valve; 12-spent catalyst lift pipe; 13-spent catalyst lifting gas inlet; 14-fluidized bed regenerator; 15-regenerator shell; 16-regenerator feed distributor; 17-regenerator gas-solid separator; 18-regenerator heat extractor; 19-flue gas outlet; 20-regenerator stripper; 21-regenerator stripping gas inlet; 22-inclined regenerated catalyst pipe; 23-regenerated catalyst sliding valve; 24-regenerated catalyst lift pipe; 25-regenerated catalyst lifting gas inlet; 26-perforated plate.

DETAILED DESCRIPTION OF THE EMBODIMENT

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

Unless otherwise specified, the raw materials and catalysts in the embodiments of the present application are commercially available.

As an embodiment of the present application, a schematic diagram of a device for producing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene is shown in FIG. 1. The device comprises the turbulent fluidized bed reactor 1, which comprises a reactor shell 2, n reactor feed distributors 3-1 to 3-n (the distributor between 3-1 and 3-n in FIG. 1 takes 3-i as an example), a reactor gas-solid separator 4, a reactor gas-solid separator 5, a reactor heat extractor 6, a product gas outlet 7 and a reactor stripper 8 and a

perforated plate, wherein the lower part of the turbulent fluidized bed reactor 1 is a reaction zone, the upper part of the turbulent fluidized bed reactor 1 is a dilute phase zone, the n reactor feed distributors of 3-1 to 3- n are arranged from bottom to top in the reaction zone, $2 \leq n \leq 11$, the perforated plate 26 is placed between the reactor feed distributor 3-1 and the reactor feed distributor 3-2, the reactor heat extractor 6 is disposed in the reaction zone or outside the reactor shell 2, the reactor gas-solid separator 4 and the reactor gas-solid separator 5 are placed in the dilute phase zone or outside the reactor shell 2, the inlet of the reactor gas-solid separator 4 is connected to a regenerated catalyst lift pipe 24, the catalyst outlet of the reactor gas-solid separator 4 is located at the bottom of the dilute phase zone, the gas outlet of the reactor gas-solid separator 4 is located in the dilute phase zone, the inlet of the reactor gas-solid separator 5 is located in the dilute phase zone, the catalyst outlet of the reactor gas-solid separator 5 is located in the reaction zone, the gas outlet of the reactor gas-solid separator 5 is connected to the product gas outlet 7, and the inlet of the reactor stripper 8 is in the reaction zone of the turbulent fluidized bed reactor 1, with the horizontal height higher than that of the first reactor distributor and higher than that of the perforated plate 26.

As shown in FIG. 1, the device comprises: a fluidized bed regenerator 14 comprising a regenerator shell 15, a regenerator feed distributor 16, a regenerator gas-solid separator 17, a regenerator heat extractor 18, a flue gas outlet 19 and a regenerator stripper 20, wherein the lower part of the fluidized bed regenerator 14 is a regeneration zone, the upper part of the fluidized bed regenerator 14 is a dilute phase zone, the regenerator feed distributor 16 is placed at the bottom of the regeneration zone, the regenerator heat extractor 18 is placed in the regeneration zone, the regenerator gas-solid separator 17 is placed in the dilute phase zone or outside the regenerator shell 15, the inlet of the regenerator gas-solid separator 17 is placed in the dilute phase zone, the catalyst outlet of the regenerator gas-solid separator 17 is placed in the regeneration zone, the gas outlet of the regenerator gas-solid separator 17 is connected to the flue gas outlet 19, and the inlet of the regenerator stripper 20 is connected to the bottom of the regenerator shell 15.

As shown in FIG. 1, the bottom of the reactor stripper 8 is provided with a reactor stripping gas inlet 9, the bottom of the reactor stripper 8 is connected to the inlet of an inclined spent catalyst pipe 10, a spent catalyst sliding valve 11 is arranged in the inclined spent catalyst pipe 10, the outlet of the inclined spent catalyst pipe 10 is connected to the inlet of a spent catalyst lift pipe 12, the bottom of the spent catalyst lift pipe 12 is provided with a spent catalyst lifting gas inlet 13, and the outlet of the spent catalyst lift pipe 12 is connected to the dilute phase zone of the fluidized bed regenerator 14.

As shown in FIG. 1, the bottom of the regenerator stripper 20 is provided with a regenerator stripping gas inlet 21, the bottom of the regenerator stripper 20 is connected to the inlet of an inclined regenerated catalyst pipe 22, a regenerated catalyst sliding valve 23 is arranged in the inclined regenerated catalyst pipe 22, the outlet of the inclined regenerated catalyst pipe 22 is connected to the inlet of the regenerated catalyst lift pipe 24, the bottom of the regenerated catalyst lift pipe 24 is provided with a regenerated catalyst lifting gas inlet 25, and the outlet of the regenerated catalyst lift pipe 24 is connected to the inlet of the reactor gas-solid separator 4.

In the above embodiment as the present application, the fluidized bed regenerator 14 may be a turbulent fluidized bed regenerator; the reactor gas-solid separator 4, the reactor gas-solid separator 5 and the regenerator gas-solid separator 17 may be cyclone separators.

As a specific embodiment of the present application, the method according to the present application for producing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene includes:

a) sending a raw material containing methanol and/or dimethyl ether and benzene from the lowermost reactor feed distributor 3-1 of the turbulent fluidized bed reactor 1 into the reaction zone of the turbulent fluidized bed reactor 1, sending methanol from the reactor feed distributors 3-2 to 3-n in the turbulent fluidized bed reactor 1 into the reaction zone of the turbulent fluidized bed reactor 1, and contacting with a catalyst, to generate a material stream containing para-xylene and light olefins product and a spent catalyst containing carbon;

b) sending the material stream discharged from the turbulent fluidized bed reactor 1 containing para-xylene and light olefins product into a product separation system, obtaining para-xylene, ethylene, propylene, butene, C₅₊ chain hydrocarbons, aromatic hydrocarbon by-products and unconverted methanol, dimethyl ether and benzene after separation, in which aromatic by-products comprising toluene, o-xylene, m-xylene, ethylbenzene and C₉₊ aromatics, sending unconverted methanol and dimethyl ether from reactor feed distributor 3-2 to 3-n into the reaction zone of the turbulent fluidized bed reactor 1, sending the aromatic by-products and unconverted benzene from the reactor feed distributor 3-1 into the reaction zone of the turbulent fluidized bed reactor 1, and contacting with a catalyst to convert to product;

c) the spent catalyst passes through the reactor stripper 8, the inclined spent catalyst pipe 10, the spent sliding valve 11 and the spent catalyst lift pipe 12 into the dilute phase zone of the fluidized bed regenerator 14;

d) a regeneration medium enters the regeneration zone of the fluidized bed regenerator 14 from the regenerator feed distributor 16, the regeneration medium reacts with the

spent catalyst to removal coke, producing a flue gas containing CO and CO₂ and a regenerated catalyst, and the flue gas is discharged after dust removal by the regenerator gas-solid separator 17;

e) the regenerated catalyst passes through the regenerator stripper 20, the inclined regenerated catalyst pipe 22, the regenerated catalyst sliding valve 23 and the regenerated catalyst lift pipe 24 into the inlet of the reactor gas-solid separator 4, and after gas-solid separation, the regenerated catalyst enters the bottom of the reaction zone of the turbulent fluidized bed reactor 1;

f) the reactor stripping gas enters the reactor stripper 8 via the reactor stripping gas inlet 9 and contacts countercurrent with the spent catalyst, and then enters the turbulent fluidized bed reactor 1; the spent catalyst lifting gas enters the spent catalyst lift pipe 12 via the spent catalyst lifting gas inlet 13 and contacts cocurrent with the spent catalyst, and then enters the dilute phase zone of the fluidized bed regenerator 14;

g) the regenerator stripping gas enters the regenerator stripper 20 via the regenerator stripping gas inlet 21 and contacts countercurrent with the regenerated catalyst, and then enters the fluidized bed regenerator 14; the regenerated catalyst lifting gas enters the regenerated catalyst lift pipe 24 via the regenerated catalyst lifting gas inlet 25 and contacts cocurrent with the regenerated catalyst, and then enters the inlet of the reactor gas-solid separator 4.

In order to better illustrate the present application and facilitate the understanding of the technical scheme of the present application, representative but non-restrictive examples of the present application are listed as follows:

EXAMPLE 1

The device shown in FIG. 1 is used, but the turbulent fluidized bed reactor 1 does not contain the reactor gas-solid separator 4 and the perforated plate 26, and the regenerated catalyst lift pipe 24 is directly connected to the dilute phase zone of the turbulent fluidized bed reactor 1. The turbulent fluidized bed reactor 1 contains one reactor feed distributor 3-1.

The reaction conditions in the reaction zone of the turbulent fluidized bed reactor 1 are as follows: the apparent linear velocity of gas is about 1.0m/s, the reaction temperature is about 500°C, the reaction pressure is about 0.15 MPa, and the bed density is about 350 kg/m³.

The reaction conditions in the regeneration zone of the fluidized bed regenerator 14 are as follows: the apparent linear velocity of the gas is about 1.0 m/s, the regeneration temperature is about 650°C, the regeneration pressure is about 0.15MPa, and the bed density is about 350kg/m³.

The catalyst contains a HZSM-5 molecular sieve. The carbon content of the regenerated catalyst is about 0.1 wt.%.

The regeneration medium is air; the reactor stripping gas, the regenerator stripping gas, the spent catalyst lifting gas and the regenerated catalyst lifting gas are water vapor.

5 In the mixture entering from the lowest reactor feed distributor 3-1 of the turbulent fluidized bed reactor, the molar ratio of the aromatics to methanol is 0.5.

The results show that the conversion rate of benzene is 20%, the conversion rate of methanol is 99%, the selectivity of para-xylene is 98%, and the mass single-pass yield of para-xylene based on aromatics is 15%, and the selectivity of light olefins (ethylene + propylene
10 + butene) in chain hydrocarbons is 66%.

EXAMPLE 2

The device shown in FIG. 1 is used, the turbulent fluidized bed reactor 1 contains three reactor feed distributors 3-1 to 3-3, the porosity of the perforated plate 26 is 10%, and the
15 reactor gas-solid separator 4 is placed inside the reactor shell 2.

The reaction conditions in the reaction zone of the turbulent fluidized bed reactor 1 are as follows: the apparent linear velocity of gas is about 1.0m/s, the reaction temperature is about 500°C, the reaction pressure is about 0.15 MPa, and the bed density is about 350 kg/m³.

The reaction conditions in the regeneration zone of the fluidized bed regenerator 14
20 are as follows: the apparent linear velocity of the gas is about 1.0 m/s, the regeneration temperature is about 650°C, the regeneration pressure is about 0.15MPa, and the bed density is about 350kg/m³.

The catalyst contains a HZSM-5 molecular sieve. The carbon content of the regenerated catalyst is about 0.1 wt.%.

25 The regeneration medium is air; the reactor stripping gas, the regenerator stripping gas, the spent catalyst lifting gas and the regenerated catalyst lifting gas are water vapor.

In the mixture entering from the lowest reactor feed distributor 3-1 of the turbulent fluidized bed reactor, the molar ratio of the aromatics to methanol is 2.

The molar ratio of the oxygen-containing compounds entering from the reactor feed
30 distributors 3-2 to 3-3 and methanol entering from the reactor feed distributor 3-1 is 3.

The results show that the conversion rate of benzene is 41%, the conversion rate of methanol is 97%, the selectivity of para-xylene is 98%, and the mass single-pass yield of para-xylene based on aromatics is 32%, and the selectivity of light olefins (ethylene + propylene
+ butene) in chain hydrocarbons is 75%.

The present example is different from Example 1 in that

① the regenerated catalyst enters the bottom of the turbulent fluidized bed reactor, while in Example 1, the regenerated catalyst enters the dilute phase zone of the turbulent fluidized bed reactor;

② methanol is separately fed from three reactor feed distributors (3-1 to 3-3), while in Example 1, methanol is fed from one reactor feed distributor 3-1.

③ the perforated plate was contained, and perforated plate was not contained in Example 1.

Comparing the present example with Example 1, it can be seen that the catalyst is first exposed to a high concentration of aromatic raw material, and the conversion rate of benzene, the yield of para-xylene and the selectivity of light olefins are greatly improved.

EXAMPLE 3

The device shown in FIG. 1 is used, the turbulent fluidized bed reactor 1 contains six reactor feed distributors 3-1 to 3-6, the porosity of the perforated plate is 5%, and the reactor gas-solid separator 4 is placed inside the reactor shell 2.

The reaction conditions in the reaction zone of the turbulent fluidized bed reactor 1 are as follows: the apparent linear velocity of gas is about 0.8 m/s, the reaction temperature is about 560°C, the reaction pressure is about 0.6 MPa, and the bed density is about 460 kg/m³.

The reaction conditions in the regeneration zone of the fluidized bed regenerator 14 are as follows: the apparent linear velocity of the gas is about 1.7 m/s, the regeneration temperature is about 600°C, the regeneration pressure is about 0.6 MPa, and the bed density is about 220 kg/m³.

The catalyst contains a HZSM-11 molecular sieve. The carbon content of the regenerated catalyst is about 0.15 wt.%.

The regeneration medium is air; the reactor stripping gas, the regenerator stripping gas, the spent catalyst lifting gas and the regenerated catalyst lifting gas are water vapor.

In the mixture entering from the lowest reactor feed distributor 3-1 of the turbulent fluidized bed reactor, the molar ratio of the aromatics to methanol is 4.

The molar ratio of the oxygen-containing compounds entering from the reactor feed distributors 3-2 to 3-6 and methanol entering from the reactor feed distributor 3-1 is 20.

The results show that the conversion rate of benzene is 45%, the conversion rate of methanol is 73%, the selectivity of para-xylene is 94%, and the mass single-pass yield of para-xylene based on aromatics is 42%, and the selectivity of light olefins (ethylene + propylene

+ butene) in chain hydrocarbons is 73%.

EXAMPLE 4

The device shown in FIG. 1 is used, the turbulent fluidized bed reactor 1 contains four reactor feed distributors 3-1 to 3-4, without the perforated plate 26, and the reactor gas-solid separator 4 is placed outside the reactor shell 2.

The reaction conditions in the reaction zone of the turbulent fluidized bed reactor 1 are as follows: the apparent linear velocity of gas is about 1.5 m/s, the reaction temperature is about 440°C, the reaction pressure is about 0.2 MPa, and the bed density is about 280 kg/m³.

The reaction conditions in the regeneration zone of the fluidized bed regenerator 14 are as follows: the apparent linear velocity of the gas is about 1.2 m/s, the regeneration temperature is about 700°C, the regeneration pressure is about 0.2 MPa, and the bed density is about 330 kg/m³.

The catalyst contains a HZSM-5 molecular sieve. The carbon content of the regenerated catalyst is about 0.15 wt.%.

The regeneration medium is air; the reactor stripping gas, the regenerator stripping gas, the spent catalyst lifting gas and the regenerated catalyst lifting gas are nitrogen.

In the mixture entering from the lowest reactor feed distributor 3-1 of the turbulent fluidized bed reactor, the molar ratio of the aromatics to methanol is 3.

The molar ratio of the oxygen-containing compounds entering from the reactor feed distributors 3-1 to 3-4 and methanol entering from the reactor feed distributor 3-1 is 10.

The results show that the conversion rate of benzene is 42%, the conversion rate of methanol is 86%, the selectivity of para-xylene is 91%, and the single-pass yield of para-xylene based on aromatics is 38%, and the selectivity of light olefins (ethylene + propylene + butene) in chain hydrocarbons is 71%.

While the present application has been described above with reference to preferred embodiments, but these embodiments are not intended to limit the claims. Without departing from the spirit of the present application, people skilled in the art will be able to make several possible variations and modifications and thus the protection scope shall be determined by the scope as defined in the claims.

What is claimed is:

1. A turbulent fluidized bed reactor for preparing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene, wherein the turbulent fluidized bed reactor comprises a reaction zone, a dilute phase zone, a first reactor feed distributor and a plurality of second reactor feed distributors, the first reactor feed distributor and the plurality of second reactor feed distributors are sequentially arranged from bottom to top in the reaction zone; wherein the reaction zone is located in a lower part of the turbulent fluidized bed reactor, and the dilute phase zone is located in an upper part of the turbulent fluidized bed reactor;

the number of the second reactor feed distributors is in a range from 2 to 10,

wherein the feed to the first reactor feed distributor comprises methanol and/or dimethyl ether and benzene;

wherein the feed to the second reactor feed distributor comprises methanol and/or dimethyl ether;

wherein the turbulent fluidized bed reactor comprises a perforated plate located between the first reactor feed distributor and at least one of the second reactor feed distributors, the porosity of the perforated plate is less than or equal to 50%;

wherein the turbulent fluidized bed reactor comprises a first reactor gas-solid separator, the first reactor gas-solid separator is placed in the dilute phase zone, the first reactor gas-solid separator is provided with a regenerated catalyst inlet, a regenerated catalyst outlet of the first reactor gas-solid separator is placed at the bottom of a reaction zone and between the first reactor feed distributor and the perforated plate, and a gas outlet of the first reactor gas-solid separator is placed in the dilute phase zone.

2. The turbulent fluidized bed reactor of claim 1, wherein the turbulent fluidized bed reactor further comprises a second reactor gas-solid separator, the second reactor gas-solid separator is placed in the dilute phase zone or outside the reactor shell;

an inlet of the second reactor gas-solid separator is placed in the dilute phase zone, a catalyst outlet of the second reactor gas-solid separator is placed in the reaction zone, and a gas outlet of the second reactor gas-solid separator is connected

to a product gas outlet of the turbulent fluidized bed reactor;

the first reactor gas-solid separator and the second reactor gas-solid separator are cyclone separators.

5 3. The turbulent fluidized bed reactor of claim 1 or 2, wherein the turbulent fluidized bed reactor comprises a reactor heat extractor, and the reactor heat extractor is arranged inside or outside the shell of the turbulent fluidized bed reactor.

 4. The turbulent fluidized bed reactor of claim 3, wherein
10 the reactor heat extractor is arranged between the plurality of reactor feed distributors;

 the turbulent fluidized bed reactor comprises a reactor stripper, the reactor stripper passes through the reactor shell from the outside to the inside at the bottom of the turbulent fluidized bed reactor and is opened in the reaction zone of the
15 turbulent fluidized bed reactor, and a reactor stripping gas inlet and a spent catalyst outlet are arranged at the bottom of the reactor stripper.

 5. The turbulent fluidized bed reactor of any one of claims 1 to 4, wherein the perforated plate is located between the first reactor feed distributor and
20 the second reactor feed distributor closest to the first reactor feed distributor, the porosity of the perforated plate is in a range from 5% to 50%.

 6. The turbulent fluidized bed reactor of claim 4 or 5, wherein the reactor stripper passes through the reactor shell from the outside to the inside at the
25 bottom of the turbulent fluidized bed reactor and is opened in the reaction zone of the turbulent fluidized bed reactor, and a reactor stripping gas inlet and a spent catalyst outlet are arranged at the bottom of the reactor stripper;

 the horizontal height of opening of the reactor stripper in the reactor shell is higher than that of the first reactor feed distributor and higher than that of the
30 perforated plate.

 7. A device for preparing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene, wherein the device comprises at

least one of the turbulent fluidized bed reactor according to any one of claims 1 to 6 and a fluidized bed regenerator for regenerating a catalyst.

8. The device of claim 7, wherein the fluidized bed regenerator is a
5 turbulent fluidized bed regenerator, and the fluidized bed regenerator comprises a regenerator shell, a regenerator gas-solid separator, a regenerator heat extractor and the regenerator stripper; the lower part of the fluidized bed regenerator is a regeneration zone, the upper part of the fluidized bed regenerator is a dilute phase zone of the regenerator, the regenerator feed distributor is placed at the bottom of the
10 regeneration zone, the regenerator heat extractor is placed in the regeneration zone, and the regenerator gas-solid separator is placed in the dilute phase zone or outside the regenerator shell;

the inlet of the regenerator gas-solid separator is placed in the dilute phase zone of the regenerator, the catalyst outlet of the regenerator gas-solid separator
15 is placed in the regeneration zone, and the regenerator stripper is opened at the bottom of the regenerator shell.

9. The device of claim 7, wherein the fluidized bed regenerator comprises a regenerator shell, a regenerator feed distributor, a regenerator gas-solid
20 separator, a regenerator heat extractor, a flue gas outlet and a regenerator stripper;

the lower part of the fluidized bed regenerator is a regeneration zone, and the upper part of the fluidized bed regenerator is a dilute phase zone;

the regenerator feed distributor is placed at the bottom of the regeneration zone, the regenerator heat extractor is placed in the regeneration zone, the regenerator
25 gas-solid separator is placed in the dilute phase zone or outside the regenerator shell, the inlet of the regenerator gas-solid separator is placed in the dilute phase zone, the catalyst outlet of the regenerator gas-solid separator is placed in the regeneration zone, the gas outlet of the regenerator gas-solid separator is connected to the flue gas outlet, and the regenerator stripper is opened at the bottom of the regenerator shell;

30 the spent catalyst outlet of the reactor stripper is connected to the inlet of an inclined spent catalyst pipe, a spent catalyst sliding valve is arranged in the inclined spent catalyst pipe, the outlet of the inclined spent catalyst pipe is connected to the inlet of a spent catalyst lift pipe, the bottom of the spent catalyst lift pipe is

provided with a spent catalyst lifting gas inlet, and the outlet of the spent catalyst lift pipe is connected to the dilute phase zone of the fluidized bed regenerator;

the bottom of the regenerator stripper is provided with a regenerator stripping gas inlet, the bottom of the regenerator stripper is connected to the inlet of an inclined regenerated catalyst pipe, a regenerated catalyst sliding valve is arranged in the inclined regenerated catalyst pipe, the outlet of the inclined regenerated catalyst pipe is connected to the inlet of a regenerated catalyst lift pipe, the bottom of the regenerated catalyst lift pipe is provided with a regenerated catalyst lifting gas inlet, the outlet of the regenerated catalyst lift pipe is connected to the regenerated catalyst inlet of the first reactor gas-solid separator, and the first reactor gas-solid separator is placed in the dilute phase zone of the fluidized bed reactor.

10. A method for preparing para-xylene and co-producing light olefins from methanol and/or dimethyl ether and benzene, wherein at least one of the turbulent fluidized bed reactors according to any one of claims 1 to 6 is used;

a raw material A containing methanol and/or dimethyl ether and benzene is fed into the reaction zone of the turbulent fluidized bed reactor from the first reactor feed distributor to be in contact with the regenerated catalyst supplied from the regenerated catalyst outlet of the first reactor gas-solid separator and a raw material B containing methanol and/or dimethyl ether is respectively fed into the reaction zone of the turbulent fluidized bed reactor from a plurality of second reactor feed distributors respectively to be in contact with the catalyst, to form a material stream C containing para-xylene and light olefins products and a spent catalyst.

11. The method of claim 10, wherein the material stream C is separated to obtain para-xylene, light olefins, C₅₊ chain hydrocarbons, aromatic by-products and unconverted methanol, dimethyl ether and benzene;

wherein the unconverted methanol and dimethyl ether are fed into the reaction zone of the turbulent fluidized bed reactor from a plurality of second reactor feed distributors, the aromatic by-products and the unconverted benzene are fed into the reaction zone of the turbulent fluidized bed reactor from the first reactor feed distributor to be in contact with a catalyst.

12. The method of claim 10 or 11, wherein the spent catalyst is regenerated by a fluidized bed regenerator and fed to the bottom of the reaction zone of the turbulent fluidized bed reactor.

5 13. The method of claim 10, wherein the method comprises the steps of:

(1) feeding a material stream A containing methanol and/or dimethyl ether and benzene into the reaction zone of the turbulent fluidized bed reactor from a first reactor feed distributor below the turbulent fluidized bed reactor to be in contact with the catalyst;

10 (2) feeding a material stream B containing methanol and/or dimethyl into the reaction zone of the turbulent fluidized bed reactor from 2 to 10 second reactor feed distributors to be in contact with the catalyst, to form a material stream C containing para-xylene and light olefins products and a spent catalyst; the 2 to 10 second reactor feed distributors are arranged in sequence above the first reactor feed
15 distributor;

(3) separating the material stream C obtained from the step (2) to obtain a material stream C-1 containing unconverted methanol and dimethyl, a material stream C-2 containing unconverted benzene and aromatic by-products; the material stream C-1 is respectively fed into the reaction zone of the turbulent fluidized bed reactor from the 2 to 10 second reactor feed distributors to be in contact with the catalyst; the
20 material stream C-2 is fed into the reaction zone of the turbulent fluidized bed reactor from the first reactor feed distributor to be in contact with the catalyst;

the aromatic by-products comprise toluene, o-xylene, m-xylene, ethylbenzene and C₉₊ aromatics;

25 (4) regenerating the spent catalyst obtained from the step (2) by the fluidized bed regenerator, the regenerated catalyst is fed to the first reactor gas-solid separator to removal the gas, and then the regenerated catalyst is fed to the bottom of the reaction zone in the fast fluidized bed reactor.

30 14. The method of claim 10, wherein in the mixture fed from the first reactor feed distributor into the turbulent fluidized bed reactor, the ratio of the molecular moles of aromatics to the carbon moles of methanol and/or dimethyl ether is greater than 0.5.

15. The method of claim 10, wherein the molar ratio of all oxygen-containing compounds in the mixture fed from a plurality of second reactor feed distributors into the turbulent fluidized bed reactor to the methanol fed from the first reactor feed distributor is greater than 1.

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16. The method of claim 13, wherein the catalyst regeneration employs at least one of the devices according to any one of claims 7 to 9.

17. The method of claim 13, wherein the spent catalyst passes through
10 the reactor stripper, the inclined spent catalyst pipe, the spent catalyst sliding valve and the spent catalyst lift pipe into the dilute phase zone of the fluidized bed regenerator;

the regeneration medium enters the regeneration zone of the fluidized bed regenerator and reacts with the spent catalyst to remove coke, producing the flue gas
15 containing CO and CO₂ and the regenerated catalyst, and the flue gas is discharged after dust removal by the regenerator gas-solid separator;

the regenerated catalyst passes through the regenerator stripper, the inclined regenerated catalyst pipe, the regenerated catalyst sliding valve and the regenerated catalyst lift pipe into the inlet of the first reactor gas-solid separator, and
20 after the gas-solid separation, the regenerated catalyst enters the bottom of the reaction zone in the turbulent fluidized bed reactor;

the reactor stripping gas enters the reactor stripper via the reactor stripping gas inlet and contacts countercurrent with the spent catalyst, and then enters the turbulent fluidized bed reactor; the spent catalyst lifting gas enters the spent
25 catalyst lift pipe via the spent catalyst lifting gas inlet and contacts cocurrent with the spent catalyst, and then enters the dilute phase zone of the fluidized bed regenerator;

the regenerator stripping gas enters the regenerator stripper via the regenerator stripping gas inlet and contacts countercurrent with the regenerated catalyst, and then enters the fluidized bed regenerator; the regenerated catalyst lifting
30 gas enters the regenerated catalyst lift pipe via the regenerated catalyst lifting gas inlet and contacts concurrent with the regenerated catalyst, and then enters the inlet of the first reactor gas-solid separator, the first reactor gas-solid separator is placed in the dilute phase zone of the fluidized bed reactor;

the carbon content of the regenerated catalyst is less than or equal to 0.5wt%.

18. The method of claim 17, wherein the regeneration medium is at least one of air, oxygen-poor air or water vapor; and/or the reactor stripping gas, the regenerator stripping gas, the spent catalyst lifting gas and the regenerated catalyst lifting gas are water vapor and/or nitrogen.

19. The method of claim 17, wherein the reaction conditions in the reaction zone of the turbulent fluidized bed reactor are: the apparent linear velocity of gas is in a range from 0.1m/s to 2.0m/s, the reaction temperature is in a range from 350°C to 600°C, the reaction pressure is in a range from 0.1MPa to 1.0MPa, and the bed density is in a range from 200kg/m³ to 1200kg/m³.

20. The method of claim 17, wherein the reaction conditions in the regeneration zone of the fluidized bed regenerator are: the apparent linear velocity of the gas is in a range from 0.1m/s to 2.0m/s, the regeneration temperature is in a range from 500°C to 750°C, the regeneration pressure is in a range from 0.1MPa to 1.0MPa, and the bed density is in a range from 200kg/m³ to 1200kg/m³.