CLAIMS

WHAT IS CLAIMED IS:

1. A coke-control reactor, wherein the coke-control reactor comprises a riser reactor and a bed reactor;

the bed reactor comprises a bed reactor shell, and the bed reactor shell encloses a reaction zone I, a transition zone, and a gas-solid separation zone I from bottom to top;

a bed reactor distributor is arranged in an inner lower part of the reaction zone I;

a coke controlled catalyst delivery pipe is arranged outside the reaction zone I;

an upper section of the riser reactor penetrates through a bottom of the bed reactor and is axially inserted in the bed reactor; and

an outlet end of the riser reactor is located in the transition zone.

2. The coke-control reactor according to claim 1, wherein at least one perforated plate is arranged in the reaction zone I; the plurality of perforated plates are axially arranged on a periphery of the riser reactor in sequence;

the outlet end of the riser reactor is located above the perforated plate;

the bed reactor distributor is located below the perforated plate;

preferably, the perforated plate has a porosity of 1% to 30%.

3. The coke-control reactor according to claim 1, wherein the gas-solid separation zone I is provided with a bed reactor gas-solid separator and a bed reactor gas collection chamber;

a gas outlet of the bed reactor gas-solid separator communicates with the bed reactor gas collection chamber;

a catalyst outlet of the bed reactor gas-solid separator is located above the perforated plate;

the bed reactor gas collection chamber communicates with a coke control product gas delivery pipe located outside the bed reactor;

preferably, the bed reactor distributor is configured to feed a bed reactor raw material; and

the bed reactor raw material comprises 0 wt% to 20 wt% of hydrogen, 0 wt% to 50 wt% of methane, 0 wt% to 50 wt% of ethane, 0 wt% to 20 wt% of ethylene, 0 wt% to 50 wt% of propane, 0 wt% to 20 wt% of propylene, 0 wt% to 90 wt% of butane, 0 wt% to 90 wt% of butene, 0 wt% to

90 wt% of pentane, 0 wt% to 90 wt% of pentene, 0 wt% to 90 wt% of hexane, 0 wt% to 90 wt% of hexene, 0 wt% to 50 wt% of methanol, 0 wt% to 50 wt% of ethanol, and 0 wt% to 50 wt% of water, and a total content of methanol, ethanol, and water is greater than or equal to 10 wt%;

preferably, the riser reactor is configured to feed a catalyst and a riser reactor raw material; and the riser reactor raw material comprises 0 wt% to 20 wt% of hydrogen, 0 wt% to 50 wt% of methane, 0 wt% to 50 wt% of ethane, 0 wt% to 20 wt% of ethylene, 0 wt% to 50 wt% of propane, 0 wt% to 20 wt% of propylene, 0 wt% to 90 wt% of butane, 0 wt% to 90 wt% of butene, 0 wt% to 90 wt% of pentane, 0 wt% to 90 wt% of pentene, 0 wt% to 90 wt% of hexane, 0 wt% to 90 wt% of hexene, 0 wt% to 50 wt% of methanol, 0 wt% to 50 wt% of ethanol, and 0 wt% to 50 wt% of water, and a total content of methanol, ethanol, and water is greater than or equal to 10 wt%.

4. A device for preparing low-carbon olefins from an oxygen-containing compound, wherein the device comprises a methanol conversion reactor and the coke-control reactor according to claim 1.

5. The device according to claim 4, wherein the methanol conversion reactor comprises a methanol conversion reactor shell and a delivery pipe;

the methanol conversion reactor shell comprises a lower shell and an upper shell;

the lower shell encloses a reaction zone II, and a methanol conversion reactor distributor is arranged in an inner lower part of the reaction zone II;

the delivery pipe is located above the reaction zone II; the delivery pipe has one end closed and the other end communicating with the reaction zone II;

the upper shell is arranged on a periphery of the delivery pipe;

the upper shell and a pipe wall of the delivery pipe enclose a cavity;

the cavity is divided into a spent catalyst zone and a gas-solid separation zone II from bottom to top;

the spent catalyst zone is provided with a spent catalyst zone gas distributor~~.~~;

preferably, the gas-solid separation zone II is provided with a first gas-solid separation unit of the methanol conversion reactor;

an upper part of the delivery pipe is connected to an inlet of the first gas-solid separation unit

of the methanol conversion reactor;

a catalyst outlet of the first gas-solid separation unit of the methanol conversion reactor is

located in the spent catalyst zone;

a gas outlet of the first gas-solid separation unit of the methanol conversion reactor communicates with a methanol conversion reactor gas collection chamber; and the methanol conversion reactor gas collection chamber is connected to a product gas delivery pipe.;

preferably, the gas-solid separation zone II is further provided with a second gas-solid separation unit of the methanol conversion reactor; a gas inlet of the second gas-solid separation unit of the methanol conversion reactor is located in the gas-solid separation zone II; a catalyst outlet of the second gas-solid separation unit of the methanol conversion reactor is located in the spent catalyst zone; and a gas outlet of the second gas-solid separation unit of the methanol conversion reactor communicates with the methanol conversion reactor gas collection chamber.;

preferably, the spent catalyst zone gas distributor is located below the first gas-solid separation unit of the methanol conversion reactor and the second gas-solid separation unit of the methanol conversion reactor; and a methanol conversion reactor cooler is further provided in the spent catalyst zone;

preferably, a spent catalyst circulation pipe and a spent catalyst inclined pipe are further arranged outside the spent catalyst zone; the spent catalyst circulation pipe is configured to connect the spent catalyst zone and the reaction zone II; and the spent catalyst inclined pipe is configured to output a spent catalyst;

preferably, the gas-solid separation zone II communicates with the bed reactor gas collection chamber through a coke control product gas delivery pipe; and the reaction zone II communicates with the reaction zone I through the coke controlled catalyst delivery pipe.

6. The device according to claim 4, wherein the device further comprises a regenerator;

the regenerator is connected to the spent catalyst inclined pipe, and a spent catalyst is delivered to the regenerator;

the regenerator is connected to the riser reactor, and a regenerated catalyst is delivered to the

coke-control reactor;

an inner bottom of the regenerator is provided with a regenerator distributor.;

preferably, a bottom of the regenerator is further provided with a regenerator stripper; an upper section of the regenerator stripper is arranged inside the regenerator, and an inlet of the upper section of the regenerator stripper is located above the regenerator distributor; a lower section of the regenerator stripper is arranged outside the regenerator, and an outlet of the lower section of the regenerator stripper is connected to the riser reactor; and the regenerator stripper is further provided with a regenerator cooler;

preferably, the regenerator is connected to the spent catalyst inclined pipe through a spent catalyst delivery pipe and a methanol conversion reactor stripper; and the regenerator is connected to an inlet of the riser reactor through the regenerator stripper and a regenerated catalyst inclined pipe.

7. The device according to claim 6, wherein the regenerator is further provided with a regenerator gas-solid separation unit and a regenerator gas collection chamber;

a catalyst outlet of the regenerator gas-solid separation unit is located above the regenerator distributor;

a gas outlet of the regenerator gas-solid separation unit is connected to the regenerator gas collection chamber; and

the regenerator gas collection chamber is connected to a flue gas delivery pipe located outside the regenerator.

8. A method for preparing low-carbon olefins from an oxygen-containing compound, comprising

5

a method for on-line modification of a dimethyl ether/methanol to olefins (DMTO) catalyst; wherein the method for on-line modification of a DMTO catalyst comprises: feeding a riser reactor raw material and a catalyst into a transition zone from a riser reactor, and feeding a bed reactor raw material into a reaction zone I; and allowing the catalyst to contact and react with the riser reactor raw material and the bed reactor raw material to generate a coke controlled catalyst and a coke control product gas, wherein the catalyst is a DMTO catalyst; and the coke controlled catalyst is a modified DMTO catalyst.

9. The method according to claim 8, wherein an active component of the catalyst is an SAPO-34 molecular sieve;

preferably, the catalyst is a regenerated catalyst; and a coke content in the regenerated catalyst is less than or equal to 3 wt%;

preferably, a coke content in the coke controlled catalyst is 4 wt% to 9 wt%; and a quartile deviation of a coke content distribution in the coke controlled catalyst is less than 1 wt%;

preferably, coke species in the coke controlled catalyst comprise polymethylbenzene and polymethylnaphthalene; a total mass of the polymethylbenzene and the polymethylnaphthalene accounts for greater than or equal to 70 wt% of a total mass of coke; a mass of coke species with a molecular weight greater than 184 accounts for less than or equal to 25 wt% of the total mass of coke; and the total mass of coke refers to a total mass of the coke species;

preferably, process operating conditions of the riser reactor are as follows: an apparent gas linear velocity is 3 m/s to 10 m/s; a reaction temperature is 400°C to 700°C; a reaction pressure is 100 kPa to 500 kPa; and a bed density is 10 kg/m^3 to 150 kg/m^3;

preferably, process operating conditions of the reaction zone I of a bed reactor are as follows: an apparent gas linear velocity is 0.1 m/s to 1.0 m/s; reaction temperature is 300°C to 650°C; a reaction pressure is 100 kPa to 500 kPa; and a bed density is 150 kg/m^3 to 800 kg/m^3.

10. The method according to claim 8, further comprising: feeding a coke control product gas into a gas-solid separation zone of a methanol conversion reactor.

11. The method according to claim 8, further comprising: feeding a coke controlled

catalyst into a reaction zone II of a methanol conversion reactor;

preferably, in the reaction zone II, a raw material with the oxygen-containing compound contacts and reacts with the coke controlled catalyst to generate a stream A with low-carbon olefins and a spent catalyst~~.~~;

preferably, the stream A is separated into a gas-phase stream B and a solid-phase stream C after being subjected to gas-solid separation in a gas-solid separation zone II of the methanol conversion reactor; the gas-phase stream B enters a methanol conversion reactor gas collection chamber; the solid-phase stream C enters a spent catalyst zone; and the gas-phase stream B comprises the low-carbon olefins, and the solid-phase stream C comprises the spent catalyst~~.~~;

preferably, a spent catalyst zone fluidizing gas is fed into the spent catalyst zone; the spent catalyst zone fluidizing gas and a coke control product gas are mixed and carry a part of the spent catalyst to produce a stream D; the stream D is separated into a gas-phase stream E and a solid-phase stream F after being subjected to gas-solid separation; the gas-phase stream E enters the methanol conversion reactor gas collection chamber; the solid-phase stream F enters the spent catalyst zone; the gas-phase stream E is a mixed gas of the spent catalyst zone fluidizing gas and the coke control product gas; and the solid-phase stream F is the spent catalyst;

preferably, the gas-phase stream B and the gas-phase stream E are mixed in the methanol conversion reactor gas collection chamber to produce a product gas, and the product gas enters a downstream working section through a product gas delivery pipe.

12. The method according to claim 11, wherein a part of the spent catalyst in the spent catalyst

zone is returned to a bottom of the reaction zone II through a spent catalyst circulation pipe; and

the remaining part of the spent catalyst is discharged through a spent catalyst inclined pipe.

13. The method according to claim 12, wherein the spent catalyst discharged through the spent catalyst inclined pipe is fed into a regenerator;

a regeneration gas is fed into the regenerator to contact and react with the spent catalyst to obtain a stream G with a flue gas and a regenerated catalyst~~.~~;

preferably, the stream G is subjected to gas-solid separation; a separated flue gas enters a regenerator gas collection chamber, and then enters a downstream flue gas treatment system through a flue gas delivery pipe; and a separated regenerated catalyst is stripped and cooled, and then enters a coke-control reactor.

14. The method according to claim-11, wherein the oxygen-containing compound comprises methanol and/or dimethyl ether (DME).

15. The method according to claim 11, wherein a coke content in the spent catalyst is 9 wt% to 13 wt%.

16. The method according to claim11, wherein the spent catalyst zone fluidizing gas comprises nitrogen and/or water vapor.

17. The method according to claim 13, wherein the regeneration gas comprises 0 wt% to 100 wt% of air, 0 wt% to 50 wt% of oxygen, 0 wt% to 50 wt% of nitrogen, and 0 wt% to 50 wt% of water vapor; and

the air, the oxygen, the nitrogen, and the water vapor are not simultaneously zero.

18. The method according to claim 8, wherein process operating conditions of the reaction zone II of the methanol conversion reactor are as follows: an apparent gas linear velocity is 0.5 m/s to 7.0 m/s; a reaction temperature is 350°C to 550°C; a reaction pressure is 100 kPa to 500 kPa; and a bed density is 100 kg/m^3 to 500 kg/m^3.

19. The method according to claim 8, wherein process operating conditions of the spent catalyst zone of the methanol conversion reactor are as follows: an apparent gas linear velocity is

0.1 m/s to 1.0 m/s; a reaction temperature is 350°C to 550°C; a reaction pressure is 100 kPa to 500 kPa; and a bed density is 200 kg/m^3 to 800 kg/m^3.

20. The method according to claim 8, wherein process operating conditions of the regenerator are as follows: an apparent gas linear velocity is 0.5 m/s to 2.0 m/s; a regeneration temperature is 600°C to 750°C; a regeneration pressure is 100 kPa to 500 kPa; and a bed density is 150 kg/m^3 to 700 kg/m^3.