Claims

1. A process for producing gasoline and propylene, comprising the steps of:

1) subjecting a feedstock oil to a first catalytic conversion reaction in a first catalytic conversion reaction device to obtain a first reaction product comprising propylene, C_4 olefin and a gasoline component;

2) separating the first reaction product to obtain a propylene fraction, a gasoline fraction, a fraction comprising C_4 olefin, optionally a light cycle oil fraction and optionally a fluidized catalytic cracking gas oil (FGO) fraction;

3) subjecting the C_4 olefin-containing fraction obtained in step 2) and optionally a C_4 olefin-containing fraction from an external source to olefin oligomerization in an oligomerization reactor to obtain an oligomerization product comprising C_{12} olefin, and separating the oligomerization product to obtain a fraction comprising C_{12} olefin, wherein the C_{12} olefin-containing fraction has a C_{12} olefin content of no less than 40 wt%, for example no less than 50 wt%, no less than 70 wt%, no less than 80 wt% or no less than 90 wt%, based on the weight of the C_{12} olefin-containing fraction, and is preferably essentially consisted of C_{12} olefin;

4) recycling the C_{12} olefin-containing fraction to the first catalytic conversion reaction device, and/or sending the C_{12} olefin-containing fraction to a second catalytic conversion reaction device for a second catalytic conversion reaction to obtain a second reaction product comprising propylene; and

5) optionally, subjecting the FGO fraction to a first hydrotreatment in a first hydrotreating reactor to obtain a hydrogenated FGO, and recycling at least a part of the hydrogenated FGO to the first catalytic conversion reaction device and/or sending it to the second catalytic conversion reaction device, and/or recycling the light cycle oil fraction to the first catalytic conversion reaction device and/or sending it to the second catalytic conversion reaction device.

2. The process according to claim 1, wherein the olefin oligomerization of step 3) is carried out in the presence of an oligomerization catalyst under conditions including:

a temperature of 50-500 ºC, a pressure of 0.5-5.0 MPa, and a weight hourly space velocity of 0.1-100 h^{-1};

the oligomerization catalyst is one or more selected from the group consisting of phosphoric acid catalysts, acidic resins, silica-alumina solid acid catalysts and zeolite solid acid catalysts;

preferably, the phosphoric acid catalyst is one or more selected from the group consisting of a catalyst formed by loading phosphoric acid on diatomite, a catalyst formed by loading phosphoric acid on activated carbon, a catalyst formed from phosphoric acid-soaked quartz sand, a catalyst formed by loading phosphoric acid on silica gel, and a catalyst formed by loading copper pyrophosphate on silica gel;

the silica-alumina solid acid catalyst is a catalyst formed by loading metal ion(s) on alumina and/or amorphous silica-alumina carrier, wherein the loaded metal ion(s) is selected from Group VIII metals, Group IVA metals or a combination thereof; and

the zeolite solid acid catalyst comprises 10-100 wt% of a zeolite and 0-90 wt% of a matrix, based on the weight of the zeolite solid acid catalyst, wherein the zeolite is one or more selected from the group consisting of one-dimensional zeolites selected from MTW (ZSM-12), MTT (ZSM-23) and TON (ZSM-22) zeolites, two-dimensional zeolites selected from FER (ferrierite), MFS (ZSM-57), MWW (MCM-22) and MOR (mordenite) zeolites, and three-dimensional zeolites selected from beta zeolites.

3. The process according to claim 1 or 2, wherein:

the optional light cycle oil fraction in step 2) has a distillation range of 190-350 ºC, preferably 190-280 ºC, and more preferably 200-260 ºC; the optional FGO fraction is a heavy fraction obtained at the bottom of the catalytic conversion fractionator, preferably a fraction having a cut point of greater than 250 ºC, more preferably a fraction having a cut point of greater than 260 ºC;

the oligomerization product obtained in step 3) has a C_{12} olefin content of no less than 20 wt%, based on the weight of the oligomerization product;

the C_4 olefin-containing fraction has a C_4 olefin content of 40-100 wt%,

based on the weight of the C_4 olefin-containing fraction; and/or

the C_{12} olefin recycled to the first catalytic conversion reaction device and/or sent to the second catalytic conversion reaction device in step 4) accounts for 0.1-80 wt%, preferably 1-70 wt%, of the total catalytic conversion feedstock.

4. The process according to any one of claims 1-3, further comprising:

subjecting the feedstock oil to a second hydrotreatment in a second hydrotreating reactor prior to step 1);

the second hydrotreatment is carried out in the presence of a second hydrotreating catalyst under conditions including:

a hydrogen partial pressure of 3.0-20.0 MPa, a reaction temperature of 300-450 ºC, a hydrogen-to-oil volume ratio of 100-2000 standard cubic meter/cubic meter, and a volume space velocity of 0.1-3.0 h^{-1};

the second hydrotreating catalyst comprises 70-99 wt% of a carrier, which is one or more selected from the group consisting of alumina, silica and amorphous silica-alumina, and 0.1-30 wt% of an active metal (calculated as oxide) selected from Group VIB metals, Group VIII metals, or a combination thereof.

5. The process according to any one of claims 1-4, wherein the first hydrotreatment is conducted in the presence of a first hydrotreating catalyst under conditions including:

a hydrogen partial pressure of 3.0-20.0 MPa, a reaction temperature of 300-450 ºC, a volume space velocity of 0.1-10.0 h^{-1}, and a hydrogen-to-oil volume ratio of 100-1500 standard cubic meters/cubic meter;

the first hydrotreating catalyst comprises 60-99 wt% of a carrier, which is one or more selected from the group consisting of alumina, silica and amorphous silica-alumina, and 5-40 wt% of an active metal (calculated as oxide) selected from Group VIB metals, Group VIII metals or a combination thereof.

6. The process according to any one of claims 1-5, wherein the first catalytic conversion reaction device and the second catalytic conversion reaction device, if present, are each independently a fluidized bed reactor,

preferably, the fluidized bed reactor is a riser reactor or a composite reactor

comprising a riser in combination with a dense phase bed, wherein the riser reactor is an equal-diameter riser reactor, a constant-linear-velocity riser reactor or a diameter-transformed riser reactor.

7. The process according to any one of claims 1-6, wherein the catalyst for the first catalytic conversion reaction and the catalyst for the second catalytic conversion reaction, if present, are each independently selected from amorphous silica-alumina catalysts, zeolite catalysts, or a combination thereof, and wherein the zeolite in the zeolite catalyst is one or more selected from the group consisting of Y zeolite, HY zeolite, ultrastable Y zeolite, ZSM-5 series zeolite, high-silica zeolite having a pentasil structure, and ferrierite.

8. The process according to any one of claims 1-7, wherein the feedstock oil is selected from petroleum hydrocarbons, other mineral oils, or a combination thereof, wherein the petroleum hydrocarbon is one or more selected from the group consisting of atmospheric gas oil, vacuum gas oil, atmospheric residue, vacuum residue, hydrogenated residue, coker gas oil, and deasphalted oil; and said other mineral oil is one or more selected from the group consisting of coal and natural gas derived liquid oil, oil sand oil, tight oil and shale oil,

preferably, the property of the feedstock oil satisfies at least one of the following criteria:

a density of 900-1000 kg/m^3, preferably 930-960 kg/m^3, a carbon residue of 4-15 wt%, preferably 6-12 wt%, a metal content of 15-600 ppm, preferably 15-100 ppm, and an acid value of 0.5-20 mg KOH/g, preferably 0.5-10 mg KOH/g.

9. The process according to claim 6, wherein the first catalytic conversion reaction device is a fluidized bed reactor and the reaction conditions in the first catalytic conversion reaction device include: a reaction temperature of 420-650 ºC, a reaction time of 0.05-20 seconds, a catalyst-to-feedstock weight ratio of 3 : 1 to 15 : 1, and a steam-to-feedstock weight ratio of 0.03 : 1 to 0.5 : 1;

preferably, the reaction conditions include: a reaction temperature of 480-600 ºC, a reaction time of 0.1-15 seconds, a catalyst-to-feedstock weight ratio of 4 : 1 to 12 : 1, and a steam-to-feedstock weight ratio of 0.05 : 1 to 0.3 : 1.

10. The process according to claim 9, wherein the first catalytic conversion reaction was carried out to the extent that a gaseous product having the following characteristics is obtained:

a mass fraction ratio of liquefied gas (C_3 and C_4) to dry gas (C_2 or lower) of not less than 7, preferably not less than 10;

a methane yield of no greater than 2.0%, preferably no greater than 1.0%;

a mass fraction ratio of propylene to propane in the liquefied gas of not less than 3.5, preferably not less than 5.0; and/or

a mass fraction ratio of isobutylene to isobutane of not less than 1.5, preferably not less than 4.0.

11. The process according to any one of claims 6 and 9-10, wherein the first catalytic conversion reaction device is a diameter-transformed riser reactor comprising a first reaction zone and a second reaction zone having a diameter greater than the first reaction zone along the direction of the reaction stream,

wherein the reaction conditions in the first reaction zone include:

a reaction temperature of 450-620 ºC, preferably 500-600 ºC, a reaction time of 0.5-2.0 seconds, preferably 0.8-1.5 seconds, and a catalyst-to-feedstock weight ratio of 3 : 1 to 15 : 1, preferably 4 : 1 to 12 : 1, a steam-to-feedstock weight ratio of 0.03 : 1 to 0.3 : 1, preferably 0.05 : 1 to 0.15 : 1;

the reaction conditions in the second reaction zone include:

a reaction temperature of 460-550 ºC, preferably 480-530 ºC, a reaction time of 2-30 seconds, preferably 3-15 seconds, a catalyst-to-feedstock weight ratio of 4 : 1 to 18 : 1, preferably 4.5 : 1 to 15 : 1, and a steam-to-feedstock weight ratio of 0.03 : 1 to 0.3 : 1, preferably 0.05 : 1 to 0.15 : 1.

12. The process according to claim 11, wherein the C_{12} olefin-containing fraction is recycled to the second reaction zone of the diameter-transformed riser reactor in step 4), at least a part of the hydrogenated FGO is recycled to the first reaction zone of the diameter-transformed riser reactor in step 5), and/or the light cycle oil fraction is recycled to the first reaction zone of the diameter-transformed riser reactor in step 5).

13. The process according to any one of claims 1-11, wherein in step 4)

said C_{12} olefin-containing fraction is sent to the second catalytic conversion reaction device, and the second catalytic conversion reaction is carried out under conditions including: a reaction temperature of 450-620 ºC, a reaction time of 0.5-20.0 seconds, a catalyst-to-oil weight ratio of 1-25, and a steam-to-oil weight ratio of 0.03-0.3.

14. A system for producing gasoline and propylene, comprising a first catalytic conversion reaction device, a regenerator, an oil-catalyst separation device, a product separation device, an oligomerization reactor, a rectifying column and optionally a second catalytic conversion reaction device, wherein:

the first catalytic conversion reaction device is configured to perform a first catalytic conversion reaction on a feedstock oil therein to obtain a reaction effluent comprising a first reaction product and a spent catalyst, wherein the first reaction product comprises propylene, C_4 olefin and a gasoline component;

the oil-catalyst separation device is configured to separate the first reaction product from the spent catalyst in the reaction effluent of the first catalytic conversion reaction device,

the regenerator is configured to regenerate the spent catalyst and recycle the regenerated catalyst to the first catalytic conversion reaction device,

the product separation device is configured to separate the first reaction product to obtain a propylene fraction, a gasoline fraction, a fraction comprising C_4 olefin, optionally a light cycle oil fraction and optionally a FGO fraction,

the oligomerization reactor is configured to conduct an olefin oligomerization on the fraction comprising C_4 olefin to obtain an oligomerization product comprising C_{12} olefin, and send the oligomerization product to the rectifying column;

the rectifying column is configured to separate the oligomerization product to obtain a fraction comprising C_{12} olefin, and recycle the fraction comprising C_{12} olefin to the first catalytic conversion reaction device or send it to the optional second catalytic conversion reaction device, wherein the C_{12} olefin-containing fraction has a C_{12} olefin content of no less than 40 wt%, for example no less than 50 wt%, no less than 70 wt%, no less than 80 wt% or no

less than 90 wt%, based on the weight of the C_{12} olefin-containing fraction, and is preferably essentially consisted of C_{12} olefin;

the second catalytic conversion reaction device, if present, is configured to perform a second catalytic conversion reaction on the C_{12} olefin-containing fraction to obtain a second reaction product comprising propylene.

15. The system according to claim 14, wherein the first catalytic conversion reaction device is provided with a catalyst inlet, at least one catalytic conversion feedstock inlet and an oil-catalyst outlet, the regenerator is provided with a spent catalyst inlet and a regenerated catalyst outlet, the oil-catalyst separator is provided with an oil-catalyst inlet, an oil-gas outlet and a spent catalyst outlet, the product separator is provided with an oil-gas inlet, a propylene fraction outlet, a gasoline fraction outlet and a C_4 olefin-containing fraction outlet, the oligomerization reactor is provided with an oligomerization feedstock inlet and an oligomerization product outlet, and the rectifying column is provided with an oil-gas inlet and a C_{12} olefin-containing fraction outlet;

the catalyst inlet of the first catalytic conversion reaction device is in fluid communication with the regenerated catalyst outlet of the regenerator, the oil-catalyst outlet of the first catalytic conversion reaction device is in fluid communication with the oil-catalyst inlet of the oil-catalyst separation device, the spent catalyst outlet of the oil-catalyst separation device is in fluid communication with the spent catalyst inlet of the regenerator, the oil-gas outlet of the oil-catalyst separation device is in fluid communication with the oil-gas inlet of the product separation device, the C_4 olefin-containing fraction outlet of the product separation device is in fluid communication with the oligomerization feedstock inlet of the oligomerization reactor, the oligomerization product outlet of the oligomerization reactor is in fluid communication with the oil-gas inlet of the rectifying column, and the C_{12} olefin-containing fraction outlet of the rectifying column is in fluid communication with one catalytic conversion feedstock inlet of the first catalytic conversion reaction device or in fluid communication with the feedstock inlet of the second catalytic conversion reaction device;

preferably, the system further comprises a first hydrotreating reactor, the product separation device is further provided with a FGO fraction outlet, the feedstock inlet of the first hydrotreating reactor is in fluid communication with the FGO fraction outlet of the product separation device, and the product outlet of the first hydrotreating reactor is in fluid communication with one catalytic conversion feedstock inlet of the first catalytic conversion reaction device or in fluid communication with the feedstock inlet of the second catalytic conversion reaction device; and

further preferably, the product separation device is further provided with a light cycle oil fraction outlet, and the light cycle oil fraction outlet of the product separation device is in fluid communication with one catalytic conversion feedstock inlet of the first catalytic conversion reaction device or in fluid communication with the feedstock inlet of the second catalytic conversion reaction device.

16. The system according to claim 14 or 15, wherein the first catalytic conversion reaction device is a diameter-transformed riser reactor comprising a first reaction zone and a second reaction zone having a diameter greater than the first reaction zone along the flow direction of the reaction stream,

wherein the bottom of the second reaction zone is provided with one or more chilling/heating medium inlets, and/or the second reaction zone is provided with a heat remover, and the height of the heat remover is 50-90% of the height of the second reaction zone.

17. The system according to any one of claims 14-16, further comprising a second hydrotreating reactor provided with a feedstock inlet and a hydrogenation product outlet, wherein the hydrogenation product outlet of the second hydrotreating reactor is in fluid communication with one catalytic conversion feedstock inlet of the first catalytic conversion reaction device.