Method and system for producing light aromatic hydrocarbons

Technical Field

The present invention relates to the field of petrochemical industry, and in particular, to a method and system for producing light aromatic hydrocarbons.

Background Art

With the development of crude oil heavyization and the rapid growth of market demand for light and high-quality products, in China, catalytic cracking technology as a heavy oil lightening process has been developed rapidly. However, the quality of catalytic cracking diesel (or referred to as catalytic cracking light cycle oil) has been poor, with high density, high aromatic hydrocarbon content, and low cetane number, making it difficult to meet increasingly stringent diesel specifications even through diesel hydro-upgrading technology. Light aromatic hydrocarbons represented by para-xylene (benzene, toluene, xylene, collectively referred to as BTX) are important raw materials for the production of chemical products, and the market demand for light aromatic hydrocarbons is increasing and is still in short supply. Therefore, producing light aromatic hydrocarbons from catalytic cracking diesel is a route that meets market demand.

US4585545A discloses a catalytic conversion method in which a full fraction of catalytic cracking light cycle oil is first hydrotreated, and the obtained hydrotreated diesel is then subjected to catalytic cracking to produce a gasoline rich in monocyclic aromatic hydrocarbons.

CN01807978.4 (CN1422327A) discloses a method for upgrading catalytic cracking light cycle oil, in which a catalytic light cycle oil produced from a first catalytic cracking unit using heavy oil as feedstock is subjected to deep hydrogenation, and the obtained hydrogenated diesel is then fed into a second catalytic cracking unit.

CN01808113.4 (CN1466619A) discloses a method for converting catalytic cracking light cycle oil, in which a catalytic cracking riser reaction zone is divided into an upstream reaction zone and a downstream reaction zone, wherein heavy oil is injected into the downstream zone, and a hydrogenated cycle oil obtained after hydrotreating a light cycle oil product from its catalytic cracking is injected into the upstream zone.

CN103923698B discloses a method for producing aromatic compounds, in which a diesel fraction obtained from catalytic cracking having a distillation range of 250-450°C is sent to a hydrotreating unit, and a catalytic cracking heavy cycle oil after hydrogenation is then subjected to catalytic cracking to produce benzene, toluene and xylene.

CN104560187B discloses a catalytic conversion method for producing a gasoline rich in aromatic hydrocarbons, in which a catalytic cracking light cycle oil is cut into a light fraction and a heavy fraction, and the heavy fraction after hydrotreating and the light fraction are separately and respectively fed into different catalytic cracking units to maximize the production of a catalytic gasoline rich in benzene, toluene and xylene.

From the documents disclosed above, it can be found that one of the important routes for processing catalytic cracking light cycle oil is to hydrotreat it and then subject it to catalytic cracking. However, it is worth noting that the selection of fractions before and after hydrogenation is crucial both for hydrogen consumption and for the yield of light aromatic hydrocarbons from catalytic cracking.

Summary of the Invention

The objective of the present invention is to provide a method and system for producing light aromatic hydrocarbons on the basis of the prior art.

In order to achieve the above objective, the present invention provides a method and system for producing light aromatic hydrocarbons, wherein a catalytic cracking cycle oil is subjected to two-stage hydrotreating in the presence of hydrogen and a hydrogenation catalyst, and the hydrogenated oil-gas mixture is passed through a first distillation column to obtain an overhead light fraction, a light aromatic hydrocarbon-rich fraction, a first recycle fraction having a specific distillation range, and a second recycle fraction having a specific distillation range; the first recycle fraction is introduced into a first reaction zone of a catalytic cracking unit for a catalytic cracking reaction, and the second recycle fraction is introduced into a second reaction zone of the catalytic cracking unit for reaction or returned to the second stage of the hydrotreating for recycling. Further, the reaction product obtained from the catalytic cracking is separated by a second distillation column to obtain a first fraction oil-gas mixture, a second fraction oil and a slurry oil, the second fraction oil is returned to the above-mentioned first stage of the hydrotreating, and optionally the first fraction oil-gas mixture together with the hydrogenated oil-gas mixture is introduced into the above-mentioned first distillation column for separation.

A first aspect of the present invention provides a method for producing light aromatic hydrocarbons, wherein the method comprises the following steps:

(1) subjecting a catalytic cracking cycle oil to two-stage hydrotreating in the presence of hydrogen and a hydrogenation catalyst to obtain a hydrogenated product;

(2) removing hydrogen from the hydrogenated product to obtain a hydrogenated oil-gas mixture, and passing the hydrogenated oil-gas mixture through a first distillation column to obtain an overhead light fraction, a light aromatic hydrocarbon-rich fraction, a first recycle fraction and a

second recycle fraction, wherein the first recycle fraction has an initial boiling point of 140-170°C and a final boiling point of 220-300°C, and the second recycle fraction has a distillation range starting point temperature of any value between 220-300°C;

(3) introducing the first recycle fraction obtained in step (2) into a first reaction zone of a catalytic cracking unit for a catalytic cracking reaction, and optionally, introducing the second recycle fraction into a second reaction zone of the catalytic cracking unit or returning it to a second stage of said hydrotreating for recycling;

(4) separating the reaction product obtained from the catalytic cracking unit by a second distillation column to obtain a first fraction, a second fraction and a slurry oil, returning the second fraction to a first stage of said hydrotreating, and optionally, introducing the first fraction oil-gas mixture together with the hydrogenated oil-gas mixture into the first distillation column of step (2) for separation;

(5) separating light aromatic hydrocarbons from the light aromatic hydrocarbon-rich fraction.

According to the method of the first aspect, wherein in step (1), the hydrotreating reaction conditions satisfy at least one of the following requirements:

the reaction conditions for the first stage of said hydrotreating are: hydrogen partial pressure 5.0-20.0 MPa, reaction temperature 300-450°C, volumetric space velocity 0.5-5.0 h^{-1}, hydrogen-to-hydrocarbon volume ratio 300-1600 Nm^3/m^3; and/or

the reaction conditions for the second stage of said hydrotreating are: hydrogen partial pressure 5.0-20.0 MPa, reaction temperature 10-50°C higher than the temperature of the first stage of said hydrotreating, volumetric space velocity 0.1-3.0 h^{-1}, hydrogen-to-hydrocarbon volume ratio 400-2000 Nm^3/m^3.

According to the method of the first aspect, wherein in step (1), in the two-stage hydrotreating reaction, the hydrogenation catalysts in the first and second stages are the same or different;

Preferably, the hydrogenation catalyst for the first stage of said hydrotreating comprises an active metal component and a carrier, and the hydrogenation catalyst for the second stage of hydrotreating comprises an active metal, a molecular sieve and a carrier, wherein the active metal component is selected from one or more of Group VIB metals and/or Group VIII non-noble metals, the carrier is selected from one or more of alumina, silica, and amorphous silica-alumina, and the molecular sieve is selected from one or more of Y-type molecular sieve and USY molecular sieve;

more preferably, the active metal component is selected from any one of the following metal combinations: nickel-tungsten, nickel-tungsten-cobalt, nickel-molybdenum, cobalt-molybdenum.

According to the method of the first aspect, wherein in step (1), the catalytic cracking cycle oil has an initial boiling point of 160-250°C and a final boiling point of 340-400°C; the hydrogenated oil-gas mixture has an initial boiling point of 0-60°C and a final boiling point of 280-360°C.

According to the method of the first aspect, wherein in step (2), the first distillation column is a divided-wall distillation column, at least one dividing wall is arranged in the divided-wall distillation column, and the dividing wall divides the divided-wall distillation column into four regions: a pre-separation zone, a public rectifying section, a public stripping section and a side-draw rectifying zone, and the feed position of the hydrogenated oil-gas mixture is located in the pre-separation zone.

According to the method of the first aspect, wherein in step (2), the pre-separation zone of the divided-wall distillation column has 10-40 trays, preferably 20-35 trays;

The public rectifying section has 2-10 trays, preferably 3-8 trays;

the public stripping section has 2-10 trays, preferably 3-8 trays;

the side-draw rectifying zone has 10-45 trays, preferably 25-40 trays;

and the divided-wall distillation column satisfies at least one of the following requirements:

the feed position of the hydrogenated oil-gas mixture is above the 4th to 22nd tray from the top of the pre-separation zone, preferably above the 8th to 20th tray;

the overhead pressure of the divided-wall distillation column is 0.1-1.0 MPa, preferably 0.2-0.7 MPa, and the overhead temperature is 50-120°C, preferably 60-100°C; and/or

the overhead reflux ratio is 3-10, preferably 4-8.

According to the method of the first aspect, wherein in step (2), the divided-wall distillation column satisfies at least one of the following requirements:

The side-draw rectifying zone of the divided-wall distillation column is provided with a middle-stage circulation heat removal, at the 20th to 45th tray of the side-draw rectifying zone, more preferably at the 30th to 40th tray;

preferably, the pre-separation zone of the divided-wall distillation column is provided with a vapor draw, with a draw position at the 20th to 40th tray of the pre-separation zone, more preferably at the 25th to 35th tray, and a vapor return position at the 30th to 45th tray of the side-draw rectifying zone, more preferably at the 35th to 40th tray, and a return position at the same tray as the draw position or one tray above or below the draw position;

the mass flow ratio of the liquid phases from the public rectifying section into the pre-separation zone and the side-draw rectifying zone is 0.1-10, preferably 0.5-5;

the mass flow ratio of the vapor phases from the public stripping section into the pre-separation zone and the side-draw rectifying zone is 1-15, preferably 2-10; and/or

the ratio of the total vapor mass flow to the drawn vapor mass flow at the draw position of the pre-separation zone is 10-150, preferably 30-120.

According to the method of the first aspect, wherein in step (2), the divided-wall distillation column satisfies at least one of the following requirements:

the overhead light fraction is separated from the public rectifying section of the divided-wall distillation column, and has an initial boiling point of 20-40°C and a final boiling point of 55-65°C;
the light aromatic hydrocarbon-rich fraction is drawn from the 5th to 30th tray of the side-draw rectifying zone of the divided-wall distillation column, preferably from the 10th to 20th tray, and has an initial boiling point of 55-65°C and a final boiling point of 140-170°C;
the first recycle fraction is drawn from any position below the draw position of the light aromatic hydrocarbon-rich fraction in the side-draw rectifying zone, preferably from the 25th to 35th tray; and/or
the second recycle fraction is separated from the lower part of the public stripping section of the divided-wall distillation column.
According to the method of the first aspect, wherein in step (3), the catalytic conversion conditions in the first reaction zone satisfy at least one of the following requirements:
the weight hourly space velocity of hydrocarbons is 4-20 h^{-1}, preferably 6-12 h^{-1};
the reaction time is 0.5-8 s, preferably 1-5 s; and/or
the catalyst-to-hydrocarbon weight ratio is 6-26, preferably 8-12.
According to the method of the first aspect, wherein in step (3), the catalytic conversion conditions in the second reaction zone satisfy at least one of the following requirements:
the weight hourly space velocity of hydrocarbons is 4-20 h^{-1}, preferably 8-12 h^{-1};
the reaction time is 1-10 s, preferably 2-6 s; and/or
the catalyst-to-hydrocarbon weight ratio is 4-14, preferably 6-10.
According to the method of the first aspect, wherein in step (4), the distillation column has an overhead pressure of 0.15-0.4 MPa, preferably 0.18-0.35 MPa, and an overhead temperature of 100-160°C, preferably 115-145°C.
According to the method of the first aspect, wherein in step (4), the second distillation column satisfies at least one of the following requirements:
the first fraction has an initial boiling point of 0-40°C and a final boiling point of 220-250°C, and the first fraction, after being combined with the hydrogenated fraction and degassed, is introduced into the first distillation column of step (2) for separation;
the second fraction oil has an initial boiling point of 220-250°C and a final boiling point of 340-360°C, and its initial boiling point is not lower than the initial boiling point of the first fraction oil-gas mixture; and/or
the slurry oil has an initial boiling point of 340-360°C.
A second aspect of the present invention provides a system for producing light aromatic hydrocarbons, comprising:
A hydrogenation unit, wherein the hydrogenation unit is a two-stage hydrogenation unit and

provided with a hydrogen inlet and a hydrogenated product outlet, a first stage of the hydrogenation unit is provided with a catalytic cracking cycle oil inlet, and a second stage of the hydrogenation unit is optionally provided with a second recycle fraction inlet;

a catalytic cracking unit, comprising a catalytic cracking reactor, wherein the catalytic cracking reactor is provided with a first reaction zone and an optional second reaction zone, the first reaction zone is provided with a first recycle fraction inlet, and the optional second reaction zone is provided with a second recycle fraction inlet;

a second distillation column, wherein the second distillation column is provided with a catalytic cracking reaction product inlet, a first fraction outlet, a second fraction outlet and a slurry oil outlet, and the second fraction outlet is in communication with the catalytic cracking cycle oil inlet of the hydrogenation unit;

a degassing separation unit, wherein the degassing separation unit is provided with an oil-gas mixture inlet, a rich gas outlet and a fraction oil outlet, and the oil-gas mixture inlet of the degassing separation unit is connected to the first fraction oil-gas mixture outlet of the second distillation column and the hydrogenated product outlet of the hydrogenation unit, so that the first fraction oil-gas mixture from the second distillation column and the hydrogenated oil-gas mixture from the hydrogenation unit are degassed and separated into a rich gas and a fraction oil;

a first distillation column, wherein the distillation column is provided with a fraction oil inlet, an overhead light fraction outlet, a light aromatic hydrocarbon-rich fraction outlet, a first recycle fraction outlet, and a second recycle fraction outlet, the fraction oil inlet is in communication with the fraction oil outlet of the degassing separation unit, the first recycle fraction outlet is in communication with the first recycle fraction inlet of the catalytic cracking unit, the second recycle fraction outlet is in communication with the second recycle fraction inlet provided for the optional second reaction zone of the catalytic cracking unit, or the second recycle fraction outlet is in communication with the second recycle fraction inlet optionally provided for the second stage of the hydrogenation unit;

a separation unit, which is provided with a light aromatic hydrocarbon-rich fraction inlet and a light aromatic hydrocarbon fraction outlet, and the light aromatic hydrocarbon-rich fraction inlet is connected to the light aromatic hydrocarbon-rich fraction outlet of the first distillation column.

In the second aspect of the present invention, the first distillation column is a divided-wall distillation column, at least one dividing wall is arranged in the divided-wall distillation column, and the dividing wall divides the divided-wall distillation column into four regions: a public stripping section arranged at the bottom of the column, a public rectifying section arranged at the top of the column, a pre-separation zone and a side-draw rectifying zone arranged in the middle section of the column; the pre-separation zone is provided with a fraction oil inlet, and the fraction

oil inlet is in communication with the fraction oil outlet of the degassing separation unit; the public rectifying section is provided with an overhead light fraction outlet, the side-draw rectifying zone is provided with a light aromatic hydrocarbon-rich fraction outlet and a first recycle fraction outlet, and the bottom of the public stripping section is provided with a second recycle fraction outlet,

here, the first recycle fraction outlet of the side-draw rectifying zone of the divided-wall distillation column is in communication with the first recycle fraction inlet of the catalytic cracking reactor,

the second recycle fraction outlet at the bottom of the public stripping section of the divided-wall distillation column is in communication with the second recycle fraction inlet provided for the optional second reaction zone of the first distillation column, or is in communication with the second recycle fraction inlet optionally provided on the second stage of the hydrogenation unit.

In the second aspect of the present invention, the system further comprises a high/low pressure separation unit, the high/low pressure separation unit is provided with a hydrogenated product inlet, a hydrogen outlet and a hydrogenated oil-gas mixture outlet, and the high/low pressure separation unit separates the hydrogenated oil-gas mixture from hydrogen in the hydrogenated product from the hydrogenation unit;

here, the hydrogenated product inlet is connected to the hydrogenated product outlet of the hydrogenation unit, and the oil-gas mixture inlet of the degassing separation unit is connected to the first fraction oil-gas mixture outlet of the second distillation column and the hydrogenated oil-gas mixture outlet of the high/low pressure separation unit.

According to the system of the second aspect, the first reaction zone and the second reaction zone of the catalytic cracking reactor are arranged on the same or different reactors;

preferably, the reactor is selected from one or more of the following: an equal-diameter riser, an equal-linear-velocity riser, a variable-diameter riser, a variable-linear-velocity riser, a fluidized bed, and a composite reactor, wherein the composite reactor is composed of an equal-diameter or variable-diameter riser and a fluidized bed, and preferably the reactor is a riser reactor;

more preferably, the first reaction zone and the second reaction zone of the catalytic cracking reactor are arranged on a dual-riser reactor, wherein the main riser is the first reaction zone, and the auxiliary riser is the second reaction zone.

Through the above technical solutions, the effective components of each specific distillation range after hydrogenation and catalytic cracking are efficiently separated and then introduced into the catalytic cracking unit or the hydrogenation unit for zone-specific conversion, thereby greatly increasing the yield of light aromatic hydrocarbons.

Other features and advantages of the present invention will be described in detail in the following Detailed Description of Embodiments.

Description of the Drawings

The accompanying drawings are used to provide a further understanding of the present invention, and constitute a part of the specification. Together with the following detailed description, they are used to explain the present invention, but do not constitute a limitation on the present invention. In the drawings:

Figure 1 is a schematic structural diagram of a system according to a preferred embodiment of the present invention.

Figure 2 is a schematic structural diagram of a system according to another preferred embodiment of the present invention.

Description of Reference Numerals

100. Hydrogenation unit; 200. Divided-wall distillation column; 300. Catalytic cracking unit; 400. Distillation column;

101. Catalytic cracking cycle oil inlet; 102. Hydrogenated product pipeline; 103. Hydrogenated oil-gas mixture pipeline;

201. Fraction oil pipeline; 202. Overhead light fraction pipeline; 203. Light aromatic hydrocarbon-rich fraction pipeline; 204. First recycle fraction pipeline; 205. Second recycle fraction pipeline;

A. Pre-separation zone; B. Public rectifying section; C. Side-draw rectifying zone; D. Public stripping section; E. Dividing wall;

301. First reaction zone of catalytic cracking unit; 302. Second reaction zone of catalytic cracking unit; 303. Feedstock oil inlet of first reaction zone; 304. Feedstock oil inlet of second reaction zone; 305. Catalytic cracking product pipeline;

401. First fraction oil-gas mixture pipeline; 402. Second fraction oil pipeline; 403. Slurry oil pipeline;

Detailed Description of Embodiments

The present application will be further described in detail below through the accompanying drawings and examples. Through these descriptions, the features and advantages of the present application will become clearer and more apparent.

The term "exemplary" and the like as used herein specifically means "serving as an example, embodiment, or illustration". Any embodiment described herein as "exemplary" is not necessarily to be construed as superior or better than other embodiments. Although various aspects of the

embodiments are shown in the drawings, the drawings are not necessarily drawn to scale unless specifically indicated.

Any specific numerical value (including the endpoints of numerical ranges) disclosed in the present application is not limited to the exact value of that numerical value, but should be understood to also cover values close to the exact value, for example, all possible numerical values within ±5% of the exact value. Furthermore, for the disclosed numerical ranges, any combination of the endpoint values of the range, the endpoint values and the specific point values within the range, and among the specific point values can be arbitrarily combined to obtain one or more new numerical ranges, and these new numerical ranges should also be considered as being specifically disclosed herein.

In the present application, the terms "upstream" and "downstream" are based on the flow direction of the reaction materials. For example, when the reaction materials flow from bottom to top, "upstream" indicates a position located below, and "downstream" indicates a position located above.

Unless otherwise stated, the terms used herein have the same meanings as commonly understood by those skilled in the art. If a term is defined herein and its definition differs from the common understanding in the art, the definition herein shall prevail.

The specific embodiments of the present invention are described in detail below. It should be understood that the specific embodiments described herein are only used to illustrate and explain the present invention, and are not intended to limit the present invention.

In addition, the technical features involved in the different embodiments of the present application described below can be combined with each other as long as they do not conflict with each other.

In the context of the present invention, unless otherwise specifically stated, the physical property values of substances (such as boiling points) are measured at normal temperature/room temperature (25°C) and atmospheric pressure (101325 Pa).

Before setting forth the technical solution of the present invention, the terms used herein are defined as follows:

The term "initial boiling point" means: the distillation range starting point temperature.

The term "final boiling point" means: the distillation range end point temperature. It is determined using the internationally accepted standard method ASTM D86.

The light aromatic hydrocarbons in the present invention refer to C6-C8 aromatic hydrocarbons, specifically benzene, toluene, xylene, and ethylbenzene.

The present invention provides a method for producing light aromatic hydrocarbons, wherein the method comprises the following steps:

(1) subjecting a catalytic cracking cycle oil to two-stage hydrotreating in the presence of hydrogen and a hydrogenation catalyst to obtain a hydrogenated product;

(2) removing hydrogen from the hydrogenated product to obtain a hydrogenated oil-gas mixture, and passing the hydrogenated oil-gas mixture through a first distillation column to obtain an overhead light fraction, a light aromatic hydrocarbon-rich fraction, a first recycle fraction and a second recycle fraction, wherein the first recycle fraction has an initial boiling point of 140-170°C and a final boiling point of 220-300°C, and the second recycle fraction has a distillation range starting point temperature of any value between 220-300°C;

(3) introducing the first recycle fraction obtained in step (2) into a first reaction zone of a catalytic cracking unit for a catalytic cracking reaction, and optionally, introducing the second recycle fraction into a second reaction zone of the catalytic cracking unit or returning it to a second stage of said hydrotreating for recycling;

(4) separating the reaction product obtained from the catalytic cracking unit by a second distillation column to obtain a first fraction oil-gas mixture, a second fraction oil and a slurry oil, returning the second fraction oil to a first stage of said hydrotreating, and optionally, introducing the first fraction oil-gas mixture together with the hydrogenated oil-gas mixture into the first distillation column of step (2) for separation;

(5) separating light aromatic hydrocarbons from the light aromatic hydrocarbon-rich fraction.

The inventors of the present invention have found that by effectively separating the catalytic cracking product and the hydrogenated product, and returning the heavy aromatic hydrocarbon fraction to the catalytic cracking riser for zone-specific conversion, the yield of light aromatic hydrocarbons from catalytic cracking can be increased. The present invention performs effective separation of the catalytic conversion product and the hydrogenated product in the same separation system and then performs zone-specific catalytic conversion, which increases the yield of light aromatic hydrocarbons while saving the equipment investment for the separation unit.

In the present invention, the catalytic cracking cycle oil used as the feedstock may be a catalytic cracking cycle oil (light cycle oil, LCO) known in the art. Preferably, the catalytic cracking cycle oil of the present invention has an initial boiling point of 160-250°C and a final boiling point of 340-400°C.

In the present invention, the hydrotreating of the catalytic cracking cycle oil may be carried out using conditions and catalysts known in the art.

In one embodiment, in step (1), the reaction conditions for the first stage of the hydrotreating are: hydrogen partial pressure 5.0-20.0 MPa, reaction temperature 300-450°C, volumetric space velocity 0.5-5.0 h^{-1}, hydrogen-to-hydrocarbon volume ratio 300-1600 Nm^3/m^3.

In one embodiment, in step (1), the reaction conditions for the second stage of the hydrotreating are: hydrogen partial pressure 5.0-20.0 MPa, reaction temperature 350-500°C, preferably 10-50°C higher than the temperature of the first stage of the hydrotreating, volumetric space velocity 0.1-3.0

h^{-1}, hydrogen-to-hydrocarbon volume ratio 400-2000 Nm^3/m^3.

According to the method of the first aspect, wherein in step (1), in the two-stage hydrotreating reaction, the hydrogenation catalysts in the first and second stages are the same or different.

According to the method of the first aspect, wherein in step (1), the hydrogenation catalyst for the first stage of said hydrotreating comprises an active metal component and a carrier, and the hydrogenation catalyst for the second stage of hydrotreating comprises an active metal, a molecular sieve and a carrier, wherein the active metal component is selected from one or more of Group VIB metals and/or Group VIII non-noble metals, the carrier is selected from one or more of alumina, silica, and amorphous silica-alumina, and the molecular sieve is selected from one or more of Y-type molecular sieve and USY molecular sieve;

more preferably, the active metal component is selected from any one of the following metal combinations: nickel-tungsten, nickel-tungsten-cobalt, nickel-molybdenum, cobalt-molybdenum.

In an embodiment, in step (1), the catalytic cracking cycle oil has an initial boiling point of 160-250°C and a final boiling point of 340-400°C.

In an embodiment, in step (2), the hydrogenated oil-gas mixture has an initial boiling point of 0-60°C and a final boiling point of 280-350°C.

In an embodiment, in step (2), the first distillation column is a divided-wall distillation column, at least one dividing wall is arranged in the divided-wall distillation column, and the dividing wall divides the divided-wall distillation column into four regions: a pre-separation zone, a public rectifying section, a public stripping section and a side-draw rectifying zone, and the feed position of the hydrogenated oil-gas mixture is located in the pre-separation zone.

In an embodiment, in step (2), the first distillation column is a divided-wall distillation column, the pre-separation zone of the divided-wall distillation column has 10-40 trays, preferably 20-35 trays;

the public rectifying section has 2-10 trays, preferably 3-8 trays;

the public stripping section has 2-10 trays, preferably 3-8 trays;

the side-draw rectifying zone has 10-45 trays, preferably 25-40 trays;

the feed position of the hydrogenated oil-gas mixture is above the 4th to 22nd tray from the top of the pre-separation zone, preferably above the 8th to 20th tray;

In an embodiment, in step (2), the overhead pressure of the divided-wall distillation column is 0.1-1.0 MPa, preferably 0.2-0.7 MPa, and the overhead temperature is 50-120°C, preferably 60-100°C; and/or

the overhead reflux ratio is 3-10, preferably 4-8.

In an embodiment, in step (2), the divided-wall distillation column satisfies at least one of the following requirements:

the side-draw rectifying zone of the divided-wall distillation column is provided with a

middle-stage circulation heat removal, with a draw position at the 20th to 45th tray of the side-draw rectifying zone, more preferably at the 30th to 40th tray, and a return position at the same tray as the draw position or one tray above or below the draw position;

the pre-separation zone of the divided-wall distillation column is provided with a vapor draw, with a draw position at the 20th to 40th tray of the pre-separation zone, more preferably at the 25th to 35th tray, and a vapor return position at the 30th to 45th tray of the side-draw rectifying zone, more preferably at the 35th to 40th tray;

the mass flow ratio of the liquid phases from the public rectifying section into the pre-separation zone and the side-draw rectifying zone is 0.1-10, preferably 0.5-5;

the mass flow ratio of the vapor phases from the public stripping section into the pre-separation zone and the side-draw rectifying zone is 1-15, preferably 2-10; and/or

the ratio of the total vapor mass flow to the drawn vapor mass flow at the draw position of the pre-separation zone is 10-150, preferably 30-120.

In an embodiment, in step (2), the divided-wall distillation column satisfies at least one of the following requirements:

the overhead light fraction is separated from the public rectifying section of the divided-wall distillation column, and has an initial boiling point of 20-40°C and a final boiling point of 55-65°C;

the light aromatic hydrocarbon-rich fraction is drawn from the 5th to 30th tray of the side-draw rectifying zone of the divided-wall distillation column, preferably from the 10th to 20th tray, and has an initial boiling point of 55-65°C and a final boiling point of 140-170°C;

the first recycle fraction is drawn from any position below the draw position of the light aromatic hydrocarbon-rich fraction in the side-draw rectifying zone, preferably from the 25th to 35th tray; and/or

the second recycle fraction is separated from the lower part of the public stripping section of the divided-wall distillation column. In an embodiment, in step (3), the catalytic conversion conditions in the first reaction zone satisfy at least one of the following requirements:

the weight hourly space velocity of oil-gas mixture is 4-20 h^{-1}, preferably 6-12 h^{-1};

the reaction time is 0.5-8 s, preferably 1-5 s; and/or

the catalyst-to-hydrocarbon weight ratio is 6-26, preferably 8-12.

In an embodiment, in step (3), the catalytic conversion conditions in the second reaction zone satisfy at least one of the following requirements:

the weight hourly space velocity of oil-gas mixture is 4-20 h^{-1}, preferably 8-12 h^{-1};

the reaction time is 1-10 s, preferably 2-6 s; and/or

the catalyst-to-hydrocarbon weight ratio is 4-14, preferably 6-10.

In the present invention, the catalytic cracking catalyst used in the catalytic cracking unit may be a

Y-type molecular sieve catalytic cracking catalyst known in the art.

In an embodiment, the second distillation column in step (4) satisfies at least one of the following requirements:

the distillation column has an overhead pressure of 0.15-0.4 MPa, preferably 0.18-0.35 MPa, and an overhead temperature of 100-160°C, preferably 115-145°C,

the first fraction has an initial boiling point of 0-40°C and a final boiling point of 220-250°C, and the first fraction, after being combined with the hydrogenated fraction and degassed, is introduced into the first distillation column of step (2) for separation;

the second fraction oil has an initial boiling point of 220-250°C and a final boiling point of 340-360°C, and its initial boiling point is not lower than the initial boiling point of the first fraction oil-gas mixture; and/or

the slurry oil has an initial boiling point of 340-360°C.

In an embodiment, in step (2), the first recycle fraction has an initial boiling point of 140-170°C and a final boiling point of 220-250°C, and the second recycle fraction has a distillation range starting point temperature of any value between 220-250°C.

In an embodiment, in step (2), the first recycle fraction has an initial boiling point of 140-170°C and a final boiling point of 250-300°C, and the second recycle fraction has a distillation range starting point temperature of any value between 250-300°C.

In an embodiment, when the distillation range starting point temperature of the second recycle fraction is any value between 220-250°C, in step (3), the second recycle fraction is introduced into the second reaction zone of the catalytic cracking unit.

In an embodiment, when the distillation range starting point temperature of the second recycle fraction is any value between 270-300°C, in step (3), the second recycle fraction is returned to the second stage of the hydrotreating for recycling.

In an embodiment, when the distillation range starting point temperature of the second recycle fraction is any value between 250-270°C, in step (3), the second recycle fraction is introduced into the second reaction zone of the catalytic cracking unit or returned to the second stage of the hydrotreating for recycling.

In an embodiment, in step (5), light aromatic hydrocarbons are separated from the light aromatic hydrocarbon-rich fraction. The separation may be a method known in the art, including but not limited to fractionation, extraction, solvent extraction, adsorption, membrane separation, etc.

The present invention further provides a system for producing light aromatic hydrocarbons, comprising:

a hydrogenation unit, wherein the hydrogenation unit is a two-stage hydrogenation unit and provided with a hydrogen inlet and a hydrogenated product outlet, a first stage of the hydrogenation

unit is provided with a catalytic cracking cycle oil inlet, and a second stage of the hydrogenation unit is optionally provided with a second recycle fraction inlet;

a catalytic cracking unit, comprising a catalytic cracking reactor, wherein the catalytic cracking reactor is provided with a first reaction zone and an optional second reaction zone, the first reaction zone is provided with a first recycle fraction inlet, and the optional second reaction zone is provided with a second recycle fraction inlet;

a second distillation column, wherein the second distillation column is provided with a catalytic cracking reaction product inlet, a first fraction oil-gas mixture outlet, a second fraction oil outlet and a slurry oil outlet, and the second fraction oil outlet is in communication with the catalytic cracking cycle oil inlet of the hydrogenation unit;

a degassing separation unit, wherein the degassing separation unit is provided with an oil-gas mixture inlet, a rich gas outlet and a fraction oil outlet, and the oil-gas mixture inlet of the degassing separation unit is connected to the first fraction oil-gas mixture outlet of the second distillation column and the hydrogenated product outlet of the hydrogenation unit, so that the first fraction oil-gas mixture from the second distillation column and the hydrogenated oil-gas mixture from the hydrogenation unit are degassed and separated into a rich gas and a fraction oil;

a first distillation column, wherein the distillation column is provided with a fraction oil inlet, an overhead light fraction outlet, a light aromatic hydrocarbon-rich fraction outlet, a first recycle fraction outlet, and a second recycle fraction outlet, the fraction oil inlet is in communication with the fraction oil outlet of the degassing separation unit, the first recycle fraction outlet is in communication with the first recycle fraction inlet of the catalytic cracking unit, the second recycle fraction outlet is in communication with the second recycle fraction inlet provided for the optional second reaction zone of the catalytic cracking unit, or the second recycle fraction outlet is in communication with the second recycle fraction inlet optionally provided for the second stage of the hydrogenation unit;

a separation unit, which is provided with a light aromatic hydrocarbon-rich fraction inlet and a light aromatic hydrocarbon fraction outlet, and the light aromatic hydrocarbon-rich fraction inlet is connected to the light aromatic hydrocarbon-rich fraction outlet of the first distillation column.

In an embodiment, in the system of the present invention, the first distillation column is a divided-wall distillation column, at least one dividing wall is arranged in the divided-wall distillation column, and the dividing wall divides the divided-wall distillation column into four regions: a public stripping section arranged at the bottom of the column, a public rectifying section arranged at the top of the column, a pre-separation zone and a side-draw rectifying zone arranged in the middle section of the column; the pre-separation zone is provided with a fraction oil inlet, and the fraction oil inlet is in communication with the fraction oil outlet of the degassing separation unit;

the public rectifying section is provided with an overhead light fraction outlet, the side-draw rectifying zone is provided with a light aromatic hydrocarbon-rich fraction outlet and a first recycle fraction outlet, and the bottom of the public stripping section is provided with a second recycle fraction outlet,

here, the first recycle fraction outlet of the side-draw rectifying zone of the divided-wall distillation column is in communication with the first recycle fraction inlet of the catalytic cracking reactor,

the second recycle fraction outlet at the bottom of the public stripping section of the divided-wall distillation column is in communication with the second recycle fraction inlet provided for the optional second reaction zone of the first distillation column, or is in communication with the second recycle fraction inlet optionally provided on the second stage of the hydrogenation unit.

In an embodiment, in the system of the present invention, the system further comprises a high/low pressure separation unit, the high/low pressure separation unit is provided with a hydrogenated product inlet, a hydrogen outlet and a hydrogenated oil-gas mixture outlet, and the high/low pressure separation unit separates the hydrogenated oil-gas mixture from hydrogen in the hydrogenated product from the hydrogenation unit;

here, the hydrogenated product inlet is connected to the hydrogenated product outlet of the hydrogenation unit, and the oil-gas mixture inlet of the degassing separation unit is connected to the first fraction oil-gas mixture outlet of the second distillation column and the hydrogenated oil-gas mixture outlet of the high/low pressure separation unit.

In an embodiment, in the system of the present invention, the first reaction zone and the second reaction zone of the catalytic cracking reactor are arranged on the same or different reactors;

preferably, the reactor is selected from one or more of the following: an equal-diameter riser, an equal-linear-velocity riser, a variable-diameter riser, a variable-linear-velocity riser, a fluidized bed, and a composite reactor, wherein the composite reactor is composed of an equal-diameter or variable-diameter riser and a fluidized bed, and preferably the reactor is a riser reactor;

more preferably, the first reaction zone and the second reaction zone of the catalytic cracking reactor are arranged on a dual-riser reactor, wherein the main riser is the first reaction zone, and the auxiliary riser is the second reaction zone.

In an embodiment, the system for producing light aromatic hydrocarbons of the present invention comprises:

a hydrogenation unit, wherein the hydrogenation unit is a two-stage hydrogenation unit and provided with a hydrogen inlet and a hydrogenated product outlet, a first stage of the hydrogenation unit is provided with a catalytic cracking cycle oil inlet;

a catalytic cracking unit, comprising a catalytic cracking reactor, wherein the catalytic cracking reactor is provided with a first reaction zone and a second reaction zone, the first reaction zone is

provided with a first recycle fraction inlet, and the second reaction zone is provided with a second recycle fraction inlet;

a second distillation column, wherein the second distillation column is provided with a catalytic cracking reaction product inlet, a first fraction oil-gas mixture outlet, a second fraction oil outlet and a slurry oil outlet, and the second fraction oil outlet is in communication with the catalytic cracking cycle oil inlet of the hydrogenation unit;

a degassing separation unit, wherein the degassing separation unit is provided with an oil-gas mixture inlet, a rich gas outlet and a fraction oil outlet, and the oil-gas mixture inlet of the degassing separation unit is connected to the first fraction oil-gas mixture outlet of the second distillation column and the hydrogenated product outlet of the hydrogenation unit, so that the first fraction oil-gas mixture from the second distillation column and the hydrogenated oil-gas mixture from the hydrogenation unit are degassed and separated into a rich gas and a fraction oil;

a first distillation column, wherein the distillation column is provided with a fraction oil inlet, an overhead light fraction outlet, a light aromatic hydrocarbon-rich fraction outlet, a first recycle fraction outlet, and a second recycle fraction outlet, the fraction oil inlet is in communication with the fraction oil outlet of the degassing separation unit, the first recycle fraction outlet is in communication with the first recycle fraction inlet of the catalytic cracking unit, the second recycle fraction outlet is in communication with the second recycle fraction inlet provided for the second reaction zone of the catalytic cracking unit;

a separation unit, which is provided with a light aromatic hydrocarbon-rich fraction inlet and a light aromatic hydrocarbon fraction outlet, and the light aromatic hydrocarbon-rich fraction inlet is connected to the light aromatic hydrocarbon-rich fraction outlet of the first distillation column.

In an embodiment, the system for producing light aromatic hydrocarbons of the present invention comprises:

a hydrogenation unit, wherein the hydrogenation unit is a two-stage hydrogenation unit and provided with a hydrogen inlet and a hydrogenated product outlet, a first stage of the hydrogenation unit is provided with a catalytic cracking cycle oil inlet, and a second stage of the hydrogenation unit is provided with a second recycle fraction inlet;

a catalytic cracking unit, comprising a catalytic cracking reactor, wherein the catalytic cracking reactor is provided with a first reaction zone, the first reaction zone is provided with a first recycle fraction inlet;

a second distillation column, wherein the second distillation column is provided with a catalytic cracking reaction product inlet, a first fraction oil-gas mixture outlet, a second fraction oil outlet and a slurry oil outlet, and the second fraction oil outlet is in communication with the catalytic cracking cycle oil inlet of the hydrogenation unit;

a degassing separation unit, wherein the degassing separation unit is provided with an oil-gas mixture inlet, a rich gas outlet and a fraction oil outlet, and the oil-gas mixture inlet of the degassing separation unit is connected to the first fraction oil-gas mixture outlet of the second distillation column and the hydrogenated product outlet of the hydrogenation unit, so that the first fraction oil-gas mixture from the second distillation column and the hydrogenated oil-gas mixture from the hydrogenation unit are degassed and separated into a rich gas and a fraction oil;

a first distillation column, wherein the distillation column is provided with a fraction oil inlet, an overhead light fraction outlet, a light aromatic hydrocarbon-rich fraction outlet, a first recycle fraction outlet, and a second recycle fraction outlet, the fraction oil inlet is in communication with the fraction oil outlet of the degassing separation unit, the first recycle fraction outlet is in communication with the first recycle fraction inlet of the catalytic cracking unit, the second recycle fraction outlet is in communication with the second recycle fraction inlet provided for the second stage of the hydrogenation unit;

a separation unit, which is provided with a light aromatic hydrocarbon-rich fraction inlet and a light aromatic hydrocarbon fraction outlet, and the light aromatic hydrocarbon-rich fraction inlet is connected to the light aromatic hydrocarbon-rich fraction outlet of the first distillation column.

With reference to Figure 1, the present invention is described. As a preferred embodiment of the present invention, a catalytic cracking cycle oil having a distillation range of 220-370°C is introduced into the hydrogenation unit 100 via inlet 101, and is brought into contact with hydrogen and a hydrogenation catalyst to carry out a two-stage hydrogenation reaction. The resulting hydrogenated product is introduced via pipeline 102 into a high/low pressure separation unit. After a recycle hydrogen and a low-pressure gas are separated, the resulting hydrogenated oil-gas mixture is introduced via pipeline 103 into a degassing separation unit. The separated fraction oil is introduced via pipeline 201 into the pre-separation zone A of the divided-wall distillation column 200 serving as the first distillation column. An overhead light fraction of less than 60°C is separated at the column top from the public rectifying section B, and leaves via pipeline 202. A second recycle fraction of greater than 240°C is separated at the column bottom from the public stripping section D, introduced via pipeline 205 to the feedstock oil inlet 304 of the secondary riser 302 of the catalytic cracking unit 300, and used as the feedstock for the second reaction zone to carry out a catalytic cracking reaction. A light aromatic hydrocarbon-rich fraction having a distillation range of 80-150°C is separated at the upper part of the side-draw rectifying zone C and leaves via pipeline 203. A first recycle fraction having a distillation range of 150-240°C is separated at the lower part of the side-draw rectifying zone C, introduced via pipeline 204 to the feedstock oil inlet 303 of the first riser 301 of the catalytic cracking unit 300, and used as the feedstock for the first reaction zone to carry out a catalytic cracking reaction. The resulting catalytic cracking product is introduced via

product pipeline 305 to the second distillation column 400 for separation. A first fraction oil-gas mixture of less than 240°C is separated, leaves via pipeline 401, combined with the hydrogenated oil-gas mixture from pipeline 103 into the degassing separation unit. A second fraction oil having a distillation range of 240-350°C is returned via pipeline 402 to the inlet 101 of the hydrogenation unit for recycling. The catalytic cracking unit employs a dual-riser reactor, wherein the first reaction zone is located in the first riser, and the second reaction zone is located in the second riser. Then, a separation unit is used to separate light aromatic hydrocarbons from the light aromatic hydrocarbon-rich fraction.

With reference to Figure 2, the present invention is described. As another preferred embodiment of the present invention, a catalytic cracking cycle oil having a distillation range of 220-370°C is introduced into the hydrogenation unit 100 via inlet 101, and is brought into contact with hydrogen and a hydrogenation catalyst to carry out a two-stage hydrogenation reaction. The resulting hydrogenated product is introduced via pipeline 102 into a high/low pressure separation unit. After a recycle hydrogen and a low-pressure gas are separated, the resulting hydrogenated oil-gas mixture is introduced via pipeline 103 into a degassing separation unit. The separated fraction oil is introduced via pipeline 201 into the pre-separation zone A of the divided-wall distillation column 200 serving as the first distillation column. An overhead light fraction of less than 60°C is separated at the column top from the public rectifying section B, and leaves via pipeline 202. A second recycle fraction of greater than 280°C is separated at the column bottom from the public stripping section D, introduced via pipeline 205 to the second stage of the hydrogenation reaction of the hydrogenation unit 100. A light aromatic hydrocarbon-rich fraction having a distillation range of 80-150°C is separated at the upper part of the side-draw rectifying zone C and leaves via pipeline 203. A first recycle fraction having a distillation range of 150-280°C is separated at the lower part of the side-draw rectifying zone C, introduced via pipeline 204 to the feedstock oil inlet 303 of the first riser 301 of the catalytic cracking unit 300, and used as the feedstock for the first reaction zone to carry out a catalytic cracking reaction. The resulting catalytic cracking product is introduced via product pipeline 305 to the second distillation column 400 for separation. A first fraction oil-gas mixture of less than 240°C is separated, leaves via pipeline 401, combined with the hydrogenated oil-gas mixture from pipeline 103 into the degassing separation unit. A second fraction oil having a distillation range of 240-350°C is returned via pipeline 402 to the inlet 101 of the hydrogenation unit 100 for recycling. The catalytic cracking unit employs a single riser reactor. Then, a separation unit is used to separate light aromatic hydrocarbons from the light aromatic hydrocarbon-rich fraction.

In some embodiments of the present invention, the following Scheme I is provided:

Scheme I-1. 1. A method for producing light aromatic hydrocarbons, characterized in that the

method comprises the following steps:

(1) introducing a catalytic cracking feedstock oil into a first reaction zone of a catalytic cracking unit to contact with a catalyst for catalytic cracking, to obtain a first reaction oil-gas mixture;

(2) after gas-solid separation of the first reaction oil-gas mixture from the catalyst, introducing it into a first distillation column to separate and obtain a first fraction oil-gas mixture, a second fraction oil and a slurry oil;

(3) subjecting the second fraction oil obtained in step (2) to hydrotreating, removing hydrogen from the hydrogenated product obtained from the hydrotreating to obtain a hydrogenated oil-gas mixture, and subjecting the hydrogenated oil-gas mixture and the first fraction oil-gas mixture to degassing separation to obtain a rich gas and a first fraction oil;

(4) introducing the first fraction oil obtained in step (3) into a second distillation column to separate and obtain a naphtha fraction rich in light aromatic hydrocarbons, a heavy aromatic hydrocarbon fraction and a diesel fraction;

(5) introducing the heavy aromatic hydrocarbon fraction obtained in step (4) into a second reaction zone of the catalytic cracking unit and the diesel fraction into the first reaction zone of the catalytic cracking unit for catalytic conversion;

Scheme I-2. The method according to Scheme I-1, characterized in that in step (1), the catalytic cracking feedstock oil is selected from at least one of hydrogenated diesel, straight-run diesel, straight-run gas oil, hydrogenated gas oil, hydrogenated cracking tail oil, hydrogenated residue oil, and the diesel fraction separated from the second distillation column.

Scheme I-3. The method according to Scheme I-1, characterized in that in step (1), the catalytic conversion conditions in the first reaction zone are:

the weight hourly space velocity of oil-gas mixture is 4-20 h^{-1}, preferably 6-12 h^{-1};

the reaction time is 1-10 s, preferably 2-6 s; and/or

the catalyst-to-hydrocarbon weight ratio is 4-14, preferably 6-10;

Scheme I-4. The method according to Scheme I-1, characterized in that in step (2), the first fraction oil-gas mixture has an initial boiling point of 0-40°C and a final boiling point of 220-250°C;

the second fraction oil has an initial boiling point of 220-250°C and a final boiling point of 340-360°C; and/or

the slurry oil has an initial boiling point of 340-360°C;

Scheme I-5. The method according to Scheme I-1, characterized in that in step (3), the hydrotreating is hydrofining, hydro-upgrading or hydrocracking; and/or

the hydrogenated oil-gas mixture has an initial boiling point of 0-60°C and a final boiling point of 280-350°C;

Scheme I-6. The method according to Scheme I-1, characterized in that in step (4), the naphtha

fraction has an initial boiling point of 20-40°C and a final boiling point of 140-160°C;

the heavy aromatic hydrocarbon fraction has an initial boiling point of 140-160°C and a final boiling point of 210-250°C;

the diesel fraction has an initial boiling point of 210-250°C and a final boiling point of 280-350°C;

Scheme I-7. The method according to Scheme I-1, characterized in that in step (5), the catalytic conversion conditions in the second reaction zone are:

the weight hourly space velocity of oil-gas mixture is 4-20 h^{-1}, preferably 8-16 h^{-1};

the reaction time is 0.1-5 s, preferably 0.5-3 s; and/or

the catalyst-to-hydrocarbon weight ratio is 6-26, preferably 10-22;

Scheme I-8. The method according to Scheme I-1, characterized in that in step (2), the first distillation column has an overhead pressure of 0.1-0.5 MPa, preferably 0.2-0.4 MPa;

Scheme I-9. The method according to Scheme I-1, characterized in that in step (4), the second distillation column is a conventional distillation column or a divided-wall distillation column;

preferably, the second distillation column has an overhead pressure of 0.1-1.0 MPa, more preferably 0.2-0.7 MPa; and/or

the second distillation column has an overhead temperature of 50-120°C, more preferably 60-100°C;

further preferably, the second distillation column has 30-50 trays, preferably 35-45 trays; and the naphtha fraction rich in light aromatic hydrocarbons is drawn from the side of the second distillation column at the 10th to 40th tray from the top, preferably at the 15th to 30th tray;

Scheme I-10. The method according to Scheme I-9, characterized in that, the second distillation column is a divided-wall distillation column, at least one dividing wall is arranged in the divided-wall distillation column, and the dividing wall divides the divided-wall distillation column into four regions: a pre-separation zone, a public rectifying section, a public stripping section and a side-draw rectifying zone, and the feed position of the oil-gas mixture is located in the pre-separation zone;

preferably, the pre-separation zone of the divided-wall distillation column has 10-40 trays, preferably 20-35 trays; the public rectifying section has 3-20 trays, preferably 5-15 trays; the public stripping section has 2-10 trays, preferably 3-8 trays; and/or the side-draw rectifying zone has 10-45 trays, preferably 25-40 trays;

more preferably, the mass flow ratio of the liquid phases from the public rectifying section into the pre-separation zone and the side-draw rectifying zone is 0.1-10, preferably 0.5-5; and/or

more preferably, the mass flow ratio of the vapor phases from the public stripping section into the pre-separation zone and the side-draw rectifying zone is 1-15, preferably 2-10;

Scheme I-11. The method according to Scheme I-1, characterized in that in step (1), the catalyst is a

catalytic cracking catalyst, and the catalyst comprises 10-60 parts by weight of a molecular sieve, 1-40 parts by weight of a binder and 1-90 parts by weight of a carrier;

wherein the molecular sieve is selected from one or more of ZSM molecular sieve, Y molecular sieve, HY molecular sieve, USY molecular sieve, and Beta molecular sieve, and optionally, the molecular sieve contains rare earth, which is one or more of La, Ce, Pr and Nd;

the binder is a silica binder and/or an alumina binder; and/or

the carrier is selected from one or more of silica, kaolin, montmorillonite, diatomaceous earth, halloysite, saponite, rectorite, sepiolite, attapulgite, hydrotalcite and bentonite;

Scheme I-12. A system for producing light aromatic hydrocarbons, comprising:

a catalytic cracking unit, comprising:

a reactor, wherein the reactor is provided with a first reaction zone and a second reaction zone, wherein the first reaction zone and the second reaction zone are arranged in series, and the second reaction zone is arranged upstream of the first reaction zone, and the feedstock oil is sequentially contacted with the catalyst in the second reaction zone and the catalyst in the first reaction zone;

a gas-solid separator, wherein the gas-solid separator is in communication with a hydrocarbon-catalyst mixture outlet of the first reaction zone of the reactor, so that the first reaction oil-gas mixture from the first reaction zone is separated from the catalyst;

a first distillation column, wherein the first distillation column is provided with a first fraction oil-gas mixture outlet, a second fraction oil outlet and a slurry oil outlet, and the inlet of the first distillation column is in communication with the outlet of the catalytic cracking unit via pipeline;

a diesel hydrogenation unit, wherein the diesel hydrogenation unit is provided with a diesel inlet, a hydrogen inlet and a hydrogenated product outlet, and the diesel inlet of the diesel hydrogenation unit is in communication with the second fraction oil outlet of the first distillation column;

a high/low pressure separation unit, wherein the high/low pressure separation unit is provided with a hydrogenated product inlet, a hydrogen outlet and a hydrogenated oil-gas mixture outlet, and the hydrogenated product inlet is connected to the hydrogenated product outlet of the diesel hydrogenation unit, and the high/low pressure separation unit separates the hydrogenated oil-gas mixture from hydrogen in the hydrogenated product from the diesel hydrogenation unit;

a degassing separation unit, wherein the degassing separation unit is provided with an oil-gas mixture inlet, a rich gas outlet and a fraction oil outlet, and the oil-gas mixture inlet of the degassing separation unit is connected to the first fraction oil-gas mixture outlet of the first distillation column and the hydrogenated product outlet of the high/low pressure separation unit, so that the first fraction oil-gas mixture from the first distillation column and the hydrogenated oil-gas mixture from the high/low pressure separation unit are degassed and separated into a rich gas and a first fraction oil;

a second distillation column, wherein the second distillation column is provided with a first fraction oil inlet, a naphtha fraction outlet, a heavy aromatic hydrocarbon fraction outlet and a diesel fraction outlet, the first fraction oil inlet is connected to the first fraction oil outlet of the degassing separation unit, the heavy aromatic hydrocarbon fraction outlet is connected to the second reaction zone of the reactor of the catalytic cracking unit, and the diesel fraction outlet is connected to the first reaction zone of the reactor of the catalytic cracking unit;

Scheme I-13. The system according to Scheme I-12, characterized in that the reactor is selected from one or two of the following: an equal-diameter riser, an equal-linear-velocity riser, a variable-diameter riser, a variable-linear-velocity riser, a fluidized bed, and a composite reactor, wherein the composite reactor is composed of an equal-diameter or variable-diameter riser and a fluidized bed;

the first reaction zone and the second reaction zone are arranged on the same or different reactors;

Scheme I-14. The system according to Scheme I-12, characterized in that the first distillation column is a conventional distillation column having 20-40 trays, preferably 25-35 trays, and the second fraction oil is drawn from the side of the first distillation column at the 5th to 30th tray from the top, preferably at the 10th to 25th tray;

preferably, the first distillation column is supplied with heat by vapor feed, and steam stripping is employed at the bottom of the column; and/or

preferably, the side of the first distillation column may be provided with one or more middle-stage circulation heat removals.

Scheme I-15. The system according to any one of Schemes I-12 to I-14, characterized in that the first distillation column and/or the second distillation column is an equal-diameter distillation column or a variable-diameter distillation column.

In some embodiments of the present invention, the following Scheme II is provided:

Scheme II-1. A catalytic conversion method for producing light aromatic hydrocarbons, characterized in that the method comprises the following steps:

(1) introducing a catalytic cracking cycle oil into a hydrogenation unit, and contacting it with hydrogen in the presence of a hydrogenation catalyst to carry out a two-stage hydrotreating reaction, to obtain a hydrogenated oil-gas mixture;

(2) introducing the hydrogenated oil-gas mixture into a divided-wall distillation column to separate and obtain a gas, a hydrogenated light naphtha, a hydrogenated heavy naphtha, a hydrogenated light diesel and a hydrogenated heavy diesel;

(3) returning the hydrogenated heavy diesel obtained in step (2) to the second stage of the hydrogenation unit for recycling, introducing the hydrogenated light diesel obtained in step (2) into a catalytic cracking unit, and subjecting it to a catalytic cracking reaction in the presence of a

catalytic cracking catalyst, and separating the catalytic cracking reaction product to obtain dry gas, liquefied gas, a catalytic naphtha rich in light aromatic hydrocarbons, a catalytic heavy aromatic hydrocarbon fraction, a catalytic cracking cycle oil and a slurry oil;

(4) returning the catalytic cracking cycle oil obtained in step (3) to the first stage of the hydrogenation unit for recycling; optionally, returning the catalytic heavy aromatic hydrocarbon fraction obtained in step (3) to the catalytic cracking unit for recycling conversion;

Scheme II-2. The catalytic conversion method according to Scheme II-1, characterized in that in step (1), the hydrotreating reaction conditions are:

preferably, the reaction conditions for the first stage of said hydrotreating are: hydrogen partial pressure 5.0-20.0 MPa, reaction temperature 300-450°C, volumetric space velocity 0.5-5.0 h^{-1}, hydrogen-to-hydrocarbon volume ratio 300-1600 Nm^3/m^3; and/or

preferably, the reaction conditions for the second stage of said hydrotreating are: hydrogen partial pressure 5.0-20.0 MPa, reaction temperature 350-500°C, volumetric space velocity 0.1-3.0 h^{-1}, hydrogen-to-hydrocarbon volume ratio 400-2000 Nm^3/m^3;

Scheme II-3. The catalytic conversion method according to Scheme II-1, characterized in that in step (1), in the two-stage hydrotreating reaction, the hydrogenation catalysts in the first and second stages are the same or different;

preferably, the hydrogenation catalyst comprises an active metal component and a carrier, wherein the active metal component is selected from one or more of Group VIB metals and/or Group VIII non-noble metals, and the carrier is selected from one or more of alumina, silica, and amorphous silica-alumina;

more preferably, the active metal component is selected from any one of the following metal combinations: nickel-tungsten, nickel-tungsten-cobalt, nickel-molybdenum, cobalt-molybdenum;

Scheme II-4. The catalytic conversion method according to Scheme II-1, characterized in that in step (2), at least one dividing wall is arranged in the divided-wall distillation column, and the dividing wall divides the divided-wall distillation column into four regions: a pre-separation zone, a public rectifying section, a public stripping section and a side-draw rectifying zone, and the feed position of the hydrogenated oil-gas mixture is located in the pre-separation zone;

Scheme II-5. The method according to Scheme II-4, characterized in that the divided-wall distillation column has an overhead pressure of 0.1-1.0 MPa, preferably 0.2-0.7 MPa;

has an overhead temperature of 50-120°C, preferably 60-100°C;

the mass flow ratio of the liquid phases from the public rectifying section into the pre-separation zone and the side-draw rectifying zone is 0.1-10, preferably 0.5-5;

the mass flow ratio of the vapor phases from the public stripping section into the pre-separation zone and the side-draw rectifying zone is 1-15, preferably 2-10; and/or

the ratio of the total vapor mass flow to the drawn vapor mass flow at the draw position of the pre-separation zone is 10-150, preferably 30-120;

Scheme II-6. The catalytic conversion method according to Scheme II-4, characterized in that the pre-separation zone of the divided-wall distillation column has 10-40 trays, preferably 20-35 trays;

the public rectifying section has 2-10 trays, preferably 3-8 trays; the public stripping section has 2-10 trays, preferably 3-8 trays; and/or

the side-draw rectifying zone has 10-45 trays, preferably 25-40 trays;

Scheme II-7. The catalytic conversion method according to Scheme II-6, characterized in that in the side-draw rectifying zone, the draw and return positions are at the 20th to 45th tray of the side-draw rectifying zone, preferably at the 30th to 40th tray; preferably, the side-draw rectifying zone is further provided with a middle-stage circulation heat removal; and/or

preferably, the pre-separation zone is provided with a vapor draw, with a draw position at the 20th to 40th tray of the pre-separation zone, preferably at the 25th to 35th tray, and a vapor return position at the 30th to 45th tray of the side-draw rectifying zone, preferably at the 35th to 40th tray;

Scheme II-8. The catalytic conversion method according to any one of Schemes II-4 to II-7, characterized in that in step (2), the hydrogenated light naphtha is separated from the public rectifying section of the divided-wall distillation column, and has an initial boiling point of 20-40°C and a final boiling point of 55-65°C;

the hydrogenated heavy naphtha is drawn from the 5th to 30th tray of the side-draw rectifying zone of the divided-wall distillation column, preferably from the 10th to 20th tray, and has an initial boiling point of 55-65°C and a final boiling point of 140-170°C;

the hydrogenated light diesel is drawn from any position below the draw position of the hydrogenated heavy naphtha in the side-draw rectifying zone, preferably from the 25th to 35th tray, and has an initial boiling point of 140-170°C and a final boiling point of 230-280°C; and/or

the hydrogenated heavy diesel is separated from the lower part of the public stripping section of the divided-wall distillation column, and has an initial boiling point of 230-280°C;

Scheme II-9. The catalytic conversion method according to Scheme II-1, characterized in that in step (3), the catalytic naphtha has an initial boiling point of 20-40°C and a final boiling point of 140-160°C;

the catalytic heavy aromatic hydrocarbon fraction has an initial boiling point of 140-160°C and a final boiling point of 210-250°C; and/or

the catalytic cracking cycle oil has an initial boiling point of 210-250°C and a final boiling point of 320-360°C;

Scheme II-10. The catalytic conversion method according to Scheme II-1, characterized in that in step (3), the catalytic cracking reaction conditions include:

the weight hourly space velocity of oil-gas mixture is 4-20 h^{-1}, preferably 8-16 h^{-1};

the catalytic cracking reaction temperature is 520-650°C, preferably 560-640°C; and/or

the catalyst-to-hydrocarbon weight ratio is 6-26, preferably 8-20;

Scheme II-11. The catalytic conversion method according to Scheme II-1, characterized in that in step (3), the catalytic cracking catalyst comprises 10-60 parts by weight of a molecular sieve, 1-40 parts by weight of a binder and 1-90 parts by weight of a carrier;

the molecular sieve is selected from one or more of ZSM molecular sieve, Y molecular sieve, HY molecular sieve, USY molecular sieve, and Beta molecular sieve, and optionally, the molecular sieve contains rare earth, which is one or more of La, Ce, Pr and Nd;

the binder is selected from a silica binder and/or an alumina binder;

the carrier is selected from one or more of silica, kaolin, montmorillonite, diatomaceous earth, halloysite, saponite, rectorite, sepiolite, attapulgite, hydrotalcite and bentonite;

Scheme II-12. A catalytic conversion system for producing light aromatic hydrocarbons, comprising:

a hydrogenation unit, wherein the hydrogenation unit is a two-stage hydrogenation unit and provided with a hydrogen inlet and a hydrogenated oil-gas mixture outlet, a first stage of the hydrogenation unit is provided with a catalytic cracking cycle oil inlet, and a second stage of the hydrogenation unit is provided with a hydrogenated heavy diesel inlet;

a divided-wall distillation column, wherein at least one dividing wall is arranged in the divided-wall distillation column, the dividing wall divides the divided-wall distillation column into four regions: a public stripping section arranged at the bottom of the column, a public rectifying section arranged at the top of the column, a pre-separation zone and a side-draw rectifying zone arranged in the middle section of the column; the pre-separation zone is provided with a hydrogenated oil-gas mixture inlet, the hydrogenated oil-gas mixture inlet is in communication with the hydrogenated oil-gas mixture outlet of the hydrogenation unit; the public rectifying section is provided with a low boiling point fraction outlet, the side-draw rectifying zone is provided with an outlet for the hydrogenated heavy naphtha rich in light aromatic hydrocarbons and an outlet for the hydrogenated light diesel, the bottom of the public stripping section is provided with a hydrogenated heavy diesel outlet, the hydrogenated heavy diesel outlet is in communication with the hydrogenated heavy diesel inlet provided for the second stage of the hydrogenation unit;

a catalytic cracking unit, wherein the catalytic cracking unit is provided with an inlet of hydrogenated light diesel, an outlet of dry gas, an outlet of liquefied gas, an outlet of catalytic naphtha rich in light aromatic hydrocarbons, an outlet of catalytic heavy aromatic hydrocarbon, an outlet of catalytic cracking cycle oil, and an outlet of slurry oil, the outlet of catalytic heavy aromatic hydrocarbon is in communication with a catalytic heavy aromatic hydrocarbon

fractionation pipeline, the outlet of catalytic cracking cycle oil is in communication with the inlet of catalytic cracking cycle oil provided for the first stage of the hydrogenation unit, optionally, the catalytic heavy aromatic hydrocarbon fractionation pipeline is in communication with the inlet of hydrogenated light diesel;

Scheme II-13. The catalytic conversion system according to Scheme II-12, characterized in that the catalytic cracking unit is a single reactor or a dual reactor;

preferably, the reactor is selected from one or more of the following: an equal-diameter riser, an equal-linear-velocity riser, a variable-diameter riser, a variable-linear-velocity riser, a fluidized bed, and a composite reactor, wherein the composite reactor is composed of an equal-diameter or variable-diameter riser and a fluidized bed.

In some embodiments of the present invention, the following Scheme III is provided by the present invention:

Scheme III-1. A method for producing light aromatic hydrocarbons, characterized in that the method comprises:

(1) subjecting a catalytic cracking cycle oil to a hydrotreating reaction in the presence of hydrogen and a hydrogenation catalyst to obtain a hydrogenated oil-gas mixture;

(2) separating the hydrogenated oil-gas mixture with a divided-wall distillation column to obtain a gas, a hydrogenated light naphtha, a hydrogenated heavy naphtha rich in light aromatic hydrocarbons, a hydrogenated light diesel and a hydrogenated heavy diesel;

(3) introducing the hydrogenated light diesel obtained in step (2) into the first reaction zone of the catalytic cracking unit, and introducing the hydrogenated heavy diesel into the second reaction zone of the catalytic cracking unit, and subjecting them respectively to the catalytic cracking reaction in the presence of the catalytic cracking catalyst, and separating the reaction product to obtain dry gas, liquefied gas, catalytic naphtha, catalytic heavy aromatic hydrocarbon fraction, catalytic cracking cycle oil and slurry oil;

(4) returning the catalytic cracking cycle oil obtained in step (3) to the hydrogenation unit for recycling; optionally, returning the catalytic heavy aromatic hydrocarbon fraction obtained in step (3) to the third reaction zone of the catalytic cracking unit for recycling conversion, wherein the third reaction zone is located upstream of the first reaction zone and/or the second reaction zone;

Scheme III-2. The method according to Scheme III-1, characterized in that in step (1), the hydrotreating reaction conditions are: hydrogen partial pressure 5.0-20.0 MPa, reaction temperature 330-450°C, volumetric space velocity 0.1-3.0 h^{-1}, hydrogen-to-hydrocarbon volume ratio 350-2000 Nm^3/m^3; and/or

the catalytic cracking cycle oil has an initial boiling point of 210-250°C and a final boiling point of 340-400°C;

Scheme III-3. The method according to Scheme III-1, characterized in that in step (2), at least one dividing wall is arranged in the divided-wall distillation column, and the dividing wall divides the divided-wall distillation column into four regions: a pre-separation zone, a public rectifying section, a public stripping section and a side-draw rectifying zone, and the feed position of the hydrogenated oil-gas mixture is located in the pre-separation zone;

Scheme III-4. The method according to Scheme III-3, characterized in that in step (2), the pre-separation zone of the divided-wall distillation column has 10-40 trays, preferably 20-35 trays;

the public rectifying section has 2-10 trays, preferably 3-8 trays;

the public stripping section has 2-10 trays, preferably 3-8 trays; and/or

the side-draw rectifying zone has 10-45 trays, preferably 25-40 trays;

preferably, the side-draw rectifying zone of the divided-wall distillation column is provided with a middle-stage circulation heat removal, with a draw position at the 20th to 45th tray of the side-draw rectifying zone, more preferably at the 30th to 40th tray; and/or

preferably, the pre-separation zone of the divided-wall distillation column is provided with a vapor draw, with a draw position at the 20th to 40th tray of the pre-separation zone, more preferably at the 25th to 35th tray, and a return position at the 30th to 45th tray of the side-draw rectifying zone, more preferably at the 35th to 40th tray;

Scheme III-5. The method according to Scheme III-3, characterized in that the divided-wall distillation column has an overhead pressure of 0.1-1.0 MPa, preferably 0.2-0.7 MPa;

has an overhead temperature of 50-120°C, preferably 60-100°C;

the mass flow ratio of the liquid phases from the public rectifying section into the pre-separation zone and the side-draw rectifying zone is 0.1-10, preferably 0.5-5;

the mass flow ratio of the vapor phases from the public stripping section into the pre-separation zone and the side-draw rectifying zone is 1-15, preferably 2-10; and/or

the ratio of the total vapor mass flow to the drawn vapor mass flow at the draw position of the pre-separation zone is 10-150, preferably 30-120;

Scheme III-6. The method according to Scheme III-3, characterized in that the hydrogenated light naphtha is separated from the public rectifying section of the divided-wall distillation column, and has an initial boiling point of 20-40°C and a final boiling point of 55-65°C;

the heavy naphtha is drawn from the 5th to 30th tray of the side-draw rectifying zone of the divided-wall distillation column, preferably from the 10th to 20th tray, and has an initial boiling point of 55-65°C and a final boiling point of 140-170°C;

the hydrogenated light diesel is drawn from any position below the draw position of the heavy naphtha in the side-draw rectifying zone, preferably from the 25th to 35th tray, and has an initial boiling point of 140-170°C and a final boiling point of 220-250°C; and/or

the hydrogenated heavy diesel is separated from the lower part of the public stripping section of the divided-wall distillation column, and has a distillation range starting point temperature of any value between 220-250°C;

Scheme III-7. The method according to Scheme III-1, characterized in that in step (3), the catalytic conversion conditions in the first reaction zone comprise:

the weight hourly space velocity of oil-gas mixture is 4-20 h^{-1}, preferably 6-12 h^{-1};

the reaction time is 0.5-8 s, preferably 1-5 s; and/or

the catalyst-to-hydrocarbon weight ratio is 6-26, preferably 8-12;

Scheme III-8. The method according to Scheme III-1, characterized in that in step (3), the catalytic conversion conditions in the second reaction zone comprise:

the weight hourly space velocity of oil-gas mixture is 4-20 h^{-1}, preferably 8-12 h^{-1};

the reaction time is 1-10 s, preferably 2-6 s; and/or

the catalyst-to-hydrocarbon weight ratio is 4-14, preferably 6-10;

Scheme III-9. The method according to Scheme III-1, characterized in that in step (4), the catalytic conversion conditions in the third reaction zone comprise:

the weight hourly space velocity of oil-gas mixture is 4-20 h^{-1}, preferably 8-15 h^{-1};

the reaction time is 0.1-5 s, preferably 0.5-3 s; and/or

the catalyst-to-hydrocarbon weight ratio is 6-26, preferably 10-22;

Scheme III-10. The method according to Scheme III-1, characterized in that in step (3), the catalytic naphtha has an initial boiling point of 20-40°C and a final boiling point of 140-160°C; and/or

the heavy aromatic hydrocarbon fraction has an initial boiling point of 140-160°C and a final boiling point of 210-250°C;

Scheme III-11. The method according to Scheme III-1, characterized in that the hydrogenation catalyst comprises an active metal and a carrier, wherein the active metal component is selected from one or more of Group VIB metals and/or Group VIII non-noble metals, and the carrier is selected from any one or more of alumina, silica, and amorphous silica-alumina;

preferably, the active metal component is selected from any one of the following metal combinations: nickel-tungsten, nickel-tungsten-cobalt, nickel-molybdenum, cobalt-molybdenum;

Scheme III-12. The method according to Scheme III-1, characterized in that the catalytic cracking catalyst comprises 10-60 parts by weight of a molecular sieve, 1-40 parts by weight of a binder and 1-90 parts by weight of a carrier;

the molecular sieve is selected from one or more of ZSM molecular sieve, Y molecular sieve, HY molecular sieve, USY molecular sieve, and Beta molecular sieve, and optionally, the molecular sieve contains a rare earth, and the rare earth is one or more of La, Ce, Pr and Nd;

the binder is selected from a silica binder and/or an alumina binder;

the carrier is selected from one or more of silica, kaolin, montmorillonite, diatomaceous earth, halloysite, saponite, rectorite, sepiolite, attapulgite, hydrotalcite and bentonite;

Scheme III-13. A system for producing light aromatic hydrocarbons, comprising:

a hydrogenation unit, wherein the hydrogenation unit is provided with a hydrogen inlet, a catalytic cracking cycle oil inlet and a hydrogenated oil-gas mixture outlet;

a divided-wall distillation column, wherein at least one dividing wall is arranged in the divided-wall distillation column, the dividing wall divides the divided-wall distillation column into four regions: a public stripping section arranged at the bottom of the column, a public rectifying section arranged at the top of the column, a pre-separation zone and a side-draw rectifying zone arranged in the middle section of the column; the pre-separation zone is provided with a hydrogenated oil-gas mixture inlet, the hydrogenated oil-gas mixture inlet is in communication with the hydrogenated oil-gas mixture outlet of the hydrogenation unit; the public rectifying section is provided with a low boiling point fraction outlet, the side-draw rectifying zone is provided with an outlet for the hydrogenated heavy naphtha rich in light aromatic hydrocarbons and an outlet for the hydrogenated light diesel, the bottom of the public stripping section is provided with a hydrogenated heavy diesel outlet;

a catalytic cracking unit, comprising:

a catalytic cracking reactor, wherein the catalytic cracking reactor is provided with a first reaction zone and a second reaction zone, the first reaction zone is provided with a hydrogenated light diesel inlet, and the second reaction zone is provided with a hydrogenated heavy diesel inlet, and the hydrogenated heavy diesel inlet is in communication with the hydrogenated heavy diesel outlet at the bottom of the public stripping section of the divided-wall distillation column;

a product separation device, wherein the product separation device is provided with a catalytic cracking reaction product inlet, a dry gas outlet, a liquefied gas outlet, a catalytic naphtha outlet, a catalytic heavy aromatic hydrocarbon outlet, a catalytic cracking cycle oil outlet and a slurry oil outlet, and the catalytic cracking cycle oil outlet is in communication with the catalytic cracking cycle oil inlet of the hydrogenation unit;

optionally, the catalytic cracking reactor is further provided with a third reaction zone, the third reaction zone is provided with a catalytic heavy aromatic hydrocarbon inlet, and the third reaction zone is located upstream of the first reaction zone and/or the second reaction zone, and the catalytic heavy aromatic hydrocarbon inlet is in communication with the catalytic heavy aromatic hydrocarbon outlet of the product separation unit;

Scheme III-14. The system according to Scheme III-13, characterized in that the third reaction zone, the first reaction zone and the second reaction zone of the catalytic cracking reactor are arranged on the same or different reactors;

preferably, the (riser) reactor is selected from one or more of the following: an equal-diameter riser, an equal-linear-velocity riser, a variable-diameter riser, a variable-linear-velocity riser, a fluidized bed, and a composite reactor, wherein the composite reactor is composed of an equal-diameter or variable-diameter riser and a fluidized bed;

more preferably, the third reaction zone, the first reaction zone and the second reaction zone of the catalytic cracking reactor are arranged on a dual-riser reactor, wherein the main riser is the first reaction zone, and the auxiliary riser is provided with the third reaction zone and the second reaction zone, and the third reaction zone is located upstream of the second reaction zone.

Examples

The present invention is further described in detail below by means of examples. Unless otherwise specified, the raw materials used in the examples were all commercially available.

The hydrogenation catalysts used in the examples and comparative examples were purchased under the trade names FZC and FC-70, which were a hydrofining catalyst and a hydrocracking catalyst, respectively.

The catalytic cracking catalyst used in the examples and comparative examples was purchased under the trade name SLA-1.

The manufacturers of the catalysts were all Sinopec Catalyst Co., Ltd.

The properties of the catalytic cracking cycle oil used in the examples and comparative examples are shown in Table 1.

Table 1

Feedstock oil	Catalytic cracking cycle oil
Density (20°C), kg/m^3	945.3
Hydrocarbon composition, wt%	100
Paraffins	13.1
Naphthenes	5.4
Aromatics	81.5
Monocyclic aromatics	26.4
Bicyclic aromatics	47.7
Other aromatics	7.4
Distillation range, °C	240-367

Example 1

Referring to Figure 1, the hydrogenation unit 100 had upper and lower two-stage series-connected fixed-bed reactors, in which FZC and FC-70 hydrogenation catalysts were respectively placed. The catalytic cracking cycle oil and hydrogen were introduced respectively through the inlet 101 and the

hydrogen inlet into the hydrogenation unit 100, and sequentially passed through the upper and lower stages to contact with the hydrogenation catalysts for a two-stage hydrogenation reaction.

One dividing wall E was installed in the divided-wall distillation column 200 to radially divide the column. The numbers of trays in the pre-separation zone A, the public rectifying section B, the public stripping section D and the side-draw rectifying zone C were 34, 6, 4 and 36, respectively. The overhead reflux ratio was 6. The feed position was at the 17th tray from the top of the pre-separation zone. The light aromatic hydrocarbon-rich fraction B was drawn at the 15th tray from the top of the side-draw rectifying zone C, and the first recycle fraction was drawn at the 32nd tray from the top of the side-draw rectifying zone C. The overhead pressure of the divided-wall distillation column 200 was 0.4 MPa, and the overhead temperature was 67°C. The side-draw rectifying zone of the divided-wall distillation column was provided with a middle-stage circulation heat removal, with side draw and return positions at the 31st and 30th trays of the side-draw rectifying zone. The pre-separation zone of the divided-wall distillation column was provided with a vapor draw, with a draw position at the 25th tray of the pre-separation zone and a return position at the 30th tray of the side-draw rectifying zone. The mass flow ratio of the liquid phases from the public rectifying section into the pre-separation zone and the side-draw rectifying zone was 2.64; the mass flow ratio of the vapor phases from the public stripping section into the pre-separation zone and the side-draw rectifying zone was 4.22; and the ratio of the total vapor mass flow to the drawn vapor mass flow at the draw position of the pre-separation zone was 69.36. The catalytic cracking unit 300 had a dual-riser reactor, and the catalyst used in two risers was the catalytic cracking catalyst SLA-1, wherein the main riser was the first reaction zone, and the auxiliary riser was the second reaction zone.

The catalytic cracking cycle oil having a distillation range of 240-367°C and hydrogen were respectively introduced through the catalytic cracking cycle oil inlet 101 and the hydrogen inlet, and sequentially passed through the hydrogenation unit 100 to contact with the hydrogenation catalyst for reaction. The hydrogen partial pressure was 8.0 MPa. The reaction temperatures of the upper and lower two stages were 360°C and 380°C, respectively. The volumetric space velocities were 2.0 h-1 and 1.0 h-1, respectively. The hydrogen-to-hydrocarbon volume ratios were 800 and 1300, respectively. The hydrogenated product was introduced via pipeline 102 into a high/low pressure separation unit. After a recycle hydrogen and a low-pressure gas were separated, the resulting hydrogenated oil-gas mixture was introduced via pipeline 103 into a degassing separation unit. The separated fraction oil was introduced via pipeline 201 into the divided-wall distillation column 200 for separation. Four fractions with increasing boiling points were separated. The overhead fraction was an overhead light fraction of less than 60°C. The bottom fraction was a second recycle fraction of greater than 240°C, which was introduced via pipeline 205 into the second reaction zone located

in the second riser of the catalytic cracking unit 300 for a catalytic cracking reaction under conditions of reaction time of 3 s, catalyst-to-hydrocarbon mass ratio of 10, and weight hourly space velocity of 10 h-1. A light aromatic hydrocarbon-rich fraction of 57-150°C was drawn from the upper part of the side-draw rectifying zone C (at the 15th tray from top to bottom), and a first recycle fraction of 150-240°C was drawn from the lower part (at the 32nd tray from top to bottom). The first recycle fraction was introduced via the first recycle fraction pipeline 204 into the first reaction zone located in the first riser of the catalytic cracking unit 300 for a catalytic cracking reaction under conditions of reaction time of 1.2 s, catalyst-to-hydrocarbon mass ratio of 10, and weight hourly space velocity of 10 h-1. The post-reaction product was introduced via the product pipeline 305 into the second distillation column 400 for separation. A first fraction oil-gas mixture of less than 240°C was separated and left via pipeline 401, and combined with the hydrogenated oil-gas mixture from pipeline 103 into the degassing separation unit. A second fraction oil having a distillation range of 240-350°C was returned via pipeline 402 to the inlet 101 of the hydrogenation unit for recycling. A slurry oil of greater than 350°C was discharged via pipeline 403. The product distribution and the distillation ranges of each fraction are shown in Tables 2 and 3.

Comparative Example 1

The procedure was the same as in Example 1, except that the catalytic cracking reaction product was separated in the second distillation column 400, and the separated first fraction oil-gas mixture had a final boiling point of 210°C, and the second fraction oil had a distillation range of 210-350°C. The product distribution and the distillation ranges of each fraction are shown in Tables 2 and 3.

Comparative Example 2

The procedure was the same as in Example 1, except that the catalytic cracking reaction product was separated in the second distillation column 400, and the separated first fraction oil-gas mixture had a final boiling point of 260°C, and the second fraction oil had a distillation range of 260-350°C. The product distribution and the distillation ranges of each fraction are shown in Tables 2 and 3.

Comparative Example 3

The procedure was the same as in Example 1, except that both the first recycle fraction and the second recycle fraction separated by the divided-wall distillation column 200 were introduced into the first reaction zone of the catalytic cracking unit 300 for a catalytic cracking reaction under conditions of reaction time of 3 s, catalyst-to-hydrocarbon mass ratio of 10, and weight hourly space velocity of 10 h-1. The product distribution and the distillation ranges of each fraction are shown in Tables 2 and 3.

Comparative Example 4

The procedure was the same as in Example 1, except that the operating conditions of the divided-wall distillation column were adjusted such that the first recycle fraction was drawn at the

36th tray from the top of the side-draw rectifying zone, and the mass ratios of the gas and liquid phases into the side-draw rectifying zone were both reduced. The light aromatic hydrocarbon-rich fraction had a distillation range of 57-180°C, the first recycle fraction had a distillation range of 180-310°C, and the second recycle fraction had a distillation range of 310-350°C. The product distribution and the distillation ranges of each fraction are shown in Tables 2 and 3.

Comparative Example 5

The procedure was the same as in Example 1, except that the column 200 was replaced by two conventional distillation columns to achieve the function of the divided-wall distillation column. A distillation column A with 40 trays and a reflux ratio of 0.8 was used, and a gas and an overhead light fraction were separated from the top of the column, a light aromatic hydrocarbon fraction was separated from the middle of the column, and a heavy fraction was separated from the bottom of the column and introduced into a distillation column B with 20 trays and a reflux ratio of 1.8, with which a first recycle fraction was separated from the top of the column, and a second recycle fraction was separated from the bottom of the column. The product distribution and the distillation ranges of each fraction are shown in Tables 2 and 3.

Comparative Example 6

After the catalytic cracking cycle oil was subjected to two-stage hydrotreating, the hydrogenated product was subjected to high-low pressure separation. After separation of recycle hydrogen, the hydrogenated oil-gas mixture was introduced via a hydrogenated oil-gas mixture pipeline into a first distillation column with 40 trays and a reflux ratio of 0.8, with which a gas and an overhead light fraction was separated from the top of the column, a light aromatic hydrocarbon-rich fraction was separated from the middle of the column, and a hydrogenated diesel fraction was separated from the bottom of the column. The hydrogenated diesel fraction was introduced into a catalytic cracking unit for a catalytic cracking reaction under conditions of reaction time of 3 s, catalyst-to-hydrocarbon mass ratio of 10, and weight hourly space velocity of 10 h-1. The post-reaction product was introduced into a second distillation column for separation. A first fraction oil-gas mixture having a distillation range of <210°C was obtained from the top of the column, a second fraction oil having a distillation range of 210-350°C was obtained from the middle of the column, and a catalytic slurry oil having a distillation range of >350°C was obtained from the bottom of the column. The second fraction oil having a distillation range of 210-350°C was returned to the hydrogenation unit for recycling. The first fraction oil-gas mixture was introduced into a third distillation column with 40 trays and a reflux ratio of 2 for separation of light aromatic hydrocarbons. A light fraction was obtained from the top of the column, a light aromatic hydrocarbon fraction was obtained from the middle of the column, and a heavy fraction was obtained from the bottom of the column. The product distribution and the distillation ranges of each

fraction are shown in Tables 2 and 3.

Example 2

Referring to Figure 2, the hydrogenation unit 100 had upper and lower two-stage series-connected fixed-bed reactors, in which two hydrogenation catalysts FZC and FC-70 were respectively provided in said upper and lower two stages. The catalytic cracking cycle oil was introduced through the catalytic cracking cycle oil inlet 101 and into the hydrogenation unit 100, and sequentially passed through the first and second stages.

One dividing wall E was installed in the divided-wall distillation column 200 to radially divide the column. The numbers of trays in the pre-separation zone A, the public rectifying section B, the public stripping section D and the side-draw rectifying zone C were 34, 6, 4 and 36, respectively. The overhead reflux ratio was 6. The feed position was at the 17th tray from the top of the pre-separation zone. The light aromatic hydrocarbon-rich fraction B was drawn at the 15th tray from the top of the side-draw rectifying zone C, and the first recycle fraction was drawn at the 34th tray from the top of the side-draw rectifying zone C. The overhead pressure of the divided-wall distillation column 200 was 0.4 MPa, and the overhead temperature was 67°C. The side-draw rectifying zone of the divided-wall distillation column was provided with a middle-stage circulation heat removal, with side draw and return positions at the 32nd and 33rd trays of the side-draw rectifying zone. The pre-separation zone of the divided-wall distillation column was provided with a vapor draw, with a draw position at the 25th tray of the pre-separation zone and a return position at the 33rd tray of the side-draw rectifying zone. The mass flow ratio of the liquid phases from the public rectifying section into the pre-separation zone and the side-draw rectifying zone was 2.64; the mass flow ratio of the vapor phases from the public stripping section into the pre-separation zone and the side-draw rectifying zone was 4.22; and the ratio of the total vapor mass flow to the drawn vapor mass flow at the draw position of the pre-separation zone was 69.36. The catalytic cracking unit 300 had a single riser reactor.

The catalytic cracking cycle oil having a distillation range of 240-367°C and hydrogen were respectively introduced through the catalytic cracking cycle oil inlet 101 and the hydrogen inlet, and passed through the upper and lower two stages of the hydrogenation unit 100 to contact with two hydrogenation catalysts for reaction. The hydrogen partial pressure was 8.0 MPa. The reaction temperatures of the upper and lower two stages were 360°C and 380°C, respectively. The volumetric space velocities were 2.0 h-1 and 1.0 h-1, respectively. The hydrogen-to-hydrocarbon volume ratios were 800 and 1300, respectively. The hydrogenated product was introduced via pipeline 102 into a high/low pressure separation unit. After a recycle hydrogen and a low-pressure gas were separated, the resulting hydrogenated oil-gas mixture was introduced via pipeline 103 into a degassing separation unit. The separated fraction oil was introduced via pipeline 201 into the

divided-wall distillation column 200 for separation. Four fractions with increasing boiling points were separated. An overhead light fraction of less than 60°C was drawn from the top of the column. A second recycle fraction of greater than 280°C was obtained from the bottom of the column, and returned via pipeline 205 to the inlet of the second stage of the hydrogenation unit 100. A light aromatic hydrocarbon-rich fraction of 57-150°C was drawn from the upper part of the side-draw rectifying zone C (at the 15th tray from top to bottom), and a first recycle fraction of 150-280°C was drawn from the lower part (at the 34th tray from top to bottom). The first recycle fraction was introduced via the first recycle fraction pipeline 204 into the first reaction zone located in the first riser of the catalytic cracking unit 300 for a catalytic cracking reaction under conditions of reaction time of 2 s, catalyst-to-hydrocarbon mass ratio of 10, and weight hourly space velocity of 10 h-1. The post-reaction product was introduced via the product pipeline 305 into the second distillation column 400 for separation. A first fraction oil-gas mixture of less than 240°C was separated and left via pipeline 401, and combined with the hydrogenated oil-gas mixture from pipeline 103 into the degassing separation unit. A second fraction oil having a distillation range of 240-350°C was returned via pipeline 402 to the inlet 101 of the hydrogenation unit for recycling. A slurry oil of greater than 350°C was discharged via pipeline 403. The product distribution is shown in Table 2.

Comparative Example 7

The procedure was the same as in Example 2, except that the operating conditions of the divided-wall distillation column were adjusted such that the first recycle fraction was drawn at the 36th tray from the top of the side-draw rectifying zone, and the mass ratios of the gas and liquid phases into the side-draw rectifying zone were both reduced. The light aromatic hydrocarbon-rich fraction had a distillation range of 57-180°C, the first recycle fraction had a distillation range of 180-310°C, and the second recycle fraction had a distillation range of 310-350°C. The product distribution and the distillation ranges of each fraction are shown in Tables 2 and 3.

Table 2

	Ex. 1	Comp. Ex. 1	Comp. Ex. 2	Comp. Ex. 3	Comp. Ex. 4	Comp. Ex. 5	Ex. 2	Comp. Ex. 6	Comp. Ex. 7
Product distribution, wt%									
C6-C8 aromatics yield, wt%	39.54	37.34	36.54	35.74	36.04	39.04	38.57	24.59	35.20
Benzene	3.72	3.53	3.44	3.36	3.18	3.62	3.61	2.31	3.24
Toluene	15.60	14.71	14.45	14.10	14.40	15.45	15.11	9.70	13.93
C8 aromatics	20.22	19.10	18.65	18.28	18.46	19.97	19.85	12.58	18.03

The separation process accuracy and energy consumption of the examples and comparative examples were calculated, with Example 1 as the benchmark, and the comparison is shown in Table 3.

Table 3

	Ex. 1	Comp. Ex. 1	Comp. Ex. 2	Comp. Exs. 4&7	Comp. Ex. 5	Ex. 2	Comp. Ex. 6
Distillation range, °C							
First fraction oil-gas mixture	20-240	20-210	20-260	20-240	20-240	20-240	20-210
Second fraction oil	240-350	210-350	260-350	240-350	240-350	240-350	210-250
Light aromatic hydrocarbon fraction	57-150	57-150	57-150	57-180	57-155	57-150	60-170
First recycle fraction	150-240	150-240	150-240	180-310	150-250	150-280	-
Second recycle fraction	220-350	220-350	220-350	310-350	230-350	280-350	-
Hydrogenated diesel							153-340
Relative energy consumption	1.0	1.0	1.0	1.0	1.3	1.0	1.3

Tray numbers and reflux ratios of the distillation columns/divided-wall distillation columns in the examples and comparative examples are shown in Table 4 below.

Table 4

	Distillation columns		Tray number, trays	Reflux ratio
Exs. 1&2	Divided-wall distillation column	Pre-separation zone	34	6
		Public rectifying section	6	
		Public stripping section	4	
		Side-draw rectifying zone	36	
Comp. Exs. 4&7	Divided-wall distillation column		Same as Ex. 1	5
Comp. Ex. 5	Distillation column A		40	0.8
	Distillation column B		20	1.8
Comp. Ex. 6	First distillation column		40	0.8
	Third distillation column		40	2

Compared with Comparative Examples 1, 2, 3 and 4, the separation energy consumption of Example 1 was comparable, and the light aromatic hydrocarbon yields were higher by 2.2, 3.0, 3.8 and 3.5 percentage points, respectively.

Compared with Comparative Example 5, under the same separation feedstock and separation products, the total energy consumption of the separation in Example 1 using the divided-wall distillation column was 30% lower.

Compared with Comparative Example 7, the separation energy consumption of Example 2 was comparable, and the yield of light aromatic hydrocarbons was improved by 3.37 percentage points.

Compared with Comparative Example 6, the separation energy consumption of Examples 1 and 2 was 30% lower and the yields for light aromatic hydrocarbons were higher by 14.95 and 13.98 percentage points, respectively.

The present application has been described above in conjunction with preferred embodiments, but these embodiments are merely exemplary and are for illustrative purposes only. On this basis, various substitutions and improvements can be made to the present application, all of which fall within the scope of protection of the present application.

Claims

1. A method for producing light aromatic hydrocarbons, characterized in that the method comprises the following steps:

(1) subjecting a catalytic cracking cycle oil to two-stage hydrotreating in the presence of hydrogen and a hydrogenation catalyst to obtain a hydrogenated product;

(2) removing hydrogen from the hydrogenated product to obtain a hydrogenated oil-gas mixture, and passing the hydrogenated oil-gas mixture through a first distillation column to obtain an overhead light fraction, a light aromatic hydrocarbon-rich fraction, a first recycle fraction and a second recycle fraction, wherein the first recycle fraction has an initial boiling point of 140-170°C and a final boiling point of 220-300°C, and the second recycle fraction has a distillation range starting point temperature of any value between 220-300°C;

(3) introducing the first recycle fraction obtained in step (2) into a first reaction zone of a catalytic cracking unit for a catalytic cracking reaction, and optionally, introducing the second recycle fraction into a second reaction zone of the catalytic cracking unit or returning it to a second stage of said hydrotreating for recycling;

(4) separating the reaction product obtained from the catalytic cracking unit by a second distillation column to obtain a first fraction oil-gas mixture, a second fraction oil and a slurry oil, returning the second fraction oil to a first stage of said hydrotreating, and optionally, introducing the first fraction oil-gas mixture together with the hydrogenated oil-gas mixture into the first distillation column of step (2) for separation;

(5) separating light aromatic hydrocarbons from the light aromatic hydrocarbon-rich fraction.

2. The method according to claim 1, characterized in that in step (1), the hydrotreating reaction conditions satisfy at least one of the following requirements:

the reaction conditions for the first stage of said hydrotreating are: hydrogen partial pressure 5.0-20.0 MPa, reaction temperature 300-450°C, volumetric space velocity 0.5-5.0 h^{-1}, hydrogen-to-hydrocarbon volume ratio 300-1600 Nm^3/m^3; and/or

the reaction conditions for the second stage of said hydrotreating are: hydrogen partial pressure 5.0-20.0 MPa, reaction temperature 350-500°C, preferably 10-50°C higher than the temperature of the first stage of said hydrotreating, volumetric space velocity 0.1-3.0 h^{-1}, hydrogen-to-hydrocarbon volume ratio 400-2000 Nm^3/m^3.

3. The method according to claim 1 or 2, characterized in that in step (1), in the two-stage hydrotreating reaction, the hydrogenation catalysts in the first and second stages are the same or different;

Preferably, the hydrogenation catalyst for the first stage of said hydrotreating comprises an active

metal component and a carrier, and the hydrogenation catalyst for the second stage of hydrotreating comprises an active metal, a molecular sieve and a carrier, wherein the active metal component is selected from one or more of Group VIB metals and/or Group VIII non-noble metals, the carrier is selected from one or more of alumina, silica, and amorphous silica-alumina, and the molecular sieve is selected from one or more of Y-type molecular sieve and USY molecular sieve;

more preferably, the active metal component is selected from any one of the following metal combinations: nickel-tungsten, nickel-tungsten-cobalt, nickel-molybdenum, cobalt-molybdenum.

4. The method according to any one of claims 1-3, characterized in that at least one of the following requirements is satisfied:

the catalytic cracking cycle oil has an initial boiling point of 160-250°C and a final boiling point of 340-400°C; and/or

the hydrogenated oil-gas mixture has an initial boiling point of 0-60°C and a final boiling point of 280-360°C.

5. The method according to claim 1, characterized in that in step (2), the first distillation column is a divided-wall distillation column, at least one dividing wall is arranged in the divided-wall distillation column, and the dividing wall divides the divided-wall distillation column into four regions: a pre-separation zone, a public rectifying section, a public stripping section and a side-draw rectifying zone, and the feed position of the hydrogenated oil-gas mixture is located in the pre-separation zone.

6. The method according to claim 5, characterized in that in step (2),

the pre-separation zone of the divided-wall distillation column has 10-40 trays, preferably 20-35 trays;

The public rectifying section has 2-10 trays, preferably 3-8 trays;

the public stripping section has 2-10 trays, preferably 3-8 trays;

the side-draw rectifying zone has 10-45 trays, preferably 25-40 trays;

and the divided-wall distillation column satisfies at least one of the following requirements:

the feed position of the hydrogenated oil-gas mixture is above the 4th to 22nd tray from the top of the pre-separation zone, preferably above the 8th to 20th tray;

the overhead pressure of the divided-wall distillation column is 0.1-1.0 MPa, preferably 0.2-0.7 MPa, and the overhead temperature is 50-120°C, preferably 60-100°C; and/or

the overhead reflux ratio is 3-10, preferably 4-8.

7. The method according to claim 6, characterized in that in step (2), the divided-wall distillation column satisfies at least one of the following requirements:

the side-draw rectifying zone of the divided-wall distillation column is provided with a middle-stage circulation heat removal, with a draw position at the 20th to 45th tray of the side-draw

rectifying zone, more preferably at the 30th to 40th tray, and a return position at the same tray as the draw position or one tray above or below the draw position;

the pre-separation zone of the divided-wall distillation column is provided with a vapor draw, with a draw position at the 20th to 40th tray of the pre-separation zone, more preferably at the 25th to 35th tray, and a vapor return position at the 30th to 45th tray of the side-draw rectifying zone, more preferably at the 35th to 40th tray;

the mass flow ratio of the liquid phases from the public rectifying section into the pre-separation zone and the side-draw rectifying zone is 0.1-10, preferably 0.5-5;

the mass flow ratio of the vapor phases from the public stripping section into the pre-separation zone and the side-draw rectifying zone is 1-15, preferably 2-10; and/or

the ratio of the total vapor mass flow to the drawn vapor mass flow at the draw position of the pre-separation zone is 10-150, preferably 30-120.

8. The method according to claim 6, characterized in that in step (2), the divided-wall distillation column satisfies at least one of the following requirements:

the overhead light fraction is separated from the public rectifying section of the divided-wall distillation column, and has an initial boiling point of 20-40°C and a final boiling point of 55-65°C;

the light aromatic hydrocarbon-rich fraction is drawn from the 5th to 30th tray of the side-draw rectifying zone of the divided-wall distillation column, preferably from the 10th to 20th tray, and has an initial boiling point of 55-65°C and a final boiling point of 140-170°C;

the first recycle fraction is drawn from any position below the draw position of the light aromatic hydrocarbon-rich fraction in the side-draw rectifying zone, preferably from the 25th to 35th tray; and/or

the second recycle fraction is separated from the lower part of the public stripping section of the divided-wall distillation column.

9. The method according to any one of claims 1-8, characterized in that

in step (2), the first recycle fraction has an initial boiling point of 140-170°C and a final boiling point of 220-250°C, and the second recycle fraction has a distillation range starting point temperature of any value between 220-250°C; or

in step (2), the first recycle fraction has an initial boiling point of 140-170°C and a final boiling point of 250-300°C, and the second recycle fraction has a distillation range starting point temperature of any value between 250-300°C.

10. The method according to any one of claims 1-8, characterized in that

when the distillation range starting point temperature of the second recycle fraction is any value between 220-250°C, in step (3), the second recycle fraction is introduced into the second reaction zone of the catalytic cracking unit;

when the distillation range starting point temperature of the second recycle fraction is any value between 270-300°C, in step (3), the second recycle fraction is returned to the second stage of the hydrotreating for recycling;

When the distillation range starting point temperature of the second recycle fraction is any value between 250-270°C, in step (3), the second recycle fraction is introduced into the second reaction zone of the catalytic cracking unit or returned to the second stage of the hydrotreating for recycling.

11. The method according to any one of claims 1-10, characterized in that in step (3), the catalytic conversion conditions in the first reaction zone satisfy at least one of the following requirements:

the weight hourly space velocity of oil-gas mixture is 4-20 h^{-1}, preferably 6-12 h^{-1};

the reaction time is 0.5-8 s, preferably 1-5 s; and/or

the catalyst-to-hydrocarbon weight ratio is 6-26, preferably 8-12;

and the catalytic conversion conditions in the second reaction zone satisfy at least one of the following requirements:

the weight hourly space velocity of oil-gas mixture is 4-20 h^{-1}, preferably 8-12 h^{-1};

the reaction time is 1-10 s, preferably 2-6 s; and/or

the catalyst-to-hydrocarbon weight ratio is 4-14, preferably 6-10.

12. The method according to any one of claims 1-11, characterized in that in step (4), the second distillation column satisfies at least one of the following requirements:

the distillation column has an overhead pressure of 0.15-0.4 MPa, preferably 0.18-0.35 MPa, and an overhead temperature of 100-160°C, preferably 115-145°C,

the first fraction has an initial boiling point of 0-40°C and a final boiling point of 220-250°C, and the first fraction, after being combined with the hydrogenated fraction and degassed, is introduced into the first distillation column of step (2) for separation;

the second fraction oil has an initial boiling point of 220-250°C and a final boiling point of 340-360°C; and/or

the slurry oil has an initial boiling point of 340-360°C.

13. A system for producing light aromatic hydrocarbons, comprising:

A hydrogenation unit, wherein the hydrogenation unit is a two-stage hydrogenation unit and provided with a hydrogen inlet and a hydrogenated product outlet, a first stage of the hydrogenation unit is provided with a catalytic cracking cycle oil inlet, and a second stage of the hydrogenation unit is optionally provided with a second recycle fraction inlet;

a catalytic cracking unit, comprising a catalytic cracking reactor, wherein the catalytic cracking reactor is provided with a first reaction zone and an optional second reaction zone, the first reaction zone is provided with a first recycle fraction inlet, and the optional second reaction zone is provided with a second recycle fraction inlet;

a second distillation column, wherein the second distillation column is provided with a catalytic cracking reaction product inlet, a first fraction oil-gas mixture outlet, a second fraction oil outlet and a slurry oil outlet, and the second fraction oil outlet is in communication with the catalytic cracking cycle oil inlet of the hydrogenation unit;

a degassing separation unit, wherein the degassing separation unit is provided with an oil-gas mixture inlet, a rich gas outlet and a fraction oil outlet, and the oil-gas mixture inlet of the degassing separation unit is connected to the first fraction oil-gas mixture outlet of the second distillation column and the hydrogenated product outlet of the hydrogenation unit, so that the first fraction oil-gas mixture from the second distillation column and the hydrogenated oil-gas mixture from the hydrogenation unit are degassed and separated into a rich gas and a fraction oil;

a first distillation column, wherein the distillation column is provided with a fraction oil inlet, an overhead light fraction outlet, a light aromatic hydrocarbon-rich fraction outlet, a first recycle fraction outlet, and a second recycle fraction outlet, the fraction oil inlet is in communication with the fraction oil outlet of the degassing separation unit, the first recycle fraction outlet is in communication with the first recycle fraction inlet of the catalytic cracking unit, the second recycle fraction outlet is in communication with the second recycle fraction inlet provided for the optional second reaction zone of the catalytic cracking unit, or the second recycle fraction outlet is in communication with the second recycle fraction inlet optionally provided for the second stage of the hydrogenation unit;

a separation unit, which is provided with a light aromatic hydrocarbon-rich fraction inlet and a light aromatic hydrocarbon fraction outlet, and the light aromatic hydrocarbon-rich fraction inlet is connected to the light aromatic hydrocarbon-rich fraction outlet of the first distillation column.

14. The system according to claim 13, characterized in that the first distillation column is a divided-wall distillation column, at least one dividing wall is arranged in the divided-wall distillation column, and the dividing wall divides the divided-wall distillation column into four regions: a public stripping section arranged at the bottom of the column, a public rectifying section arranged at the top of the column, a pre-separation zone and a side-draw rectifying zone arranged in the middle section of the column; the pre-separation zone is provided with a fraction oil inlet, and the fraction oil inlet is in communication with the fraction oil outlet of the degassing separation unit; the public rectifying section is provided with an overhead light fraction outlet, the side-draw rectifying zone is provided with a light aromatic hydrocarbon-rich fraction outlet and a first recycle fraction outlet, and the bottom of the public stripping section is provided with a second recycle fraction outlet,

here, the first recycle fraction outlet of the side-draw rectifying zone of the divided-wall distillation column is in communication with the first recycle fraction inlet of the catalytic cracking reactor,

the second recycle fraction outlet at the bottom of the public stripping section of the divided-wall distillation column is in communication with the second recycle fraction inlet provided for the optional second reaction zone of the first catalytic cracking reactor, or is in communication with the second recycle fraction inlet optionally provided on the second stage of the hydrogenation unit.

15. The system according to claim 13 or 14, characterized in that the system further comprises a high/low pressure separation unit, the high/low pressure separation unit is provided with a hydrogenated product inlet, a hydrogen outlet and a hydrogenated oil-gas mixture outlet, and the high/low pressure separation unit separates the hydrogenated oil-gas mixture from hydrogen in the hydrogenated product from the hydrogenation unit;

here, the hydrogenated product inlet is connected to the hydrogenated product outlet of the hydrogenation unit, and the oil-gas mixture inlet of the degassing separation unit is connected to the first fraction oil-gas mixture outlet of the second distillation column and the hydrogenated oil-gas mixture outlet of the high/low pressure separation unit.

16. The system according to any one of claims 13-15, characterized in that the first reaction zone and the second reaction zone of the catalytic cracking reactor are arranged on the same or different reactors;

the reactor is selected from one or more of the following: an equal-diameter riser, an equal-linear-velocity riser, a variable-diameter riser, a variable-linear-velocity riser, a fluidized bed, and a composite reactor, wherein the composite reactor is composed of an equal-diameter or variable-diameter riser and a fluidized bed, and preferably the reactor is a riser reactor;

more preferably, the first reaction zone and the second reaction zone of the catalytic cracking reactor are arranged on a dual-riser reactor, wherein the main riser is the first reaction zone, and the auxiliary riser is the second reaction zone.