

A PROCESS FOR PRODUCING AN AVIATION FUEL COMPONENT

TECHNICAL FIELD

5 The present disclosure relates to a process for producing a renewable aviation fuel component. The disclosure relates particularly, though not exclusively, to a process for producing a renewable aviation fuel component, comprising subjecting a renewable feedstock to a decarboxylation and/or decarbonylation (DCO) reaction, hydrotreatment and hydroisomerisation.

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BACKGROUND

This section illustrates useful background information without admission of any technique described herein representative of the state of the art.

15 Presently, there is an ongoing need to reduce greenhouse gas emissions and/or carbon footprint in transportation, especially in aviation. Accordingly, the interest towards renewable aviation fuels and aviation fuel components is and has been growing.

Processes for producing aviation fuel components from renewable raw materials have been proposed. However, the carbon and hydrogen efficiency of aviation fuel components, as well as yield of aviation fuel has been relatively low in said processes. Therefore, there is
20 an ongoing need to improve the yield of renewable aviation fuel components in the processes producing renewable aviation fuel components that could be used in aviation fuels.

Many of the renewable feedstocks of biological origin comprise C18 and heavier fatty acids, which restricts their efficient use in a renewable aviation fuel production, due to the
25 properties of formed n-paraffins and i-paraffins affecting, for example, the boiling point range, cold flow properties, density, and cold viscosity of the final aviation fuel component. Therefore, there is an ongoing need to utilize renewable feedstocks comprising C18 and heavier fatty acids in renewable aviation fuel production without unnecessary yield losses.

30 SUMMARY

The present application concerns the inventions defined in the appended independent claims, and their embodiments disclosed below. The appended claims define the present

invention. Any examples and technical descriptions of apparatuses, products and/or methods in the description and/or drawings not covered by the claims are presented not as embodiments of the invention but as background art or examples useful for understanding the invention.

- 5 In view of the above, an object of the invention is to provide a process for producing a renewable aviation fuel component from a renewable feedstock. An aim is to increase carbon and hydrogen efficiency of the renewable aviation fuel component, and to enable increasing the yield of the aviation fuel component in the process for producing said aviation fuel component. Another aim is to improve utilisation of the carboxylic acids of the
- 10 renewable feedstock with a carbon number C18 or longer, in production of an aviation fuel component. Another aim is to minimize hydrogen consumption in the process for producing a renewable aviation fuel component. Another aim is to minimize production of C1-C7 hydrocarbons which are not suitable for an aviation fuel component. However, in practice this means the content of C1-C4 hydrocarbons is reduced the most, due to lower and/or
- 15 upper end tailing of the distillation cut points. A further aim of the invention is to provide a process for producing other hydrocarbon fractions from renewable sources.

According to a first example aspect there is provided a process for producing a renewable aviation fuel component, the process comprising:

- i) providing a renewable feedstock comprising free carboxylic acids (FCAs), esters of
- 20 carboxylic acids, triglycerides, or combinations thereof;
- ii) subjecting the feedstock to a decarboxylation and/or decarbonylation (DCO) reaction in a DCO zone, for removal of one carbon from a carbon chain of a carboxylic acid moiety of the renewable feedstock, in the presence of a DCO catalyst, wherein the DCO zone:
- the DCO deoxygenation selectivity is at least 75 wt-% of the total weight of
- 25 deoxygenated hydrocarbons, and
- the deoxygenation conversion is at least 50 wt-% of the total weight of the feedstock, thereby obtaining a DCO effluent;
- iii) subjecting at least a portion of the DCO effluent from step ii) to a hydrotreatment (HT) reaction in the presence of hydrogen and a hydrotreatment catalyst in a HT zone, to obtain
- 30 a hydrotreated effluent;
- iv) subjecting the hydrotreated effluent from step iii) to a gas-liquid separation to obtain a degassed hydrotreated effluent;
- v) subjecting at least a portion of the degassed hydrotreated effluent from step iv) to hydroisomerisation (H-ISO), thereby obtaining a hydroisomerised effluent; and

vi) subjecting at least a portion of the hydroisomerised effluent from step v) to a fractionation and recovering at least the renewable aviation fuel component.

According to a second example aspect there is provided a use of the renewable aviation fuel component, obtainable from the process of the first aspect, as a renewable aviation fuel
5 blend component in an aviation fuel blend, wherein the aviation fuel blend also comprises a fossil aviation fuel blend component.

Different non-binding example aspects and embodiments have been illustrated in the foregoing. The embodiments in the foregoing are used merely to explain selected aspects or steps that may be utilized in different implementations. Some embodiments may be
10 presented only with reference to certain example aspects. It should be appreciated that corresponding embodiments may apply to other example aspects as well.

BRIEF DESCRIPTION OF THE FIGURES

Some example embodiments will be described with reference to the accompanying figures,
15 in which:

Fig. 1 illustrates schematically the present process for producing a renewable aviation fuel component, according to a first example embodiment;

Fig. 2 illustrates schematically the present process for producing the renewable aviation fuel component, according to a second example embodiment;

20 Fig. 3 illustrates schematically the present process for producing the renewable aviation fuel component, according to a third example embodiment;

Fig. 4 illustrates schematically the present process for producing the renewable aviation fuel component, according to a fourth example embodiment;

25 Fig. 5 illustrates schematically the present process for producing the renewable aviation fuel component, according to a fifth example embodiment; and

Fig. 6 illustrates schematically the present process for producing the renewable aviation fuel component, according to a sixth example embodiment.

DETAILED DESCRIPTION

30 DEFINITIONS

The invention relates to a process for producing a renewable aviation fuel component. The disclosure relates to a process comprising subjecting a renewable feedstock to a decarboxylation and/or decarbonylation (DCO) reactions for removal of one carbon from carbon chains of carboxylic acids of the renewable feedstock. As used herein, the term

5 "decarboxylation/decarbonylation" means the removal of one carbon atom together with covalently bound carboxyl oxygen through CO₂ (decarboxylation) or through CO (decarbonylation) from a carboxyl group. Decarboxylation/decarbonylation can take place completely without the influence of molecular hydrogen. Decarboxylation and decarbonylation reactions either together or alone are referred to as decarb-reactions

10 (DCO). A liquid product, comprising mainly odd or uneven carbon number n-paraffins, is recovered from the DCO reaction.

In the following description, like reference signs denote like elements or steps.

All standards referred to herein are the latest revisions available at the filing date, unless otherwise mentioned.

15 As used in the context of this disclosure, the properties of the renewable aviation fuel component comply with the specifications set out in the ASTM D7566-22 Annex A2. In an embodiment, the renewable aviation fuel component obtained from the process comprises C8-C17 alkanes.

Unless otherwise stated, regarding distillation characteristics and boiling ranges, reference

20 is made to EN ISO 3405:2019. For boiling point distribution, reference may also be made to gas chromatography -based methods like ASTM D2887-19e1. Fatty acid distribution for the feed may be determined according to ISO 12966-4:2015, or measured using known analysis methods based on e.g. GC-FID or GC-AED.

As used herein, with the term "feed" or "feedstock" is meant any feedstock which is fed into

25 a particular reaction. As used herein, the term "renewable" refers to compounds or compositions that are obtainable, derivable, or originating from plants and/or animals, including compounds or compositions obtainable, derivable, or originating from fungi and/or algae, in full or in part. As used herein, renewable compounds or compositions may comprise gene manipulated compounds or compositions. Renewable feeds, components,

30 compounds or compositions may also be referred to as biological compounds or compositions, or as biogenic compounds or compositions.

Renewable organic compounds can be differentiated chemically from those of fossil origin, including hydrocarbons, by suitable method for analysing the content of carbon from renewable sources, such as DIN 51637 (2014), ASTM D6866 (2020) and EN 16640 (2017).

Said methods are based on the fact that carbon atoms of renewable or biological origin comprise a higher number of unstable radiocarbon (^{14}C) atoms compared to carbon atoms of fossil origin. Therefore, it is possible to distinguish between carbon compounds derived from renewable or biological sources or raw material and carbon compounds derived from fossil sources or raw material by analysing the ratio of ^{12}C and ^{14}C isotopes. Thus, a particular ratio of said isotopes can be used as a "tag" to identify a renewable carbon compound and differentiate it from non-renewable carbon compounds. The isotope ratio does not change in the course of chemical reactions. Therefore, the isotope ratio can be used for identifying renewable compounds, components, and compositions and distinguishing them from non-renewable, fossil materials in reactor feeds, reactor effluents, separated product fractions and various blends thereof.

Numerically, the biogenic carbon content can be expressed as the amount of biogenic carbon in the material as a weight percent of the total carbon (TC) in the material (in accordance with ASTM D6866 (2020) or EN 16640 (2017)). In the present context, the term renewable preferably refers to a material having a biogenic carbon content of more than 95 wt-%, even more preferably about 100 wt-%, based on the total weight of carbon in the material (EN 16640 (2017)).

As used herein the term "hydrocarbons" refers to compounds comprising carbon and hydrogen, especially to paraffins, n-paraffins, i-paraffins, monobranched i-paraffins, multiple-branched i-paraffins, olefins, naphthenes and aromatics. Oxygenated hydrocarbons specifically refer herein to hydrocarbons comprising covalently bound oxygen.

As used herein paraffins refer to non-cyclic alkanes, i.e., non-cyclic, open chain saturated hydrocarbons that are linear (normal paraffins, n-paraffins) or branched (isoparaffins, i-paraffins). In other words, paraffins refer herein to n-paraffins and/or isoparaffins. In the context of the present disclosure, isoparaffins refer to branched open chain alkanes, i.e., non-cyclic, open chain saturated hydrocarbons having one or more alkyl side chains. Isoparaffins having one alkyl side chain or branch are herein referred to as monobranched isoparaffins and isoparaffins having two or more alkyl side chains or branches are herein referred to as multiple-branched isoparaffins.

In the context of this disclosure, CX+ carboxylic acids, CX+ fatty acids, CX+ hydrocarbons, CX+ paraffins, or CX+ isoparaffins refer to carboxylic acids, fatty acids, hydrocarbons, paraffins, or isoparaffins, respectively, having a carbon number of at least X, where X is any feasible integer.

As used herein, the term "free carboxylic acids (FCAs)" refers to an organic acid that contains a carboxyl group (-COOH) attached to an R alkyl group which has one or more carbons. As used herein, the term "esters of carboxylic acids" refers to a derivative of a carboxylic acid in which the hydrogen atom of the hydroxyl group has been replaced with
5 an alkyl group R', the ester of carboxylic acid having the structure R-COO-R', wherein the R' is an alkyl chain comprising one or more carbons. As used herein, the term "triglycerides" refers to an ester derived from glycerol and three fatty acids.

In the context of the present disclosure, weight-% is abbreviated as wt-%, referring to the indicated weight of a liquid stream or effluent in question obtained through gas-liquid
10 separation, relative to the (total) weight of the feed, stream, effluent, product, component, or sample in question. Any known method can be used for the analysis. One example of a usable method includes the PIONA method (method to determine n-paraffins, iso-paraffins, olefins, naphthenes and aromatics), which is a GCxGC analysis method, as published e.g., by Pyl et al. in Journal of Chromatography A, 1218 (2011) 3217–3223 for the GCxGC
15 description. The weight-% of fatty acids, fatty acid methyl esters and trans fatty acid isomers in a liquid animal and/or vegetable fats and oils can be analysed, for example, according to ISO 12966-1:2014.

As used herein, the term "hydrotreatment (HT)" or "hydroprocessing", means a catalytic process for treating organic material by means of molecular hydrogen. In the context of the
20 present disclosure, hydrotreatment can comprise at least removal of oxygen from organic oxygen compounds as water i.e. hydrodeoxygenation (HDO), removal of sulfur from organic sulfur compounds as dihydrogen sulphide (H₂S), i.e. hydrodesulfurisation, (HDS), removal of nitrogen from organic nitrogen compounds as ammonia (NH₃), i.e. hydrodenitrogenation (HDN), removal of halogens, for example chlorine from organic chloride compounds as
25 hydrochloric acid (HCl), i.e. hydrodechlorination (HDCI), removal of metals by demetallization, removal of phosphorous through dephosphorization, hydroisomerisation (H-ISO) of the feed, hydrodearomatisation of the feed, and/or hydrogenation of olefinic bonds if present in the feed.

As used herein, the term "hydrodeoxygenation (HDO)" means removal of covalently bound
30 oxygen as water from the carboxylic acids of the feedstock using hydrogen. From an HDO reaction, a liquid product, comprising mainly even carbon number n-paraffins, is recovered.

As used herein, the term "hydroisomerisation" or "H-ISO" refers to isomerisation process in the presence of hydrogen, wherein the properties of the feedstock are improved by transforming normal/linear hydrocarbons to branched ones having the same carbon
35 number.

As used herein, the term "hydrocracking" refers to catalytic decomposition of organic hydrocarbon materials using molecular hydrogen at high pressure. In hydrocracking, the feedstock is catalytically converted to lower molecular weight compounds (i.e., lower Mw than in the compounds of the initial feedstock). The cracking conversion to lower molecular weight compounds is relatively unselective, and therefore carbon chains with various chain lengths are obtained through hydrocracking reactions.

As used herein, the term "deoxygenation conversion" refers to the wt-% of feedstock in question which has gone through removal of oxygen from the total weight of the initial feedstock. For example, deoxygenation conversion of 85 wt-% for C18 carboxylic acids indicates that 85 wt-% of the C18 carboxylic acids in the feedstock have undergone deoxygenation. The term "deoxygenation conversion" does not specify alone through which reaction oxygen is removed. Therefore, in the context of DCO reaction, such as that of step ii) of the current process, the deoxygenation conversion may in theory take place through DCO and/or HDO reactions. DCO selectivity, on the other hand, is calculated from deoxygenated products and therefore quantifies the amount of desired DCO reactions occurred compared to other deoxygenation reactions, excluding the unconverted feed (i.e., the non- deoxygenated feed) from the calculation.

As used herein, the term "selectivity" or "DCO deoxygenation selectivity" in context of a DCO reaction refers to a wt-% of deoxygenated product undergone deoxygenation through DCO reactions from the total amount of said deoxygenated product. In practice this means, the remaining wt-% of deoxygenated product/hydrocarbons have undergone deoxygenation through HDO reactions. For example, the specific DCO selectivity wt-% for C18 carboxylic acids can be calculated from the wt-%s of the C17 and C18 hydrocarbon components (i.e., HC17 and HC18) in the deoxygenated product with the formula $HC17/(HC17 + HC18) * 100$, thereby indicating the wt-% portion of C18 carboxylic acids which have undergone DCO reaction and converted to C17 hydrocarbons. The total DCO selectivity for both C18 and C16 carboxylic acids can be calculated from wt-%s of the C15 - C18 hydrocarbon components in the deoxygenated product with the formula $(HC15 + HC17)/(HC15 + HC16 + HC17 + HC18) * 100$.

A yield of a component, such as an aviation fuel intermediate component, that can be recovered from the DCO reaction of the step ii) of the current process, can be calculated as conversion multiplied by selectivity of the process in question. The aviation fuel intermediate component yield from the DCO reaction provides the potential for the aviation fuel component yield. Isomerised C17 paraffins boil in the range of aviation fuel boiling range

100 - 300°C (~C8-C17-fraction), whereas C18 paraffins boil at 317°C and should not be used as aviation fuel component.

As used herein, "degassed" refers to an effluent that has been subjected to gas-liquid separation and from which at least species that are gaseous at NTP (normal temperature and pressure) have been separated or removed. For example, such degassed effluents include degassed hydrotreated effluent. Further, for the purpose of analyses, any stream, effluent, product or sample analysed for any physico-chemical or a compositional characteristic, is in practice degassed prior to conducting any analysis. In practical language, they would be understood as a "liquid" stream, effluent, product or sample, respectively. The fraction separated from said degassed effluents, typically C1- C4 such as propane, may be referred to as "gaseous phase" or "gaseous fraction" of the respective effluent. Further, where any stream, effluent, product or sample is characterized by corresponding parameters, the numbers are given relative to the degassed weight or volume.

As used herein, the term "catalyst deactivation" refers to decreased activity of the catalyst, reflected by the amount of unconverted feed in the reactor effluent, and/or decreased selectivity of the catalyst, reflected by decreased amount of desired reaction products in the reactor effluent, at a given time point, compared to the activity and/or selectivity of the catalyst in the beginning of the process of the present disclosure. As used herein, the term catalyst deactivation is not limited to any specific deactivation type or mechanism, although the catalyst deactivation observed in the process of the present disclosure is generally believed to be attributed to poisoning and fouling phenomena and encompasses both reversible and irreversible deactivation.

As used herein, the term "ASA" refers to "amorphous silica alumina". In an embodiment, the terms "ASA" and "amorphous silica alumina" are synonymous to each other.

As used herein, the term "noble metal" refers to metallic chemical elements that can resist corrosion and typically are found in its elemental form, such as Au, Pt, Ru, Rh, Pd, Os, and Ir.

As used herein, the term "DCO zone" or "HT zone" refers to a DCO/HT reaction space wherein the DCO/HT reactions take place, respectively. The DCO/HT zone is an area or space delimited by the process conditions and/or by the process equipment, or the DCO/HT zone is a separate reactor dedicated for DCO/HT reactions, respectively. The DCO/HT zone can also be a separate catalyst bed dedicated for DCO/HT reactions, respectively, inside a

reactor wherein also other catalyst beds are located, or a separate catalyst bed part dedicated for DCO/HT reactions, in a catalyst bed wherein also other reactions take place.

As used herein "naphtha" refers to hydrocarbon components suitable for use in fuel compositions meeting standard specifications for gasoline fuels, such as specifications laid down in EN 228:2012 + A1:2017. Typically, such gasoline fuel components boil, i.e., have IBP and FBP, within a range from about 25 °C to about 210 °C, as determined according to EN ISO 3405:2019.

As used herein "diesel component" refers to hydrocarbon compositions suitable for use in fuel compositions meeting standard specifications for diesel fuels, such as specifications laid down in EN 590:2022 or in EN 15940:2016 + A1:2018 + AC:2019. Typically, such diesel fuel components boil, i.e. have IBP and FBP, within a range from about 160 °C to about 380 °C, as determined according to EN ISO 3405:2019.

THE OVERALL PROCESS

The present disclosure provides a process for producing a renewable aviation fuel component.

In an embodiment, a step i) of the process for producing a renewable aviation fuel component comprises providing a renewable feedstock comprising free carboxylic acids (FCAs), esters of carboxylic acids, triglycerides, or combinations thereof.

In the process, a renewable feedstock or bio-oil and/or fats are used which originate from renewable sources, such as fats and oils from plants and/or animals and/or fish and compounds derived from them. The structural units of a typical plant or vegetable or animal oil, or fat, useful as the feedstock are units comprising free fatty acids (FFAs) and/or free carboxylic acids (FCAs), and/or esters of carboxylic acids such as triglycerides, (which are carboxylic acid esters of glycerol, such as triester of glycerol with three fatty acid moieties), diglycerides and monoglycerides.

In an embodiment, the renewable feedstock comprises free carboxylic acids, esters of carboxylic acids, triglycerides, free fatty acids, derivatives of said fatty acids, such as esters of fatty acids, as well as triglycerides of fatty acids, metal salts of said fatty acids, or combinations of thereof.

In an embodiment, the renewable feedstock comprises at least 70 wt-%, preferably at least 80 wt-%, more preferably at least 90 wt-% of free carboxylic acids (FCAs), esters of

carboxylic acids, triglycerides or combinations thereof, of the total weight of the renewable feedstock.

In an embodiment, the renewable feedstock comprises natural fats or derivatives thereof.

In an embodiment, the renewable feedstock is selected from a feedstock consisting of plant
5 fats, plant oils, plant waxes, animal fats, animal oils, animal waxes, fish fats, fish oils, fish waxes, animal and/or fish and/or plant waste and residue materials such as used cooking oils, or any combinations thereof.

The feedstock may include, but is not limited to, plant oils, vegetable oils, microbial oils like
10 babassu oil, palm seed oil, carinata oil, olive oil, coconut butter, soybean oil, canola oil, coconut oil, muscat butter oil, rapeseed oil, peanut oil, sesame oil, maize oil, sunflower oil, poppy seed oil, cottonseed oil, soy oil, laurel seed oil, crude tall oil, tall oil, tall oil fatty acid, tall oil pitch, crude palm oil, palm oil, palm oil fatty acid distillate, jatropha oil, palm kernel oil, camelina oil, archaeal oil, bacterial oil, fungal oil, protozoal oil, algal-based oils, muscat butter oil, seaweed oil, mustard seed oil, oils from halophiles, lauric-myristic acid group
15 (C12-C14) including milk fats, palmitic acid group (C16) including earth animal fats, stearic acid group (C18) including earth animal fats, linoleic acid group (unsaturated C18) including whale and fish oils, erucic acid group (mono unsaturated carboxylic acid, C22:1) including whale and fish oils, oleo stearic acid group (conjugated unsaturated C18) including whale and fish oils, fats with substituted fatty acids (ricin oleic acid, C18) such as castor oil, or
20 mixtures of any two or more thereof.

In one embodiment, the renewable feedstock is selected from suitable feedstocks, preferably wastes and residues, listed in Annex IX, Part A or Annex IX, Part B of the Renewable Energy Directive (EU) 2018/2001, and mixtures thereof.

An exemplary renewable feedstock preferably includes waste and residue materials
25 originating from animal fat/oil, plant fat/oil and/or fish fat/oil. These may comprise sludge palm oil, such as palm effluent sludge (PES) or palm oil mill effluent (POME), used cooking oil (UCO), acid oils (ASK), brown grease (BG), sludge palm oil, spent bleaching earth oil (SBEO), technical corn oil (TCO) or lignocellulosic based oils, municipal solid waste-based oils, and/or algae-based oils. Most preferably, the feedstock includes UCO, sludge palm oil,
30 TCO and/or algae-based oils.

The carboxylic acids found in natural triglycerides are almost solely fatty acids of even carbon number. Therefore, the renewable feedstock typically comprises C8-C24 carboxylic acids, mainly C16-C22 carboxylic acids, more typically C16-C18 carboxylic acids, most typically C18 carboxylic acids.

The renewable feedstock may contain compounds comprising carbon-carbon double bonds, and thus be either saturated, unsaturated or polyunsaturated. In an embodiment, at least 40 wt-%, at least 50 wt-%, at least 60 wt-%, or at least 80 wt-% of the of the free carboxylic acids, esters of carboxylic acids and triglycerides of the renewable feedstock are
5 unsaturated.

In an embodiment, the aviation fuel component according to the disclosure comprises C8-C17 hydrocarbons. C18 hydrocarbons do not really fit within the required boiling range which is set for the aviation fuel according to ASTM D7566-2022 Annex A2, so only a very limited amount of C18 hydrocarbons may be comprised in the aviation fuel component,
10 depending on its desired properties. Nonetheless, the aviation fuel component, fulfils the required properties of ASTM D7566-2022 Annex A2. Decarboxylation/decarbonylation (DCO) is a selective reaction capable of reducing the carbon number of a carboxylic acids with one carbon, such as from C18 to C17, the DCO reaction thereby allowing the utilisation of C18 carboxylic acids in an aviation fuel component.

15 In an embodiment, the renewable feedstock comprises free carboxylic acids, esters of carboxylic acids and triglycerides with a carbon number of C18 or more. In an embodiment, the renewable feedstock comprises at least 40 wt-%, preferably at least 50 wt-%, more preferably at least 60 wt-%, more preferably at least 70 wt-%, even more preferably at least 80 wt-%, most preferably at least 90 wt-% of FCAs and esters of carboxylic acids having a
20 carbon number of at least C18, of the total weight of the renewable feedstock.

In an embodiment, at least 40 wt-%, preferably at least 50 wt-%, more preferably at least 70 wt-%, even more preferably at least 80 wt-% of the carbons of the renewable feedstock are contained in C18 FCAs and esters of carboxylic acids.

In an embodiment, the renewable feedstock is of plant or animal origin, or combinations
25 thereof.

In an embodiment, a step ii) of the process for producing a renewable aviation fuel component comprises subjecting the feedstock to a decarboxylation and/or decarbonylation (DCO) reaction in a DCO zone, for removal of one carbon from a carbon chain of a carboxylic acid moiety of the renewable feedstock, in the presence of a DCO catalyst,
30 thereby obtaining a DCO effluent 21.

In Fig. 1 is schematically presented an exemplary embodiment of a process sequence presenting the current process for producing a renewable aviation fuel component. In the process of Fig. 1, the renewable feedstock 10 is subjected to the DCO reaction 20 at the step ii), from which the obtained DCO effluent 21 is subjected to a hydrotreatment (HT)

reaction 30 at a step iii), from which a hydrotreated effluent 31 is obtained and subjected to a gas-liquid separation 40 at a step iv), from which a degassed hydrotreated effluent 41 is obtained and subjected to a hydroisomerisation (H-ISO) 50 at a step v), from which a degassed hydroisomerised effluent 51 is obtained and subjected to a fractionation 60 at a step vi), from which at least the renewable aviation fuel component is recovered.

The DCO reaction 20 of the step ii) is carried out in the liquid phase. Thereby, the free carboxylic acids (FCAs) and the esters of carboxylic acids with a carbon number C18 in the renewable feedstock can be converted to C17 hydrocarbons through the DCO reaction 20 without reducing the carbon number of said C17 hydrocarbons further, unlike through, for example, a less controllable cracking process, thereby losing yield. Therefore, the current process is beneficial, as it results in surprisingly low yield loss and low hydrogen consumption, compared to a process converting C18 hydrocarbons to C17 hydrocarbons, or smaller, by hydrocracking. Therefore, the current process is also beneficial, as it produces a renewable aviation fuel component without the need for an additional hydrocracking step or without the need to expose the entire reaction effluent to hydrocracking, thereby allowing higher yield of the renewable aviation fuel component to be obtained.

In an embodiment, the renewable feedstock 10 subjected to DCO reaction 20 at step ii) comprises less than 50 wt-ppm, preferably less than 10 wt-ppm, more preferably less than 5 wt-ppm of sulfur. In an embodiment, wherein the renewable feedstock is at least partly of animal origin, the feedstock 10 subjected to DCO reaction 20 at step ii) may comprise more than 5 wt-ppm, or more than 50 wt-ppm of sulfur. In such embodiments, catalyst deactivation can be avoided through catalyst selection, for example by selecting a non-noble metal catalyst.

In an embodiment, the renewable feedstock 10 subjected to DCO reaction 20 at the step ii) comprises less than 50 wt-ppm, preferably less than 10 wt-ppm, more preferably less than 5 wt-ppm of nitrogen. In an embodiment, wherein the renewable feedstock is at least partly of animal origin, the feedstock 10 subjected to DCO reaction 20 at the step ii) may comprise more than 5 wt-ppm, or more than 50 wt-ppm of nitrogen. In such embodiments, catalyst deactivation can be avoided through catalyst selection, for example by selecting a non-noble metal catalyst.

In an embodiment, the renewable feedstock 10 entered into the DCO reaction 20 of the step ii) comprises 8 - 12 wt-% of oxygen of the total weight of the renewable feedstock.

In an embodiment, the feedstock is subjected to the decarboxylation and/or decarbonylation (DCO) reaction 20 in the DCO zone, wherein the DCO deoxygenation selectivity is at least 75 wt-% of the total weight of deoxygenated hydrocarbons, and the deoxygenation conversion is at least 50 wt-% of the total weight of feedstock.

- 5 In an embodiment, the feedstock is provided to the DCO zone so that a backflow of the feedstock stream is prevented. In an embodiment, the process conditions of the DCO zone, such as H₂ flow and weight hourly space velocity (WHSV), can be adjusted independently of the process conditions within any other catalyst bed located inside the same reactor.

The deoxygenation conversion of the feedstock is an adjustable process parameter-which
10 can be adjusted by modifying one or more of the operating conditions of the process step in question, selected from temperature, pressure, WHSV, or combination thereof. The deoxygenation conversion of the feedstock can, for example, be increased by increasing the temperature, or decreasing the WHSV and reducing diluent circulation into the process step (increased residence time). The WHSV of a process can be increased, for example,
15 by increasing the flow rate of the feed or decreasing the amount of catalyst. Increased WHSV raises the ratio of DCO/HDO reactions.

In an embodiment, the DCO deoxygenation selectivity in the step ii) is at least 80 wt-%, preferably at least 85 wt-%, more preferably at least 90 wt-%, even more preferably at least 95 wt-%, most preferably at least 97 wt-%, of the total weight of deoxygenated
20 hydrocarbons. In an embodiment, the DCO reaction of the step ii) comprises a DCO deoxygenation selectivity of at least 90 wt-%, at least 91 wt-%, at least 92 wt-%, at least 93 wt-%, at least 94 wt-%, at least 95 wt-%, at least 96 wt-%, at least 97 wt-%, at least 98 wt-%, at least 99 wt-%, or 100 wt-%, of the total weight of deoxygenated hydrocarbons in the DCO effluent 21.

25 In an embodiment, the renewable feedstock mainly comprises C16 and C18 carboxylic acids, and therefore the DCO deoxygenation selectivity in the step ii) in practice means selectivity for C18 and C16 carboxylic acids. In an embodiment, the DCO deoxygenation selectivity for C18 and C16 carboxylic acids is at least 75 wt-%, preferably at least 80 wt-% of the total weight of deoxygenated hydrocarbons and/or deoxygenated C18 and C16
30 hydrocarbons. In an embodiment, the DCO deoxygenation selectivity for C18 and C16 carboxylic acids in the step ii) is at least 85 wt-%, preferably at least 90 wt-%, more preferably at least 95 wt-%, most preferably at least 97 wt-%, of the total weight of deoxygenated hydrocarbons.

In an embodiment, the deoxygenation conversion of the step ii) is at least 55 wt-%, preferably at least 60 wt-%, more preferably at least 70 wt-%, more preferably at least 80 wt-%, at least 90 wt-%, at least 95 %, or at least 99 wt-%, of the total weight of the feedstock.

5 In an embodiment, the DCO reaction 20 of the step ii) comprises a deoxygenation conversion of at least 55 wt-%, at least 60 wt-%, at least 65 wt-%, at least 70 wt-%, at least 75 wt-%, at least 80 wt-%, at least 85 wt-%, at least 90 wt-%, or at least 95 wt-%, or at least 99 wt-%, of the total weight of the feed.

10 In an embodiment, the DCO reaction 20 of the step ii) comprises a deoxygenation conversion of C18 and C16 carboxylic acids of at least 50 wt-% of the total weight of the feed. In an embodiment, the deoxygenation conversion of C18 and C16 carboxylic acids is at least 55 wt-%, at least 60 wt-%, at least 70 wt-%, at least 80 wt-%, at least 90 wt-%, at least 95 %, or at least 99 wt-%, of the total weight of the feed.

The deoxygenation conversion may be adjusted by e.g., altering the processing temperature and/or the feed inlet flow rate.

15 In an embodiment, the deoxygenation conversion of the DCO reaction of the step ii) is however, preferably not adjusted to 100 wt-% of the total weight of the feedstock, as this can compromise the DCO deoxygenation selectivity of the process. In some embodiments, the deoxygenation conversion of the DCO reaction of the step ii) is nonetheless adjusted up to 99.5 wt-% of the total weight of the feed.

20 In an embodiment, at least 50 wt-%, preferably at least 70 wt-%, more preferably at least 85 wt-%, even more preferably at least 95 wt-% of the total weight of the free carboxylic acids, esters of carboxylic acids, triglycerides, or combinations thereof comprised by the renewable feedstock 10 are deoxygenated in the DCO reaction 20 of step ii).

25 In an embodiment, the DCO effluent from step ii) is subjected at least partly to a separation of at least CO and/or CO₂ gases, before directing the DCO effluent to step iii). In an embodiment, the DCO effluent 21 from the step ii) is directed to the separation of at least CO and/or CO₂ gases 25, wherein majority of carbon oxides present in the DCO effluent are removed, thereby obtaining a carbon oxide deprived DCO effluent 22 (Fig. 2). In another embodiment, the CO and/or CO₂ gasses are removed from the DCO effluent completely, 30 thereby obtaining a carbon oxide depleted DCO effluent 22. In an embodiment, subjecting the DCO effluent to a gas-liquid separation 25 comprises removal of CO₂ and CO in a gaseous phase. In an embodiment, the DCO effluent 21 from step ii) is subjected to a degassing 25 of all gaseous components, thereby obtaining a degassed DCO effluent 22. In context of this application, the term DCO effluent can refer to DCO effluent wherefrom

carbon oxides are or are not separated from. In a preferable embodiment, the carbon oxides are separated from the DCO effluent.

In an embodiment, the DCO effluent 22 (e.g., Fig. 2) wherefrom carbon oxides are separated from, comprises at least 85 wt-%, preferably at least 90 wt-% of hydrocarbons with uneven carbon number, of the total weight of the DCO effluent 22, provided that the DCO deoxygenation selectivity and the deoxygenation conversion are set to 100 wt-%.

The separation of at least CO and/or CO₂ gases from the DCO effluent is beneficial prior to the hydrotreatment (HT) of the consecutive step iii), as it prevents/minimises occurrence of methanation reaction ($\text{CO} + 3\text{H}_2 \leftrightarrow \text{CH}_4 + \text{H}_2\text{O}$), reverse water gas reaction (rWGS) ($\text{CO}_2 + \text{H}_2 \leftrightarrow \text{CO} + \text{H}_2\text{O}$), and/or Boudouard reaction ($\text{CO} + \text{CO} \leftrightarrow \text{C} + \text{CO}_2$) in the DCO effluent during the hydrotreatment reaction 30 of the step iii). Occurrence of the methanation reaction and the rWGS are harmful, as they increase the consumption of hydrogen needed for the reaction. Occurrence of the Boudouard reaction in the DCO effluent is harmful, as it might increase coking of the HT/HDO catalyst.

In another embodiment, the DCO effluent 21 from the step ii) is not subjected to a separation of CO/CO₂ gases before the step iii).

In an embodiment, carbon oxides are separated from the DCO effluent 21 by flashing, or gas stripping, preferably carbon oxides are separated by reducing the pressure in the DCO effluent i.e., by flashing. In an embodiment, the CO and/or CO₂ gases are separated from the DCO effluent 21 by reducing the pressure of said effluent by 0.2 - 1 MPa after leaving the DCO reaction zone.

In an embodiment, carbon oxides comprising CO and/or CO₂ gases are separated from the DCO effluent 21 by gas stripping, wherein a gas flow comprising inert gas such as steam, methane, nitrogen, or combinations thereof may be used for removing gaseous products. In such embodiments, the gas flow comprising inert gas may be combined with the DCO effluent 21, or it may be led to the DCO catalyst bed, or it may also be led to other parts of the reactor. In an embodiment, the resulting gas flow comprising stripped CO and/or CO₂ gases is fed to an amine wash and then released to the atmosphere or burned.

The separation of carbon oxides and other gases can take place between reactors, between zones, or in between catalyst beds within the same reactor system. The separation of carbon oxides and other gases can also take place at the bottom of the reactor, even when the reactor also comprises other catalyst beds besides a DCO catalyst bed.

In a preferable embodiment, the concentration of gases in the DCO effluent 21 is reduced through degassing. In an embodiment, at least part of the DCO effluent 21 is deprived from

all gaseous elements, including, in addition to the carbon oxides, also redundant/superfluous gaseous hydrogen, which is needed in the subsequent steps of the process. However, the amount of hydrogen in the DCO effluent 21 is small, as the amount of hydrogen added to the DCO reaction 20 is typically low. In some embodiments, some of
5 the gaseous components which are condensed after the removal from the DCO effluent, may be returned to the DCO effluent at the hydrotreatment (HT) of the step iii).

In an embodiment, the total oxygen content of the carbon oxide deprived DCO effluent 22 is less than 5 wt-% of oxygen, preferably less than 3 wt-%, more preferably less than 2 wt-%, of the total weight of said carbon oxide deprived DCO effluent 22.

10 The wt-% of decarboxylated/decarbonylated hydrocarbons in the DCO effluent can be calculated, for example, as a wt-% sum of hydrocarbons with an odd carbon number. In an embodiment, at least 55 wt-%, preferably at least 70 wt-%, more preferably at least 80 wt-%, more preferably at least 85 wt-%, even more preferably at least 90 wt-%, most preferably at least 95 wt-%, or at least 99 wt-% of the free carboxylic acids (FCAs), esters of carboxylic
15 acids, and triglycerides comprised in the renewable feedstock are deoxygenated through the DCO reaction 20 in the step ii).

In an embodiment, at least 55 wt-%, preferably at least 70 wt-%, more preferably at least 80 wt-%, even more preferably at least 90 wt-%, most preferably at least 95 wt-%, or at least 99 wt-% of the FCAs and esters of carboxylic acids comprised be the renewable
20 feedstock are converted to hydrocarbons with odd carbon number in the DCO reaction 20 of the step ii). For example, natural vegetable oils mainly comprise even carbon number carboxylic acids.

In an embodiment, the DCO catalyst of the DCO reaction 20 of the step ii) is a heterogeneous catalyst comprising at least a metal and a catalyst support. In an
25 embodiment, the DCO catalyst is a heterogeneous catalyst comprising at least one transition metal or a combination of transition metals. In an embodiment, the DCO catalyst comprises at least one metal from the group 7 to 11 metals of the Periodic Table of the elements, or any combination thereof.

In an embodiment, the metal of the heterogeneous DCO catalyst is an elemental metal, or
30 a metal compound, depending on the metal(s) comprised by the DCO catalyst. In an embodiment, the DCO reaction 20 of step ii) is done in the presence of a metal catalyst, such as elemental metal catalyst. In an embodiment, the DCO reaction of step ii) is done in the presence of an elemental noble metal catalyst. In an alternative embodiment, the DCO reaction of step ii) is done in the presence of a metal compound catalyst.

In an embodiment, the DCO catalyst of the step ii) is a heterogeneous catalyst comprising a metal selected from nickel, cobalt, copper, zinc, molybdenum, manganese, ruthenium, rhodium, rhenium, iridium, palladium, platinum and any combinations thereof.

In an embodiment, the heterogeneous DCO catalyst is a monometallic catalyst. In an embodiment, the heterogeneous DCO catalyst is a bimetallic catalyst. In an embodiment, the heterogeneous DCO catalyst is a trimetallic catalyst. In an embodiment, the DCO catalyst is a metal compound.

In an embodiment, the heterogeneous DCO catalyst comprises a noble metal selected from ruthenium, rhodium, platinum, palladium, rhenium, iridium, and any combinations thereof, more preferably the heterogeneous DCO catalyst is a non-sulfided catalyst comprising a noble metal selected from ruthenium, rhodium, platinum, palladium and any combinations thereof. In an embodiment, the heterogeneous DCO catalyst comprising at least one noble metal preferably contains platinum and/or palladium.

In an embodiment, the heterogeneous DCO catalyst comprises nickel, platinum and/or palladium, as these catalysts are selective towards deoxygenation via DCO reactions. In an embodiment, the selectivity of the catalysts towards DCO reaction decreases in the following order $\text{Pd} > \text{Pt} > \text{Ni} > \text{Rh} > \text{Ir} > \text{Ru}$.

Renewable feedstocks, such as various vegetable oils, do not typically comprise significant amounts of sulfur components, with the exception of e.g. tall oil feed originating from the Kraft processing and selected animal fats, and therefore it is possible to use noble metal catalysts with such feedstocks. In an embodiment, the heterogeneous DCO catalyst is preferably a non-sulfided catalyst. Non-sulfided catalysts are preferred because they do not require sulfidation prior to or during operation and hence, no additional process steps for sulfidation of the catalyst or addition of sulfur into feed for maintaining the catalyst activity are required. Further, the sulfur content of various process streams may be kept low and less efficient H_2S separation and recovery is needed from various process streams, and therefore no additional equipment, such as amine wash or sulfur unit, are required. Especially non-sulfided catalysts comprising noble-metals may be active at lower temperatures and show higher selectivity for isomerisation reactions, compared to sulfided catalysts, but are sensitive to deactivation by H_2S . The sulfur content of a feed may be determined according to ISO 20846-2011 from liquids or according to ASTM-D6667 from gaseous fractions. Non-sulfided noble metal DCO catalysts are beneficial, as they are efficient in catalysing specific DCO reactions.

In an embodiment, the function of catalysts containing noble metals such as Pt and/or Pd is impaired in the presence of sulfur. In such embodiment, sulfur functions as a catalyst inhibitor and reduces the catalysts life, necessitating more frequent catalyst regeneration or change. In an embodiment, the feedstock subjected to the DCO reaction 20 at the step ii) comprises less than 50 wt-ppm, preferably less than 10 wt-ppm, more preferably less than 5 wt-ppm of sulfur of the total weight of the feedstock. In an embodiment, any sulfur present in the feedstock in the DCO reaction is not added sulfur, but sulfur which was initially present in the renewable feedstock 10. In an embodiment, wherein the renewable feedstock 10 comprises more than 50 wt-ppm of sulfur, the sulfur in the feedstock is removed in a pretreatment 80 prior to the DCO reaction (e.g., Fig. 2). In an embodiment, the sulfur in the feedstock is removed in a pretreatment 80 prior to the DCO reaction wherein the DCO catalyst comprises noble metals such as Pt and/or Pd.

In an embodiment, the DCO catalyst comprises nickel, preferably oxidized Ni. In an embodiment, the DCO catalyst comprises cobalt. In an embodiment, the DCO catalyst comprises copper. In an embodiment, the DCO catalyst comprises zinc. In an embodiment, the DCO catalyst comprises molybdenum. In an embodiment, the DCO catalyst comprises manganese.

DCO catalysts comprising nickel (Ni) are beneficial, as in the presence of hydrogen such catalysts efficiently hydrogenate double bonds of fatty acids as well as catalyse specific DCO reactions. In an embodiment, catalysts comprising nickel can function in the presence of sulfur. Therefore, when the DCO reaction is done in the presence of a heterogeneous catalyst containing Ni, the feedstock may comprise higher amount of sulfur without compromising the catalyst function, or the conversion and/or selectivity of the DCO reaction 20. In an embodiment, the DCO reaction 20 of the step ii) is done in the presence of a heterogeneous catalyst containing Ni and the feedstock subjected to the DCO reaction comprises at most 100 wt-ppm, preferably at most 50 wt-ppm, more preferably at most 25 wt-ppm of sulfur of the total weight of the feedstock. DCO reactions comprising these low amounts of sulfur are beneficial, as low sulfur content of the feedstock allows improved hydrogenation of double bonds.

In an embodiment, wherein the DCO catalyst comprises a noble metal such as platinum or palladium, a pretreatment of the noble metal catalyst with hydrogen is beneficial, said pretreatment by hydrogenation of the catalyst being carried out before subjecting the renewable feedstock 10 to the DCO reaction 20 in the DCO zone.

In an embodiment, loading of the active catalyst metal varies in the range of 0.1 - 20 wt-% of the total catalyst weight. In the case nickel is used as an active catalyst metal, the

beneficial loading varies in the range of 2 - 55 wt-%, preferably 10 - 30 wt-% of the total catalyst weight. In the case Pt and/or Pd is used as an active catalyst metal, the beneficial loading varies in the range of 0.1 – 0.4 wt-%, preferably 0.2 wt-% of the total catalyst weight.

In an embodiment, the DCO catalyst further comprises at least one support, selected from
5 alumina; silica; zirconia; titania; carbon, such as activated carbon or graphite; molecular sieve; and any combinations thereof.

In a preferable embodiment, the DCO catalyst support is a regenerable catalyst support, such as alumina, silica or zeolite. A regenerable catalyst support is beneficial as the lifetime of such support is long. Alumina as a catalyst support is beneficial, as it is a stable support,
10 and it tolerates carbon oxides present in the DCO reaction well. Zeolite as a catalyst support is beneficial, as it has a high surface area, high porosity, high adsorption capacity, and it is easily separated from reactants and products.

In an embodiment, the DCO catalyst is supported on oxides and/or mesoporous materials.

In an embodiment, the DCO catalyst support is a carbonaceous support, such as activated
15 carbon, carbon fibres, carbon nanotubes attached to monoliths, or carbon cloths. In an embodiment, carbonaceous catalyst support is beneficial as it enhances the adsorption of acidic CO₂ on the basic sites of the catalyst support's surface, thereby inhibiting carbon deposition on the catalyst and thus enhancing the catalytic stability.

In an embodiment, the DCO reaction 20 of step ii) is done in the presence of an elemental
20 noble metal catalyst containing at least one noble metal selected from platinum, palladium, ruthenium and rhodium, the said catalyst being supported on regeneratable support, preferably the support is alumina and/or silica.

In an embodiment, the DCO catalyst is a heterogeneous catalyst comprising a metal selected from Ni, Co, Cu, Zn, Mo, Mn, Ru, Rh, Re, Ir, Pd, Pt or any combinations thereof,
25 and a support, selected from alumina, silica, zirconia, titania, carbon, such as activated carbon or graphite, molecular sieve, or any combinations thereof.

In an embodiment, the DCO catalyst metal is impregnated or deposited on the catalyst support and optionally converted into its sulphides.

The DCO reaction 20 may be carried out in batch, semi-batch or continuous mode of
30 reaction, in reactors such as a trickle-bed reactor, continuous tubular or continuous stirred tank reactor, also allowing separation of the gaseous CO/CO₂, as well as any light hydrocarbons having carbon number < C₄.

In an embodiment, the DCO reaction of the step ii) is carried out in the presence of a H₂ feed ratio of less than 300 nl H₂/ liter of the feedstock or completely without added H₂. By the notation "nl H₂/ liter of feedstock" is herein meant normal liters of hydrogen gas per liter of feedstock. In an embodiment, the DCO reaction 20 is carried out in the presence of a H₂ feed ratio of less than 250 nl H₂/ liter of the feedstock, less than 200 nl H₂/ liter of the feedstock, less than 150 nl H₂/ liter of the feedstock, less than 100 nl H₂/ liter of the feedstock, less than 50 nl H₂/ liter of the feedstock, or completely without added H₂.

The DCO reaction does not require any added hydrogen, and the absence of hydrogen in the DCO reaction ensures that the feedstock is deoxygenated through DCO reactions instead of hydrodeoxygenation (HDO). Therefore, in an embodiment, no external H₂ is added into the DCO reaction 20 of step ii). No hydrogen is consumed in side-reactions or in direct reduction of carboxylic groups at the DCO reaction of the step ii).

In an embodiment, the DCO reaction of the step ii) is done in the presence of steam, nitrogen, and/or methane (CH₄), preferably the DCO reaction of the step ii) is performed in the presence of methane (CH₄). In such embodiment, no external H₂ needs to be added into the DCO reaction of step ii).

In an embodiment, the DCO reaction 20 is carried out in the presence of a very low amount of hydrogen, i.e., H₂ feed ratio of 0.01 - 300 nl H₂/ liter of the feedstock, such as 0.01 - 150 nl H₂/ liter of the feedstock, such as 0.01 - 100 nl H₂/ liter of the feedstock, such as 0.01 - 50 nl H₂/ liter of the feedstock, such as 0.01 - 20 nl H₂/ liter of the feedstock, or even 0.01 - 10 nl H₂/ liter of the feedstock.

Said very low amount or no H₂ in the DCO step is beneficial as the consumption of hydrogen in the overall process is thereby minimized. Moreover, said very low amount or no added hydrogen in the DCO reaction suppresses the possible methanation reaction, i.e., the conversion of carbon oxides to methane. In an embodiment, a very low hydrogen amount is required at the DCO reaction 20 only for the reduction of the DCO catalyst. The DCO catalyst can, however, be optionally pretreated with hydrogen prior to directing the renewable feed in contact with the catalyst. The pretreatment of the DCO catalyst is preferred as it ensures the activity of the catalyst. Therefore, no hydrogen is necessarily required for the reduction of the DCO catalyst in the DCO reaction 20.

However, in some embodiments, it is beneficial to carry out the DCO reaction of the step ii) in the presence of a slightly higher amount of H₂, as hydrogenation of the feedstock allows hydrogenation of possible double bonds present in the feedstock which also releases heat to be utilized in the endothermic DCO reaction 20. Therefore, in certain embodiments,

the DCO reaction 20 of the step ii) is carried out in the presence of a H₂ feed ratio of 50 - 300 nl H₂/ liter of the feedstock, or 50 - 200 nl H₂/ liter of the feedstock, or 50 - 150 nl H₂/ liter of the feedstock, which means, the amount of H₂ is still low. Nevertheless, advantageously the DCO reaction 20 is carried out in the presence of said very low amount
5 of hydrogen, and the hydrogenation of double bonds present in the feedstock is done primarily in the HT reaction 30 step iii).

In an embodiment, wherein the DCO reaction 20 of the step ii) is carried out in the presence of a H₂ feed ratio of 0.01 - 300 nl H₂/ liter of the feedstock, preferably 0.01 -150 nl H₂/ liter of the feedstock, the DCO deoxygenation selectivity is at least 75 wt-%, preferably at least
10 80 wt-% of the total weight of deoxygenated hydrocarbons, and the deoxygenation conversion is at least 50 wt-%, preferably at least 55 wt-% of the total weight of feed.

In an embodiment, deoxygenation of the renewable feedstock 10 through the DCO reaction 20, is beneficial as the deoxygenation of the feedstock and thereby the overall process of producing a renewable aviation fuel has a reduced consumption of hydrogen. In an
15 embodiment, deoxygenation of the feedstock through the DCO reaction 20 reduces the required amount of hydrogen in any consecutive deoxygenation step utilizing hydrogen, such as hydrodeoxygenation (HDO), as the need for further deoxygenation of the feed after the DCO reaction of the step ii) is reduced or excluded.

In an embodiment, the DCO reaction 20 of the step ii) is carried out at a temperature of 50
20 - 450 °C, preferably the temperature is 200 – 450 °C, more preferably the temperature is 250 - 400 °C; and/or at a pressure of 0.1 - 10 MPa, preferably the pressure is 0.1 - 2 MPa. In an embodiment, the DCO reaction 20 of the step ii) is carried out at a temperature of 350 – 400 °C and/or at a pressure of 0.1 - 2 MPa or 0.1 - 0.5 MPa.

In an embodiment, the upper limit of the reaction temperature of the DCO reaction is set
25 sufficiently low, i.e., less than 400 °C, and thus no undesired decomposition of the feedstock occurs. In an embodiment, the reaction temperature of the DCO reaction is sufficiently high, and the pressure of the DCO reaction is sufficiently low, for the deoxygenation to take place through DCO reactions instead of HDO reactions.

In an embodiment, the DCO reaction 20 of the step ii) is carried out at a Weight Hourly
30 Space Velocity (WHSV) of 0.1 - 15 h⁻¹, preferably a WHSV of 0.25 - 5 h⁻¹. In an embodiment, the DCO reaction 20 of the step ii) comprises a Weight Hourly Space velocity (WHSV) of 0.1 - 15 h⁻¹, preferably the WHSV is 0.1 - 1 h⁻¹. In an embodiment, the upper limit of the WHSV of the DCO reaction is set sufficiently low, to achieve the desired deoxygenation

conversion wt-% and to avoid nonspecific side reactions, especially if the DCO reaction temperature is high.

In an embodiment, the DCO reaction 20 of the step ii) is carried out:

- in the presence of the heterogeneous DCO catalyst comprising a metal selected from nickel, cobalt, copper, zinc, molybdenum, manganese, ruthenium, rhodium, rhenium, iridium, palladium, platinum and any combinations thereof, wherein the DCO catalyst comprises at least one support selected from alumina, silica, zirconia, titania, carbon such as activated carbon or graphite, molecular sieve, and any combinations thereof;
 - in the presence of a H₂ feed ratio of less than 300 nl H₂/ liter of the feedstock, or completely without added H₂;
 - at a temperature of 50 – 450 °C, preferably the temperature is 200 – 450 °C, more preferably the temperature is 250 – 400 °C;
 - at a pressure of 0.1 – 10 MPa, preferably the pressure is 0.1 – 2 MPa;
 - at a Weight Hourly Space Velocity (WHSV) of 0.1 – 15 h⁻¹, preferably a WHSV of 0.25 – 5 h⁻¹;
- or any combination thereof.

In an embodiment, the step iii) of the process comprises subjecting at least a portion of the DCO effluent from step ii) to a hydrotreatment (HT) reaction in the presence of hydrogen and a hydrotreatment catalyst in a HT zone, to obtain a hydrotreated effluent.

- The DCO reaction 20 of the step ii) and the HT reaction 30 of the step iii) are two separate reactions taking place in their respective reaction zones. Consequently, the feed entering the HT reaction zone for HT reaction 30 is the DCO effluent 21 which has already undergone the DCO reaction 20 in the DCO zone.

- In an embodiment, the HT zone is a HT reaction space wherein the HT reactions 30 take place. In an embodiment, the feed is provided to the HT zone so that a backflow of the feedstock stream is prevented. In an embodiment, the process conditions of the HT zone, such as H₂ flow and WHSV can be adjusted independently of the process conditions within any other catalyst bed located inside the same reactor.

- In an embodiment, the hydrotreatment (HT) reaction of the step iii) is selected from hydrodeoxygenation (HDO), hydrogenation of double bonds, hydrodenitrogenation (HDN), hydrodesulfurization (HDS), and any combinations thereof, preferably the hydrotreatment reaction of the step iii) comprises HDN.

In an embodiment, the hydrotreatment (HT) reaction 30 of the step iii) is configured to remove any heteroatoms present in the DCO effluent.

In an embodiment, the HT reaction 30 of the step iii) comprises removal of heteroatoms and other impurities through hydrotreatment, such impurities being selected from halogens, metals, phosphorus, aromatics, or any combination thereof. In an embodiment, said removal of impurities is selected from the group consisting of hydrodehalogenation (HDH),
5 demetallization, dephosphorization, hydrodearomatization (HDA) or any combinations thereof. In an embodiment, the HT reaction 30 comprises hydrogenation of double bonds from the DCO effluent. In an embodiment, subjecting the DCO effluent to the HT reaction 30 hydrogenates any remaining double bonds and removes impurities, preferably HT
10 reaction 30 comprises HDN and/or HDS. The embodiment, wherein HT reaction 30 comprises hydrogenation of double bonds is beneficial, as the presence of olefinic bonds in the feedstock entering the H-ISO 50 of the step v) can lead to formation of aromatics (even polyaromatics), oligomerisation of olefins into high molecular weight compounds (i.e., compounds not included within aviation fuel and/or diesel fuel boiling range) and may increase the temperature of the H-ISO 50. In an embodiment, the HT reaction 30 of the step
15 iii) comprises also hydroisomerization.

In an embodiment, the HT reaction 30 is advantageously performed by avoiding hydrocracking (HC), as this would unnecessarily reduce the final yield of the aviation fuel component.

In a preferred embodiment, the HT reaction 30 of the step iii) comprises at least HDN, as
20 catalysts used in the H-ISO of the step v) may be sensitive to the presence of nitrogen, which is therefore optimally removed from the feed prior to the step v).

In an embodiment, the HT reaction 30 of the step iii) comprises HDO. In an embodiment, wherein the hydrotreatment comprises HDO, subjecting at least a fraction of the DCO effluent to the HT reaction removes oxygen possibly remaining in the DCO effluent. The
25 embodiment wherein the HT reaction comprises HDO is beneficial, as the overall process thereby comprises separate DCO and HDO reactions, which can be separately optimized through the process conditions.

All the reactions that may take place in the hydrotreatment of the step iii) can be conducted in the same reactor in different catalyst beds, or even in the same catalyst bed,
30 simultaneously or successively.

In an embodiment, the HT reaction 30 of the step iii) is carried out in the presence of the HT catalyst comprising at least one group VIII and/or VIB metal of the Periodic Table, preferably selected from nickel, molybdenum, tungsten, cobalt, and any combinations thereof, preferably selected from NiMo, CoMo, or NiW.

In an embodiment, the hydrotreatment of the step iii) comprises HDO and/or HDN and/or HDS and the reaction of the step iii) is carried out in the presence of a hydrotreatment catalyst comprising Ni, NiMo, CoMo, or NiW.

5 In an embodiment, the hydrotreatment reaction of the step iii) is done in the presence of a sulfided hydrotreatment catalyst. Sulfided catalyst is beneficial in maintaining activity of the catalyst.

10 In an embodiment, the hydrotreatment of the step iii) comprises HDO and/or HDN and/or HDS and the reaction of the step iii) is carried out in the presence of a sulfided hydrotreatment catalyst. The sulfided catalyst favours the catalysis of deoxygenation reactions.

In an embodiment, the HT catalyst of the step iii) further comprises at least one support selected from zeolite, silica, alumina, amorphous silica alumina (ASA), and any combination thereof.

15 In an embodiment, the hydrotreatment of the step iii) is done in the presence of the sulfided hydrogenation catalyst containing at least one supported Ni, NiMo, CoMo, and/or NiW catalyst, the support being zeolite, silica, alumina and/or ASA, or combinations thereof. In an embodiment, the hydrogenation catalyst is a sulfided NiMo supported on ASA and/or alumina.

20 In an embodiment, the hydrotreatment of the step iii) comprises HDN, and the HT reaction 30 of the step iii) is preferably carried out in the presence of a hydrotreatment catalyst comprising NiMo supported on alumina.

25 In an embodiment, less than 50 wt-%, preferably less than 40 wt-%, more preferably less than 30 wt-%, even more preferably less than 20 wt-%, even more preferably less than 15 wt-%, even more preferably less than 10 wt-%, most preferably less than 5 wt-% of the total weight of the of the free carboxylic acids, esters of carboxylic acids, triglycerides, or combinations thereof, comprised by the initial renewable feedstock 10 are deoxygenated in the HT reaction 30 of the step iii).

30 Majority or all oxygen in the renewable feedstock is removed at the DCO reaction 20 of the step ii), therefore HDO in the hydrotreatment of step iii) may not be necessary. In some embodiments, wherein the deoxygenation conversion of the DCO reaction 20 at the step ii) is not 100 wt-%, but instead, for example 70 wt-%, 80 wt-%, 90 wt-%, or 95 wt-% of the total weight of the renewable feedstock, a low concentration of oxygen remains in the DCO effluent. In such embodiments, it is beneficial when the hydrotreatment of the step iii) comprises HDO, for removal of the remaining oxygen from the DCO effluent. In such

embodiments however also, the amount of hydrogen needed for the HDO of the DCO effluent is considerably lower when compared to deoxygenation of the renewable feedstock based solely on of HDO. Therefore, even if the HT reaction 30 of the step iii) comprises HDO, the process is beneficial in that it utilises very little hydrogen. Therefore, the DCO reaction 20 being performed prior to the HT reaction 30 in the current process is beneficial, as the total amount of hydrogen needed for both the DCO and HT reactions is minimized. For example, in other processes wherein a HT reaction, such as HDO reaction, is prior to a DCO reaction, or wherein DCO and HDO reactions occur simultaneously, the consumption of hydrogen may be much higher when hydrogen is made available (for reference, see reaction 5 in the Example 2).

Furthermore, by feeding only a low amount of H_2 into the HT reaction 30 of the step iii), and the process conditions allowing, the absence of hydrogen will favour deoxygenation through DCO reactions. In an embodiment, the process conditions favouring DCO reactions at the HT reaction 30 of the step iii) comprise a pressure below 10 MPa, H_2 feed ratio below 300 nl H_2 / liter of feed, and a temperature of about 350 °C. In an embodiment, at least 30 % preferably at least 40 % more preferably at least 50 % of the deoxygenation reactions during the HT reaction 30 of the step iii) occur through DCO reactions. In an embodiment, the HT reaction comprises approximately 30 % of deoxygenation through DCO.

In an embodiment, the HT reaction 30 of the step iii) is carried out at a H_2 feed ratio of 50 - 2000 nl H_2 / liter of feed, preferably 100 - 1000 nl H_2 / liter of feed, more preferably 150 - 500 nl H_2 / liter of feed. In an embodiment, the HT reaction 30 of the step iii) is carried out at a H_2 feed ratio of at least 200 nl H_2 / liter of feed, at least 250 nl H_2 / liter of feed, at least 300 nl H_2 / liter of feed, at least 350 nl H_2 / liter of feed, at least 400 nl H_2 / liter of feed, at least 450 nl H_2 / liter of feed, or at least 500 nl H_2 / liter of feed. The required amount of H_2 at the HT reaction of the step iii) depends mainly on the deoxygenation conversion wt-% at the DCO reaction 20 at the step ii). For example, if the deoxygenation conversion wt-% at the DCO reaction is 50 wt-%, the amount of H_2 needed at the HT reaction is relatively higher than if the deoxygenation conversion wt-% at the DCO reaction is 90 wt-%, for complete removal of oxygen from the feed. In comparison to the HDO reactions, the other hydrotreatment reactions, such as HDN and the HDS reactions, do not consume significant amounts of H_2 due to the low concentration of S and N compounds in the feed, when compared to oxygenates in the renewable feedstocks. In an embodiment, the HT reaction 30 of the step iii) is carried out at a H_2 feed ratio of 300 – 500 nl H_2 / liter of feed, preferably 400 – 500 nl H_2 / liter of feed.

To ensure adequate amount of H_2 at the HT reaction 30 of the step iii), a 3 - 4 times excess amount of H_2 compared to the theoretical consumption is fed into the reaction. In an embodiment, the HT reaction 30 of the step iii) mainly comprises HDN and/or HDS and is carried out at a H_2 feed ratio of 50 - 500 nl H_2 / liter of feed, preferably 50 -200 nl H_2 / liter of feed.

For a HDO reaction, a H_2 feed ratio of at least 250 nl H_2 / liter of feed is preferred. Therefore, in an embodiment, the HT reaction 30 of the step iii) comprises at least HDO and is carried out at a H_2 feed ratio of 250 - 2000 nl H_2 / liter of feed, preferably 300 - 500 nl H_2 / liter of feed.

10 In an embodiment, the HT reaction 30 of the step ii) is carried out in the presence of at least 1.5 times or 2 times higher H_2 feed ratio than the H_2 feed ratio used in the DCO reaction 20 of the step ii) of the process. In an embodiment, the DCO reaction of the step ii) is carried out at a H_2 feed ratio of 50 - 150 nl H_2 / liter of feed, and the HT reaction of the step iii) is carried out at a H_2 feed ratio of 250 - 500 nl H_2 / liter of feed. In an embodiment, wherein at least 80 wt-% of oxygen comprised by the renewable feedstock is removed at the DCO reaction 20 of the step ii), the HT reaction 30 of the step iii) can be carried out at a H_2 feed ratio of 250 -500 nl H_2 / liter of feed.

In an embodiment, the HT reaction 30 of the step iii) is carried out at a temperature of 250 - 450 °C, preferably the temperature is 280 - 350 °C, and/or a pressure of 1 - 20 MPa, preferably the pressure is 2 - 10 MPa. In an embodiment, the HT reaction of the step iii) is carried out at a pressure of 2 - 20 MPa, or 4 - 10 MPa, or 4 - 5 MPa.

In an embodiment, the HT reaction 30 of the step iii) is carried out at a weight hourly space velocity (WHSV) of 0.25 - 5 h^{-1} , preferably the WHSV is 0.5 - 2 h^{-1} .

In an embodiment, the HT reaction 30 of the step iii) is carried out:

- 25 - in the presence of the HT catalyst comprising at least one group VIII and/or VIB metal of the Periodic Table, preferably selected from nickel, molybdenum, tungsten, cobalt, and any combinations thereof, preferably selected from NiMo, CoMo, and NiW; the HT catalyst further comprising at least one support selected from zeolite, silica, alumina, amorphous silica alumina (ASA), and any combination thereof;
- 30 - at a H_2 feed ratio of 50 - 2000 nl H_2 / liter of feed, preferably 100 - 1000 nl H_2 / liter of feed, more preferably 150 - 500 nl H_2 / liter of feed;
- at a temperature of 250 - 450 °C, preferably the temperature is 280 - 350 °C;
- at a pressure of 1 - 20 MPa, preferably the pressure is 2 - 10 MPa;

- at a weight hourly space velocity (WHSV) of $0.25 - 5 \text{ h}^{-1}$, preferably the WHSV is $0.5 - 2 \text{ h}^{-1}$;

or any combination thereof.

5 In an embodiment, 50-5000 wt-ppm, preferably 100 – 2000 wt-ppm of sulfur is added to the DCO effluent 21 directed to step iii). In an embodiment, 50-5000 wt-ppm, preferably 100 - 2000 wt-ppm of sulfur, is added to the carbon oxide deprived DCO effluent 22, i.e., the DCO effluent wherefrom carbon oxides are at least partly removed. Adding sulfur to the DCO effluent prior to the step iii) is beneficial for the function of the hydrotreatment catalyst.

10 In an embodiment, the hydrotreatment 30 of step iii) is preferably done in the presence of a sulfided catalyst. In an embodiment, the hydrotreatment catalyst is sulfided at the hydrotreatment catalyst bed prior to the hydrotreatment of the step iii) and prior to bringing at least a portion of the DCO effluent from step ii) in contact with the catalyst. In such embodiments, it is not necessary to add any additional sulfur to the DCO effluent subjected to the hydrotreatment.

15 Even though the reaction conditions of the DCO and HDO reactions may appear similar, at least the combination of used catalysts with the appropriate DCO or HDO reaction conditions are different from each other, the DCO reaction 20 fulfilling the required DCO deoxygenation selectivity and conversion criteria.

20 In an embodiment, a step iv) of the process comprises subjecting the hydrotreated effluent 31 from the step iii) to a gas-liquid separation 40 to obtain a degassed hydrotreated effluent 41.

25 In an embodiment, the gas-liquid separation 40 of step iv) comprises removal of at least gaseous sulfur and nitrogen from the hydrotreated effluent 31. In an embodiment, gaseous sulfur and nitrogen are removed as gaseous H_2S and NH_3 at the step iv). In an embodiment, the gas-liquid separation 40 of step iv) comprises removal of at least CO and/or CO_2 gases from the hydrotreated effluent. This would especially be the case, if the DCO effluent 21 from the step ii) is not subjected to a separation of CO/ CO_2 gases 25 before the step iii). In an embodiment, the gas-liquid separation 40 of step iv) comprises removal of also other gaseous components, selected from steam, hydrogen, methane and combinations thereof.

30 In an embodiment, the degassed hydrotreated effluent 41 comprises less than 30 wt-ppm, preferably less than 20 wt-ppm, more preferably less than 10 wt-ppm, even more preferably less than 5 wt-ppm, most preferably less than 2 wt-ppm of sulfur of the total respective effluent (ppm by weight, calculated as elemental S).

A very low sulfur content, such as less than 2 wt-ppm, in the degassed hydrotreated effluent 41 entering the hydroisomerisation (H-ISO) 50 of the step v) is beneficial in that only very low amounts or no H₂S is present in the H-ISO 50. This is beneficial especially in embodiments wherein the H-ISO of the step v) comprises a noble metal catalyst, as the function of noble metal catalysts is impaired in the presence of sulfur. Furthermore, since less H₂S is present, also less corrosion of the equipment is foreseen in the long-run, and potentially even less stringent corrosion resistance requirements could be applicable for some of the equipment materials.

Removal of nitrogen (N) from the feed prior to hydroisomerization 50 is important, as nitrogen is harmful for the isomerisation catalyst, the catalyst passivation and/or deactivation being accelerated in presence of increasing concentrations of nitrogen. Hence, a very low nitrogen content of the stream entering the H-ISO 50 of the step v) is beneficial in that passivation and/or deactivation of the H-ISO catalyst can be delayed and/or avoided.

In an embodiment, the degassed hydrotreated effluent 41 from the step iv) comprises 10 wt-ppm or less, preferably 8 wt-ppm or less, more preferably 5 wt-ppm or less, even more preferably 2 wt-ppm or less, most preferably 1 wt-ppm or less, such as even 0.5 wt-ppm of nitrogen, of the total weight of the degassed hydrotreated effluent 41, calculated as elemental N.

In an embodiment, the gas-liquid separation 40 of step iv) is conducted as an integral step within the reactor wherein the HT reaction 30 of the step iii) takes place, even if said reactor comprises also other catalyst beds besides a HT catalyst bed. In an embodiment, the gas-liquid separation 40 of step iv) is conducted between reactors.

In an embodiment, the process comprises:

- a. subjecting at least a portion of
 - the DCO effluent 21 from the step ii), or
 - the hydrotreated effluent 31 from the step iii), or
 - the degassed hydrotreated effluent 41 from the step iv), or
 - the hydroisomerised effluent 51 from the step v)to a fractionation (FRAC) 70, and recovering at least a first fraction 71 comprising hydrocarbons with a carbon number > C17 and a second fraction 72 comprising hydrocarbons with a carbon number ≤ C17;
- b. directing the hydrocarbons with a carbon number > C17 (71) to a hydrocracking (HC) 75 in the presence of a hydrocracking catalyst, thereby obtaining a hydrocracked effluent 76;
- c. combining the hydrocracked effluent 76 together with the second fraction 72; and

d. directing the combined hydrocracked effluent 76 and the second fraction 72 to step iii), iv), v) or vi), respectively.

The Fig. 2 shows an exemplary embodiment, wherein some of the possible alternative positions of the fractionation (FRAC) 70 and hydrocracking (HC) 75 steps are included in the current process sequence. In an embodiment, the process comprises a FRAC and HC steps 70,75 after the DCO reaction 20 of the step ii) and before the HT reaction 30 of the step iii), or after the HT reaction 30 of the step iii) and before the gas-liquid separation 40 of the step iv), or after the gas-liquid separation 40 of the step iv) and before the hydroisomerization (H-ISO) 50 of the step v). In an embodiment, the HC reaction 75 is conducted simultaneously with the H-ISO 50 of the step v). In an alternative embodiment, the process comprises a FRAC and HC steps 70, 75 after the H-ISO 50 of the step v). Preferably the process comprises a FRAC and HC steps 70, 75 before the H-ISO 50 of the step v).

In an embodiment, the hydrocracking 75 of the step b. is carried out in the presence of a bifunctional hydrocracking catalyst comprising a metal site and an acid site. In an embodiment, the bifunctional hydrocracking catalyst is selected from one or more of: platinum, palladium, nickel, molybdenum, cobalt, tungsten, or any combination thereof. In an embodiment, the bifunctional hydrocracking catalyst comprises a noble metal or a group VIA metal, such as molybdenum or tungsten, together with a group VIIIA metal, such as cobalt or nickel. In an embodiment, the hydrocracking catalyst is Ni/W, Ni/Mo, Co/Mo, Pt or Pd.

In an embodiment, the HC 75 of the step b. is carried out in the presence of an acidic hydrocracking catalyst support. Acidity of the hydrocracking catalyst support is important for the function of the HC 75 process. In an embodiment, the said acidic support is selected from one or more of alumina, amorphous silica alumina, zeolite and a binder. In an embodiment, the said acidic support is selected from one or more of SiO₂ and Al₂O₃.

Some renewable feedstocks may comprise a large relative portion of carboxylic acids with a carbon number > C18, such as rapeseed oil, Brassica Carinata, and some fish oils, in which case hydrocarbons suitable for aviation fuel component cannot be obtained through DCO reactions only. Therefore, by including the FRAC and HC steps 70, 75 to the process after the DCO reaction 20, also the deoxygenated hydrocarbons with a carbon number ≥ C18 (i.e., a first fraction 71) can be utilized in the aviation fuel fraction, whereas a lighter, second fraction 72 comprising deoxygenated ≤ C17 hydrocarbons does not need to be unnecessarily reduced and exposed to hydrocracking (Fig. 3).

In an exemplary embodiment of Figure 3, is provided a more detailed exemplary embodiment, wherein the fractionation (FRAC) step 70 is performed after the gas-liquid separation 40 of the step iv). The first fraction 71 comprising hydrocarbons with a carbon number $> C17$ and a second fraction 72 comprising hydrocarbons with a carbon number $\leq C17$ are obtained, but only the first fraction 71 is exposed to the HC step 75, thereby obtaining a hydrocracked effluent 76. The second fraction 72 must not be exposed to hydrocracking and thereby unnecessary yield losses.

For example, rapeseed oil fatty acids (95% C18FA and 5% C16FA) may be introduced to the pretreatment step 80 where impurities of feedstock are removed. The removed impurities can comprise alkali metals, such as sodium, potassium, and/or alkaline earth metals such as magnesium and calcium. Pretreatment step also removes impurities such as sulfur-, phosphorus-, silicon-, and chloride containing compounds and/or polyethene wax. The pretreatment step 80 can comprise filtration, degumming, heat treatment, solvent extraction, distillation/evaporation and/or bleaching procedures.

A pretreated effluent 85, for example purified rapeseed oil fatty acids (FFA's and TRIGs), can be directed to the DCO reaction 20 of the step ii), where from removal of oxygen is done without added hydrogen or using a minimal amount of hydrogen. The selected DCO catalyst can be, for example, a Pd/C and the DCO reaction 20 is carried out at reaction conditions comprising a temperature of about 350 °C, a pressure of about 5 bar, a weight hourly space velocity (WHSV) of about 0.25 h⁻¹, and a H₂ feed ratio of about 100 nl H₂/ liter of the feedstock. Using said catalyst and reaction conditions, DCO reactions 20 (deoxygenation via CO₂/CO removal, resulting in production of odd numbered n-paraffins) are favoured over hydrodeoxygenation (HDO, deoxygenation via hydrogen-consuming H₂O removal, resulting in production of even numbered n-paraffins). In such DCO reactions 20 also, the deoxygenation conversion of fatty acids is about 85 wt-% of the total weight of the pretreated effluent 85 and the DCO deoxygenation selectivity to C15 and C17 n-paraffins is about 95 wt-% of the total weight of deoxygenated hydrocarbons. A complete deoxygenation conversion (i.e., 100 wt-%) of the feedstock is preferably avoided during the DCO reaction 20 step ii), as this allows a higher DCO deoxygenation selectivity into DCO products. Most of the double bonds present in the fatty acids of the renewable feedstock can also be hydrogenated during the DCO reaction 20 with the indicated low H₂ feed ratio, hydrogenation of double bonds being an exothermic reaction produces thermal energy which may be utilized in the endothermic DCO reaction 20. Moreover, as the endothermic DCO reaction 20 takes up thermal energy, the exothermic hydrogenation reaction of double bonds remains well controlled. The DCO reaction 20 of the step ii) not only decreases the

hydrogen consumption of the entire process, but also converts C18 fatty acids into hydrocarbons which fit within the boiling range set for aviation fuel components.

Carbon oxides are removed from the DCO effluent 21 at the gas-liquid separation step 25, after which the formed carbon oxide deprived DCO effluent 22 can be introduced into the
5 hydrotreatment (HT) reaction 30 of the step iii) wherein all the remaining oxygenates can be deoxygenated. The hydrotreatment reaction 30, consisting optionally mainly of hydrodeoxygenation (HDO) reactions, can be carried out at a pressure of about 40-50 bar, a temperature of about 300-330 °C, a weight hourly space velocity (WHSV) of about 1 h⁻¹, and a H₂ feed ratio of about 250-300 nl H₂/ liter of the feed. Typically, also some amount of
10 DCO reactions occur during the HT reaction 30. It is possible to select conditions favouring DCO reactions during the HT reaction 30 of the step iii), as opposed to the HT reaction 30 step iii) consisting mostly of HDO reactions. This leads to even lower hydrogen consumption of the entire process and increased C17 paraffin content of the hydrotreated effluent 31, thereby increasing the potential aviation fuel component yield. It is also important to convert
15 and remove the nitrogen and sulfur containing compounds in the feed at the step iii) and iv), in order to prevent a hydrocracking catalyst, and possibly also the H-ISO catalyst, neutralisation/deactivation, thereby allowing an efficient catalyst performance.

As indicated in Fig. 3, the hydrotreated effluent 31 can then be fractionated at a fractionation (FRAC) step 70. Prior to fractionation, gasses are separated from the hydrotreated effluent
20 31 with a gas/liquid separation 40 of the step iv), thereby obtaining a the degassed hydrotreated effluent 41, which can be introduced into fractionation equipment which separates the effluent 41 into at least two fractions, namely the first fraction 71 comprising normal paraffins with a carbon number 18 and higher, and the second fraction 72, comprising normal paraffins with a carbon number 17 and lower. Cracking is not needed for
25 the second fraction 72, as it already has a renewable aviation fuel component boiling range. Therefore, by not exposing the second fraction 72 to a cracking reaction, the aviation fuel component yield losses are minimised. The first fraction 71, comprising heavier than C17 paraffins (C18 and heavier), can be introduced into the hydrocracking (HC) 75, wherein mild hydrocracking of heavier paraffinic feed is done with a bifunctional HC catalyst. Bifunctional
30 HC catalyst contains a metal, such as Pd, Pt, Ni, W, Mo, and an acidic function, such as an amorphous silica alumina (ASA), zeolite, alumina/chlorided alumina. The HC reaction of paraffins 75 can be carried out at mild HC reaction conditions, such as at a pressure of about 35-70 bar, a temperature of about 325-375 °C, a WHSV of about 0.5-3.0 h⁻¹, and a H₂ feed ratio of about 250-500 nl H₂/ liter of the feed. Hydrocracking conversion can be set
35 to about 95 wt-% of the total weight of the feed to be hydrocracked. Hydrocracking selectivity

to aviation fuel boiling range is set to about 75wt-% of the total weight of converted hydrocracked effluent. Moreover, the converted hydrocracked effluent can comprise about 7 wt-% of naphtha, and about 18 wt-% of Liquefied petroleum gas (LPG, containing C3 and C4 hydrocarbons) (of which about 2 wt-% are light gases, such as C1, C2 hydrocarbons),
5 from the total weight of the converted hydrocracked effluent.

Prior to the hydroisomerization (H-ISO) step 50, the hydrocracked effluent 76 is exposed to separation of gases from the hydrocracked effluent 76 with a gas-liquid separation 90, thereby obtaining the degassed liquid effluent 91 (e.g., Fig. 3). The degassed liquid effluent 91 can be combined with the lighter second fraction 72 (which bypasses the hydrocracking
10 reaction 75) and introduced together into the hydroisomerisation (H-ISO) 50. Hydroisomerisation 50 can be carried out at a pressure of about 30-50 bar, a temperature of about 320-340 °C, a WHSV of about 0.5-3.0 and H₂ feed ratio of about 250-500 nl H₂/ liter of the feed. In the H-ISO 50 reaction up to 99 wt-% of the n-paraffins present in the liquid effluent 91 are converted to isoparaffins, whereas cracking to naphtha and LPG
15 fraction may take place in only up to 5 wt-% of the paraffins in the liquid effluent 91. Isomerised C17 paraffins have a boiling point of less than 300 °C whereas n-C18-paraffins boil at 317 °C and n-C17 at 302 °C. Moreover, n-C18 paraffins have a higher melting point of 28 °C compared to n-C17 paraffins (22 °C), indicating isomerisation of n-C17 paraffins, comprised by the degassed liquid effluent 91, is to some extent easier than isomerisation
20 of n-C18 paraffins.

In an embodiment, reaction conditions of the HC 75 of the step b. are selected from one or more of: a temperature of 300 - 450 °C, preferably 350 - 400 °C; a pressure of 5 MPa - 25 MPa, preferably 6 MPa – 15 MPa; a H₂ feed ratio of 600 - 1500 nl H₂/ liter of feed, preferably 500 - 850 nl H₂/ liter of feed; and a WHSV of 0.25 – 5 h⁻¹, preferably 0.5 - 2 h⁻¹.
25 In an embodiment, the HC 75 of the step b. is a mild hydrocracking (MHC), resulting in MHC conversion-% of the hydrocracked effluent 76 of less than 50 wt-%, preferably less than 30 wt-%, of the total weight of the first fraction 71.

The portion of C17 hydrocarbons in the aviation fuel component is limited by the boiling range set for the aviation fuel component by the ASTM D7566-22 standard. Therefore, in
30 some embodiments, the carbon number of also at least a portion of the C17 hydrocarbons comprised in the deoxygenated second fraction 72 (comprising hydrocarbons with a carbon number ≤ C17) is further reduced by hydrocracking.

For the hydrotreated effluent to fulfil the requirements of the aviation fuel fraction, the freezing point must be adjusted. The freezing point is adjusted by isomerisation of n-

paraffins to i-paraffins, by directing the effluent 41 from step iv), or the effluent 91 from the gas-liquid separation 90 (according to Fig. 3) to contact a material catalytically active in hydroisomerisation.

5 In an embodiment, the step v) of the process comprises subjecting at least a portion of the degassed hydrotreated effluent 41 from step iv) to hydroisomerisation (H-ISO) 50, thereby obtaining a hydroisomerised effluent 51. In an embodiment, the step v) of the process comprises subjecting at least a portion of the degassed liquid effluent 91 to hydroisomerisation (H-ISO) 50, thereby obtaining a hydroisomerised effluent 51.

10 In an embodiment, the H-ISO 50 can be carried out in a conventional hydroisomerisation unit. In an embodiment, the H-ISO 50 of the step v) comprises a H_2 flow of 100-800 nl H_2 / liter of feed, preferably the H_2 flow is 200 - 650 nl H_2 / liter of feed. In an embodiment, the H-ISO 50 of the step v) comprises a WHSV of 0.5 - 3 h^{-1} , preferably the WHSV is 0,5 - 2 h^{-1} .

15 In an embodiment, in the H-ISO 50 of the step v), hydroisomerisation catalysts known in the art may be used. In an embodiment, the hydroisomerisation (H-ISO) 50 of the step v) is done in the presence of a hydroisomerisation catalyst comprising:

- at least one metal selected from Group VIII of the Periodic Table, preferably from nickel, platinum and palladium, more preferably from platinum and palladium; and/or
 - a carrier, preferably selected from Al_2O_3 or SiO_2 , and/or
 - a molecular sieve, preferably selected from SAPO-11, SAPO-41, ZSM-22, ZSM-23 and
- 20 ferrierite.

Bifunctional hydrocracking catalysts and hydroisomerisation catalysts have similarities in the sense that both contain metal sites that are capable of catalysing (de)hydrogenation of n/i-paraffins to corresponding n/i-olefins, and acid sites that are capable of catalysing protonation of the n/i-olefins to n/i-carbocations, isomerisation of n-carbocations, or i-carbocations further, and/or cracking of n/i-carbocations into lighter n/i-olefin and lighter n/i-carbocation, and deprotonation of n/i-carbocations to n/i-olefins, and hydrogenation of the various n/i-olefins is catalysed again by the metal sites of these bifunctional catalysts to form n/i-paraffins. Whether hydroisomerisation or cracking reactions prevail at given operating conditions and given feed composition at the H-ISO 50 of the step v), can be

25 influenced especially by the characteristics of the hydroisomerisation catalyst. Such characteristics of the catalyst include, for example, total acidity of the catalyst, number of Brønsted acid sites, strength and/or density of the acid sites, and content of the metal(s) in the catalyst.

30

In an embodiment, H-ISO 50 of the step v) is done in the presence of a hydroisomerisation catalyst comprising at least one noble metal, preferably platinum and/or palladium. A hydroisomerisation catalyst comprising at least one noble metal is beneficial, as it may provide higher selectivity towards isomerisation reactions, and is highly active at lower
5 operating temperatures, compared to catalysts comprising only non-noble metals.

In an embodiment, the hydroisomerisation catalyst is supported on a catalyst support comprising one or more porous acidic materials having microporous, mesoporous or hierarchical (micro-mesoporous) structure functioning as a molecular sieve. Various SAPOs and zeolites are available providing desired acidity and porosity characteristics. The
10 mentioned SAPOs and zeolites are commercially available with acidity and porosity characteristics that allow hydroisomerisation, including multiple-branching of n-paraffins, even of long-chained n-paraffins, such as C16+ paraffins.

In an embodiment, the hydroisomerisation catalyst of the step v) is a non-sulfided catalyst. In an embodiment, the hydroisomerisation catalyst is selected from Pt/SAPO-11/Al₂O₃,
15 Pt/ZSM-22/Al₂O₃, Pt/ZSM-23/Al₂O₃ and Pt/SAPO-11/SiO₂. In an embodiment, the hydroisomerisation catalyst requires the presence of hydrogen to maintain the catalyst stability.

In an embodiment, the H-ISO 50 of the step v) is conducted at: a hydroisomerisation temperature of 200 - 500 °C, preferably 230 - 500 °C, more preferably 250 - 450 °C, even
20 more preferably 280 - 400 °C; and a pressure of 2 - 15 MPa, preferably 1 - 10 MPa, more preferably 3 - 10 MPa.

In an embodiment, the renewable feedstock 10 is subjected to a pretreatment 80 prior to the step ii), wherein the pretreatment comprises:

- 25 a. removal of impurities, such as alkaline- and non-alkaline earth metals, and/or phosphorus from the renewable feedstock;
- b. hydrolysis of the renewable feedstock;
- c. fractionation of the renewable feedstock;
- d. hydrogenation of the renewable feedstock, or
- e. any combination thereof.

30 In an embodiment, the renewable feedstock 10 is subjected to a pretreatment 80 prior to the step ii), wherein the pretreatment comprises:

- a. removal of impurities, such as metals and/or sulfur and/or phosphorus, from the renewable feedstock;
- b. hydrolysis of the renewable feedstock;

- c. fractionation of the renewable feedstock;
- d. hydrogenation of the renewable feedstock, or
- e. any combination thereof.

For example, in the Figures 2 and 3, the renewable feedstock 10 is exposed to pretreatment 80 before the DCO reaction 20 of the step ii), wherein the pretreated effluent 85 is then directed to the DCO reaction 20.

In an embodiment, the pretreatment 80 comprises removal of elemental metals and/or metal compounds. In an embodiment, the pretreatment 80 comprises removal of alkali metals and/or alkaline earth metals. In an embodiment, the pretreatment 80 comprises removal of non-alkaline earth metals.

In an embodiment, the pretreated renewable effluent 85 comprises one or more of:

- less than 10 wt-ppm, preferably less than 5 wt-ppm, more preferably less than 1 wt-ppm of alkaline and alkaline earth metals, calculated as elemental alkaline and alkaline earth metals;
- less than 10 wt-ppm, preferably less than 5 wt-ppm, more preferably less than 1 wt-ppm of other than alkaline and alkaline earth metals, calculated as elemental metals; and/or
- less than 30 wt-ppm, preferably less than 15 wt-ppm, more preferably less than 5 wt-ppm of phosphorus, calculated as elemental phosphorus, of the total weight of the pretreated effluent 85.

In an embodiment, the pretreated renewable effluent 85 comprises one or more of:

- less than 10 wt-ppm, preferably less than 5 wt-ppm, more preferably less than 1 wt-ppm of alkali metals and/or alkaline earth metals, calculated as elemental metals;
- less than 10 wt-ppm, preferably less than 5 wt-ppm, more preferably less than 1 wt-ppm of other metals than alkali metals and alkaline earth metals, calculated as elemental metals; and/or
- less than 30 wt-ppm, preferably less than 15 wt-ppm, more preferably less than 5 wt-ppm of phosphorus, calculated as elemental phosphorus, of the total weight of the pretreated effluent 85.

In an embodiment, the pretreated effluent 85 comprises less than 1 wt-ppm of metals or metal-containing compounds. In an embodiment, the pretreated effluent 85 comprises less than 1 wt-ppm of alkali metals and alkaline earth metals, and other than alkali metals and alkaline earth metals, and less than 5 wt-ppm of phosphorus. In an embodiment, the pretreatment 80 of the renewable feedstock comprises removal of sulfur from the feed, the

pretreated effluent 85 comprising 50 wt-ppm or less, preferably 10 wt-ppm or less, more preferably 5 wt-ppm or less of sulfur.

In an embodiment, the pretreatment 80 comprises:

- a. subjecting the renewable feedstock to a hydrolysis; and
- 5 b. subjecting at least a portion of the hydrolysed effluent from the step a) to a fractionation and recovering a first fraction 82 comprising free fatty acids (FFAs) with a carbon number $> C17$, and a second fraction 83 comprising FFAs with a carbon number $C17$ or less;

wherein the FFAs with a carbon number $> C17$ (82) are directed to the DCO reaction 20 of the step ii) and the FFAs with a carbon number $C17$ or less 83 are directed to the hydrotreatment reaction 30 of the step iii).

Fig. 4 shows an exemplary embodiment of the process comprising the pretreatment steps a. (hydrolysis) and b. (fractionation). In an embodiment the hydrolysis and fractionation of the renewable feedstock to said first 82 and second 83 fractions is advantageous in that the carbon number of only the free fatty acids with a carbon number $> C17$ is reduced in the DCO reaction 20 (whereas the shorter $\leq C17$ FFAs 83 are not reduced). In an embodiment the hydrolysis and fractionation of the renewable feedstock to said first 82 and second 83 fractions is advantageous in that the pretreatment 80 avoids unnecessary reduction of the carbon number of the carboxylic acids already within the aviation fuel fraction limits, thereby increasing the carbon efficiency of the initial renewable feedstock 10, and thereby the final yield of the at least the renewable aviation fuel fraction.

In an embodiment, wherein the process comprises a pretreatment 80 comprising hydrolysis and fractionation of the feedstock to the first 82 and second 83 fractions, the HT reaction 30 of the step iii) comprises HDO. This is because all the FFAs comprised in the first fraction 82 (comprising FFAs with a carbon number $C17$ or less) bypass the DCO reaction 20 step and are only deoxygenated in the HDO reaction 30 of the step iii).

In an embodiment, wherein the process comprises a pretreatment 80 comprising fractionation of the feedstock to the first 82 and second 83 fractions, the feed entering the HT reaction 30 of the step iii) comprises preferably only compounds with a carbon number 17 or less.

In an embodiment, the pretreatment 80 of the renewable feedstock 10 comprises hydrogenation of the feedstock. In an embodiment the pretreatment hydrogenation of the renewable feedstock is advantageous in removing double bonds from the free carboxylic acids, esters of carboxylic acids, and/or triglycerides of the renewable feedstock.

As indicated in the Fig. 5, the different effluents obtained from the process steps can be recycled back to upstream process steps.

In an embodiment, the process comprises:

- a. recycling at least a portion of the DCO effluent 21 back to the DCO reaction 20 of the step ii),
- b. recycling at least a portion of the hydrotreated effluent 31 back to the HT reaction 30 of the step iii),
- c. recycling at least a portion of the hydrotreated effluent 31 back to the DCO reaction 20 of the step ii); or
- d. any combination thereof.

In an exemplary embodiment of Figure 5, is presented various alternative and/or parallel routes to recycle the feed during the process.

In an embodiment, the process comprises recycling at least a portion of pretreated effluent 85 back to the pretreatment 80 (Fig. 5). Recycling the pretreated effluent 85 back to the pretreatment 80 is beneficial, as this ensures efficient removal of impurities from the feedstock. Recycling at least a portion of the pretreated effluent back to the pretreatment 80 is also beneficial when the renewable feedstock 10 comprises high amount of olefinic bonds. Olefinic bonds can dimerise/oligomerize during the acid treatment procedure of pretreatment 80 step, forming heavier compounds not suitable for aviation fuels or diesel fuels.

In an embodiment, the DCO effluent 21 or the hydrotreated effluent 31 can be recycled back into the feed entering the pretreatment 80 (not shown in the Fig. 5). This is beneficial, as the DCO effluent 21 and/or the hydrotreated effluent 31, when recycled into the feed entering the pretreatment 80, can for example, lower the viscosity of the renewable feedstock 10, thereby facilitating mixture of chemicals and filtration during bleaching, thereby improving the efficiency of the pretreatment purification.

In an embodiment, the DCO effluent 21 and/or the hydrotreated effluent 31 can be recycled back into the DCO reaction 20, optionally to the pretreated effluent 85 (Fig. 5). In an embodiment, recycling at least a portion of the DCO effluent 21 back into the pretreated effluent 85 and thereby to the DCO reaction 20 of the step ii), improves the DCO conversion efficiency by deoxygenating any remaining oxygen containing compounds from the DCO effluent, thereby increasing the conversion of the unconverted oxygen components.

In an embodiment, recycling the DCO effluent 21 and/or the hydrotreated effluent 31 back to the endothermic DCO reaction 20 delivers additional thermal energy to the DCO reaction

20 of the step ii). Dilution of reactive feed entering the DCO reaction 20 with an inert (already converted) feed controls also the DCO reaction temperature, making it more stable. In an embodiment, recycling at least a portion of the hydrotreated effluent 31 back to the DCO reaction 20 of the step ii) is beneficial, as the hydrotreated effluent also dilutes the concentration of impurities, such as nitrogen and sulfur, present in the renewable feedstock 10.

In an embodiment, the hydrotreated effluent 31 can be recycled into the DCO effluent 21 or the carbon oxide deprived DCO effluent 22 (Fig. 5). The hydrotreatment reaction 30 is very exothermic and therefore recycling the hydrotreated effluent 31 back into the stream 21 or 22 stabilises the reaction temperature of the HT reaction 30. The n-paraffins produced during HT step iii) are practically inert at HT conditions. The stabilisation of the HT reaction temperature and dilution of the DCO effluent 21 or 22 (HT feed) also prevents side reactions such as oligomerisation of double bonds and ketonisation reaction of free fatty acids. The reaction temperature control during the HT step iii) is however easier due to the deoxygenation and at least partial hydrogenation of double bonds already carried out in the DCO step ii), decreasing the reactions needed during the HT step iii), and therefore also the exothermal heat formed. Recycling at least a portion of the hydrotreated effluent 31 back to the HT reaction 30 of the step iii) is also beneficial wherein the first HT reaction 30 has been inadequate, for example when the level of heteroatoms in the hydrotreated effluent 31 is too high.

In an embodiment, the step vi) of the process comprises subjecting at least a portion of the hydroisomerised effluent 51 from step v) to a fractionation 60 and recovering at least the renewable aviation fuel component.

In an embodiment, the renewable aviation fuel component fulfils the specifications set out in the ASTM D7566-22 Annex A2. In an embodiment, the renewable aviation fuel component obtained from the process comprises C8-C17 alkanes.

In an embodiment, the present process enables recovering from the fractionation 60 of the step iv) an increased yield of the renewable aviation fuel component, compared to processes without the DCO reaction 20 of the step ii), said increase in yield depending on the fatty acid distribution of the renewable feedstock 10. In some embodiments, the increased overall yield of the renewable aviation fuel component is contemplated to be due to the high content of C18 fatty acids in the renewable feedstock subjected to DCO reaction 20 of the step ii).

The present process for producing a renewable aviation fuel component provides an increased yield of the renewable aviation fuel component, without the need for cracking or comprising only mild hydrocracking conditions. Without any harsh cracking conditions, and thereby unpredictable cracking of the hydrocarbons, a larger relative portion of the initial
5 liquid renewable feedstock can be acquired as the final renewable components, including the renewable aviation fuel component.

In an embodiment, the renewable aviation fuel component is mixed with a fossil aviation fuel component, to be used as an aviation fuel blend.

In an embodiment, the renewable aviation fuel component obtained from the process
10 according to the disclosure is used as a renewable aviation fuel blend component in an aviation fuel blend, wherein the aviation fuel blend also comprises a fossil aviation fuel blend component.

In an embodiment, the renewable aviation fuel component of the present disclosure may further be used as such i.e. 100% as an aviation fuel product depending on the prevailing
15 specification requirements at the given time or on e.g. the aviation original equipment manufacturers (OEMs) needs.

In an embodiment, the fractionation 60 of the step vi) further comprises recovering a renewable diesel component, and/or a renewable naphtha component.

In an embodiment, the present process for producing a renewable aviation fuel component
20 provides similarly a method to produce the renewable diesel component and/or the renewable naphtha component.

In an embodiment, the fractionation 60 of the step vi) comprises recovering at least one further component comprising hydrocarbons having a carbon number of C18 or more, and recycling said at least one further component back to the DCO reaction 20 of the step ii), to
25 the HC step 75, or any combination thereof, as indicated in the Fig. 6.

In an embodiment, the at least one further component comprising hydrocarbons having a carbon number of C18 or more is recycled from the fractionation 60 of the step vi) back to the DCO reaction 20 of the step ii), to be used as a solvent bringing thermal energy to the DCO reaction. Since all the hydrocarbons are already deoxygenated at the step vi) of the
30 process, no further carbons could be removed from the hydrocarbons of the at least one further component through the DCO reaction of the step ii). However, the $\geq$ C18 hydrocarbons may be shortened in a HC step 75 following the DCO step later in the process. According to some embodiments, the at least one further component is recycled from the

step vi) back to the HC step 75, to further reduce the carbon number of the hydrocarbons present in the at least one further component.

EXAMPLES

5 **Example 1. Decarboxylation/Decarbonylation (DCO) of a renewable feedstock**

Background and equipment

Test runs are carried out in the conventional catalytic testing unit, which has a feeding tank on a scale, feeding pump, catalytic fixed bed reactor and pressure control unit. All the reactions are carried out in continuous mode of reaction. Products are collected into the product tank, which works also as gas/liquid separation. Gas volume is measured with measuring equipment and samples from the gas streams are taken and analysed separately. Hydrogen consumption is measured from GC results, and hydrocarbon product (liquid) and water are separated manually and analysed with GC equipment. GC analysis of hydrocarbons are well known methods.

15 *Renewable feedstock*

The renewable liquid feedstock used as a renewable feedstock in all DCO reactions according to the current disclosure, is alkali raffinated rapeseed oil (although other feedstock having suitable fatty acid distribution could be used just as well). The feedstock comprises approximately 95 wt-% of C18 fatty acids, and 5 wt-% of C16 fatty acids, from the total weight of the fatty acids of the feedstock. Therefore, said rapeseed oil comprises a considerable portion of C18 fatty acids which can be utilized in the renewable aviation fuel component prepared according to the current process. The initial rapeseed oil feedstock comprises approximately 5 wt-ppm of sulfur and 10 wt-ppm of nitrogen.

Pretreatment of the feedstock

25 Rapeseed oil fatty acids (95% C18FA and 5% C16FA) are introduced to the pretreatment step where impurities of the renewable feedstock are removed. The renewable feedstock is first filtered, thereafter heat treated at 280 °C, and finally bleached with conventional bleaching procedures. Impurities removed comprise alkali metals such as sodium and potassium, and/or alkaline earth metals such as magnesium and calcium. This pretreatment step also removes most of the phosphorus, silicon, chloride components and polyethylene wax.

DCO catalyst

The catalyst used in the DCO reaction is either palladium supported on carbon (Pd/C)

(reactions 1-3), or sulfided NiMo supported on Alumina (sulfided NiMo/ Al_2O_3) (reaction 4). In reaction 4, sulfur is added in the DCO feedstock at a concentration 1000 wt-ppm. NiMo catalyst is sulfided prior to the DCO reaction for it to contain about 6-7 wt-% of sulfur, calculated from the total weight of the catalyst. In all reactions 1-4, 10-25 g of the catalyst is charged into the reactor, followed by its reduction under hydrogen flow at 200 °C (Pd catalyst) or the sulfidation (NiMo catalyst).

However, the DCO catalyst used in the DCO reaction can be different from the above disclosed Pd and NiMo catalysts. For example, patent EP168133781B1 presents in the Table 1 other possible DCO catalysts, which can also be used in the DCO reaction of the present process. The catalysts presented in the Table 1 of EP168133781B1 are not optimized for the present DCO process and the table 1 therefore does not present the optimized %-values for DCO conversion and selectivity. Nevertheless, other catalysts can be used in the present process as well, and the DCO reaction and the catalysts as presented in the Example 1 of the EP168133781B1 are hereby incorporated by reference.

DCO reaction

The DCO reactions are conducted in parallel reactions 1-4. The reactions 1-3 observe the effect of altering the Weight Hourly Space velocity values on the resulting DCO product distribution using Pd/C-catalyst. Hence, the reaction conditions for the reactions 1-3 are otherwise the same, except the WHSV of the feedstock is adjusted to different levels in each reaction 1-3, in reaction 1 to 0.25 h^{-1} , in reaction 2 to 0.5 h^{-1} , and in reaction 3 to 1 h^{-1} , respectively. The hydrogen (H_2) /oil feed ratio is set to 150 nl H_2 / liter of feed in each parallel DCO reactions 1-3, and the DCO reactions 1-3 are carried out at a temperature of 350°C at a reactor pressure of 0.5 MPa. The reaction 4 is catalysed by sulfided NiMo/ Al_2O_3 catalyst, and the reaction conditions comprise hydrogen (H_2) feed ratio of 150 nl H_2 / liter of feed, temperature of 350°C , a reactor pressure 0.5 MPa, and a WHSV of 1 h^{-1} . Hydrogen is added to the feed in all reactions 1-4 in relatively low concentration, as it is only needed for hydrogenation of double-bonds present in the carboxylic acids of the feedstock.

Results

The gaseous product comprised by the DCO effluent after the DCO reaction includes mainly carbon oxides (CO/CO_2) and light hydrocarbons having carbon number $< \text{C}_4$. The gaseous fraction is removed from the liquid fraction through gas-liquid separation after the DCO reaction. The liquid product distribution from the DCO reactions 1-4 is shown in Table 1. The table 1 shows the individual components comprised by the liquid fraction of the DCO reaction product, i.e., liquid DCO effluent. All the reactions 1-4 result in a high wt-% of deoxygenated C_{17} and C_{15} hydrocarbons (HCs). As the amount of odd numbered

carboxylic acids in natural feedstocks is normally very low or totally absent, the C17 and C15 HCs present in the liquid DCO effluent mostly result from deoxygenation of C18 and C16 carboxylic acids, the resulting HCs having one carbon atom less than the original carboxylic acids in the starting feedstock. From the reactions 1-4, the highest wt-% of C17 HCs is comprised in the DCO effluent of the reaction 1 (74,93 wt-%). The DCO effluent of each reaction 1-3 comprises a significant portion of deoxygenated C17 HCs, which can be utilized for the aviation fuel component (as opposed to C18 HCs). Therefore, the wt-% of C17 hydrocarbons directly reflects the potential yield increase of the intermediate aviation fuel component. Not all the deoxygenation reactions of the feedstock occur through the DCO reactions, which is demonstrated by the small fractions of C18 and C16 hydrocarbons present in the DCO effluents. This is because hydrogen is present in all reactions at least in small amounts and the used catalysts also catalyse HDO reactions. However, most of the deoxygenation is carried out through DCO reactions, which can be seen when comparing the product distributions of reactions 1-4 with product distribution of sample wherein deoxygenation has taken place through HDO only (see Table 5, reaction 5). Low concentration of hydrogen is however beneficial due to the double bonds present in the fatty acids of the renewable feed. Hydrogenation of double bonds can be efficiently carried out with noble metal DCO catalysts, and removal of double bonds takes place more readily than HDO reactions during the DCO reaction.

20

Table 1. The liquid product distribution from the DCO reactions 1-4, presented as the wt-% from the total weight of the liquid DCO reaction products.

	Reaction 1	Reaction 2	Reaction 3	Reaction 4
C18FA	14.22	33.18	52.00	47.27
C16FA	0.75	1.75	2.74	2.49
C18-HC	5.64	3.70	2.13	11.82
C17-HC	74.93	57.92	40.42	35.45
C16 HC	0.30	0.19	0.11	0.62
C15 HC	3.94	3.05	2.40	2.14
<C15 HC	0.10	0.10	0.10	0.10
>C18 HC/FA	0.12	0.12	0.11	0.11

The total sulfur and nitrogen content in the resulting DCO effluents from the reactions 1-3 is only slightly lower than in the initial feedstock, as the used DCO catalysts are not designed to catalyse HDS/HDN. Therefore, a HT reaction step after the DCO reaction step is required in order to ensure efficient removal of nitrogen and sulfur components prior to hydrocracking or hydroisomerisation steps with bifunctional catalysts (including metal active sites and

25

acidic active sites). In the reaction 4 also, which is conducted with a sulfided feedstock, the sulfur is more efficiently removed, but due to the conditions favourable for DCO, HDN/HDS is not complete and HT step is also needed (Table 2).

- 5 **Table 2.** The concentration of sulfur (S) and nitrogen (N) in the DCO effluent, presented as wt-ppm.

DCO product (wt-ppm)	Reaction 1	Reaction 2	Reaction 3	Reaction 4
N	3	4	5	3
S	8	9	10	5

- The DCO reactions 1-4 do not deoxygenate the initial feedstocks completely (i.e., 100%), which is demonstrated by the presence of C18 and C16 fatty acids (FAs) in the liquid DCO effluent (Table 1). From the reactions 1-4, the highest deoxygenation conversion rate is in the reaction 1, wherein 85 wt-% of the C18 carboxylic acids and C16 carboxylic acids in the initial DCO feedstock have undergone removal of oxygen. For the reaction 2 the conversion rate is 65 wt-% for both C16 and C18, for the reaction 3 the rate is the lowest, 45 wt-% for both C16 and C18, and for the reaction 4 the rate is 50 wt-% for both C16 and C18.
- 15 Therefore, it can be concluded from the deoxygenation conversion rates, that the reaction 1 with the lowest WHSV of 0,25 h⁻¹ (longer residence time) yields the highest deoxygenation conversion in combination with Pd/C catalyst, whereas the reactions 2-3 with higher WHSV values with the same catalyst, or the reaction 4 with WHSV of 1 h⁻¹ in combination with sulfided NiMo catalyst have much lower deoxygenation conversion rates.
- 20 The deoxygenation selectivity rates of the reactions 1-3 correlate conversely with the respective conversion rates, indicating the DCO deoxygenation selectivity decreases when the deoxygenation conversion approaches 100 wt-%. To keep the DCO deoxygenation selectivity high, the deoxygenation conversion is preferably set below 100 wt-%. Therefore, the HT reaction 30 step is necessary after the DCO reaction 20 step, to complete the deoxygenation and to remove sulfur and nitrogen from the feed. From the reactions 1-4, the highest DCO deoxygenation selectivity is in the reaction 3, wherein DCO selectivity of 95 wt-% is obtained for both C17 and C15 hydrocarbons, which have undergone deoxygenation reaction through the DCO reaction, from the total amount of deoxygenated C18 and C17, or C16 and C15 hydrocarbons in the DCO effluent. For the reaction 2 the selectivities are 94 wt-% for both C17 and C15 HCs, and for the reaction 1 the selectivities are 93 wt-% for both C17 and C15 HCs. Therefore, it can be concluded that by increasing
- 25
- 30

the residence time (lowering WHSV) of the feedstock, the DCO deoxygenation selectivity of the reaction decreases, although not at the same rate as the total deoxygenation conversion rate increases. For the reaction 4 the selectivities are 75 wt-% for both C17 and C15 HCs, from the total amount of deoxygenated C18 and C17, or C16 and C15 hydrocarbons in the DCO effluent.

When considering the sum of hydrocarbons with an odd carbon number in the DCO effluent in the Table 1, as well as total deoxygenation rates of the DCO effluents obtained from the reactions 1-4 presented in the Table 3, it can be concluded, that majority of the deoxygenation reactions take place through DCO reactions.

Table 3. The total deoxygenation rates of the DCO effluents, calculated with a formula “wt-% of deoxygenated DCO effluent = 1- (sum of non-deoxygenated FAs in the DCO effluent)”.

Deoxygenation (wt-%)	Reaction 1	Reaction 2	Reaction 3	Reaction 4
wt-% of deoxygenated DCO effluent	87.5	69.9	51.2	56.5

The total deoxygenation rates of Table 3 and DCO conversion rates correlate with the amount of carbon oxides in the gaseous product comprised by the DCO effluent after the DCO reaction, in the reactions 1 – 3, whereas in the reaction 4 the deoxygenated DCO effluent comprises proportionally less carbon oxides due to lower DCO conversion rate (although having a higher total deoxygenation rate due to HDO). The gaseous product is removed from the liquid fraction through gas-liquid separation after the DCO reaction. The wt-% of carbon oxide compounds in the gaseous product, including at least CO₂, and CO are presented in the Table 4.

Table 4. The gaseous carbon oxide content of the DCO effluent from the reactions 1-4, presented as a wt-% of the total weight of the respective DCO effluent (DCO reaction product).

DCO product (wt-%)	Reaction 1	Reaction 2	Reaction 3	Reaction 4
COx	11.9	9.2	6.4	5.6

Example 2. Hydrotreatment of the DCO-effluent*Feeds used in the hydrotreatment (HT) reactions*

The liquid fractions obtained from the DCO reactions 1-4, presented in the Table 1, are further used in hydrotreatment (HT) reactions 1-4, respectively. Additionally, a comparative reaction 5 is included, wherein pure alkali raffinated rapeseed oil is exposed to HT reaction, without exposing the feedstock to a DCO reaction first. Each liquid product 1-4 from the DCO reaction is introduced into the same reactor set-up to the hydrotreatment (HT) step with a hydrotreatment catalyst.

Pretreatment of the feedstock/catalyst

The catalyst used in the HT reaction is commercially available sulfided NiMo supported on Alumina (sulfided NiMo/ Al_2O_3) in all reactions 1-5. Sulfur is added in the HT feeds at a concentration of 500 wt-ppm (Reactions 1-5), for keeping the catalyst in sulfided form. In all reactions 25 g of the catalyst is loaded into the reactor, followed by its drying and sulfiding with DMDS procedures well known in the oil refining industry. NiMo-catalyst has about 6-7 wt-% sulfur after sulfidation, calculated from the total weight of the catalyst.

HT reaction conditions

The HT reactions are conducted in parallel reactions 1-5. The HT reactions 1-3 are carried out at a temperature of 350°C and at a reactor pressure of 4 MPa and WHSV of the feedstock is adjusted to 1 h⁻¹. In the reaction 1 the hydrogen (H₂) feed ratio of is set to 250 nl H₂/ liter of the feedstock, whereas in the reaction 2 the hydrogen (H₂) feed ratio of is set to 300 nl H₂/ liter of the feedstock and in the reaction 3 the hydrogen (H₂) feed ratio of is set to 350 nl H₂/ liter of the feedstock. Thus, the reaction conditions for the reactions 1-3 are otherwise the same. The hydrogen (H₂) feed ratio in the reaction 2, and 3 is set higher, due to higher number of oxygenated compounds in the hydrotreatment feed (DCO effluent). The reaction 4, is carried out at a temperature of 330°C and at a reactor pressure of 4 Mpa, WHSV of the feedstock adjusted to 1 h⁻¹, and the hydrogen (H₂) feed ratio of is set to 500 nl H₂/ liter of the feedstock, in order to ensure efficient deoxygenation and HDS/HDN reactions.

The reaction 5 (comparative example for single step deoxygenation of rapeseed oil fatty acids), is carried out at conventional hydrotreatment reaction conditions: a temperature of 310 °C and at a reactor pressure of 5 Mpa, WHSV of the feedstock adjusted to 1 h⁻¹, and the hydrogen (H₂) feed ratio of is set to 1000 nl H₂/ liter of the feedstock. Hydrogen is added to the feed at a high concentration, as the feedstock of reaction 5 comprises all the oxygenated compounds present in the initial alkali raffinated rapeseed oil. In this case hydrotreatment needs 2-3 times excess H₂ in feed (calculated from the theoretical

consumption of hydrogen based on 100% HDO + hydrogenation of double bonds).

Results

The gaseous product of the HT effluent after the HT reaction includes carbon oxides (CO/CO₂), and light hydrocarbons having carbon number < C₄, but also H₂O released in
5 the HDO reactions. The gaseous fraction is removed from the liquid fraction through gas-liquid separation after the HT reaction. All the reactions 1-5 result in 100% deoxygenation conversion, and thus no oxygen-containing fatty acids remain in the HT effluent.

The analysis of the liquid HT reaction products from the reactions 1-5, are shown in Table 5. In each reaction 1-4, the wt-% of C₁₇ HCs has increased significantly from the amount
10 present in the corresponding DCO effluents, indicating the HT reaction comprises also a significant portion of deoxygenation through the DCO reaction route. The liquid product distribution Table 5 indicates that the HT-reactions of the reactions 1-3 comprise the largest portion of deoxygenation through DCO, whereas the reactions 5 and 4 primarily comprise deoxygenation through HDO. This is also confirmed by the DCO selectivity rates (C₁₅ and
15 C₁₇ selectivity rates) of the HT reactions 1-5. The DCO deoxygenation selectivity rates for reactions 1, 2, 3, 4 and 5 are 50 wt-%, 50 wt-%, 47 wt-%, 35 wt-%, and 30 wt-%, respectively (calculated as the wt-% of deoxygenated C₁₇ or C₁₅ hydrocarbons undergone deoxygenation through DCO, from the total amount of deoxygenated C₁₈ and C₁₇, or C₁₆ and C₁₅ hydrocarbons in the HT effluent). From the DCO deoxygenation selectivity rates
20 for the C₁₅ and C₁₇ hydrocarbons can be concluded, that in the HT reactions 1-3 about half of the deoxygenation reactions of unreacted FA's occurs via DCO reaction (50/50/47 wt-%, respectively). Each effluent from the reactions 1-5 also comprises a significant wt-% of deoxygenated C₁₈ HCs, indicating all the HT reactions 1-5 comprise deoxygenation through HDO, the reactions 4 and 5 presenting the highest HDO deoxygenation rates (Table
25 5) and the lowest DCO selectivity rates of the HT reactions.

The HT effluent of the reaction 1 has the lowest amount of C₁₈ HCs (12.45 wt-%), indicating the HT effluent from reaction 1 would require only very mild hydrocracking, or only very small portion of the HT effluent (after fractionation) would require hydrocracking, the entire effluent to fit within the boiling range set for aviation fuel components. Consequently, the
30 HT effluent of the reaction 1 has the highest amount of ≤ C₁₇ HCs (82.54 wt-%), indicating that the HT effluent from reaction 1 comprises the largest portion of hydrocarbons, which after fractionation could be used as an intermediate aviation fuel component without the need for hydrocracking, the reaction 1 thereby providing the highest potential yield. This is further elucidated by Table 6, presenting the wt-% of renewable aviation fuel component
35 intermediate products (HCs < C₁₈) obtained from the initial rapeseed oil feedstock.

Table 5. The liquid product distribution after the HT step (reactions 1-5), presented as the wt-% from the total weight of the liquid HT reaction products.

HT product distribution (wt-%)	Reaction 1	Reaction 2	Reaction 3	Reaction 4	Reaction 5
C18 HC	12.45	19.95	29.82	43.04	66.02
C17 HC	82.54	75.50	65.92	52.82	28.28
C16 HC	0.37	0.27	0.22	0.77	3.47
C15 HC	4.27	3.93	3.69	3.02	1.49
C4-C14 HC	0.24	0.23	0.23	0.23	0.50
>C18 HC/FA	0.12	0.12	0.12	0.12	0.24

5 **Table 6.** wt-% of renewable aviation fuel component intermediate hydrocarbon yield (<C18) obtained from the initial alkali raffinated rapeseed oil feedstock after DCO and HT steps.

Products < C18 (wt-%)	Reaction 1	Reaction 2	Reaction 3	Reaction 4	Reaction 5
	87	80	70	57	34

The total sulfur and nitrogen contents of the HT effluents from all the reactions 1-5 are lower than detection limit of analysis, namely <1 wt-ppm, indicating that practically complete HDN and HDS reactions are also comprised in the HT reaction.

Table 7. shows the total hydrogen consumption of the reactions 1-5 during the DCO reaction (example 1) and hydrotreatment (HT) reaction (example 2). The H₂ quantities consumed in the reactions 1 - 4 are significantly lower than the hydrogen consumption of the comparison reaction 5 (single step HT reaction at conventional HT reaction conditions, no DCO reaction). The quantities of hydrogen required in reactions 1-5 are inversely proportional to the amount of deoxygenation reactions occurring via DCO reaction (DCO selectivity).

Table 7. The combined total hydrogen consumption of the DCO and hydrotreatment (HT) reactions 1-5.

H ₂ (nl H ₂ / liter of the feedstock)	Reaction 1	Reaction 2	Reaction 3	Reaction 4	Reaction 5
H ₂ consumption	110	137	189	230	315

Hydrotreatment liquid product is thereafter optionally introduced to the hydrocracking step.

5 Prior to this, the hydrotreatment liquid product can be obtained through gas-liquid separation, which liquid product can then be distilled into two fractions, namely C17 and lower hydrocarbons, and C18 and heavier hydrocarbons, from which only the latter is introduced into the hydrocracking step. Hydrocracking product effluent together with the lighter fraction from the distillation step is then introduced into gas/liquid separation prior the

10 introduction of liquid into isomerisation step. Experimental studies of these steps may be done separately using conventional catalytic reactor systems and distillation/evaporation units.

Various embodiments have been presented. It should be appreciated that in this document, words comprise, include, and contain are each used as open-ended expressions with no

15 intended exclusivity.

The foregoing description has provided by way of non-limiting examples of particular implementations and embodiments a full and informative description of the best mode presently contemplated by the inventors for carrying out the invention. It is however clear to a person skilled in the art that the invention is not restricted to details of the embodiments

20 presented in the foregoing, but that it can be implemented in other embodiments using equivalent means or in different combinations of embodiments without deviating from the characteristics of the invention.

Furthermore, some of the features of the afore-disclosed example embodiments may be used to advantage without the corresponding use of other features. As such, the foregoing

25 description shall be considered as merely illustrative of the principles of the present invention, and not in limitation thereof. Hence, the scope of the invention is only restricted by the appended patent claims.

CLAIMS

1. A process for producing a renewable aviation fuel component, the process comprising:
- 5 i) providing a renewable feedstock (10) comprising free carboxylic acids (FCAs), esters of carboxylic acids, triglycerides, or combinations thereof;
- ii) subjecting the feedstock to a decarboxylation and/or decarbonylation (DCO) reaction (20) in a DCO zone, for removal of one carbon from a carbon chain of a carboxylic acid moiety of the renewable feedstock, in the presence of a DCO catalyst,
- 10 wherein the DCO zone:
- the DCO deoxygenation selectivity is at least 75 wt-% of the total weight of deoxygenated hydrocarbons, and
 - the deoxygenation conversion is at least 50 wt-% of the total weight of the feedstock,
- 15 thereby obtaining a DCO effluent (21);
- iii) subjecting at least a portion of the DCO effluent from step ii) to a hydrotreatment (HT) reaction (30) in the presence of hydrogen and a hydrotreatment catalyst in a HT zone, to obtain a hydrotreated effluent (31);
- iv) subjecting the hydrotreated effluent from step iii) to a gas-liquid separation (40)
- 20 to obtain a degassed hydrotreated effluent (41);
- v) subjecting at least a portion of the degassed hydrotreated effluent from step iv) to hydroisomerisation (H-ISO) (50), thereby obtaining a hydroisomerised effluent (51); and
- vi) subjecting at least a portion of the hydroisomerised effluent (51) from step v) to
- 25 a fractionation (60) and recovering at least the renewable aviation fuel component.
2. The process of claim 1, wherein the DCO deoxygenation selectivity in the step ii) is at least 80 wt-%, preferably at least 85 wt-%, more preferably at least 90 wt-%, even more preferably at least 95 wt-%, most preferably at least 97 wt-%, of the total weight of deoxygenated hydrocarbons.
- 30 3. The process of claim 1 or 2, wherein the deoxygenation conversion of the step ii) is at least 55 wt-%, preferably at least 60 wt-%, more preferably at least 70 wt-%, more

preferably at least 80 wt-%, at least 90 wt-%, at least 95 %, or at least 99 wt-%, of the total weight of the feedstock.

4. The process of any preceding claim, wherein the DCO effluent (21) from step ii) is subjected at least partly to a separation of at least CO and/or CO₂ gases (25), before directing the DCO effluent to step iii).
5. The process of any preceding claim, wherein the DCO catalyst of the step ii) is a heterogeneous catalyst comprising a metal selected from nickel, cobalt, copper, zinc, molybdenum, manganese, ruthenium, rhodium, rhenium, iridium, palladium, platinum and any combinations thereof.
- 10 6. The process of any preceding claim, wherein the DCO catalyst further comprises at least one support, selected from alumina; silica; zirconia; titania; carbon, such as activated carbon or graphite; molecular sieve; and any combinations thereof.
7. The process of any preceding claim, wherein the DCO reaction (20) of the step ii) is carried out in the presence of a H₂ feed ratio of less than 300 nl H₂/ liter of the feedstock, or completely without added H₂.
- 15 8. The process of any preceding claim, wherein the DCO reaction (20) of the step ii) is carried out at a temperature of 50 – 450 °C, preferably the temperature is 200 – 450 °C, more preferably the temperature is 250 – 400 °C; and/or at a pressure of 0.1 – 10 MPa, preferably the pressure is 0.1 – 2 MPa.
- 20 9. The process of any preceding claim, wherein the DCO reaction (20) of the step ii) is carried out at a Weight Hourly Space Velocity (WHSV) of 0.1 – 15 h⁻¹, preferably a WHSV of 0.25 – 5 h⁻¹.
10. The process of any preceding claim, wherein the HT reaction (30) of the step iii) is selected from hydrodeoxygenation (HDO), hydrogenation of double bonds, hydrodenitrogenation (HDN), hydrodesulfurization (HDS), and any combinations thereof, preferably the hydrotreatment reaction of the step iii) comprises HDN.
- 25 11. The process of any preceding claim, wherein the HT reaction (30) of the step iii) is carried out in the presence of the HT catalyst comprising at least one group VIII and/or VIB metal of the Periodic Table, preferably selected from nickel, molybdenum, tungsten, cobalt, and any combinations thereof, preferably selected from NiMo, 30 CoMo, and NiW.

12. The process of any preceding claim, wherein the HT catalyst of the step iii) further comprises at least one support selected from zeolite, silica, alumina, amorphous silica alumina (ASA), and any combination thereof.
13. The process of any preceding claim, wherein the HT reaction (30) of the step iii) is carried out at a H₂ feed ratio of 50 - 2000 nl H₂/ liter of feed, preferably 100 - 1000 nl H₂/ liter of feed, more preferably 150 - 500 nl H₂/ liter of feed.
14. The process of any preceding claim, wherein the HT reaction (30) of the step iii) is carried out at a temperature of 250 - 450 °C, preferably the temperature is 280 - 350 °C; and/or a pressure of 1 - 20 MPa, preferably the pressure is 2 - 10 MPa.
15. The process of any preceding claim, wherein 50 – 5000 wt-ppm, preferably 100 – 2000 wt-ppm of sulfur is added to the DCO effluent (21) directed to the step iii).
16. The process of any preceding claim, comprising:
- a. subjecting at least a portion of:
 - the DCO effluent (21) from the step ii), or
 - the hydrotreated effluent (31) from the step iii),
 - the degassed hydrotreated effluent (41) from the step iv) or
 - the hydroisomerised effluent (51) from the step v)to a fractionation (FRAC) (70), and recovering at least a first fraction (71) comprising hydrocarbons with a carbon number > C17 and a second fraction (72) comprising hydrocarbons with a carbon number ≤ C17;
 - b. directing the hydrocarbons with a carbon number > C17 (71) to a hydrocracking (HC) (75) in the presence of a hydrocracking catalyst, thereby obtaining a hydrocracked effluent (76);
 - c. combining the hydrocracked effluent (76) together with the second fraction (72); and
 - d. directing the combined hydrocracked effluent (76) and the second fraction (72) to step iii), iv), v) or vi), respectively.
17. The process of any preceding claim, wherein the H-ISO (50) of the step v) is done in the presence of a hydroisomerisation catalyst comprising:
- at least one metal selected from Group VIII of the Periodic Table, preferably from nickel, platinum and palladium, more preferably from platinum and palladium; and/or
 - a carrier, preferably selected from Al₂O₃ or SiO₂; and/or

- a molecular sieve, preferably selected from SAPO-11, SAPO-41, ZSM-22, ZSM-23 and ferrierite.

18. The process of any preceding claim, wherein the H-ISO of the step v) is conducted at:

5 a hydroisomerisation temperature of 200 - 500 °C, preferably 230 - 500 °C, more preferably 250 - 450 °C, even more preferably 280 - 400 °C; and
a pressure of 2 - 15 MPa, preferably 1 - 10 MPa, more preferably 3 - 10 MPa.

19. The process of any preceding claim, wherein the renewable feedstock (10) is subjected to a pretreatment (80) prior to the step ii), wherein the pretreatment
10 comprises:

- a. removal of impurities, such as metals and/or sulfur and/or phosphorus, from the renewable feedstock;
- b. hydrolysis of the renewable feedstock;
- c. fractionation of the renewable feedstock;
- 15 d. hydrogenation of the renewable feedstock, or
- e. any combination thereof.

20. Use of the renewable aviation fuel component obtainable from the process of any one of claims 1-19, as a renewable aviation fuel blend component in an aviation fuel blend, wherein the aviation fuel blend also comprises a fossil aviation fuel blend component.