

Process for Production of Renewable Fuels from Alcohols

RELATED APPLICATIONS

[0001] This disclosure claims priority to United States Provisional Patent Application Serial Number 63/490,708, titled Process for production of renewable fuels from Alcohols, filed 16 March 2023, the disclosure of which is incorporated herein by reference as if set out in full.

FIELD

[0002] This disclosure relates to systems and methods for producing renewable alcohols, hydrocarbons, and further processing for production of transportation fuels, such as, gasoline, jet fuel, and diesel fuels. The disclosure also relates to processes for conversion of C₁-C₅ alcohols to a mixture of C₂-C₇ olefins for use in the production of transportation fuels. Processes for conversion of bio-based ethanol to ethylene and subsequently to mixtures of C₃-C₈ olefins in a single-step are also disclosed.

BACKGROUND

[0003] Petroleum is a nonrenewable resource and its combustion results in carbon being released into the environment. There is an increasing demand for the use of biomass sources for replacing petroleum as the starting point for the synthesis of fuels, specifically transportation fuels, such as gasoline, jet fuel, and diesel fuel (generically referred to as fuels or transportation fuels herein). With the increased availability and reduced cost of biomass-derived alcohols, biomass-derived alcohols are inexpensive and a renewable feedstock for making a variety of olefins for use in producing downstream hydrocarbons.

[0004] Current biogas production and spent grain processing technologies produce waste products, such as proteins that could be used to produce other products, e.g. animal feed, and thus are not fully efficient technologies. Furthermore, current technologies fail to address the issues created by high nitrogen concentrations in anaerobic digestion systems. Current integrated systems and processes for improving yield and efficiency of alcohol production also suffer from inefficiencies and limitations. For example, there is a need for systems that produce stillage specifications that have fully efficient and low cost anaerobic digestion. There is also a need for production of high energy density fuels, such as hydrocarbon fuels or transportation fuels, especially those with a low or no carbon footprint (i.e. carbon neutral). Further, there is also a need for an improved method for processing of the renewable alcohols and hydrocarbons for production of transportation fuels.

[0005] Current technologies employing petroleum fuels inherently have a high carbon footprint, in part, because they are produced by recovering oil from the earth which ultimately ends up in the atmosphere once the fuel is combusted, such as in an engine. Use of biodiesel can pose a problem as well, because it is not a hydrocarbon and its chemical properties make its fuel properties inferior and its blending limits low. Renewable diesel is a hydrocarbon that is commonly used, but it is produced from fats, greases, and animal- and plant-derived oils that are in relatively low supply considering the high global demand for transportation fuels. Additionally, the coproducts produced from sources like renewable diesel and propane have a low octane value and high vapor pressure, and therefore very little value for recycling into transportation fuel markets, such as, the gasoline market or SAF market, to reduce net carbon footprint or net energy use.

[0006] Today there is a demand for zero or low carbon (i.e. carbon neutral) and negative carbon (i.e. carbon sequestering) footprint, high energy density transportation fuels such as jet fuel

(SAF), diesel fuel, bunker fuel, or gasoline. These fuels can be used in existing assets and blended at high levels with conventional fuels, or may be used directly without blending. Fuels such as ethanol have low energy density and blend limits due to incompatibility with existing assets. The ethanol production assets currently existing were not created to be carbon neutral. Renewable diesel is an example of a high energy lower carbon fuel, but the carbon footprint is much higher than zero, renewable diesel only serves the diesel fuel market, and the amounts of renewable feedstocks for production of renewable diesel are low compared to the market size and demand.

[0007] To date, the necessary steps to deliver a carbon neutral commercially viable transportation fuel with high energy density made from carbohydrates has not been identified. As can be seen in the California Low Carbon Fuel System pathway reports, there are no viable carbohydrate based fuels with high energy density (i.e. greater than 110,000 BTU/gallon) nor hydrocarbon fuels with zero carbon score.

[0008] Current commercial ethanol processes rely on distillation and evaporation as fermentation product recovery unit operations, which require large amounts of energy, even when heat integrated to minimize energy. For example, according to the Renewable Fuels Association (RFA) in a 2016 publication, a typical ethanol plant uses approximately 26,700 BTU/gallon of process energy (thermal energy plus electrical energy) from outside energy sources (hereafter referred to as “external energy”). External energy as used herein refers to energy sources brought in from outside ethanol plant limits. Ethanol has a low energy density of 76,300 BTU/gallon, which means production process energy represents approximately 35% of the output energy. This process energy, natural gas and grid electricity, represent a large carbon footprint and therefore prevent the processes from achieving carbon neutrality. Moreover, ethanol has an energy density approximately 65% of that of petroleum gasoline (116,000 BTU/gallon). Ethanol therefore poses

a limit on the transportation range in a vehicle, and ethanol cannot be used in jet fuel, diesel fuel, nor bunker fuel. Furthermore, its blend ratio in gasoline is limited.

[0009] Alcohol to hydrocarbon fuels conversion processes have not been commercially deployed for various reasons with varying success. The desire for renewable fuels with high energy density has only been a recent occurrence. The ability to convert alcohols to transportation fuels meeting fuel specifications, which requires control over the molecular architecture of the products, has not been developed heretofore. Therefore, there has been an unmet need for the supply of renewable alcohols having a carbon footprint low enough to enable the resultant hydrocarbon to have an attractive carbon footprint and be commercially viable for use as a transportation fuel.

[0010] Bio-based alcohol is a significant commodity chemical product. With the increased availability and reduced cost of bio-based ethanol, researchers have explored bioethanol as a feedstock for making a variety of downstream hydrocarbons, including the Jet Fuel and/or Diesel Fuel. Gevo Inc. researchers have developed a fermentative process facilitating bio-isobutanol production and provided herein a new route to bio-based fuels based on bio-based alcohols and subsequent conversions thereof.

[0011] The process of oligomerization of gaseous mono-olefins so as to form gasoline-type hydrocarbons is known, and has been disclosed in the patent literature (U.S. Pat. Nos. 4,613,719; and 9,688,590). However, there is an ever present need to develop new oligomerization processes employing more effective and/or less expensive methods.

[0012] The process of oligomerization of olefins has been carried out by using acid catalysts such as supported phosphoric acid, and olefin dimers have been generally obtained for gasoline additive after hydrogenation of the dimers (U.S. Pat. Nos. 6,689,927, and 6,284,938).

[0013] There are reported examples that use cation exchange resins for the oligomerization. It has been claimed that a cation exchange resin can be used in a dimerization (USP 20050119111A1). U.S. Pat. No. 5,789,643 taught that oligomerization could be catalyzed by zeolites, aluminas and ion exchange resins. Tetramers or pentamers could be obtained by the oligomerization of pre-formed dimers with ion exchange resins (U.S. Pat. No. 6,239,321).

[0014] Oligomerization of lower olefins (*e.g.*, C₂-C₈) to jet fuel and/or diesel fuel from bio-based alcohols utilizing catalytic conversion processes employing a first catalyst such as tungstated zirconium, molybdenum zirconium, or a combination thereof and a second catalyst such as an acidic ZSM-5 catalyst has been recently disclosed in U.S. Pat. No. 11,078,433 (Gevo Inc.), which is incorporated herein by reference in its entirety.

[0015] Provided herein are, *inter alia*, solutions to the above discussed and other problems in the field.

SUMMARY

[0016] This disclosure describes, among other things, a system for producing renewable hydrocarbons. The system includes a fractionation subsystem for processing a biomass that contains a carbohydrate; a fermentation subsystem for converting the carbohydrate to a fermentation product; a water treatment subsystem for receiving a first portion of the fermentation product; an alcohol enrichment subsystem for receiving a second portion of the fermentation product; a hydrocarbon production subsystem for receiving renewable alcohols and producing the renewable hydrocarbons; processing of the renewable alcohols and hydrocarbons for production of transportation fuel; and an energy management subsystem for receiving a fuel stream from the system and for delivering power to the system.

[0017] In some embodiments, one or more of the following features may be included in any feasible combination. For example, the fractionation subsystem may include at least one of: a storage vessel for storing the biomass; a pulverizer for breaking apart the biomass; a pretreatment vessel for pretreating the biomass; a treatment vessel for treating the biomass to produce the carbohydrate; and a high-temperature-short-time (HTST) vessel for pasteurizing the carbohydrate.

[0018] In some embodiments, the fermentation subsystem may include at least one of: a fermenter for converting the carbohydrate to the fermentation product; a nutrient addition subsystem; a pH adjustment subsystem; and an inoculum propagation subsystem.

[0019] In some embodiments, the water treatment subsystem may include at least one of: a beer well for receiving the first portion of the fermentation product; a microorganism separation device for removing microorganisms from the first portion of the fermentation product; a first distillation column for separating the first portion of the fermentation product into an alcohol and a bottom product; a digester for receiving a portion of the bottom product and for producing a biogas; and a biogas removal subsystem for removing a pollutant from the biogas.

[0020] In some embodiments, the alcohol enrichment subsystem may include at least one of: a flash tank for separating an condensate from the second portion of the fermentation product; a separation vessel for separating the condensate into a light phase and a heavy phase; an ion exchange vessel for purifying the light phase; a membrane separator for separating the light phase into an alcohol-high retentate and a water-rich permeate; a second distillation column for separating water from the alcohol-high retentate; and an alcohol storage vessel.

[0021] In some embodiments, the hydrocarbon production subsystem may include at least one of: a denitrogenation subsystem for separating nitrogen from the alcohol; a dehydration subsystem for converting the alcohol to an olefin; a hydrocarbon pretreatment subsystem for conditioning an

olefin feed; a hydrogen supply subsystem for providing hydrogen; and a hydrocarbon processing subsystem for converting the olefin feed to at least one of an iso-octane fraction, a C₁₂ alkane fraction, and/or C₁₆ alkane fraction.

[0022] In some embodiments, the hydrocarbon processing subsystem may include at least one of: a first reactor; a debutanizer; a second reactor; a separator; a first splitter; and/or a second splitter.

[0023] In some embodiments, the energy management subsystem may include at least one of: a fuel gas system for distributing fuel gas received from the hydrocarbon processing subsystem; a low pressure boiler for generating steam; a high pressure boiler for generating steam; a combined heat and power unit for generating steam and electricity; a wind turbine for generating electricity; an electric boiler for receiving renewable electricity and for heating water to produce steam; a biomass boiler for combusting biomass for heating water to produce steam; and/or a steam turbine for generating electricity.

[0024] In some embodiments, the biomass is corn, wheat, and/or sorghum. In some embodiments, the biomass has a negative carbon footprint. In some embodiments, the biomass is grown using strip till or no till. In some embodiments, the pulverizer is a mill. In some embodiments, the biomass is grown using agricultural practices to increase soil organic carbon. In some embodiments, the pretreatment vessel may include a recycled water input. In some embodiments, pretreating the biomass may include pretreating the biomass with an enzyme. In some embodiments, the treatment vessel may include a non-fermentable solids output. In some embodiments, the non-fermentable solids output may include dried distillers grain output and/or corn oil output. In some embodiments, a portion of the corn oil output may be used as fuel for a boiler.

[0025] In some embodiments, the fermenter is a continuous fermenter. In some embodiments, the fermentation product may include an alcohol. In some embodiments, the alcohol is isobutanol.

[0026] In some embodiments, the inoculum propagation subsystem is configured to provide a microorganism to the fermenter. In some embodiments, the microorganism is yeast. In some embodiments, the first portion of the fermentation product may include an alcohol. In some embodiments, the first portion of the fermentation product may include isobutanol. In some embodiments, the first portion of the fermentation product may include water. In some embodiments, the microorganism separation device may include a centrifuge. In some embodiments, the microorganism separation device may include a filter. In some embodiments, the microorganism separation device may include a settling tank. In some embodiments, the microorganisms are yeast.

[0027] In one aspect of the present disclosure, the first distillation column includes at least one vapor recompression subsystem. In another aspect of the present disclosure, the first distillation column includes at least one mechanical vapor recompression (MVR). In another aspect of the present disclosure, the first distillation column includes at least one thermal vapor recompression (TVR).

[0028] In one aspect of the present disclosure, the bottom product may include stillage. In some embodiments, the digester is an anaerobic digester. In another aspect of the present disclosure, the digester is a continuous digester. In another aspect of the present disclosure, the biogas removal subsystem may include a scrubber. In another aspect of the present disclosure, the pollutant is hydrogen sulfide.

[0029] In some embodiments, the flash tank receives the second portion of the fermentation product from the fermentation subsystem and returns part of the second portion of the fermentation

product back to the fermentation subsystem. In some embodiments, the second portion of the fermentation product is a broth and an isobutanol lean broth is returned back to the fermentation subsystem. In some embodiments, the flash tank operates at atmospheric pressure. In some embodiments, the flash tank operates under a vacuum at a reduced pressure below atmospheric pressure. In some embodiments, the flash tank operates at a temperature below a temperature of the second portion of the fermentation product received from the fermentation subsystem. In some embodiments, the separation vessel is a liquid/liquid separator for separating the condensate into the light phase and the heavy phase. In some embodiments, the light phase contains a greater amount of alcohol than the heavy phase. In some embodiments, the alcohol is isobutanol and the liquid/liquid separator form a biphasic system. In some embodiments, the ion exchange vessel contains an ion exchange resin and receives the light phase received from the separation vessel and removes impurities from the light phase by trapping ions in the ion exchange resin. In some embodiments, the ion exchange vessel receives a caustic solution to regenerate the ion exchange resin by flushing the impurities to a beer well.

[0030] In some embodiments, the hydrocarbon production subsystem further includes at least one of: an oil and water separator for separating wastewater from the hydrocarbon processing subsystem; and a hydrocarbon storage vessel for storing at least one of iso-octane, C₁₂ blend, or C₁₆ blend from the hydrocarbon processing subsystem.

[0031] In some embodiments, the denitrogenation subsystem receives the alcohol from the alcohol enrichment subsystem. In some embodiments, the alcohol is a fuel grade isobutanol. In some embodiments, the dehydration subsystem receives the alcohol from the denitrogenation subsystem and sends water and/or purge products to a beer well. In some embodiments, the olefin is a four carbon olefin.

[0032] In some embodiments, the hydrocarbon pretreatment subsystem may include at least one of: a coalescer for receiving the olefin from the dehydration subsystem; an adsorber for removing unreacted alcohols, water, and nitrogen; and a feed receiver.

[0033] In some embodiments, the coalescer further receives wash water and sends waste water to a surge tank and/or a dehydration unit. In some embodiments, the coalescer further receives wash water and sends waste water to a surge tank and/or a dehydration unit. In some embodiments, the olefin is a C₄ olefin. In some embodiments, the unreacted alcohol is isobutanol.

[0034] In some embodiments, the feed receiver receives the olefin from the adsorber. In some embodiments, the first reactor is a solid acid catalyzed oligomerization (e.g. Polynaphtha™ as one commercial example) reactor. In some embodiments, the first reactor receives the olefin from the feed receiver and oligomerizes the olefin to a polyolefin. In some embodiments, the debutanizer receives a polyolefin from the first reactor. In some embodiments, the debutanizer recycles an unreacted olefin and/or paraffin to the first reactor. In some embodiments, the second reactor is a hydrogenation reactor. In some embodiments, the second reactor receives a polyolefin from the debutanizer. In some embodiments, the second reactor further receives hydrogen from a hydrogen supply subsystem and hydrogenates the polyolefin to a C₈-C₁₆ alkane. In some embodiments, the separator removes an unreacted olefin and/or paraffin from the C₈-C₁₆ alkane. In some embodiments, the unreacted olefins and/or paraffins include C₄ products. In some embodiments, the separator recycles a portion of the C₈-C₁₆ alkane to the second reactor. In some embodiments, the first splitter separates the isooctane fraction from the C₈-C₁₆ alkane.

[0035] In some embodiments, the isooctane fraction may include about 85%, 90%, 91%, 92%, 93%, 94%, 95, 96%, 97%, 98%, 99% or 100% wt% C₈ alkanes, about 1%, 2%, 3%, 4% or 5 vol% olefins, and/or about 2 ppm, 4 ppm, 6 ppm, 8 ppm or 10 ppm sulfur.

[0036] In other embodiments, the isooctane fraction may include greater than 95 wt% C₈ alkanes, less than 5 vol% olefins, and/or less than 10 ppm sulfur. In some embodiments, the first splitter separates an unreacted olefin and/or paraffin from the C₈-C₁₆ alkane. In some embodiments, the unreacted olefins and/or paraffins include C₄ products. In some embodiments, the second splitter separates the C₁₂ alkane fraction from the C₁₈ fraction.

[0037] In some embodiments, the fuel gas system provides fuel gas to the low pressure boiler. In some embodiments, the fuel gas system provides fuel gas to the high pressure boiler. In some embodiments, the fuel gas system provides fuel gas to a fired heater. In some embodiments, the low pressure boiler is further configured to received natural gas and/or biogas from an anaerobic digester. In some embodiments, the combined heat and power unit is further configured to received natural gas and/or biogas from an anaerobic digester. In some embodiments, the low pressure boiler in the fuel gas system is configured to receive corn oil, biomass, or products from the pyrolysis of biomass as low carbon intensity fuels.

[0038] In one embodiment, there is provided a process for producing renewable transportation fuels, such as, for example, gasoline, jet fuel (SAF), and diesel fuel from renewable C₁-C₅ alcohols, comprising: a) converting C₁-C₅ alcohols to a mixture of C₂-C₇ olefins to generate stream A; b) optionally separating a predominantly C₂ olefin fraction from stream A for recycle to the alcohol conversion step to create or separate oligomerization conversion to generate a predominantly C₄+ stream B; c) optionally purifying stream A and/or stream B and combining to create a stream C suitable for oligomerization into fuel range olefins; d) oligomerizing stream C into fuel range olefins to create a stream D; e) optionally hydrogenating stream D to create a predominantly paraffin stream E; and f) separating stream E into fractions appropriate for use as gasoline, jet, and diesel fuels.

[0039] In another embodiment, there is provided a process for producing a renewable alkylate from renewable C₁-C₅ alcohols, comprising: a) converting C₁-C₅ alcohols to a mixture of C₂-C₇ olefins to generate stream A; b) optionally separating a predominantly C₂ olefin fraction from stream A for recycle to the alcohol conversion step to create or separate oligomerization conversion to generate a predominantly C₄+ stream B; c) optionally purifying stream A and/or stream B and combining with renewable or petrochemical isobutane to create a stream C suitable for alkylation; d) subjecting stream C to an alkylation process in order to produce a crude renewable alkylate stream D; and e) optionally purifying stream D such that it is suitable for direct use as a renewable gasoline.

[0040] Yet in another embodiment, there is provided a process of converting bio-based ethanol to higher octane gasoline and jet fuel fractions, comprising: a) converting ethanol to ethylene and subsequently to mixtures of C₃-C₈ olefins in a single-step; and b) the C₃-C₈ olefins, linear butenes and/or fractions thereof, are co-oligomerized with isobutylene, thereby increasing the branching and providing a octane gasoline and jet fuel fractions.

[0041] It should be appreciated that all combinations of the foregoing concepts and additional concepts discussed in greater detail below (provided such concepts are not mutually inconsistent) are contemplated as being part of the inventive subject matter disclosed herein. In particular, all combinations of claimed subject matter of this disclosure are contemplated as being part of the embodiments disclosed herein. It should also be appreciated that terminology explicitly employed herein that also may appear in any disclosure incorporated by reference should be accorded a meaning most consistent with the particular concepts disclosed herein.

BRIEF DESCRIPTION OF THE DRAWINGS

[0042] FIG. 1 is a flow chart illustrating a low energy production process consistent with embodiments of the current disclosure.

[0043] FIG. 2 is a flow chart illustrating a process for conversion of carbohydrates to hydrocarbons consistent with embodiments of the current disclosure.

[0044] FIG. 3 is a flow chart illustrating a ethanol pretreatment and processing system consistent with embodiments of the current disclosure.

[0045] FIG. 4 is a diagram illustrating the integration of alcohol to hydrocarbons waste heat with biomass processing consistent with embodiments of the current disclosure.

[0046] FIG. 5 is a process flow diagram illustrating a typical ethanol distillation system.

[0047] FIG. 6 is a process flow diagram illustrating a conventional triple-effect evaporator used in ethanol purification.

[0048] FIG. 7 is a process flow diagram illustrating the use of multiple MVR fans in ethanol purification consistent with embodiments of the current disclosure.

[0049] FIG. 8 is a diagram illustrating the evaporator system integrated with FIG. 7 consistent with embodiments of the current disclosure.

[0050] FIG. 9 is a graph showing a comparison of carbon intensities for different petroleum-based fuels as well as the carbon intensity reductions by employing various aspects of the low energy process to produce biofuels.

[0051] FIG. 10 is a flow chart illustrating an isobutanol production process consistent with embodiments of the current disclosure.

[0052] FIG. 11 is a flow chart illustrating a carbon neutral, zero net energy isobutanol process consistent with embodiments of the current disclosure.

[0053] FIG. 12 is a flow chart illustrating a continuation of the process of FIG. 11 consistent with embodiments of the current disclosure.

[0054] FIG. 13 is a flow chart illustrating a hydrocarbon pretreatment and processing system consistent with embodiments of the current disclosure.

[0055] FIG. 14 is a diagram illustrating the process of recycling ETO with C₂ to fuel products consistent with embodiments of the current disclosure.

[0056] FIG. 15 is a diagram illustrating the process of dimerization of ETO with C₂ to fuel products consistent with embodiments of the current disclosure.

[0057] FIG. 16 is a diagram illustrating the process of ethanol dehydration and solid-acid alkylation to alkylate product consistent with embodiments of the current disclosure.

[0058] FIG. 17 is a diagram illustrating the process involving ETO with liquid-acid or ionic liquid alkylation to alkylate product consistent with embodiments of the current disclosure.

[0059] FIG. 18 is a graph depicting the increase in octane as a result of co-oligomerizing isobutylene with linear butenes versus blending iso-octane.

DETAILED DESCRIPTION

[0060] Disclosed herein is a process for producing high energy density, zero carbon footprint hydrocarbon fuels, using a feedstock that is in large supply throughout the world and can be procured with zero carbon footprint itself: carbohydrates. In the systems and processes described herein, carbohydrates (*e.g.*, corn, wheat, and/or sorghum) are converted to hydrocarbons. In general, the systems and processes achieve a net zero energy use and zero carbon footprint by

using a low carbon footprint carbohydrate-based feedstock, a feedstock fractionation process which can be employed prior to or post-fermentation, to remove most protein and water-insoluble solids, a fermentation process, a biogas generation process using fermentation stillage, and an alcohol-to-hydrocarbon conversion process.

[0061] In the systems and processes disclosed herein, the total external process energy requirements are less than about 20,000 BTU/gallon, 15,000 BTU/gallon, 10,000 BTU/gallon, 8,000 BTU/gallon, 6,000 BTU/gallon, 4,000 BTU/gallon, 2,000 BTU/gallon; and the product is a hydrocarbon having an energy density of at least 100,000 BTU/gallon, 105,000 BTU/gallon, 110,000 BTU/gallon, 115,000 BTU/gallon, 120,000 BTU/gallon, 125,000 BTU/gallon, 130,000 BTU/gallon, 135,000 BTU/gallon, 140,000 BTU/gallon, 145,000 BTU/gallon, or 150,000 BTU/gallon. Therefore the external process energy represents no more than 6%, 8%, 10%, or 12% of the product energy. Achieving a zero carbon footprint, *i.e.* carbon neutrality, is possible due to this low external process energy requirement plus proper selection of the carbohydrate feedstock. Achieving this for ethanol or any other low energy density fuel is easier than for other fuels, given the lower energy density of the low energy density fuel. However, demand exists for a fuel that has an energy density closer to that of petroleum fuels, for example energy density of at least 65%, 70%, 75%, 80%, or 85% of petroleum gasoline or jet fuel.

[0062] In the hydrocarbon processes described herein, any gasoline distillation range hydrocarbon products provides an octane rating of at least 50, 55, 60, 65, 70, 75, or 80 to ensure economical blending with other components to produce gasoline. This contrasts renewable diesel production today, which gives a coproduct (propane) with an octane value less than 50, 55, 60, 65, 70, 75, or 80.

[0063] In the processes described herein, carbon intensity (CI) (gCO₂e/MJ) is used as a measure of a high energy density carbohydrate derived fuel using reduced tillage corn feedstock, stillage biogas, and wind power.

[0064] The “Carbon Intensity” (CI), as described herein, is calculated based on industry standard Argonne National Laboratory’s (Argonne) Greenhouse gases, Regulated Emissions, and Energy use in Technologies (GREET®) Model. However, other CI frameworks, although not used for calculating the CI values in this application, also can be used, including but not limited to: The California Low-Carbon Fuel Standard (CA-GREET3.0), The Carbon Offsetting and Reduction Scheme for International Aviation by the UN International Civil Aviation Organization (ICAO CORSIA), The Environmental Protection Agency’s Renewable Fuel Standard (EPA RFS), The European Renewable Energy Directive (EU REDII), The Canadian Clean Fuel Standard (CFS), and a carbon reduction regulation used by Brazil (RenovaBio). Argonne’s GREET® Model was developed by Argonne with the support of the U.S. Department of Energy (DOE), to consider carbon intensity related to the energy and environmental effects of fuels and vehicle technologies. Argonne GREET® model can be applied to all aspects of the fuel supply chain, enabling a true apples-to-apples comparison of greenhouse gas emissions for suppliers, consumers, industry, and regulators. Argonne GREET® Model quantifies emissions over the entire life cycle, and is deemed superior to other life-cycle analysis (LCA) tools available to date. Argonne GREET® Model includes the Feedstock Carbon Intensity Calculator (FD-CIC) and the Carbon Calculator for Land Use Change from Biofuels (CCLUB). These models can be downloaded from <https://greet.es.anl.gov/>.

[0065] Described and illustrated herein are carbon neutral, zero net energy systems systems, equipment, processes, features, and functions processes to address the issues in current systems.

In the following description, certain specific details are set forth in order to provide a thorough understanding of various embodiments. In some instances, well-known structures have not been shown or described in detail to avoid unnecessarily obscuring descriptions of the embodiments. Unless the context requires otherwise, throughout the specification and claims which follow, the word “comprise” and variations thereof, such as, “comprises” and “comprising” are to be construed in an open, inclusive sense, that is, as “including, but not limited to.” Further, headings provided herein are for convenience only and do not interpret the scope or meaning of the claimed disclosure.

[0066] Reference throughout this specification to “some embodiments”, “one embodiment” or “an embodiment” means a particular feature, structure or characteristic described in connection with the embodiment is included in at least one embodiment. Thus, the appearances of the phrases “in some embodiments”, “in one embodiment” or “in an embodiment” in various places throughout this specification are not necessarily all referring to the same embodiment. Furthermore, the particular features, structures, or characteristics may be combined in any suitable manner in one or more embodiments. Also, as used in this specification and the appended claims, the singular forms “a,” “an,” and “the” include plural referents unless the content clearly dictates otherwise. It should also be noted that the term “or” is generally employed in its sense including “and/or” unless the content clearly dictates otherwise.

[0067] The word “about” when immediately preceding a numerical value means a range of plus or minus 10 % of that value, e.g., “about 50” means 45 to 55, “about 25,000” means 22,500 to 27,500, etc. Furthermore, the phrases “less than about” a value or “greater than about” a value should be understood in view of the definition of the term “about” provided herein.

[0068] The C₁₂- oligomers can be use directly for the production of renewable diesel fuel and renewable jet fuel post hydrogenation, for example, for the isobutanol/iC₄ process.

[0069] *Process scheme for a Low Energy Production Process for Producing Biofuels*

[0070] A schematic of a low energy production process 100 is shown in FIG. 1. The low energy production process 100 is sometimes referred to as a net zero carbon footprint process (sometimes “Net Zero” or “NZ”). Regarding the feedstock and feedstock selection, it is now known that certain growing practices allow for carbohydrates having zero carbon footprint or lower. For example, reduce tillage corn grown with reasonable yields and no or low irrigation has been estimated to have a low or zero carbon footprint.

[0071] Referring to Step 1 of FIG. 1, the low energy roduction process 100 begins with preparing a fermentation feedstock. The fermentation feedstock is a biomass source that contains carbohydrates and, optionally, reducing water insoluble solids and proteins once the feedstock is introduced to water to make a slurry. It has been determined that a carbohydrate source (e.g., a high sugar corn) can impact the ability to ultimately meet a net zero energy, carbon neutral target. As mentioned in the FIG. 1, the feedstock is fractionated as needed for the process.

[0072] Step 2 of FIG. 1 includes conducting a fermentation to convert the carbohydrate to fermentation product. Alternatively, Step 1 can be carried out after Step 2 process.

[0073] Step 3 of FIG. 1 includes subjecting the fermentation broth containing a fermentation product to one or more fermentation product separation processes, which produces a first stream that contains the fermentation product and a second stream consisting of fermentation spent broth. The fermentation product is purified such that it can meet product specifications using one or more energy efficient processes consistent with the technology described in the present application.

Optionally, the fermentation product is subjected to one or more chemical transformations that result in one or more hydrocarbons that, in certain embodiments described herein, meet specifications for use in fuels and/or chemicals. A low thermal energy demanding unit operates to recover the fermentation product including, as described by the present application, distillation processes and equipment fitted with mechanical vapor recompression (MVR) and/or membranes, filters, or centrifuges, which enable an overall production process thermal energy that is less than the zero carbon energy generated by the fuel gas system.

[0074] In step 4 of FIG. 1, the fermentation product is converted to a hydrocarbon product having an energy density of at least 100,000 BTU/gallon, 105,000 BTU/gallon, 110,000 BTU/gallon, 115,000 BTU/gallon, 120,000 BTU/gallon, 125,000 BTU/gallon, 130,000 BTU/gallon, 135,000 BTU/gallon, 140,000 BTU/gallon, 145,000 BTU/gallon, or 150,000 BTU/gallon using a combination of dehydration and oligomerization followed by hydrogenation. The heat output of the oligomerization step can be integrated with the first conversion step, which includes dehydration, such that the overall net energy required to convert alcohol to hydrocarbon is minimized to achieve a low energy process.

[0075] In step 5 of FIG. 1, the hydrocarbon product is separated into transportation fuels, such as, jet fuel, diesel fuel, bunker fuel, and/or gasoline distillation range hydrocarbons. The hydrocarbon product, produced from the fermented product, generally consists of a range of C₆ to C₂₀ hydrocarbons that are useful for gasoline, jet, diesel, and/or bunker fuels.

[0076] At step 6 of FIG. 1, the stillage has a composition which is substantially free of proteins and suspended solids, thereby allowing for a water treatment system which is efficient and low cost. Specifically, the stillage composition have the following characteristics: greater than 40,000 chemical oxygen demand (COD), 45,000 COD, 50,000 COD, 55,000 COD, 60,000 COD, 65,000

COD, 70,000 COD, 75,000 COD, 80,000 COD, 85,000 COD, or 90,000 COD; less than 0.2%, 0.4%, 0.8%, 1%, 1.2%, or 1.4% Total Suspended Solids (TSS), less than 1000 ppm, 1300 ppm, 1600 ppm, 1900 ppm, 2200 ppm, or 2500 ppm Total Kjeldahl Nitrogen (TKN); less than 2000 ppm, 4000 ppm, 6000 ppm, 8000 ppm, 10,000 ppm, or 12,000 ppm protein.

[0077] Step 7 of FIG. 1 may include subjecting the stillage to one or more filtration processes to reduce the suspended solids and proteins, such as to achieve these specifications. In certain embodiments, step 7 may be optional depending on the compositions.

[0078] In step 8 of FIG. 1, the fermentation spent broth is subjected to an anaerobic digestion process that, in certain embodiments, produces at least 6,000 BTU, 8,000 BTU, 10,000 BTU, 12,000 BTU, 18,000 BTU, 22, 000 BTU, 25,000 BTU, 30,000 BTU, 35,000 BTU, or 40,000 BTU of methane per gallon of hydrocarbon product. This represents at least 6%, 8%, 10%, 12%, 18%, 22%, 25%, 30%, 35%, or 40% of the energy content of the product being produced. In some embodiments, the anaerobic digestion process can produce at least 20,000 BTU of methane per gallon of hydrocarbon product, which can represent at least 20% of the energy content of the product being produced. In some embodiments, the amount of methane generated is at least as great as the external process energy required. As an alternative to using the methane produced in step 8 as an energy source, at least some of the oils produced during biomass fractionation (Step 1) may be used as an energy source to drive the entire production process, or at least a portion of the production process, such that less external energy sources are required, which reduces the overall carbon footprint, at least in part.

[0079] Subsequent use of at least some of this methane as an energy source to drive the entire production process, whether in all or part of the production process, results in the need for less external energy sources. For example, with a COD of 82,000 mg/L in a 58 mmgpy isobutanol

plant, there would be approximately 233 MMBTU/hr, or 500 MMBTU/day, of methane generated in an anaerobic digester. This equates to approx. 2 million MMBTU/yr, assuming 24 hours per day and 350 days per year. The overall process thermal energy required is ~3300 MMBTU/day, not including energy required to dry any of the solids streams, and approximately 5500 MMBTU/day if 90% of the solids are dried to approx. 88% dry matter, with the rest being supplied to the feed market on a wet basis.

[0080] At step 9 of FIG. 1, water treatment processes are used to produce water. The recovered water can be recycled back to the front end of the process.

[0081] At step 10 of FIG. 1, one or more boilers can accept the gas outlet of the anaerobic digester (AD) system. The heat or energy from the AD system, in certain embodiments, produces steam that drives some of process. Using the heat/energy from the AD system also serves to lower overall process external energy requirements.

[0082] Referring to FIG. 2, a flow chart illustrating a process for converting carbohydrates to hydrocarbons via isobutanol 1000 is shown. A carbohydrate, such as corn 1001, is subjected to fractionation 1005. Fractionation generally refers to the conversion of the carbohydrate into constituent components. The fractionation 1005 products are then used for either feed and foodstock 1010, for example although other alternative uses 1010 may be available for different biomasses, or used in isobutanol production 1015. Although described as isobutanol, isobutanol is an example of one possible alcohol that may be used in the overall process. The isobutanol production 1015 products can be used for hydrocarbons productions 1020 or water treatment 1030. The isobutanol production 1015 products used for hydrocarbons production 1020 become isooctane jet fuel 1025. The isobutanol production 1015 product used for water treatment 1030 becomes, in part, a biogas 1035. The biogas 1035 and/or natural gas 1055 may be used to create an energy supply 1040.

Optionally, corn oil and/or biomass can also be used as part of the the energy supply 1040. In certain embodiments, any corn oil byproducts may be sold or otherwise disposed. The energy supply 1040 includes aggregate biogas, natural gas, corn oil and biomass. The energy 1040, in certain embodiments produces electricity 1050 and steam 1045, which, optionally, uses wind power, CHP, electric boilers, combinations thereof, and the like.

[0083] *Process scheme for a Energy Efficient Hydrocarbon Production Process Using Ethanol as an Alcohol Intermediate*

[0084] A schematic of a low energy production process well-tailored to Jet and Diesel fuels derived from an ethanol intermediate is shown in FIG. 3. The feedstock selection is critical to ensure that the input feedstock has zero or a negative carbon footprint.

[0085] FIG. 3 is a process and equipment diagram 50. The process and equipment diagram 50 shows a fractionation and fermentation unit 51 receives a carbohydrate 52, such as corn 52. The fractionation and fermentation unit 51 outputs a fermentation broth 53 that a distillation unit 54 receives. The fractionation and fermentation unit 51 also outputs corn fiber 55 and corn oil 56. The corn fiber 55 may be provided to an anaerobic digester (explained elsewhere herein). The corn oil 56 may be used as fuel or otherwise disposed. The distillation unit 54 outputs an alcohol 57, such as ethanol 57, that is received by dehydration unit 58. The distillation unit 54 outputs fusel oils 59 that are combined with other combustibles and sent to a boiler or the like. The distillation unit 54 also outputs a feed co-product 60 and wastewater 61. The dehydration unit 58 dehydrates the alcohol 57 and outputs ethylene 62 that a dimerization unit 63 receives. The dehydration unit 58 also outputs ethane 64 that is combined with the fusel oils 59 and other combustibles and additional wastewater 61. The dimerization unit 63 receives the ethylene 62, catalysts and/or chemicals 65 to facilitate operation of dimerization unit 63, and water 66 to produce C₄-C₆ olefins 67 and C₈+

olefins 67. The C₄-C₆ olefins 67 are received by a oligomerization unit 68. The dimerization unit 63 outputs lights purge 69 that are combined with the other combustibles. The dimerization unit 63 also outputs byproducts that include spent caustic agents 70 and additional wastewater 61. The oligomerization unit 68 receives the C₄-C₆ olefins 67 to produce C₆₊ olefins 71 received by a hydrogenation and fractionation unit 72. The oligomerization unit 68 also products additional lights purge 69 that are combined with other combustibles. The hydrogenation and fractionation unit 72 receives the C₄-C₆ olefins 67 and C₈₊ olefins 67 to produce transportation fuels 73. The hydrogenation and fractionation unit 72 also outputs lights purge 69 that are combined with other combustibles.

[0086] As disclosed herein, low carbon intensity (CI) corn or some other low CI carbohydrate, carbohydrate-rich feedstock is fed into the process as shown in FIG. 3. In the case of corn, for example, the feedstock is simply milled into a flour of desired particle size before being fed forward. The flour is mixed with warm water and enzymes (for example, alpha amylase and/or other suitable enzymes) before being fed into a liquefaction system that provides enough residence time for the starch in the corn flour to liquefy into polysaccharides. In some aspects, enzymes, such as cellulolytic enzymes, cellulase, protease, pectinase, xylanase, α -l-arabinofuranosidase, and/or phytase can be used. In conventional corn-based ethanol dry mill facilities, heating the corn flour/water mixture before liquefaction is a utility-intensive process in that large amounts of steam are directly injected into the slurry tank to achieve a target temperature. To help minimize steam and freshwater usage, warm water from the ethanol distillation system, warm water from the evaporator system, and warm thin stillage (referred to as “backset”) are recycled to the slurry tank from downstream. Additional water is often needed though to achieve the target percent total solids in the slurry tank. CO₂ scrubber bottoms is a common cold-water stream that is fed into the slurry

tank. If anaerobic digestion is being used for the project, the treated AD effluent water can also be fed to the slurry tank. In some cases, fresh make-up water such as city water could also be fed into the slurry tank. At their sources, these water streams are relatively cold and will increase the load of steam needed in the slurry tank. For this low energy process to produce biofuels, design features are being included to use excess heat from other parts of the NZ plant to heat the cold-water streams feeding the slurry tank thereby decreasing steam usage to the point of only being needed during start-up. FIG. 4 shows an example of the heat integration scheme being used to pre-heat the water feeding the slurry tank. Other configurations of FIG. 4 to integrate heat generated from one part of the NZ plant to another are possible and within the scope of the present application.

[0087] FIG. 4 shows a process and equipment diagram 75 for cook water heating integration. The diagram 75 shows one possible equipment and process configuration and others are possible and within the spirit and scope of the present technology. Liquid, heated by other processes associated with the net zero carbon footprint plant to produce transportation fuels and described elsewhere herein, are received by a process water tank 76. The process water tank 76 contains heated water 77 or process water 77 for re-use by the plant, which conserves the heat energy. The process water 77 may be used direction from the process water tank 76 or it can be further heated, and pressurized, by one or more heat exchangers 78, described further below as exemplary exchangers, where the heat exchangers 78 heat the process water 77 to superheated process water 79. To be safely stored, a slurry heat exchanger 80 uses the heat from superheated process water 79 to heat the slurry 81 in slurry tank 82. The slurry heat exchanger 80 cools and depressurizes the superheated process water 79 to cook water 83 that is stored in cook water tank 84. A slurry mixer 85 uses the cook water to combine cook water 83, corn flour (or the like), backset (described elsewhere herein) and enzymes to produce the slurry 81.

[0088] The anaerobic digester (AD) described elsewhere herein is optional. As the AD is optional, it may not be available to produce an effluent water stream as shown. A ethanol reactor creates an ethanol-to-ethylene purge water stream, which reactor is optional. The ethanol reactor converts ethanol to ethylene and water. This stream can be used to offset the freshwater fed to the CO₂ scrubber or it can be fed directly into the process water tank as shown in FIG. 4. An ethylene dimerization wash water is generated from washing the butenes formed from dimerizing ethylene in preparation for sending the butenes to the oligomerization process that produces C₈, C₁₂, and C₁₆ olefins. The ethanol-to-ethylene process receives 190 proof ethanol as a feed and, for safety reasons, it is cooled prior to going into intermediate storage by a water chiller in this example, other other heat exchangers could be used. The hot 190 proof ethanol is cross-exchanged with cold fresh make-up water to that is heated by the hot 190 proof ethanol, which conserves the heat drawn from the hot 190 proof ethanol. The heated water is provided to the process water tank. The stripper column bottoms is a relatively clean water stream that is generated in the production of 190 proof ethanol. It is warm when it comes from ethanol distillation and is fed directly into the process water tank. The evaporator condensate water is generated in the process of removing water from thin stillage to produce syrup. The syrup is ultimately blended with wet cake and either sold as-is for animal feed or, optionally, fed to a dryer to produce DDGS (Dry Distillers' Grains with Solubles). The evaporator condensate water is warm and contains low levels of impurities such as organic acids. It is fed directly to the process water tank.

[0089] The combined water streams in the process water tank are warm, but additional heat may be needed to achieve the target temperature in the slurry tank. A broad set of cooling services in the ethanol-to-jet process were evaluated where the heat removed by a cooling water cooled exchanger or air-cooled exchanger would instead be used to heat the water leaving the process

water tank. Cooling services were evaluated based on cooling load, temperature, operability, and safety. Although there were other services with higher cooling loads or useful temperature ranges, the three services that satisfied all criteria including operability and safety are shown in FIG. 4. Cooling services related to the exothermic heat of reaction generated from dimerizing ethylene and oligomerizing butenes were evaluated but were ultimately eliminated as heat sources for cook water due to being too cold or presenting safety/operability concerns in managing reaction temperatures. The three cooling services shown in FIG. 4 are arranged in a way that allows for the process water to be heated from about 67°C to about 112°C in three steps. The process water leaving the cross-exchanger with saturated recycle from hydrogenation is above the boiling point of water at atmospheric pressure. The process water will be back-pressure controlled at this point making it superheated. It is necessary to accumulate the hot process water in a surge tank to manage process fluctuations, but the water cannot be stored superheated safely or economically and a portion will vaporize if it is stored at atmospheric pressure. For this reason, the superheated process water stream is cross-exchanged with a circulation loop from the slurry tank to decrease the temperature of the process water to below the atmospheric boiling point before being fed to the cook water tank which is at atmospheric pressure. The cook water tank, as shown in FIG. 4, will operate at approximately 95°C. The cook water is fed to the slurry tank where it is combined with corn flour, backset, and enzymes. In total the heat integration scheme, as shown in FIG. 4, saves at least 33 MMBTU/hr of steam for a 100 million gallon per year ethanol capacity (anhydrous basis, 8000 annual operating hours) which equates to 39 MMBTU/hr of natural gas assuming a boiler efficiency of 85%.

[0090] Optionally, the stream leaving liquefaction can undergo additional processing to reduce the particle size of corn particles before being fed downstream. This is advantageous because it

helps liberate additional corn oil from the corn particles making more corn oil available for recovery downstream, although recovering energy via the corn oil is optional as the corn oil is usable in other carbon reducing manners. In the case of this low energy process for producing biofuels, the corn oil will be recovered and used as a fuel for the boiler system which helps reduce the carbon intensity of the final jet product as well as the naphtha and diesel coproducts.

[0091] The stream leaving liquefaction feeds into the fermenter tanks where it combines with additional enzymes and yeast. The fermentation is carried out in a conventional way. The liquified starch is simultaneously saccharified to form monosaccharides which are then consumed by the yeast to form ethanol and CO₂ gas. The CO₂ gas is vented from the fermenters to a scrubber system that uses water to scrub any ethanol or other water-soluble impurities that have been carried with the CO₂ vapor. The fermentations are carried out in batches. When a fermentation batch is complete the monosaccharides are nearly gone with a small number of residual sugars left behind. The ethanol concentration is about 13.0 wt%, 13.5 wt%, 14.0 wt%, 14.5 wt%, 15.0 wt%, or 15.5 wt%. In general the yeast does not consume the corn fiber, corn oil, corn protein, ash, and other corn-based components that are in the fermenter broth with the exception of a small amount of nutrients consumed as the yeast continue to multiply during fermentation. The ethanol, water, yeast, and corn-based components are transferred to a large surge tank referred to as the beer well. A stream of beer is continuously fed out of the beer well downstream into ethanol distillation where the ethanol is recovered while most of the water, corn components, yeast, and enzymes are sent downstream in what is called whole stillage.

[0092] Ethanol distillation can be carried out in several ways. At corn ethanol dry mill facilities that produce fuel-grade ethanol, the ethanol distillation system is almost always heat-integrated with the evaporator system in some way to reduce the amount of fresh steam needed to run the

process. FIG. 5 shows one example of a commercially available technology for distilling fuel-grade ethanol.

[0093] Like all available technologies, the one in FIG. 5 has features to facilitate operation. The beer from the beer well is almost always cross-exchanged in exchanger 101 with the stream leaving liquefaction to pre-heat the beer on its way to distillation in the beer stripper 102 (or beer distillation tower 102) and to save on cooling water demand when cooling the liquified mash before fermentation. FIG. 5 shows the wet ethanol 103 vapor removed at a top of the beer stripper 102 and being condensed in a first evaporator of a multi-effect evaporator system 104, shown in FIG. 6. The heat removed from the wet ethanol vapor is used to drive water vapor away from the thin stillage in the first evaporator, as shown in FIG. 6. The water removed in the first evaporator is used to drive the separation in the second-effect evaporator and so on. Evaporation is an energy intensive process, but the scheme shown in FIGs. 5 and 6 is such that the evaporator system requires no fresh steam. FIG. 6 shows a typical triple-effect evaporator configuration that can be used. FIG. 6 shows additional distillation components including, a beer reboiler 105, a rectifier column 106 (rectifier distillation tower 106), where the condensed wet ethanol 103 is revaporized by the rectifier column 106 and used to provide heat to the beer reboiler 105. A rectifier overhead condenser 107 uses the cooler output of the beer reboiler 105 to further condense the output to a 190 proof ethanol 108. A fusel oil wash and separation unit 109 is used by the rectifier column 106 and provides fusel output 110 that is combined, in this example, with the 190 proof ethanol 108. A side stripper 111 (side distillation tower 111 is driven by fresh steam 112 in a side stripper reboiler 113 to provide additional energy into the system 114.

[0094] FIG. 7 shows an exemplary embodiment of the multi-effect evaporation system 104 in more detail. The multiple-effect evaporation 104 receives the wet-ethanol vapor 103 and a thin

stillage from a stillage separation, which is explained elsewhere herein. Although the multi-effect evaporation system 104 is shown with three (3) evaporators 120, 121, and 122, more or less evaporators are possible depending on cost, design, and recovery constraints. The condensed wet ethanol out 103 is provided to the rectifier column 106. An energy conservation improvement to the distillation system shown in FIG. 5 consistent with the technology of the present application I shown in FIG. 7, below. In particular, it has been found that condensing the overhead vapor from the rectifier column 106 allows the pressure and temperature of the overhead vapor/liquid to be heated using one or more pumps, such as the shown mechanical vapor recompression fans. The pressure can be such that the condensed overhead vapor of the rectifier column 106 operates at a higher pressure and temperature such that the 190 proof ethanol can run the beer reboiler 105. The beer reboiler 105 has a substantial duty, so this heat integration saves on fresh steam and reduces the overall external energy required by the NZ plant. The issue with condensing the vapor at the top of the beer stripper column 102 before feeding into the rectifier column 106 is that most of the feed into the rectifier now needs to be re-vaporized to drive the ethanol to the top of the rectifier column 106. The energy for re-vaporization comes from a large amount of fresh steam 112 used for the side stripper reboiler 113.

[0095] For this lower energy process to produce biofuels, it is desirable to reduce all fresh steam and natural gas usage as much as possible, so even a scheme like the one in FIGs. 5 and 6 have opportunities for energy reductions. For this lower energy process to produce biofuels, a scheme like the one shown in FIG. 7 may be deployed. FIG. 7 shows a possible configuration although other configurations are within the spirit and scope of the present application.

[0096] FIG. 7 shows the distillation system with MVR fans, see beer column with MVR fans 3195 in FIG. 11 below. The MVR fans shown in FIG. 7 may be considered a mechanical vapor

recompression (MVR) system 123. FIG. 7 has some similar components to the distillation system of FIG. 5, and the similarities will not be re-explained. As shown in FIG. 7, the wet-ethanol 103 is provided directly to the rectifier column 106. The vapor from the top of the rectifier column 106 is fed to the MVR system 123, which has a series of MVR fans 124₁₋₅, of which MVR system 123 has five (5) MVR fans 124. The MVR fans 124 increase the pressure and temperature of the vapor from the top of the rectifier column 106 to drive the beer reboiler 105. The discharge of the beer reboiler is fed to a multi-effect finisher evaporator 125, which replaces the multi-effect evaporation system 104 above. A draw trap (not shown) in rectifier column 106 is used to draw 190 proof ethanol 108 directly from the rectifier column 106, which removes fermentation byproducts from the 190 proof ethanol 108 stream that would present issues in the ethanol dehydration reaction and purification. The 190 proof ethanol 108 is pulled from a draw tray one or more stages down from the top of the rectifier column (notice, columns as used in the present application in context may be distillation towers). This enables acetaldehyde to accumulate in the rectifier overhead, away from the 190 proof, and be purged out in the lights purge stream shown in the FIG. 7. The acetaldehyde can form unwanted byproducts like CO₂ during the ethanol dehydration reaction, so keeping it out of the 190 proof ethanol is desired. As shown in FIG. 7, the fusel stream is not recombined with the 190 proof ethanol 108 stream as it is in a conventional corn dry mill plant that produces fuel-grade ethanol. The fusel oil stream contains high concentrations of C₄ and C₅ alcohols. These C₄ and C₅ alcohols react in the ethanol dehydration reaction to form their respective mono-olefins. These olefins are ultimately removed from the ethylene produced along with other oxygenated byproducts and then burned as fuel to offset other heating demands of the process. The fusels have more heating value as a fuel before being converted to olefins, so for this low energy process to produce biofuels they are not combined with

the 190 proof ethanol. Table 1 below shows examples of the fusel oil stream composition and lights purge composition shown in FIG. 7.

[0097] Table 1.

Component	Lights Purge wt%	Fusels wt%
Ethanol	70.8905	25.6346
Water	3.6641	27.8144
Carbon Dioxide	8.2980	
Acetaldehyde	14.8326	0.0032
Acetal	0.0950	0.0089
Ethyl Acetate	2.0923	0.0012
Methanol	0.1249	0.0010
N-propanol		0.0897
N-butanol		0.9071
2-butanol		0.0696
Isobutanol		7.8546
Isopentyl Alcohol		22.7177
Amyl Alcohol		15.0954
Acetone	0.1096	0.00003
Isopropanol	0.0001	0.0001
1-pentanol		0.0011
Diethyl Ether	0.0916	
Tert-butanol	0.0008	0.0012
Total	100.000	100.000

[0098] The lights purge plus fusels combined are able to offset as much as 27 MMBTU/hr (higher heating value) of thermal demand for the project. Burning the corn oil as fuel provides another 62 MMBTU/hr (higher heating value) for producing steam.

[0099] The ethanol distillation for this lower energy process to produce biofuels differs from what a conventional corn ethanol dry mill may do in a number of ways. One clear distinction, as shown in FIG. 7, is the addition of the MVR system 123, centered around five (5) MVR fans 124 placed in series in between the rectifier column 106 overhead vapor and the beer reboiler 105. The MVR system 123 mechanically compresses a vapor which simultaneously increases its temperature to a more useful range and increases the pressure such that the vapor will condense at a higher temperature. This effect makes available the latent heat of condensation of the vapor to drive something like the beer column reboiler as shown in FIG. 7. At least three, four, and five or more MVR fans are shown because both temperature and pressure rise across an MVR fan is in

the range of 11°C and 6.5 psi, respectively. It is believed, a total pressure rise of 20 psi, 25 psi, 30 psi, 35 psi, 40 psi, 45 psi, 50 psi or more is preferred across the fans for the vapor to have a condensing range that is higher temperature than what the bottom of the beer column will be operating. The temperature rise across the fans is controlled by injecting high proof liquid from the rectifier pump into the feed to each fan. In one embodiment, the total number of MVR fan is preferably five. A single MVR fan provides benefits. In general, certain embodiments of the present technology will include at least three (3) MVR fans. Other embodiments of the present technology will include no more than five (5) MVR fans. In still other embodiments of the present technology, three (3) to five (5) MVR fans are provided.

[0100] The exemplary system configuration shown in FIG. 7 provides five (5) MVR fans to increase the pressure and temperature of the rectifier overhead vapor to a useful range where it can be used to run the reboiler of the beer stripper.

[0101] The five (5) stage MVR, for example, as depicted in FIG. 7, is found to be the most cost effective to get the required thermal energy savings, for the reasons, for example:

- i. MVRs utilize Hydrocarbon Ethanol to drive the reboiler for the beer stripper,
- ii. More MVRs drove the cost up,
- iii. Less MVRs did not provide the energy at the preferred pressure,
- iv. A compressor costs nearly three times the cost of five MVRs, and
- v. Thermal Energy Savings is at least about 7327 BTU/gal of ethanol, or at least about 11,780 BTU/gal of Hydrocarbon (based on 100 MMGPY ETOH and 62.2 MMGPY of Hydrocarbon).

[0102] An added water stripper takes up the hydraulic load from the bottom of the rectifier column and reduces the hydraulic load on the beer stripper column. This reduces the size

(diameter) of the beer stripper and enables the column to be fabricated in the shop and transported to the field. This provides significant cost savings over a field fabricated column.

[0103] Another way is to generate low pressure steam from an evaporator condensate or boiler water to drive an auxiliary reboiler for the water stripper that reduces the Natural Gas Thermal load by about 1,884 BTU/ gal of ethanol, or about 3,028 BTU/ gal of hydrocarbon.

[0104] The rectifier column 106 overhead vapor is partially condensed in the beer reboiler 105. Additional heat can be recovered after the beer reboiler 105 by the multi-effect finisher evaporator 125, which is shown in more detail in one exemplary configuration in FIG. 8.

[0105] FIG. 8 also shows thin stillage being concentrated into syrup. A mechanical vapor recompression (MVR) evaporator 126 uses a single MVR fan 127 to increase the water vapor pressure and temperature from a draw from the MVR evaporator 126 to produce the heat necessary to drive water off the thin stillage being fed to it. The discharge of the single MVR fan 126 is used for the reflux of a distillation system. The multi-effect finisher evaporator 125 is shown with three (3) finisher evaporators 127, 128, 129. The first finisher evaporator 127 receives the output of the beer reboiler 105. As can be appreciated from FIG. 8, the configuration of the vapor and condensate of the MVR evaporator 126 and the multi-effect finisher evaporator 125 is to recover/conserves as much energy as possible to reduce the overall carbon footprint of the NZ plant into which it is incorporated.

[0106] The mid-stillage produced from the MVR evaporator 126 is about 35 wt% to 45 wt% solids and is sent to a corn oil separation system 130 where corn oil is removed via centrifugation. Optionally, the distiller corn oil generated by the corn oil separation may be consumed by the NZ plant as energy to reduce the plant's footprint. Alternatively, the distiller corn oil may be disposed of in other carbon reducing manners.

[0107] The mid-stillage is then sent to the finisher evaporators 127 and 128 in this exemplary configuration where the output of the beer reboiler 105 is used to provide the heat for removing the rest of the water to achieve a syrup solids content of about 45 wt% or 45 wt% .

[0108] Output from beer reboiler 105 is used by finisher evaporators 127 and 128 to increase the syrup solids. This allows for better wet cake solids for shipping and sales, and reduces the energy in a rotary dryer by about 1.8 MMBTU/hr.

[0109] The condensed rectifier vapor is then sent back to a rectifier overhead tank 131. The contents of the rectifier overhead tank 11 may be used as a reflux injection to the rectifier column 106 to facilitate the distillation process.

[0110] Heat Integration in ethanol and hydrocarbon results in energy reduction of about 3,446 BTU/gal of ethanol, or about 5,540 BTU/ gal of hydrocarbon.

[0111] Another feature apparent from FIG. 7 and 8 is that the disclosed scheme improves efficiency because the beer stripper vapor 103, weth ethanol 103 vapor in this exemplary embodiment, is not condensed before feeding into the rectifier. With this feature, high rates of fresh steam 112 are not needed for the side stripper reboiler 113 to re-vaporize the feed to the rectifier column 106. The scheme in FIG. 7 incorporates the evaporator system into a heat-integration strategy. After the beer reboiler 105, a portion of the rectifier column 10 overheads is still vapor. The condensing load of this remaining vapor is used to drive the multi-effect finisher evaporator 125 that is part of the evaporator system that produces syrup. FIG. 8 shows an exemplary configuration of the mutli-effect finisher evaporator 125 scheme that integrates the rectifier column 106 overheads in FIG. 7 into the multi-effect finisher evaporator 125.

[0112] The scheme in FIG. 8 uses an MVR fan 127 to drive the MVR evaporator 126 to remove a portion of the water in the thin stillage. One benefit of MVR fan 127 is that it increases the overall efficiency of the evaporator and it eliminates the need for fresh steam in exchange for electricity from a renewable source like wind. The thin stillage feed to the evaporator system, as shown in FIG. 8, originates at the bottom of the beer column 102, as shown in FIG. 7, as whole stillage. The whole stillage is processed through a system to remove most of the insoluble solids (also known as suspended solids) from the liquid. Some type of centrifugation is typically employed for this. The two streams that come from the separation of whole stillage are referred to as thin stillage and wet cake. An example of the whole stillage, thin stillage, and wet cake flows and compositions are shown in Table 2. These correspond to over a 100 million gallon per year anhydrous ethanol capacity (8000 operating hours per year).

[0113] Table 2.

Component	Whole Stillage	Thins Stillage	Wet Cake
Large fibers + ash, lb/hr	16,482	2,472	14,010
Fine fibers + ash, lb/hr	12,034	4,212	7,822
Sugar (dextrose), lb/hr	778	597	181
Fine starch, lb/hr	3,328	1,165	2,163
Fine protein, lb/hr	22,423	7,848	14,575
Large protein, lb/hr	0	-	-
Corn glycerides, lb/hr	10,606	6,576	4,030
Corn FFA, lb/hr	442	274	168
Water, lb/hr	354,949	272,477	82,472
Lactic acid, lb/hr	78	60	18
Glycerol, lb/hr	2,969	2,279	690
Amylase protein, lb/hr	9	7	2
Yeast, lb/hr	3176	476	2,700
Urea, lb/hr	134	103	31
Total, lb/hr	427,410	298,547	128,862

[0114] Often a portion of the thin stillage is recycled back to the slurry tank to mix with incoming corn flour, cook water, and enzymes. This recycled thin stillage is called backset. The backset serves multiple functions. It helps reduce fresh water usage to the plant and reduces the size and utilities associated with the evaporator system. It also provides nutrients that are utilized by the yeast during fermentation ensuring that the yeast are healthy and perform quickly and efficiently.

[0115] The diagram in FIG. 8 shows the thin stillage feeding into an MVR evaporator 126. A constant circulation of thickened stillage is pumped from the bottom of the evaporator to the top. The thin stillage feed mixes with this recirculation flow as it travels to the top of the evaporator. This type of evaporator is referred to as a falling-film type of a evaporator because the feed to the

top is distributed over an array of tubes. The stillage flows as a film down the inside of the tubes and is heated by the hot vapor flowing over the outside of the tubes. Water vapor evaporates from the thin stillage as it travels down the tubes and is allowed to disengage from the liquid at the bottom of the evaporator. In FIG. 8, the water vapor that separates from the stillage at the bottom is fed to the MVR fan 127 that compresses the water vapor and forces it across the outside of the tubes where the majority of it condenses as it heats the stillage on the inside of the tubes. With the MVR fan 127, it is believed no fresh steam is needed to run multi-effect finisher evaporator 125. The stillage stream produced in the MVR evaporator 126 contains about 35 wt% or about 45 wt% total solids. This concentration is necessary to remove distillers' corn oil from the stillage stream in between the MVR evaporator 125 and the finisher evaporators 127, 128, 129. The process for removing distillers' corn oil typically involves the addition of a chemical to the stillage that acts as a demulsifier followed by centrifugation. The corn oil, being less dense than water, can be separated quickly and efficiently under the high relative centrifugal forces imposed by centrifugation.

[0116] In the case of this lower energy process to produce biofuels, the distillers' corn oil is used as a fuel to run the boiler system.

[0117] The de-oiled mid-stillage from the oil separation block in FIG. 8 is sent to the finisher evaporators 127, 128, 129. The heat to drive the first finisher evaporator 127 is supplied by the output of the beer reboiler 105. The stillage is circulated up to the top and back down through the tubes in the same manner as the MVR evaporator 126, above, except in this case the high proof rectifier vapor fed to the shell side of the first finisher evaporator 127 condenses as it transfers heat through the tube walls into the stillage. The separated water vapor at the bottom of the first finisher evaporator 127 is used to drive the second finisher evaporator 128. The concentrated stillage

produced by the second finisher evaporator 128 is referred to as syrup and is mixed with the wet cake from whole stillage separation. The wet cake can be sold as-is as animal feed and is referred to as wet distillers' grains with solubles (WDGS). The wet cake can also be dried and then sold as animal feed referred to as dry distillers' grains with solubles (DDGS). The water vapor from the second finishing evaporator 128 is condensed and mixed with the other cook water streams to be used in the slurry tank, see FIG. 4. The Table 3 below compares the energy usage for a conventional ethanol distillation and evaporation system, as shown in FIGs. 5 and 6 versus the system with energy reductions shown in FIGs. 7 and 8.

[0118] Table 3^a.

Parameter	Case 1 ^b Conventional	Case 2 ^c Low Energy
MVR Fan Power, Megawatts	0	6.4
MVR Fan Power, MMBtu/hr	0	21.8
Steam Usage, MMBtu/hr	117	15.8
Natural Gas Equiv. MMBtu/hr ^d	137.6	18.6

^aAll values based on a 100 million gallon per year anhydrous ethanol capacity and 8000 operating hours;

^bFor FIGs. 5 and 6;

^cFor FIGs. 7 and 8; and

^dAssumes a boiler efficiency of about 85%.

[0119] It is evident from Table 3, steam demand drops dramatically when using the MVR based system by about 86.5%. In addition, when placing the MVR power requirement in the same units as the steam, it is not a one-to-one trade from steam to electricity. The MVR realizes an energy reduction of about 67% compared to the conventional independent of heat source.

[0120] In the disclosed lower energy process to produce biofuels, all electrical demands are satisfied with renewable power generated from wind. It is also feasible to use other forms of renewable electricity such as solar, nuclear or hydroelectric. For the disclosed lower energy

process to produce biofuels, thermal loads that is normally be satisfied by natural gas can be satisfied with fuel streams generated in-situ by the process itself or from a blend of natural gas and renewable natural gas whose net carbon intensity is zero.

[0121] The disclosed low energy process to produce biofuels, for the jet fuel, naphtha, and renewable diesel products are marketable and valuable as renewable fuels, since their carbon intensity is low compared to their conventional petroleum-based counterparts. This low energy process to produce biofuels is designed to produce liquid transportation fuels possessing a carbon intensity of about 45 or less using the methodologies defined by the California Air Resource Board's (CARB) Low Carbon Fuel Standard (LCFS). FIG. 9 shows a comparison of carbon intensities for different petroleum-based fuels as well as the carbon intensity reductions by employing various aspects of the this low energy process to produce biofuels design to ultimately achieve a carbon intensity of about 45 or less for the liquid transportation fuel products.

[0122] FIG. 10 shows an isobutanol process and system 2000. At the beginning of the isobutanol process and system 2000, a storage vessel 2001, e.g. corn bins, contains a carbohydrate, e.g. corn or high sugar corn, which may be sent to a pulverizer such as a hammer mill 2005, or alternatively carbohydrate thin stillage 2040 may be subjected to thin stillage high-temperature-short-time (HTST) processing 2045. Referring to the carbohydrate that is sent to the pulverizer, it next goes to a slurry tank 2010, followed by liquid tanks 2015 and subsequently treatment tanks 2020. The product of the treatment tanks 2020 undergoes separation such as feed and oil separation 2025, to produce end products such as corn oil and animal feed 2030 for storage, and waste products that are recycled into the process such as via Mash HTST 2035.

[0123] Referring to the thin stillage 2040, it is subjected to thin stillage HTST 2045 before use in isobutanol yeast propagation 2050. The product of the isobutanol yeast propagation 2050 then

goes to a fermentation unit such as isobutanol fermenter 2055, along with products of Mash HTST 2060. The products of isobutanol fermenters 2055 are sent to modified GIFT columns 2065 and a portion is recycled back to the isobutanol fermenters 2055 or sent to a beer well 2090. What is not recycled from the modified GIFT columns 2065 back into the isobutanol fermenters 2055/beer well 290 is sent to a liquid/liquid separator 2070. Products of the beer well 2090 are sent to a first distillation column 2085, such as beer column 2085. A product of the distillation column 2085 is fed to the liquid/liquid separator 2070, and a product of the distillation column 2085 is fed to a filter unit 2115. The liquid/liquid separator 2070 produces heavy phase 2075 and light phase 2080 liquids. The heavy phase 2075 liquid is sent back to the beer column 2085, such as for a reflux operation. The light phase 2080 liquid is sent to impurity removal 2095, through membranes 2100 for filtration and the like, and into a lights removal unit 2105. The lights removal unit 2105 produces fuel grade isobutanol 2110 for storage or hydrocarbon processing. The product from beer column 2085 fed to filter unit 2115 is combined with lights removal unit 2105 product and filtered either at filter unit 2115, optionally a filter prior to the beer column 2085 may be provided. The output of the filter 2115 is sent to an anaerobic digester 2120. High protein products 2125 from filter 2115 can be used for animal feed, recycled as fermentation nutrients, or otherwise disposed. The anaerobic digester 2120 produces biogas 2130 which can be sent to the boiler of another subsystem, and recycle water 2135, which is recycled into other subsystems.

[0124] FIG. 11 is a flowchart illustrating an isobutanol production process and system 3000 consistent with a net zero production plant consistent with the technology of the present application that produces, in certain embodiments, transportation fuels. The isobutanol production process and system 3000 flowcharts includes systems and subsystems that having particular structures that, taken together describes a NZ plant consistent with the technology of the represent application.

While certain structures, processes, and features of the NZ plant are described with relation to one or more of the identified subsystems, the specific structures, processes, and features may be spread across or contain in other subsystems of the NZ plant without departing from the scope of the present technology. The isobutanol production process and system 3000 includes a fractionation subsystem 3002, a fermentation subsystem 3003, a water treatment subsystem 3004, and an alcohol enrichment system 3006. Products from these subsystems, along with the subsystems shown in FIG. 12, are combined to create the carbon neutral, zero net energy overall production plant that uses the described processes and equipment.

[0125] In the fractionation subsystem 3002, recycle water 3001 is put through cook 3005 to create cook water 3010. Additionally, a carbohydrate such as corn 3030 moves through a receiving and storage system 3035, and subsequently milling 3025 to make milled corn 3020. The milled corn 3020 and cook water 3010, along with added enzyme 3045, are placed into slurry pretreatment 3015. The resultant slurry 3050, along with biogas 3035, goes into clean sugar process 3055. The clean sugar process 3055 has a dryer vent to atmosphere 3060, and produces products 3070, such as dried distiller's grain (DDG) and corn oil for storage and load out 3075, and sugar/mash 3080 for Mash HTST 3085 processor, which treats the sugar/mash 3080. The sterilized mash 3090 from the Mash HTST 3085 is then put into the fermentation subsystem 3003.

[0126] In the fermentation subsystem 3003, there is sterile air 3135, and additionally, air from a rich CO₂ vent collection header 3095. The rich CO₂ air is scrubbed by a rich CO₂ vent system scrubber 3100 and the resultant air is vented to the atmosphere 3105.

[0127] The sterilized mash 3090, received from fractionation subsystem 3002, along with nutrients 3115 and a pH adjustment agent 3120 are placed into fermentation 3110 structure. Additionally, recycle water 3250 is directed into stillage sterilization 3255, and the resulting sterile

stillage 3260, along with nutrients 3280, a pH adjustment agent 3290, and a microorganism 3275, such, for example, as yeast cream 3275, from a microorganism propagation unit 3270, such as, for example, yeast seed system 3270, are placed into propagation 3265. The propagation 3265 produces an inoculum 3285. The inoculum 3285 is fed to fermentation 3110, along with the above constituent parts. The fermentation 3110 unit produces at least a first portion and a second portion. The fermentation 3110 unit second portion includes at least a fermentation broth 3125 and. The fermentation 3110 unit first portion includes at least a dilute beer 3165 (or an alcohol 3165), such as isobutanol beer 3265. The fermentation broth 3125 is provided to the modified/enhanced GIFT 3140 of the alcohol enrichment subsystem 3006, described below.

[0128] A portion of the dilute beer 3165, such as for example, a dilute isobutanol beer 3165, from fermentation 3110 may be used for water recovery in the water treatment subsystem 3004, described below. A portion of the dilute beer 316 may be used by the alcohol enrichment system 3006, as described below. Prior to water recover or alcohol enrichment, however, beer well 3170 receives the dilute beer 3165. HC water from a DHYD unit 3175, and intermittent heavy component purge from the DHYD unit 3180, are fed to beer well 3170 to create beer 3185 (or undiluted beer 3185). The beer 3185 moves through yeast separation 3190 unit a beer column with MVR 3195, such as, for example, an isobutanol (IBA) beer column with MVR 3195. The the beer column with MVR 3195 receives a pH adjustment agent 3205. The resulting beer column with MVR 3195 condensate goes to the alcohol enrichment subsystem 3006 via a light/light separation unit 3215, such as, for example, an IBA light/light separation unit 3215. The heavy phase 3210 from the light/light separation unit 3210 is recycled back into the beer column with MVR 3195. The beer column with MVR 3195 produces BC bottoms 3245, which is used in the stillage

sterilization 3255. A portion of the BC bottoms 3245 is used in the water treatment subsystem 3004, explained below.

[0129] The alcohol enrichment system 3004 receives the fermentation broth 3125 from the fermentation subsystem 3003. An enhanced GIFT 3140 receives the fermentation broth 3125. The enhanced GIFT passes chilled water 3145 and hot water 3150 back and forth between GIFT chillers 3155. The enhanced GIFT 3140 produces lean broth 313, such as, IBA-lean broth 3130, that is provided to the fermentation 3110 of the fermentation subsystem described above and a GIFT condensate 3160. A separation unit 3125, such as IBA light/light separation unit 3215, receives the GIFT condensate 3160 to produce, among other things, a light phase IBA 3220 that goes to ion exchange 3225, as well as the heavy phase 3210 received by the beer column with MVR 3195 described above. The ion exchange 3225 may be upstream or downstream of drying membranes 3300, such as IBA membranes 3300. Caustic agents 3230 are added to the ion exchange 3225, and resulting impurities 3235, optionally, are recycled to the beer well 3170. IBA 3240 from the ion exchange 3225 also moves through IBA membranes 3300 to create water-rich permeate 3306 that is recycled to IBA light/light separation 3215 and IBA-rich retentate 3305 that is sent to an IBA light column 3310. The IBA light column 3310 creates fuel grade IBA 3315, which can be sent to IBA storage 3340 before being sent to the HC unit as fuel grade IBA 3345. Light column overhead 3320 from the IBA light column 3310 is fed to the anaerobic digester (AD) 3350 of the water treatment subsystem 3004, described below. The alcohol enrichment subsystem 3006 may include a lean CO₂ vent collection header 3325 that is vented to atmosphere 3335 through a lean CO₂ vent system scrubber 3330.

[0130] The water treatment subsystem 3004, as described above, receives BC bottom 3245 from the fermentation subsystem 3003 and light column overhead 3320 from the alcohol

enrichment subsystem 3006 at AD 3350. The AD 3350 also uses reject water from the HC unit 3375, and creates crude biogas 3355, recovered water 3380, and reject waste products 3370. The rejected waste products are disposed, such as, for example, by being sent to wetlands or land applications. The crude biogas 3355 is sent to biogas H₂S removal 3360 unit, to create biogas 3365 that may be sent to the LP broiler, CHP unit, DDG dryer, a combination thereof, or the like. The recovered water 3380 is sent to recycled water handling 3385 to create recycle water 3390.

[0131] Shown in FIG. 12 is a hydrocarbon production process 4000. The hydrocarbon production process and 4000 includes a hydrocarbon production subsystem 4002 and an energy management subsystem 4003. The hydrocarbon production process and system 4000 also provides certain ancillary component parts and processes that may or may not be specifically incorporated into the hydrocarbon production subsystem 4002 and energy management subsystem 4003. The hydrocarbon production process and system 4000 is coupled with the isobutanol process and system 3000 to form part of a carbon neutral, net zero energy overall process and plant. The hydrocarbon production subsystem 4002 intakes a water supply 4001, which includes recycled water from other subsystems such as recycle water 3390, described above, into a hydrogen supply system 4005 to form hydrogen 4010. Additionally, fuel grade IBA 4050, which includes IBA from another subsystem, such as fuel grade IBA 3345, described above, undergoes nitrogen removal 4045 to produce IBA 4040. A dehydration unit 4035 receives the IBA 4040 and produces hydrocarbon water 4055 and intermittent heavies purge 4060 to be sent to the beer well, such as beer well 3170 as described above. The dehydration unit 4035 also produces a feed 4030, such as, for example C₄ Olefins 4030, to hydrocarbon pre-treatment 4025. The hydrocarbon pre-treatment derives olefin feed 4020 and wash water 4065, which may be, optionally, used in various systems described herein. The olefin feed 4020, along with hydrogen 4010, described above, are sent to a

hydrocarbon processing unit 4015. The hydrocarbon processing unit 4015 produces hydrocarbon wastewater 4120, which is sent to a hydrocarbon oily water separator 4130 to create hydrocarbon reject water 4135 to be sent to the AD 3350 described above. The hydrocarbon processing unit 4015 also produces hydrocarbon vent products 4070 that leave the hydrocarbon production subsystem 4002 via a fuel gas system 4075 to produce fuel gas 4080, described below, and the vent products may be provided to a PSV header 4095 and subsequently received in a flare 4100 system. Additionally, the hydrocarbon processing unit 4015 produces isooctane 4115 and C₁₂-C₁₆ jet fuel 4110, both of which are sent to hydrocarbon storage and loadout 4125. Isooctane 4140 and C₁₂-C₁₆ jet fuel 4145 can then be obtained from hydrocarbon storage and loadout 4125.

[0132] The fuel gas system 4075, described above, also provides fuel gas 4080 to a fired heater 4085 and a HP boiler 4090, where the fuel gas 400 is used.

[0133] Excess fuel gas 4105 from fuel system 4075, to the extent there is excess fuel gas, may be provided to the energy management subsystem 4003 via an LP boiler 4155. The LP boiler 4155 also intakes, as needed, renewable natural gas 4150, biogas 4160 from the AD 3355, described above, and natural gas 4165 from a utility system, again as needed. A CHP unit 4170 intakes biogas 4160, natural gas 4165, as needed, and, optionally, corn oil from fractionation. The output of the LP boiler 415 and the CHP unit 4170, create LP steam 4175. LP Steam 4170, which may be a steam header, may receive input from a biomass boiler/ turbine 4195 and an electric boiler 4180, which electric boiler 4180 may use renewable energy sources to power the electric boiler 4180. The renewable energy sources include wind power 4190, photovoltaic panels (not shown) or energy created by the turbine of the biomass boiler/turbine 4195. The hydrocarbon production process and system 4000 also includes some ancillary systems and process. For example, the hydrocarbon production process and system 400 includes, among other things, a vent gas

collection and volatile organic compound (VOC) reduction unit 4210 which intakes from IBA vents 4200 and dryer vent 4205. Other components of the hydrocarbon production process and system 4000 include a flammable vapor catch tank 4215, nitrogen 4220, plant and instrument air 4225, raw water treatment 4230, reverse osmosis (RO) water 4235, cooling water 4240, sewers and pumps 4245, a clean in place (CIP) unit 4250, flush water 4255, chemicals 4260, and firewater 4265. The hydrocarbon pre-treatment 4025 and hydrocarbon processing unit 4015 are described in further detail in FIG. 13.

[0134] FIG. 13 illustrates more detail of the hydrocarbon pre-treatment 4025 and hydrocarbon processing unit 4015 as a hydrocarbon pretreatment and processing scheme 5000, which describes features, processes, and structure. Initially, a wash water column/coalescer 5010 receives wash water 5001 and a C₄ olefin feed 5005. The wash water column/coalescer 5010 feeds to a hydrocarbon waste water surge tank or dehydration unit (TBC) 5015 and to IBA water and nitrogen adsorbers 5020. The IBA water and nitrogen adsorbers 5020 feeds a feed receiver 5025 prior to combining with a C₄ olefins recycle 5045, a debutanizer recycle 5040, and a splitter bottoms recycle 5120. Polynaptha reactors 5030 received the aforementioned combination and feeds product to a debutanizer 5035. Products of the debutanizer 5035 either enter the debutanizer recycle 5040 or combine with C₄ raffinate recycle for RVP control 5055, hydrogen made up of 99.999 mol% H₂ 5060, and hydrogenated liquid recycle 5080 to enter total hydro reactors 5065.

[0135] The total hydro reactors 5065 feed a separator 5070 that produces a reactor purge 5075 and the hydrogenated liquid recycle 5080, some of which is provided to the total hydro reactors 5065 (above) and some of which is provided to splitter 5085. The reactor purge 5075, an output of the splitter 5085, and C₄ olefins recycle 5045 combine to produce C₄ raffinate product to fuel gas 5050. The output of separator 5070 not used as hydrogenated liquid recycle 5080 is received by

splitter 5085 that creates a splitter bottoms recycle 5120, which feeds back into the debutanizer recycle 5040 and feeds to C₁₂-C₁₆ splitter 5090, and isooctane 5110. The C₁₂-C₁₆ splitter 5090 creates a C₁₂ blend or jet fuel 5105 or C₁₂/C₁₆ jet fuel 5115. Furthermore, a first C₁₂ Blend 5095 can be created from isooctane 5110 and C₁₂ blend or jet fuel 5105, and a second C₁₂ blend 5100 can be created from C₁₂ blend or jet fuel 5105 and C₁₂/C₁₆ jet fuel 5115.

[0136] ETO with C₂ Recycle to Fuel Products

[0137] FIG. 14 is a process flow and equipment diagram 135, which shows equipment, processes, features, and functions according to certain aspects of the present technology. Referring to Step 1 of FIG. 14, ethanol, fusel oils, or a combination thereof is fed to ETO dehydration unit 136. The ETO dehydration unit 136 processes the feed to convert ethanol and/or fusel oils to ethylene and other light olefins. Next, an ethylene distillation tower 137 distills the ethylene. The ethylene is recycled as shown, and the C₃₊ olefins are fed to aromatics distillation tower 138, though alternatively the propylene can be distilled off as a chemical product and the C₄₊ olefins fed to aromatics distillation 138. Alternatively, all C₂ and C₄₊ olefins can be recycled to the ETO dehydration unit 136 to produce propylene at higher yield, in which case Steps 3 and 4, below, would not be present in FIG. 14. Ethane is purged from the Ethylene Recycle stream and used as low carbon intensity (CI) fuel gas. The ETO process inherently maximizes thermal efficiency and reduces energy requirements of ancillary systems by combining dehydration (endothermic) and conversion to light olefins (exothermic) to a single unit process, thereby reducing the total energy requirement as compared to a conventional 2-step process of ethanol dehydration to ethylene followed by ethylene dimerization to butenes. This combined with the large exotherm from hydrogenation results in a low-CI net-thermal-energy surplus for ETJ which can be used to satisfy part of the thermal demand of ethanol production and displace natural gas usage.

[0138] In Step 2 of FIG. 14, the aromatics and heavy olefins are removed from the light olefin stream as a purification step prior to oligomerization. The aromatic stream is then sold as a gasoline blendstock, and the aromatic distillation overhead stream comprised of primarily C₃₊ olefins is fed to an oligomerization unit 139.

[0139] In Step 3 of FIG. 14, The C₃₊ olefins are fed to the oligomerization unit 139 and converted to primarily C₆₊ olefins, by a reactor tower 140. The light olefins in the reactor tower 140 effluent stream are recycled, and the C₆₊ olefins are fed to a hydrogenation unit 141. A splitting tower 142 purges light paraffins from the recycle stream and used as low-CI fuel gas. Optionally, to increase gasoline product octane rating, linear butenes can be separated prior to oligomerization and isomerized to iso-butene then fed to oligomerization or dimerized separately to isooctene.

[0140] In Step 4 of FIG. 14, mixed olefins are fed to the hydrogenation 141 along with hydrogen and converted to paraffins. The hydrogen is produced by electrolysis powered by electricity produced by wind turbines and/or solar panels. The splitting tower 142 purges light paraffins and excess hydrogen from the recycle stream and used as low-CI fuel gas. The paraffinic mixture is split by distillation in the splitting tower 142 into lights, gasoline, sustainable aviation fuel (SAF), and diesel products. This separation can occur prior to hydrogenation if certain olefin products are desired.

[0141] System-wide measures to further reduce the system CI score include: use of electricity produced by wind turbines and/or solar panels, using surplus thermal energy and/or produced fuel gas for heating of upstream processing and/or utilities.

[0142] *ETJ NZI HC Ethanol Specification*

[0143] HC ethanol was developed to reduce the costs of producing pure ethanol to feed the HC unit. The process requires 2-3 more distillation columns (about \$10-20 MM) and about 3000-4000 BTU/ gal of ethanol.

[0144] The process allows removal of dehydration equipment and energy.

[0145] Appropriate ethanol concentration (for example, containing a minimum of about 85 wt%, 86 wt%, 87 wt%, 88 wt%, 89 wt% or 90 wt% ethanol content and a maximum of about 10 wt%, 11 wt%, 12 wt%, 13 wt%, 14 wt%, or 15 wt% water content) was developed through testing to find the most energy efficient concentration for the HC facility and allow for the removal of contaminants that would create catalyst life and product quality issues, such as:

- i. Lower contaminant level to prevent formation of acids in the HC process that would lead to serious corrosion issues or require more expensive metallurgy (costs 2-3 times),
- ii. Lower contaminant level to prevent catalyst activity loss,
- iii. Lower contaminant level to formation of by-products that would cause catalyst activity loss and possible product quality issues, and
- iv. This includes choosing the right ethanol concentration to prevent nitrogen concentrations in the ethanol that would ruin the catalyst;
 - a. Water concentration affects contaminant levels in the ethanol.

[0146] Furthermore, pure ethanol increases the CI.

[0147] *ETO with C₂ Dimerization to Fuel Products*

[0148] FIG. 15 is a process flow and equipment diagram 145, which shows equipment, processes, features, and functions according to certain aspects of the present technology. FIG. 15 Referring to Step 1 of FIG. 15, ethanol is fed to an ETO dehydration unit 146 that converts ethanol, fusel oils, or a combination thereof to ethylene and other light olefins. An ethylene distillation

tower 147 distills the ethylene that fed to dimerization unit 148 while the C₃₊ olefins are fed to aromatics distillation unit 149, though alternatively the propylene can be distilled off as a chemical product and the C₄₊ olefins fed to aromatics distillation. Ethane is purged from the distilled ethylene stream and used as low-CI fuel gas. The ETO process inherently maximizes thermal efficiency and reduces energy requirements of ancillary systems by combining dehydration (endothermic) and conversion to light olefins (exothermic) to a single unit process, thereby reducing the total energy requirement as compared to a conventional 2-step process of ethanol dehydration to ethylene followed by ethylene dimerization to butenes. This combined with the large exotherm from hydrogenation results in a low CI net-thermal-energy surplus for ETJ which can be used to satisfy part of the thermal demand of ethanol production and displace natural gas usage.

[0149] In Step 2 of FIG. 15, the distilled ethylene stream is fed to dimerization unit 148 and converted to primarily C₄₊ olefins. The light paraffins are purged and used as low-CI fuel gas. The C₄₊ olefin stream is distilled and an oligomerization unit 150 receives the C₄₊ olefin stream. The C₈₊ stream being fed directly to a hydrogenation unit 151 with the lighter C₆₊ olefin stream from the oligomerization unit 150. The C₄₊ olefins are combined with the C₃₊ olefins described below..

[0150] In Step 3 of FIG. 15, the C₃₊ olefins from the ethylene distillation tower 147 are distilled by aromatics distillation tower 149. The heavy olefins are removed from the light olefin stream in this step. The aromatic distillation tower 149 feeds the light olefin stream, which is available as a gasoline blendstock, comprised of primarily C₃₊ olefins is fed to oligomerization unit 150 along with the C₄₊ olefins described above.

[0151] In Step 4 of FIG. 15, The C₃₊ olefins from aromatics distillation tower 149 and olefins from dimerization unit 148 are fed to oligomerization unit 150 and converted to primarily

C₆₊ olefins by reactor tower 152. The light olefins in the reactor tower 152 stream are recycled, and the C₆₊ olefins are fed to a hydrogenation unit 151. The splitting tower 153 receives the output of the hydrogenation unit 151 and light paraffins are purged from the recycle stream and used as low-CI fuel gas. Optionally, to increase gasoline product octane rating, linear butenes can be separated prior to oligomerization and isomerized to iso-butene then fed to oligomerization or dimerized separately to isooctene.

[0152] In Step 5 of FIG. 15, mixed olefins are fed to the hydrogenation unit 151 along with hydrogen and converted to paraffins. The hydrogen can be produced by electrolysis powered by electricity produced by wind turbines and/or solar panels. Light paraffins and excess hydrogen are purged from the recycle stream by splitting tower 153 and used as low-CI fuel gas. The paraffinic mixture also is separated by the splitting tower 153 into lights, gasoline, SAF, and diesel products. This separation can occur prior to hydrogenation if certain olefin products are desired.

[0153] System-wide measures to further reduce the system CI score include: use of electricity produced by wind turbines and/or solar panels, using surplus thermal energy and/or produced fuel gas for heating of upstream processing and/or utilities.

[0154] *Ethanol Dehydration and Solid-Acid Alkylation to Alkylate Product*

[0155] FIG. 16 is a process flow and equipment diagram 155, which shows equipment, processes, features, and functions according to certain aspects of the present technology. Referring to Step 1 of FIG. 16, ethanol, fusel oils, or a combination thereof can be fed to a first dehydration unit 156 to convert the ethanol, fusel oils, or the combination into ethylene and other light olefins, such as C₂₊ olefins. Simultaneously, iso-butanol is fed to a second dehydration unit 157 that outputs a olefin stream that an aromatics distillation tower 158 receives to produce isobutylene. A hydrogenation unit 159 produces iso-butane where the C₂₊ light olefins from the first dehydration

unit 156 and the isobutane is combined. A alkylation unit 160 receives the C₂₊ light olefins and iso-butane. The C₂₊ light olefins and iso-butane stream is available as a gasoline blendstock and the aromatic distillation overhead stream comprised of primarily C₂₊ olefins is fed to alkylation unit 160.

[0156] In Step 2 of FIG. 16, iso-butane is produced from low-CI isobutanol (IBA) production described herein followed by dehydration to iso-utylene and then hydrogenation to iso-butane. It is notable that renewable iso-butane also can be produced from a variety of feedstocks, including ethanol. Alternatively, fossil-based iso-butane can be used though it typically carries a higher CI score.

[0157] In Step 3 of FIG. 16, alkylation unit 160 produces alkylate gasoline product. A final distillation tower 161 separates the light paraffins from the alkylate gasoline. The light paraffins are purged and used as low-CI fuel gas and iso-butane is recycled to the process.

[0158] The system-wide measures to further reduce the system CI score include: use of electricity produced by wind turbines and/or solar panels, using surplus thermal energy and/or produced fuel gas for heating of upstream processing and/or utilities.

[0159] *ETO with Liquid-Acid or Ionic Liquid Alkylation to Alkylate Product*

[0160] FIG. 17 is a process flow and equipment diagram 165, which shows equipment, processes, features, and functions according to certain aspects of the present echnology. Referring to Step 1 of FIG. 17, ethanol, fusel oils, or a combination thereof is fed to the ETO dehydration unit 166 that converts ethanol, fusel oils, or a combination to ethylene and other light olefins. A ethylene distillation tower 167 distills the output of the ETO dehydration unit 16 to ethylene and the C₃₋ olefins. A first aromatics distillation tower 168 receives the C₃₊ olefins. Optionally, the

ethylene may be dimerized into predominantly butenes and mixed with the C₃₊ olefin stream after the first aromatics distillation tower 168.

[0161] In Step 2 of FIG. 17, The first aromatics distillation tower 168 removes the aromatics and heavy olefins from the light olefin stream, such as the C₃₊ olefins steam. The aromatic stream is available as a gasoline blendstock, and the aromatic distillation overhead stream comprised of primarily C₃₊ olefins is fed to an alkylation unit 169, below

[0162] In Step 3 of FIG. 17, iso-butane is produced from low-CI IBA production described herein is received by a iso-butanol dehydration unit 170. A second aromatics distillation tower 171 produces iso-butene that is received by a hydrogenation unit 172 to produce iso-butane. It is notable that renewable iso-butane can also be produced from a variety of feedstocks, including ethanol. Alternatively, fossil-based iso-butane can be used though it typically carries a higher CI score.

[0163] In Step 4 of FIG. 17, the C₃₊ olefins (light olefins from the aromatics distillation tower 168) are combined with the iso-butane from the hydrogenation unit 172 and a alkylation unit 169 receives the the combination to output an alkylate gasoline product. A reactor tower 173 produces light paraffins, which are purged and used as low-CI fuel gas, and iso-butane is recycled to the process. The reactor tower 173 also produces alkylate gasoline.

[0164] System-wide measures to further reduce the system CI score include: use of electricity produced by wind turbines and/or solar panels, using surplus thermal energy and/or produced fuel gas for heating of upstream processing and/or utilities.

[0165] *High Octane Gasoline and Jet Fuel Fractions*

[0166] Bio-based ethanol can be converted to ethylene and subsequently to mixtures of C₃-C₈ olefins in a single-step. The resulting C₃-C₈ olefins, linear butenes and/or fractions thereof, may be co-oligomerized with isobutylene derived from bio-based isobutanol in proportions that increase branching to such an extent that mixtures of high octane and jet fuel fractions are maximized. While it is well known that increased branching results in higher octane fuels, it was surprising to find that the addition of amounts of isobutylene resulted in significantly higher octane values as compared to simply blending in more branched paraffinic compounds after oligomerization. FIG. 18 depicts the increase in octane as a result of co-oligomerizing isobutylene with linear butenes versus blending iso-octane.

[0167] Thus, the oligomerization of mixtures of isobutylene and/or linear butenes, derived from isobutanol and ethanol respectively, results in a higher octane gasoline and jet fuel fractions as compared to the oligomerization of linear butenes followed by the blending of iso-octane to increase octane values. Additionally, the side-by-side, or co-located, fermentation of isobutanol and ethanol results in surprising synergies and cost savings with regard to nutrient reduction, toxic impurity dilution, and energy savings. The exemplary oligomerization reaction step encompasses oligomerizing a mixture of isobutylene and a linear olefin stream over appropriate acidic catalysts at between 80°C to 260°C and pressures between 250 psig to 500 psig.

[0168] Single Stage Oligomerization Conditions: T=175°C in reactor, WHSV = 1.7, Feed: 14% isobutylene, 55% linear butenes (n-butene, cis/trans-2-butene), 31% pentene mixture; P=250 psig; Catalyst mixture: Tungstated Zirconia + ZSM5 Zeolite (Si/Al=55).

[0169] Selectivity of Oligomers is shown in Table 4.

[0170] **Table 4.**

	% C4	% C5- C7	% C8	% C9 - C11	% C12	% C13 - C15	C16	% C20	% C24
W2r+5S24 50nC4, 50nC4 + C5's; 175C; 250psi; T= 0-1h40'	17.25	7.95	25.26	7.93	22.17	4.09	9.56	3.65	2.14
W2r+5S24 50nC4, 50nC4 + C5's; 175C; 250psi; T= 1h40'- 2h40'	9.79	9.21	21.21	10.07	22.80	5.53	13.26	5.14	2.99
W2r+5S24 50nC4, 50nC4 + C5's; 175C; 250psi; T= 6h2'- 21h10'	18.41	9.53	24.15	9.57	21.71	4.69	8.78	2.38	0.78
W2r+5S24 50nC4, 50nC4 + C5's; 175C; 250psi; T= 21h10'- 22h10'	15.58	9.24	27.05	9.99	23.98	4.75	7.39	1.69	0.44

- [0171] Isobutylene/linear C4+linear C5 ratio= 0.16.
- [0172] Actual RON Measurement of hydrogenated C8 Fraction= 96.6.
- [0173] No isobutylene; Oligomerization of Linear C4+ linear C5.
- [0174] Actual RON Measurement of hydrogenated C8 Fraction= 72.7.
- [0175] The specific examples presented herein are intended to be illustrative and should not be construed as limiting in scope of the claims.
- [0176] The foregoing detailed description has been given for clearness of understanding only and no unnecessary limitations should be understood there from as modifications will be obvious to those skilled in the art.
- [0177] While described in connection with specific embodiments thereof, it will be understood that the principles described herein is capable of further modifications and this application is intended to cover any variations, uses, or adaptations following, in general, the principles disclosed herein and including such departures from the present disclosure as come within known or customary practice within the art to which the technology pertains and as may be applied to the essential features hereinbefore set forth and as follows in the scope of the appended claims.
- [0178] The disclosures, including the claims, figures and/or drawings, of each and every patent, patent application, and publication cited herein are hereby incorporated herein by reference in their entireties.
- [0179] *Exemplary Aspects/Embodiments*

[0180] Certain aspects, including embodiments/aspects of the present subject matter described above, may be beneficial alone or in combination, with one or more other aspects recited hereinbelow. In addition, while the present subject matter has been disclosed with reference to certain aspects recited below and, in the claims, numerous modifications, alterations, and changes to the described aspects/embodiments are possible without departing from the sphere and scope of the present disclosure. Accordingly, it is intended that the present disclosure not be limited to the described embodiments, aspects, and claims, but that it has the full scope defined by the language of this disclosure and equivalents thereof. While the present technology has been described with reference to the specific aspects/embodiments thereof, it should be understood by those skilled in the art that various changes may be made and equivalents may be substituted without departing from the true spirit and scope of the disclosure. In addition, many modifications may be made to adapt a particular situation, material, composition of matter, process, and/or process step or steps, to the objective, spirit and scope of the present disclosure. All such modifications are intended to be within the scope of the claims appended hereto.

[0181] Herein below are examples of aspects of the present technology.

1. A system for producing renewable hydrocarbons, comprising:
 - a fractionation subsystem for processing a biomass that contains a carbohydrate;
 - a fermentation subsystem for converting the carbohydrate to a fermentation product;
 - an alcohol enrichment subsystem for receiving a second portion of the fermentation product;
 - a hydrocarbon production subsystem for receiving an alcohol and producing the renewable hydrocarbons; an energy management subsystem for receiving fuel stream from the system and for delivering power to the system; and further comprising at least one of the following to reduce the energy demand or CI of the ethanol and/or fuels:

- a) a water treatment subsystem for receiving a first portion of the fermentation product;
 - b) combustion of light oxygenate, fusel oil, vegetable oil, biomass or derivative, and/or hydrocarbon co-products;
 - c) utilizing warm water sources and waste heat from the fermentation and hydrocarbon production systems for moderate temperature biomass processing;
 - d) utilizing an alcohol purification system that uses a mechanical vapor recompression (MVR) system to pressurize the rectifier column overheads, the condensation of which provides heat to the beer stripper reboiler as well as produces a purified alcohol suitable for conversion to hydrocarbons; and
 - e) a thin stillage evaporation system that utilizes the heat from a partially condensed stream of 190 proof ethanol from the rectifier overheads as well as the MVR system to produce a syrup that is at least about 35% to at least about 45% dry solids without any additional application of steam.
2. The system of aspect 1, wherein the fractionation subsystem further comprises at least one of:
- a storage vessel for storing the biomass;
 - a pulverizer for breaking apart the biomass;
 - a pretreatment vessel for pretreating the biomass;
 - a treatment vessel for treating the biomass to produce the carbohydrate; and
 - a high-temperature-short-time (HTST) vessel for pasteurizing the carbohydrate.
3. The system of aspect 2, wherein the biomass is corn.
4. The system of aspect 2, wherein the biomass has a negative carbon footprint.
5. The system of aspect 2, wherein the biomass grown using strip till or no till.
6. The system of aspect 2, wherein the pulverizer is a mill.
7. The system of aspect 2, wherein the pretreatment vessel includes a recycled water input.

8. The system of aspect 2, wherein pretreating the biomass includes pretreating the biomass with an enzyme.
9. The system of aspect 2, wherein the treatment vessel includes a non-fermentable solids output.
10. The system of aspect 9, wherein the non-fermentable solids output includes dried distillers grain output and/or corn oil output.
11. The system of any preceding aspect, wherein the fermentation subsystem comprises at least one of:
 - a fermenter for converting the carbohydrate to the fermentation product;
 - a nutrient addition subsystem;
 - a pH adjustment subsystem; and
 - an inoculum propagation subsystem.
12. The system of aspect 11, wherein the fermenter is a continuous fermenter.
13. The system of aspect 11, wherein the fermentation product includes an alcohol.
14. The system of aspect 13, wherein the alcohol is isobutanol.
15. The system of aspect 11, wherein the inoculum propagation subsystem is configured to provide a microorganism to the fermenter.
16. The system of aspect 15, wherein the microorganism is yeast.
17. The system of any preceding aspect, wherein the water treatment subsystem comprises at least one of:
 - a beer well for receiving the first portion of the fermentation product;
 - a microorganism separation device for removing microorganisms from the first portion of the fermentation product;
 - a first distillation column for separating the first portion of the fermentation product into an alcohol and a bottom product;
 - a digester for receiving a portion of the bottom product and for producing a biogas; and

a biogas removal subsystem for removing a pollutant from the biogas.

18. The system of aspect 17, wherein the first portion of the fermentation product includes an alcohol.
19. The system of aspect 17, wherein the first portion of the fermentation product includes isobutanol.
20. The system of aspect 17, wherein the first portion of the fermentation product includes water.
21. The system of aspect 17, wherein the microorganism separation device includes a centrifuge.
22. The system of aspect 17, wherein the microorganism separation device includes a filter.
23. The system of aspect 17, wherein the microorganism separation device includes a settling tank.
24. The system of aspect 17, wherein the microorganisms are yeast.
25. The system of aspect 17, wherein the first distillation column include vapor recompression subsystem.
26. The system of aspect 17, wherein the first distillation column include mechanical vapor recompression (MVR).
27. The system of aspect 17, wherein the first distillation column include thermal vapor recompression (TVR).
28. The system of aspect 17, wherein the bottom product includes stillage.
29. The system of aspect 17, wherein the digester is an anaerobic digester.
30. The system of aspect 17, wherein the digester is a continuous digester.
31. The system of aspect 17, wherein the biogas removal subsystem includes a scrubber.
32. The system of aspect 17, wherein the pollutant is hydrogen sulfide.
33. The system of any preceding aspect, wherein the alcohol enrichment subsystem comprises at least one of:

a flash tank for separating an condensate from the second portion of the fermentation product;

a separation vessel for separating the condensate into a light phase and a heavy phase;

an ion exchange vessel for purifying the light phase;

a membrane separator for separating the light phase into an alcohol-high retentate and a water-rich permeate;

a second distillation column for separating water from the alcohol-high retentate; and

an alcohol storage vessel.

34. The system of aspect 33, wherein the flash tank receives the second portion of the fermentation product from the fermentation subsystem and returns back part of the second portion of the fermentation product to the fermentation subsystem.

35. The system of aspect 33, wherein the second portion of the fermentation product is a broth and an isobutanol lean broth is returned back to the fermentation subsystem.

36. The system of aspect 33, wherein the flash tank operates at atmospheric pressure.

37. The system of aspect 33, wherein the flash tank operates under a vacuum at a reduced pressure below atmospheric pressure.

38. The system of aspect 33, wherein the flash tank operates at a temperature below a temperature of the second portion of the fermentation product received from the fermentation subsystem.

39. The system of aspect 33, wherein the separation vessel is a liquid/liquid separator for separating the condensate into the light phase and the heavy phase.

40. The system of aspect 39, wherein the light phase contains a greater amount of alcohol than the heavy phase.

41. The system of aspect 40, wherein the alcohol is isobutanol and the liquid/liquid separator form a biphasic system.

42. The system of aspect 33, wherein the ion exchange vessel contains an ion exchange resin and receives the light phase received from the separation vessel and removes impurities from the light phase by trapping ions in the ion exchange resin.
43. The system of aspect 42, wherein the ion exchange vessel receives a caustic solution to regenerate the ion exchange resin by flushing the impurities to a beer well.
44. The system of aspect 33, wherein the hydrocarbon production subsystem comprises at least one of:
- a denitrogenation subsystem for separating nitrogen from the alcohol;
 - a dehydration subsystem for converting the alcohol to an olefin;
 - a hydrocarbon pretreatment subsystem for conditioning an olefin feed;
 - a hydrogen supply subsystem for providing hydrogen; and
 - a hydrocarbon processing subsystem for converting the olefin feed to at least one of an iso-octane fraction, a C₁₂ alkane fraction, and/or C₁₆ alkane fraction.
45. The system of aspect 44, wherein the hydrocarbon production subsystem further comprises at least one of:
- an oil and water separator for separating wastewater from the hydrocarbon processing subsystem; and
 - a hydrocarbon storage vessel for storing at least one of iso-octane, C₁₂ blend, or C₁₆ blend from the hydrocarbon processing subsystem.
46. The system of aspect 44, wherein the denitrogenation subsystem receives the alcohol from the alcohol enrichment subsystem.
47. The system of aspect 46, wherein the alcohol is a fuel grade isobutanol.
48. The system of aspect 44, wherein the dehydration subsystem receives the alcohol from the denitrogenation subsystem and sends water and/or purge products to a beer well.
49. The system of aspect 44, wherein the olefin is a four carbon olefin.
50. The system of aspect 44, wherein the hydrocarbon pretreatment subsystem comprises at least one of:

a coalescer for receiving the olefin from the dehydration subsystem;
an adsorber for removing unreacted alcohols, water, and nitrogen; and
a feed receiver.

51. The system of aspect 50, wherein the coalescer further receives wash water and sends waste water to a surge tank and/or a dehydration unit.
52. The system of aspect 50, wherein the coalescer further receives wash water and sends waste water to a surge tank and/or a dehydration unit.
53. The system of aspect 50, wherein the olefin is a C₄ olefin.
54. The system of aspect 50, wherein the unreacted alcohol is isobutanol.
55. The system of aspect 50, wherein the feed receiver receives the olefin from the adsorber.
56. The system of aspect 44, wherein the hydrocarbon processing subsystem comprises at least one of:
- a first reactor;
 - a debutanizer;
 - a second reactor;
 - a separator;
 - a first splitter; and/or
 - a second splitter.
57. The system of aspect 56, wherein the first reactor is a polynaptha reactor.
58. The system of aspect 56, wherein the first reactor receives the olefin from the feed receiver and oligomerizes the olefin to a polyolefin.
59. The system of aspect 56, wherein the debutanizer receives a polyolefin from the first reactor.
60. The system of aspect 56, wherein the debutanizer recycles an unreacted olefin and/or paraffin to the first reactor.
61. The system of aspect 56, wherein the second reactor is a hydrogenation reactor.

62. The system of aspect 56, wherein the second reactor receives a polyolefin from the debutanizer.
63. The system of aspect 56, wherein the second reactor further receives hydrogen from a hydrogen supply subsystem and hydrogenates the polyolefin to a C₈-C₁₆ alkane.
64. The system of aspect 56, wherein the separator removes an unreacted olefin and/or paraffin from the C₈-C₁₆ alkane.
65. The system of aspect 64, wherein the unreacted olefins and/or paraffins include C₄ products.
66. The system of aspect 56, wherein the separator recycles a portion of the C₈-C₁₆ alkane to the second reactor.
67. The system of aspect 56, wherein the first splitter separates the isooctane fraction from the C₈-C₁₆ alkane.
68. The system of aspect 67, wherein the isooctane fraction comprises about 85%, 90%, 91%, 92%, 93%, 94%, 95, 96%, 97%, 98%, 99% or 100% wt% C₈ alkanes, about 1%, 2%, 3%, 4% or 5 vol% olefins, and/or about 2 ppm, 4 ppm, 6 ppm, 8 ppm, or 10 ppm sulfur.
68. The system of aspect 67, wherein the isooctane fraction comprises greater than 95 wt% C₈ alkanes, less than 5 vol% olefins, and/or less than 10 ppm sulfur.
70. The system of aspect 56, wherein the first splitter separates an unreacted olefin and/or paraffin from the C₈-C₁₆ alkane.
71. The system of aspect 70, wherein the unreacted olefins and/or paraffins include C₄ products.
72. The system of aspect 56, wherein the second splitter separates the C₁₂ alkane fraction from the C₁₈ fraction.
73. The system of any preceding aspect, wherein the energy management subsystem comprises at least one of:

a fuel gas system for distributing fuel gas received from the hydrocarbon processing subsystem;

a low pressure boiler for generating steam;

a high pressure boiler for generating steam;

a combined heat and power unit for generating steam and electricity;

a wind turbine for generating electricity;

an electric boiler for receiving renewable electricity and for heating water to produce steam;

a biomass boiler for combusting biomass for heating water to produce steam; and/or

a steam turbine for generating electricity.

74. The system of aspect 73, wherein the fuel gas system provides fuel gas to the low pressure boiler.

75. The system of aspect 73, wherein the fuel gas system provides fuel gas to the high pressure boiler.

76. The system of aspect 73, wherein the fuel gas system provides fuel gas to a fired heater.

77. The system of aspect 73, wherein the low pressure boiler is further configured to received natural gas and/or biogas from an anaerobic digester.

78. The system of aspect 73, wherein the combined heat and power unit is further configured to received natural gas and/or biogas from an anaerobic digester.

79. A system for producing renewable alcohols, comprising:

a fractionation subsystem for processing a biomass that contains a carbohydrate;

a fermentation subsystem for converting the carbohydrate to a fermentation product;

an alcohol enrichment subsystem for receiving at least a portion of the fermentation product;

an energy management subsystem for receiving fuel stream from the system and for delivering power to the system; and at least one of the following to reduce the energy demand or CI of the alcohol:

- a) a water treatment subsystem for receiving a first portion of the fermentation product;
- b) combustion of light oxygenate, fusel oil, vegetable oil, biomass or derivative, and/or co-products;
- c) utilizing an alcohol purification system that uses a mechanical vapor recompression system to pressurize the rectifier column overheads, the condensation of which provides heat to the beer stripper reboiler; and
- d) a thin stillage evaporation system that utilizes the heat from a partially condensed stream of 190 proof ethanol from the rectifier overheads as well as a mechanical vapor recompression system to produce a syrup that is at least about 35% to at least about 45% dry solids without any additional application of steam.

80. A process for producing renewable transportation fuels, such as, for example, gasoline, jet fuel, and diesel fuel from renewable C₁-C₅ alcohols, comprising:

- a) converting C₁-C₅ alcohols to a mixture of C₂-C₇ olefins to generate stream A;
- b) optionally separating a predominantly C₂ olefin fraction from stream A for recycle to the alcohol conversion step to create or separate oligomerization conversion to generate a predominantly C₄+ stream B;
- c. optionally purifying stream A and/or stream B and combining to create a stream C suitable for oligomerization into fuel range olefins;
- d. oligomerizing stream C into fuel range olefins to create a stream D;
- e. optionally hydrogenating stream D to create a predominantly paraffin stream E; and

- f. separating stream E into fractions appropriate for use as gasoline, jet, and diesel fuels.
81. The process of aspect 80 comprising separating a predominantly C₂ olefin fraction from stream A for recycle to the alcohol conversion step to create or separate oligomerization conversion to generate a predominantly C₄⁺ stream B.
82. The process of aspect 80 comprising purifying stream A and/or stream B and combining to create a stream C suitable for oligomerization into fuel range olefins.
83. The process of aspect 80 comprising hydrogenating stream D to create a predominantly paraffin stream E.
84. A process for producing a renewable alkylate from renewable C₁-C₅ alcohols, comprising:
- a) converting C₁-C₅ alcohols to a mixture of C₂-C₇ olefins to generate stream A;
 - b) optionally separating a predominantly C₂ olefin fraction from stream A for recycle to the alcohol conversion step to create or separate oligomerization conversion to generate a predominantly C₄⁺ stream B;
 - c) optionally purifying stream A and/or stream B and combining with renewable or petrochemical isobutane to create a stream C suitable for alkylation;
 - d) subjecting stream C to an alkylation process in order to produce a crude renewable alkylate stream D; and
 - e) optionally purifying stream D such that it is suitable for direct use as a renewable gasoline.
85. The process of aspect 84 comprising separating a predominantly C₂ olefin fraction from stream A for recycle to the alcohol conversion step to create or separate oligomerization conversion to generate a predominantly C₄⁺ stream B.

86. The process of aspect 84 comprising optionally purifying stream A and/or stream B and combining with renewable or petrochemical isobutane to create a stream C suitable for alkylation.
87. The process of aspect 84 comprising purifying stream D such that it is suitable for direct use as a renewable gasoline.
88. A process of converting bio-based ethanol to higher octane gasoline and jet fuel fractions, comprising:
- a) converting ethanol to ethylene and subsequently to mixtures of C₃-C₈ olefins in a single-step; and
 - b) the C₃-C₈ olefins, linear butenes and/or fractions thereof, are co-oligomerized with isobutylene, thereby increasing the branching and providing a octane gasoline and jet fuel fractions.
89. The process of aspect 88, wherein the isobutylene is co-oligomerized with linear butenes.
90. The process of aspect 88, wherein a mixture of isobutylene and a linear olefin are co-oligomerized with one or more suitable acidic catalysts at a temperature between 80°C to 260°C and under pressures between 250 to 500 psig.
91. The process of aspect 90, wherein the mixtures of isobutylene and linear butenes are derived from isobutanol and ethanol, respectively.
92. The process of aspect 90, wherein the catalysts are selected from a group consisting of: tungstated zirconium catalysts, nickel and/or cobalt doped tungstated zirconium catalysts, zeolites, or metal (Group IB-VIIB) doped zeolites and/or a any combination thereof.
93. The process of aspect 90, wherein the catalysts are a mixture of catalysts comprising: Tungstated Zirconia + ZSM5 Zeolite (Si/Al=55).
94. The process of aspect 88, wherein the oligomerization is carried out under the conditions of T=175°C in reactor, WHSV = 1.7, with 14% isobutylene, 55% linear butenes (n-butene, cis/trans-2-butene), 31% pentene mixture, and under a pressure of about 250 psig.

95. The process of any one of claim 80, 84, or 88, wherein the alcohols comprise a minimum of about 85 wt% to 90 wt% ethanol content and a maximum of 10 wt% to 15 wt% water content.
96. The process of claim 95, wherein the alcohols comprise a minimum of about 85 wt%, 86 wt%, 87 wt%, 88 wt%, 89 wt%, or 90 wt% ethanol content.
97. The process of claim 95 or 96, wherein the alcohols comprise a a maximum of about 10 wt%, 11 wt%, 12 wt%, 13 wt%, 14 wt%, or 15 wt% water content.
98. The process of any one of claim 80, 84, or 88, wherein the alcohols comprise a minimum of about 88 wt% ethanol content and a maximum of 12 wt% water content.

WHAT IS CLAIMED IS:

1. A process for producing renewable transportation fuels from renewable C₁-C₅ alcohols, comprising:
 - a. converting C₁-C₅ alcohols to a mixture of C₂-C₇ olefins to generate stream A;
 - b. optionally separating a predominantly C₂ olefin fraction from stream A for recycle to an alcohol conversion step to create or separate oligomerization conversion to generate a predominantly C₄₊ stream B;
 - c. optionally purifying stream A and/or stream B and combining purified stream A and/or stream B to create a stream C suitable for oligomerization into fuel range olefins;
 - d. oligomerizing stream C into fuel range olefins to create a stream D;
 - e. optionally hydrogenating stream D to create a predominantly paraffin stream E; and
 - f. separating stream E into fractions appropriate for use as transportation fuels.
2. The process of claim 1 comprising separating a predominantly C₂ olefin fraction from stream A for recycle to the alcohol conversion step to create or separate oligomerization conversion to generate a predominantly C₄₊ stream B.
3. The process of claim 1 comprising purifying stream A and/or stream B and combining to create a stream C suitable for oligomerization into fuel range olefins.
4. The process of claim 1 comprising hydrogenating stream D to create a predominantly paraffin stream E.
5. A process for producing a renewable alkylate from renewable C₁-C₅ alcohols, comprising:
 - a. converting C₁-C₅ alcohols to a mixture of C₂-C₇ olefins to generate stream A;

- b. optionally separating a predominantly C₂ olefin fraction from stream A for recycle to the alcohol conversion step to create or separate oligomerization conversion to generate a predominantly C₄₊ stream B;
 - c. optionally purifying stream A and/or stream B and combining with renewable or petrochemical isobutane to create a stream C suitable for alkylation;
 - d. subjecting stream C to an alkylation process in order to produce a crude renewable alkylate stream D; and
 - e. optionally purifying stream D such that it is suitable for direct use as a renewable gasoline.
- 6. The process of claim 5 comprising separating a predominantly C₂ olefin fraction from stream A for recycle to the alcohol conversion step to create or separate oligomerization conversion to generate a predominantly C₄₊ stream B.
- 7. The process of claim 5 comprising optionally purifying stream A and/or stream B and combining with renewable or petrochemical isobutane to create a stream C suitable for alkylation.
- 8. The process of claim 5 comprising purifying stream D such that it is suitable for direct use as a renewable gasoline.
- 9. A process of converting bio-based ethanol to higher octane gasoline and jet fuel fractions, comprising:
 - a. converting ethanol to ethylene and subsequently to mixtures of C₃-C₈ olefins in a single-step; and
 - b. the C₃-C₈ olefins, linear butenes and/or fractions thereof, are co-oligomerized with isobutylene, thereby increasing the branching and providing a octane gasoline and jet fuel fractions.
- 10. The process of claim 9, wherein the oligomerization is carried out under the conditions of T=175°C in reactor, WHSV = 1.7, with 14% isobutylene, 55% linear butenes (n-butene, cis/trans-2-butene), 31% pentene mixture, and under a pressure of about 250 psig.

11. The process of aspect 1, 5, or 9, wherein the alcohols comprise a minimum of about 85 wt% to 90 wt% ethanol content and a maximum of 10 wt% to 15 wt% water content.
12. The process of aspect 11, wherein the alcohols comprise a minimum of about 85 wt%, 86 wt%, 87 wt%, 88 wt%, 89 wt%, or 90 wt% ethanol content.
13. The process of aspect 11 or 12, wherein the alcohols comprise a a maximum of about 10 wt%, 11 wt%, 12 wt%, 13 wt%, 14 wt%, or 15 wt% water content.
14. The process of aspect 1, 5, or 9, wherein the alcohols comprise a minimum of about 88 wt% ethanol content and a maximum of 12 wt% water content.