



(51) International Patent Classification:

C10G 45/00 (2006.01) *C10G 11/18* (2006.01)
C10G 47/00 (2006.01) *C10G 65/12* (2006.01)
C10G 21/00 (2006.01) *C10G 67/04* (2006.01)
C10G 65/02 (2006.01) *C10G 49/00* (2006.01)

(21) International Application Number:

PCT/US2018/038941

(22) International Filing Date:

22 June 2018 (22.06.2018)

(25) Filing Language:

English

(26) Publication Language:

English

(30) Priority Data:

62/530,523 10 July 2017 (10.07.2017) US

(71) Applicant: **EXXONMOBIL RESEARCH AND ENGINEERING COMPANY** [US/US]; 1545 Route 22 East, P.O. Box 900, Annandale, NJ 08801-0900 (US).

(72) Inventors: **BROWN, Stephen H.**; 2 Country Place, Labanon, NJ 08833 (US). **CUNNINGHAM, Brian A.**; Bu-

reau Takanawa 603, 4-6-10 Takanawa, Minato-ku, Tokyo, 108-0074 (JP). **SMILEY, Randolph J.**; 1835 Viola Lane, Hellertown, PA 18055 (US). **ILIAS, Samia**; 2706 Johnson Circle, Bridgewater, NJ 08807 (US). **MCMANUS, Jesse R.**; 1049 W. 22nd Street, #G, Houston, TX 77008 (US).

(74) Agent: **WRKICH, Joseph E.** et al.; Exxonmobil Research and Engineering Company, 1545 Route 22 East, P.O. Box 900, Annandale, NJ 08801-0900 (US).

(81) Designated States (unless otherwise indicated, for every kind of national protection available): AE, AG, AL, AM, AO, AT, AU, AZ, BA, BB, BG, BH, BN, BR, BW, BY, BZ, CA, CH, CL, CN, CO, CR, CU, CZ, DE, DJ, DK, DM, DO, DZ, EC, EE, EG, ES, FI, GB, GD, GE, GH, GM, GT, HN, HR, HU, ID, IL, IN, IR, IS, JO, JP, KE, KG, KH, KN, KP, KR, KW, KZ, LA, LC, LK, LR, LS, LU, LY, MA, MD, ME, MG, MK, MN, MW, MX, MY, MZ, NA, NG, NI, NO, NZ, OM, PA, PE, PG, PH, PL, PT, QA, RO, RS, RU, RW, SA, SC, SD, SE, SG, SK, SL, SM, ST, SV, SY, TH, TJ, TM, TN, TR, TT, TZ, UA, UG, US, UZ, VC, VN, ZA, ZM, ZW.

(54) Title: HYDROPROCESSING OF HIGH DENSITY CRACKED FRACTIONS

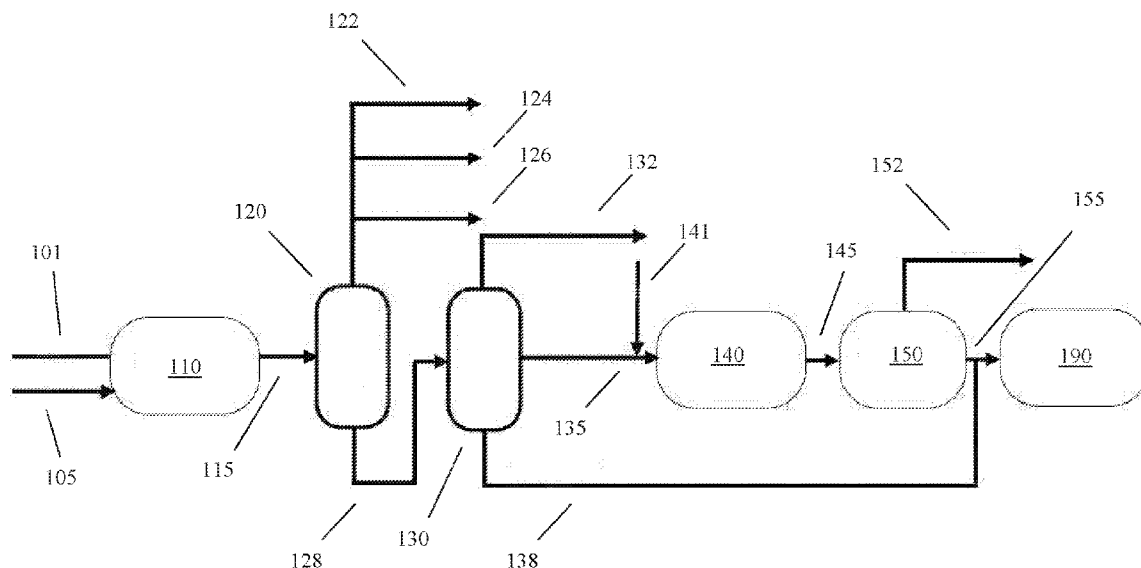


FIG. 1

(57) Abstract: Systems and methods are provided for upgrading a high density cracked feedstock, such as a catalytic slurry oil, by hydroprocessing. The upgrading can further include performing a separation on the effluent from hydroprocessing of the cracked feedstock, such as a distillation (i.e., separation based on boiling point) or a solvent-based separation. The separation on the hydroprocessed effluent can allow for separation of an aromatics-enriched fraction and an aromatics-depleted fraction from the hydroprocessed effluent. The aromatics-enriched fraction and aromatics-depleted fraction can then be separately used and/or separately undergo further processing.



(84) Designated States (*unless otherwise indicated, for every kind of regional protection available*): ARIPO (BW, GH, GM, KE, LR, LS, MW, MZ, NA, RW, SD, SL, ST, SZ, TZ, UG, ZM, ZW), Eurasian (AM, AZ, BY, KG, KZ, RU, TJ, TM), European (AL, AT, BE, BG, CH, CY, CZ, DE, DK, EE, ES, FI, FR, GB, GR, HR, HU, IE, IS, IT, LT, LU, LV, MC, MK, MT, NL, NO, PL, PT, RO, RS, SE, SI, SK, SM, TR), OAPI (BF, BJ, CF, CG, CI, CM, GA, GN, GQ, GW, KM, ML, MR, NE, SN, TD, TG).

Declarations under Rule 4.17:

- *as to applicant's entitlement to apply for and be granted a patent (Rule 4.17(ii))*
- *as to the applicant's entitlement to claim the priority of the earlier application (Rule 4.17(iii))*

Published:

- *without international search report and to be republished upon receipt of that report (Rule 48.2(g))*

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HYDROPROCESSING OF HIGH DENSITY CRACKED FRACTIONS

FIELD

[0001] Systems and methods are provided for hydroprocessing of main column bottoms from FCC processing to form hydroprocessed product fractions.

BACKGROUND

[0002] Fluid catalytic cracking (FCC) processes are commonly used in refineries as a method for converting feedstocks, without requiring additional hydrogen, to produce lower boiling fractions suitable for use as fuels. While FCC processes can be effective for converting a majority of a typical input feed, under conventional operating conditions at least a portion of the resulting products can correspond to a fraction that exits the process as a “bottoms” fraction. This bottoms fraction can typically be a high boiling range fraction, such as a $\sim 650^{\circ}\text{F}+$ ($\sim 343^{\circ}\text{C}+$) fraction. Because this bottoms fraction may also contain FCC catalyst fines, this fraction can sometimes be referred to as a catalytic slurry oil.

SUMMARY

[0003] In various aspects, a method for processing a heavy cracked feedstock is provided. The method can include exposing a feedstock comprising a density at 15°C of 1.06 g/cm^3 or more and at least 50 wt% of one or more $343^{\circ}\text{C}+$ cracked fractions to a hydroprocessing catalyst under fixed bed hydroprocessing conditions to form a hydroprocessed effluent. The one or more $343^{\circ}\text{C}+$ cracked fractions can have an aromatics content of 40 wt% or more relative to a weight of the one or more $343^{\circ}\text{C}+$ cracked fractions. A $343^{\circ}\text{C}+$ portion of the hydroprocessed effluent can have a density at 15°C of 1.04 g/cm^3 or less. The hydroprocessed effluent can then be separated in one or more separation stages to form an aromatics-enriched fraction and an aromatics-depleted fraction. At least a portion of the aromatics-enriched fraction can then be exposed to a second hydroprocessing catalyst under second fixed bed hydroprocessing conditions to form a second hydroprocessed effluent. As an alternative, the separation can be used to separate, from the hydroprocessed effluent, a first fraction comprising a T10 distillation point of at least 260°C and a T90 distillation point of 454°C or less and a second fraction comprising a T10 distillation point of at least 427°C . Optionally, in a single stage configuration, the at least a portion of the aromatics-enriched fraction (or the second fraction) can be recycled and combined with the feedstock, so that the second hydroprocessing catalyst corresponds to the first hydroprocessing catalyst. In such an option, the second hydroprocessed effluent can represent a portion of the first hydroprocessing effluent.

[0004] The separating of the hydroprocessed effluent to form the aromatics-enriched fraction and the aromatics-depleted fraction can correspond to a separation based on boiling point, a

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solvent-based separation, or a combination thereof. For a boiling point separation, some convenient cut points can correspond to forming an aromatics-enriched fraction with a T10 distillation point of 371°C or more, or 454°C or more, with a corresponding aromatics-depleted fraction having a T90 distillation point of 371°C or less, or 454°C or less.

[0005] In some aspects, the method can further include exposing at least a portion of the aromatics-depleted fraction (or at least a portion of the first fraction) to a distillate hydroprocessing catalyst under distillate fixed bed hydroprocessing conditions to form a distillate hydroprocessing effluent. In such aspects, a 177°C – 371°C portion of the distillate hydroprocessing effluent can optionally have a sulfur content of 50 wppm or less.

[0006] The one or more 343°C+ cracked fractions in the feedstock can correspond to any convenient type of heavy cracked fraction. Suitable examples of 343°C+ cracked fractions include a catalytic slurry oil, a coker bottoms fraction, a steam cracker tar fraction, a coal tar, a visbreaker gas oil, or a combination thereof. Optionally, the one or more 343°C+ cracked fractions can consist essentially of a catalytic slurry oil, so that a catalytic slurry oil corresponds to all of the 343°C+ cracked fraction. In aspects where a catalytic slurry oil is part of the one or more cracked fractions, the method can further include settling the catalytic slurry oil prior to exposing the feed to the hydroprocessing catalyst. A settled catalytic slurry oil can typically have a minimal catalyst fines content, such as 1 wppm or less.

[0007] In various aspects, the one or more 343°C+ cracked fractions can include about 2 wt% or more n-heptane insolubles. In such aspects, the hydroprocessed effluent can include about 1 wt% or less n-heptane insolubles. Additionally or alternately, the one or more 343°C+ cracked fractions can include at least a first amount of micro carbon residue, and the hydroprocessed effluent can include less than about half of the first amount of micro carbon residue. Additionally or alternately, the one or more 343°C+ cracked fractions can include at least 3 wt% of a 566°C+ portion. In such aspects, the effective hydroprocessing conditions can be effective for 55 wt% or more conversion of the feedstock relative to 566°C.

[0008] In various aspects, an insolubility number (I_N) of at least one of the first hydroprocessed effluent and the second hydroprocessed effluent can be 10 or more lower than an I_N of the feedstock. Additionally or alternately, a difference between a solubility number (S_{BN}) of the hydroprocessed effluent and the I_N of the hydroprocessed effluent can be at least 30.

[0009] In various aspects, the feedstock can include 4.0 wt% or more of micro carbon residue. Additionally or alternately, the hydroprocessed effluent can include 4.0 wt% or less of micro carbon residue. Additionally or alternately, the feedstock can include at least 1.0 wt% of organic sulfur, with the hydroprocessed effluent including 1000 wppm or less of organic sulfur.

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[0010] In various additional aspects, a system for processing a cracked feedstock is provided. The system can include a first hydroprocessing reactor comprising a first hydroprocessing inlet, a first hydroprocessing outlet, and a fixed bed comprising a first hydroprocessing catalyst. During operation of the system, the first hydroprocessing inlet can contain a feedstock comprising a density at 15°C of 1.06 g/cm³ or more and at least 50 wt% of one or more 343°C+ cracked fractions. The one or more 343°C+ cracked fractions can correspond to any of the types of 343°C+ cracked fractions described above in association with the method aspects. During operation of the system, the first hydroprocessing outlet can contain a hydroprocessed effluent. The system can further include a separation stage comprising a separation inlet, a first separation outlet, and a second separation outlet. The first separation inlet can be in fluid communication with the first hydroprocessing outlet. During operation of the system, the first separation outlet can contain a hydroprocessed effluent fraction having a T90 distillation point of 454°C or less. During operation of the system, the second separation outlet can contain a hydroprocessed effluent fraction having a T10 distillation point of at least 427°C. The system can further include a second hydroprocessing reactor comprising a second hydroprocessing inlet, a second hydroprocessing outlet, and a fixed bed comprising a second hydroprocessing catalyst. The second hydroprocessing inlet can be in fluid communication with the first separation outlet. The first hydroprocessing inlet can optionally be in fluid communication with the second separation outlet to allow for recycle of a heavier portion of the hydroprocessed effluent for further hydroprocessing. The system can optionally further include a fluid catalytic cracking reactor in indirect fluid communication with the second hydroprocessing outlet.

BRIEF DESCRIPTION OF THE FIGURES

[0011] FIG. 1 shows an example of a two stage reaction system for processing of heavy cracked feedstocks.

[0012] FIG. 2 shows an example of another reaction system for processing of heavy cracked feedstocks.

[0013] FIG. 3 shows an example of a two stage reaction system for processing of heavy cracked feedstocks.

[0014] FIG. 4 shows an example of another reaction system for processing of heavy cracked feedstocks.

[0015] FIG. 5 shows results related to solubility number and insolubility number from hydrotreatment of a feedstock containing one or more cracked fractions corresponding to catalytic slurry oils.

DETAILED DESCRIPTION

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[0016] In various aspects, systems and methods are provided for upgrading a high density cracked feedstock, such as a catalytic slurry oil, by hydroprocessing. The upgrading can further include performing a separation on the effluent from hydroprocessing of the cracked feedstock, such as a distillation (i.e., separation based on boiling point) or a solvent-based separation. The separation on the hydroprocessed effluent can allow for separation of an aromatics-enriched fraction and an aromatics-depleted fraction from the hydroprocessed effluent. The aromatics-enriched fraction and aromatics-depleted fraction can then be separately used and/or separately undergo further processing. In aspects where hydroprocessing is followed by distillation, the hydroprocessing can allow for distillation of a hydroprocessed effluent for feeds where distillation would not normally be practical under conventional distillation conditions. In aspects where hydroprocessing is followed by a solvent-based separation, the recycle of the aromatics-enriched fraction as part of the feed for hydroprocessing can enhance run lengths, due in part to the recycled portion of the feed providing an increase in the S_{BN} of the total feed to hydroprocessing.

[0017] It has been unexpectedly discovered that heavy cracked fractions, such as catalytic slurry oils, can be hydroprocessed with reduced or minimized amounts of coking by limiting the amount of conventional, non-cracked fractions included in the feedstock. While this discovery can allow for hydroprocessing of heavy cracked fractions under conventional, fixed bed processing conditions, further improvements are still desirable. For example, although heavy cracked fractions can be hydroprocessed with reduced or minimized coking and/or reactor plugging, achieving a desired target sulfur content in the hydroprocessed effluent can require low space velocities. It has been discovered that this is due in part to limitations in performing aromatics saturation at high temperatures, where the equilibrium processes for aromatic formation / saturation can tend to favor higher levels of aromatics. As another example, it can be desirable to increase the amount of conventional and/or non-cracked feeds that can be included in the feedstock while maintaining coking and/or reactor plugging at a reduced or minimized level. It has been discovered that increasing the quantity of aromatics in a feedstock can reduce the amount of coking. Without being bound by any particular theory, it is believed that increasing the aromatics content can provide an increased solubility number (S_{BN}) for a feedstock. It is further believed that increased coking and/or reactor plugging often occurs when the I_N of the feedstock / products in a reactor environment approaches near to (or possibly exceeds) the S_{BN} value of the feedstock / products. Increasing the S_{BN} of a feedstock can reduce the likelihood that the S_{BN} will approach near to the insolubility number (I_N) of the feedstock / products in the hydroprocessing environment.

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[0018] Some additional difficulties in processing heavy cracked feeds can be related to difficulties in performing distillation on the feeds. Conventionally, one of the strategies for processing a challenging feedstock can be to use distillation to separate a more favorable portion of a feed from a typically higher boiling less favorable portion. Under such a conventional strategy, an atmospheric distillation can be used to separate a feed into lower boiling portions and a higher boiling portion at a distillation cut point between about 600°F (~316°C) and about 700°F (~371°C). The higher boiling portion can then correspond to a roughly 316°C+ portion, or a roughly 343°C+ portion, or a roughly 371°C+ portion. Conventionally, a further distillation can be performed on this higher boiling portion under reduced pressure or vacuum distillation conditions. This can produce one or more vacuum distillate fractions and a bottoms fraction. Unfortunately, heavy cracked feeds such as catalytic slurry oils can often have a density of about 1.04 g/cm³ or more, or about 1.06 g/cm³ or more, or about 1.08 g/cm³ or more, such as up to 1.14 g/cm³ or possibly still higher. At such higher density values, performing a vacuum distillation under conventional vacuum distillation conditions becomes increasingly difficult and/or inefficient. In particular, such high density fractions can tend to have poor separation characteristics under conventional vacuum distillation conditions. As a result, either substantial amounts of undesirable components can remain in the “desired” distillate fraction(s), and/or substantial amounts of the desired components can remain in the bottoms fraction.

[0019] In some aspects, one or more of the difficulties in processing a heavy cracked feed can be reduced or mitigated by first performing hydroprocessing on a heavy cracked feed, and then using distillation to separate the resulting hydroprocessed effluent. After hydroprocessing, the 343°C+ portion of the hydroprocessed effluent can have a reduced density, such as 1.02 g/cm³ or less, or 1.0 g/cm³ or less, or 0.99 g/cm³ or less. This can allow the 343°C+ portion of the hydroprocessed effluent to be readily separated under conventional vacuum distillation conditions. By performing an initial hydroprocessing step, the 343°C+ portion of the hydroprocessed effluent can be separated to form various lower boiling fractions and a bottoms fraction.

[0020] After an initial hydroprocessing, a separation can be performed at any convenient cut point to assist with modifying the conditions for equilibrium conversion / formation of aromatics. For example, one convenient type of separation can be to perform a separation at a distillation cut point of about 454°C. Practically, this may lead to formation of a first fraction with a T90 or T95 distillation point of roughly 454°C and a second fraction with a T5 or T10 distillation point of roughly 454°C. At a distillation cut point of about 454°C, a substantial majority of the 4-ring naphthenes in the hydroprocessed effluent can be separated into the lower boiling fraction, along

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with a portion of the 5-ring naphthenes. By contrast, a majority of the 4-ring aromatics can be separated into the higher boiling fraction. Different processing strategies can then be used for the lower and higher boiling fractions. For example, if additional hydroprocessing is desired for the lower boiling fraction, lower processing temperatures can be used, as fewer of the larger multi-ring aromatic compounds remain in the lower boiling fraction. For the higher boiling fraction, the same (or higher severity) processing conditions can be used, since the equilibrium will now drive additional conversion of the aromatics in the higher boiling fraction toward formation of the now depleted naphthenes.

[0021] The processing and/or other use of the various lower boiling fractions can also be different from the further processing of the bottoms fraction. This can allow, for example, recycle of the bottoms fraction for use as part of the initial feedstock, which can further enhance the aromatics content of the feedstock. Additionally or alternately, the hydroprocessed bottoms can be used as part of a feed for fluid catalytic cracking.

[0022] In some additional or alternative aspects, one or more of the difficulties in processing a heavy cracked feed can be reduced or mitigated by first performing hydroprocessing on a heavy cracked feed, and then performing a solvent-based separation on the hydroprocessed effluent. During hydroprocessing, the amount of feed conversion relative to 566°C will typically be substantially less than 100%. For example, the conversion relative to 566°C can be 30 wt% to 80 wt%. This means that 20 wt% to 70 wt% of the 566°C+ portion of the feed is unconverted. Such an unconverted portion of the feed can include substantial amounts of polynuclear aromatics. For a cracked feed such as a catalytic slurry oil, unconverted polynuclear aromatics can also be present in lower boiling portions of the hydroprocessed effluent. A solvent separation process can allow polynuclear aromatics to be selectively removed from the hydroprocessed effluent for recycle while allowing higher boiling, non-aromatic compounds to be selectively passed on to subsequent processing stages and/or uses. The recycled polynuclear aromatics can then be hydroprocessed again. This can provide various advantages, including increasing the aromatic content (and therefore S_{BN}) of the feedstock, and decreasing the severity of hydroprocessing that is needed to remove lower value polynuclear aromatics while still (eventually) allowing substantially complete conversion of the polynuclear aromatics to higher value compounds.

[0023] Separation, recycle, and substantially complete conversion of polynuclear aromatics is generally understood by those skilled in the art to be impractical using conventional methods. Conventionally, polynuclear aromatic (PNA) recycle is understood to result in the accumulation of incompatible PNAs that accelerate catalyst deactivation/coking, plug reactor beds, and foul equipment with carbonaceous deposits. It has been discovered, however, that using feeds with

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high S_{BN} values can reduce or minimize the conventional difficulties that occur when attempting to recycle PNAs. It has further been unexpectedly discovered that mixtures of PNAs and naphthenes distill more readily than pure PNAs.

[0024] As defined herein, the term “hydrocarbonaceous” includes compositions or fractions that contain hydrocarbons and hydrocarbon-like compounds that may contain heteroatoms typically found in petroleum or renewable oil fraction and/or that may be typically introduced during conventional processing of a petroleum fraction. Heteroatoms typically found in petroleum or renewable oil fractions include, but are not limited to, sulfur, nitrogen, phosphorous, and oxygen. Other types of atoms different from carbon and hydrogen that may be present in a hydrocarbonaceous fraction or composition can include alkali metals as well as trace transition metals (such as Ni, V, or Fe).

[0025] In some aspects, reference may be made to conversion of a feedstock relative to a conversion temperature. Conversion relative to a temperature can be defined based on the portion of the feedstock that boils at greater than the conversion temperature. The amount of conversion during a process (or optionally across multiple processes) can correspond to the weight percentage of the feedstock converted from boiling above the conversion temperature to boiling below the conversion temperature. As an illustrative hypothetical example, consider a feedstock that includes 40 wt% of components that boil at 700°F (~371°C) or greater. By definition, the remaining 60 wt% of the feedstock boils at less than 700°F (~371°C). For such a feedstock, the amount of conversion relative to a conversion temperature of ~371°C would be based only on the 40 wt% that initially boils at ~371°C or greater. If such a feedstock could be exposed to a process with 30% conversion relative to a ~371°C conversion temperature, the resulting product would include 72 wt% of ~371°C- components and 28 wt% of ~371°C+ components.

[0026] In various aspects, reference may be made to one or more types of fractions generated during distillation of a feedstock or effluent. Such fractions may include naphtha fractions, kerosene fractions, diesel fractions, and other heavier (gas oil) fractions. Each of these types of fractions can be defined based on a boiling range, such as a boiling range that includes at least ~90 wt% of the fraction, or at least ~95 wt% of the fraction. For example, for many types of naphtha fractions, at least ~90 wt% of the fraction, or at least ~95 wt%, can have a boiling point in the range of ~85°F (~29°C) to ~350°F (~177°C). For some heavier naphtha fractions, at least ~90 wt% of the fraction, and preferably at least ~95 wt%, can have a boiling point in the range of ~85°F (~29°C) to ~400°F (~204°C). For a kerosene fraction, at least ~90 wt% of the fraction, or at least ~95 wt%, can have a boiling point in the range of ~300°F (~149°C) to ~600°F (~288°C).

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For a kerosene fraction targeted for some uses, such as jet fuel production, at least ~90 wt% of the fraction, or at least ~95 wt%, can have a boiling point in the range of ~300°F (~149°C) to ~550°F (~288°C). For a diesel fraction, at least ~90 wt% of the fraction, and preferably at least ~95 wt%, can have a boiling point in the range of ~350°F (~177°C) to ~700°F (~371°C). For a (vacuum) gas oil fraction, at least ~90 wt% of the fraction, and preferably at least ~95 wt%, can have a boiling point in the range of ~650°F (~343°C) to ~1100°F (~593°C). Optionally, for some gas oil fractions, a narrower boiling range may be desirable. For such gas oil fractions, at least ~90 wt% of the fraction, or at least ~95 wt%, can have a boiling point in the range of ~650°F (~343°C) to ~1000°F (~538°C), or ~650°F (~343°C) to ~900°F (~482°C). A residual fuel product can have a boiling range that may vary and/or overlap with one or more of the above boiling ranges. A residual marine fuel product can satisfy the requirements specified in ISO 8217, Table 2.

[0027] A method of characterizing the solubility properties of a petroleum fraction can correspond to the toluene equivalence (TE) of a fraction, based on the toluene equivalence test as described for example in U.S. Patent 5,871,634 (incorporated herein by reference with regard to the definition for toluene equivalence, solubility number (S_{BN}), and insolubility number (I_N)). The calculated carbon aromaticity index (CCAI) can be determined according to ISO 8217. BMCI can refer to the Bureau of Mines Correlation Index, as commonly used by those of skill in the art.

[0028] In this discussion, the effluent from a processing stage may be characterized in part by characterizing a fraction of the products. For example, the effluent from a processing stage may be characterized in part based on a portion of the effluent that can be converted into a liquid product. This can correspond to a C_3+ portion of an effluent, and may also be referred to as a total liquid product. As another example, the effluent from a processing stage may be characterized in part based on another portion of the effluent, such as a C_5+ portion or a C_6+ portion. In this discussion, a portion corresponding to a " C_x+ " portion can be, as understood by those of skill in the art, a portion with an initial boiling point that roughly corresponds to the boiling point for an aliphatic hydrocarbon containing " x " carbons.

[0029] In this discussion, a low sulfur fuel oil can correspond to a fuel oil containing about 0.5 wt% or less of sulfur. An ultra low sulfur fuel oil, which can also be referred to as an Emission Control Area fuel, can correspond to a fuel oil containing about 0.1 wt% or less of sulfur. A low sulfur diesel can correspond to a diesel fuel containing about 500 wppm or less of sulfur. An ultra low sulfur diesel can correspond to a diesel fuel containing about 15 wppm or less of sulfur, or about 10 wppm or less.

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[0030] In this discussion, reference may be made to catalytic slurry oil, FCC bottoms, and main column bottoms. These terms can be used interchangeably herein. It is noted that when initially formed, a catalytic slurry oil can include several weight percent of catalyst fines. Any such catalyst fines can be removed prior to incorporating a fraction derived from a catalytic slurry oil into a product pool, such as a naphtha fuel pool or a diesel fuel pool. In this discussion, unless otherwise explicitly noted, references to a catalytic slurry oil are defined to include catalytic slurry oil either prior to or after such a process for reducing the content of catalyst fines within the catalytic slurry oil.

Feedstocks for Hydroprocessing – Cracked Fractions

[0031] A catalytic slurry oil is an example of a suitable cracked fraction for incorporation into a feedstock. It is conventionally understood that conversion of $\sim 1050^{\circ}\text{F}+$ ($\sim 566^{\circ}\text{C}+$) vacuum resid fractions by hydroprocessing and/or hydrocracking can be limited by incompatibility. Under conventional understanding, at somewhere between ~ 30 wt% and ~ 55 wt% conversion of the $\sim 1050^{\circ}\text{F}+$ ($\sim 566^{\circ}\text{C}+$) portion, the reaction product during hydroprocessing can become incompatible with the feed. For example, as the $\sim 566^{\circ}\text{C}+$ feedstock converts to $\sim 1050^{\circ}\text{F}-$ ($\sim 566^{\circ}\text{C}-$) products, hydrogen transfer, oligomerization, and dealkylation reactions can occur which create molecules that are increasingly difficult to keep in solution. Somewhere between ~ 30 wt% and ~ 55 wt% $\sim 566^{\circ}\text{C}+$ conversion, a second liquid hydrocarbon phase separates. This new incompatible phase, under conventional understanding, can correspond to mostly polynuclear aromatics rich in N, S, and metals. The new incompatible phase can potentially be high in micro carbon residue (MCR). The new incompatible phase can stick to surfaces in the unit where it cokes and then can foul the equipment. Based on this conventional understanding, catalytic slurry oil can conventionally be expected to exhibit properties similar to a vacuum resid fraction during hydroprocessing. A catalytic slurry oil can have an I_N of about 70 to about 130, ~ 1 -6 wt% n-heptane insolubles, a density of 1.04 g/cm^3 or more, or 1.06 g/cm^3 or more, and a boiling range profile that includes about 3 wt% to about 12 wt% or less of $\sim 566^{\circ}\text{C}+$ material. Based on the above conventional understanding, it can be expected that hydroprocessing of a catalytic slurry oil would cause incompatibility as the asphaltenes and/or $\sim 566^{\circ}\text{C}+$ material converts.

[0032] In contrast to conventional understanding, it has been discovered that hydroprocessing can be performed while reducing or minimizing the above difficulties by using a feed composed of a substantial portion of a catalytic slurry oil, with a minor amount (or less) of a conventional vacuum resid feed. A catalytic slurry oil can be processed as part of a feed where the catalytic slurry oil corresponds to at least about 25 wt% of the feed to a process for forming fuels, such as

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at least about 50 wt%, at least about 75 wt%, at least about 90 wt%, or at least about 95 wt%. Optionally, the feed can correspond to at least about 99 wt% of a catalytic slurry oil, therefore corresponding to a feed that consists essentially of catalytic slurry oil. In particular, a feed can comprise about 25 wt% to about 100 wt% catalytic slurry oil, or about 25 wt% to about 99 wt%, or about 50 wt% to about 90 wt%. In contrast to many types of potential feeds for production of fuels, the asphaltenes in a catalytic slurry oil can apparently be converted on a time scale comparable to the time scale for conversion of other aromatic compounds in the catalytic slurry oil. In other words, without being bound by any particular theory, the asphaltene-type compounds in a catalytic slurry oil that are susceptible to precipitation / insolubility can be converted at a proportional rate to the conversion of compounds that help to maintain solubility of asphaltene-type compounds. This can have the effect that during hydroprocessing, the rate of decrease of the S_{BN} for the catalytic slurry oil can be similar to the rate of decrease of I_N , so that precipitation of asphaltenes during processing can be reduced, minimized, or eliminated. As a result, it has been unexpectedly discovered that catalytic slurry oil can be processed at effective hydroprocessing conditions for substantial conversion of the feed without causing excessive coking of the catalyst. This can allow hydroprocessing to be used to at least partially break down the ring structures of the aromatic cores in the catalytic slurry oil. In a sense, hydroprocessing of a catalytic slurry oil as described herein can serve as a type of “hydrodeasphalting”, where the asphaltene type compounds are removed by hydroprocessing rather than by solvent extraction. In various aspects, the 566°C+ conversion during hydroprocessing for a feed including catalytic slurry oil can be at least 55 wt%, or at least 65 wt%, or at least 75 wt%, such as up to about 95 wt% or still higher.

[0033] Typically the cut point for forming a catalytic slurry oil can be at least about 650°F (~343°C). As a result, a catalytic slurry oil can have a T5 distillation (boiling) point or a T10 distillation point of at least about 288°C, or at least about 316°C, or at least about 650°F (~343°C), as measured according to ASTM D2887. In some aspects the D2887 10% distillation point (T10) can be greater, such as at least about 675°F (~357°C), or at least about 700°F (~371°C). In some aspects, a broader boiling range portion of FCC products can be used as a feed (e.g., a 350°F+ / ~177°C+ boiling range fraction of FCC liquid product), where the broader boiling range portion includes a 650°F+ (~343°C+) fraction that corresponds to a catalytic slurry oil. The catalytic slurry oil (650°F+ / ~343°C+) fraction of the feed does not necessarily have to represent a “bottoms” fraction from an FCC process, so long as the catalytic slurry oil portion comprises one or more of the other feed characteristics described herein.

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[0034] In addition to and/or as an alternative to initial boiling points, T5 distillation point, and/or T10 distillation points, other distillation points may be useful in characterizing a feedstock. For example, a feedstock can be characterized based on the portion of the feedstock that boils above 1050°F (~566°C). In some aspects, a feedstock (or alternatively a 650°F+ / ~343°C+ portion of a feedstock) can have an ASTM D2887 T95 distillation point of 1050°F (~566°C) or greater, or a T90 distillation point of 1050°F (~566°C) or greater. If a feedstock or other sample contains components that are not suitable for characterization using D2887, ASTM D1160 may be used instead for such components.

[0035] In various aspects, density, or weight per volume, of the catalytic slurry oil can be characterized. The density of the catalytic slurry oil (or alternatively a 650°F+ / ~343°C+ portion of a feedstock) can be at least about 1.06 g/cc, or at least about 1.08 g/cc, or at least about 1.10 g/cc, such as up to about 1.20 g/cc. The density of the catalytic slurry oil can provide an indication of the amount of heavy aromatic cores that are present within the catalytic slurry oil.

[0036] Contaminants such as nitrogen and sulfur are typically found in catalytic slurry oils, often in organically-bound form. Nitrogen content can range from about 50 wppm to about 5000 wppm elemental nitrogen, or about 100 wppm to about 2000 wppm elemental nitrogen, or about 250 wppm to about 1000 wppm, based on total weight of the catalytic slurry oil. The nitrogen containing compounds can be present as basic or non-basic nitrogen species. Examples of nitrogen species can include quinolones, substituted quinolones, carbazoles, and substituted carbazoles.

[0037] The sulfur content of a catalytic slurry oil feed can be at least about 500 wppm elemental sulfur, based on total weight of the catalytic slurry oil. Generally, the sulfur content of a catalytic slurry oil can range from about 500 wppm to about 100,000 wppm elemental sulfur, or from about 1000 wppm to about 50,000 wppm, or from about 1000 wppm to about 30,000 wppm, based on total weight of the heavy component. Sulfur can usually be present as organically bound sulfur. Examples of such sulfur compounds include the class of heterocyclic sulfur compounds such as thiophenes, tetrahydrothiophenes, benzothiophenes and their higher homologs and analogs. Other organically bound sulfur compounds include aliphatic, naphthenic, and aromatic mercaptans, sulfides, di- and polysulfides.

[0038] Catalytic slurry oils can include n-heptane insolubles (NHI) or asphaltenes. In some aspects, the catalytic slurry oil feed (or alternatively a ~650°F+ / ~343°C+ portion of a feed) can contain at least about 1.0 wt% of n-heptane insolubles or asphaltenes, or at least about 2.0 wt%, or at least about 3.0 wt%, or at least about 5.0 wt%, such as up to about 10 wt% or more. In particular, the catalytic slurry oil feed (or alternatively a ~343°C+ portion of a feed) can contain

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about 1.0 wt% to about 10 wt% of n-heptane insolubles or asphaltenes, or about 2.0 wt% to about 10 wt%, or about 3.0 wt% to about 10 wt%. Another option for characterizing the heavy components of a catalytic slurry oil can be based on the amount of micro carbon residue (MCR) in the feed. In various aspects, the amount of MCR in the catalytic slurry oil feed (or alternatively a $\sim 343^{\circ}\text{C}+$ portion of a feed) can be at least about 5 wt%, or at least about 8 wt%, or at least about 10 wt%, or at least about 12 wt%, such as up to about 20 wt% or more.

[0039] Based on the content of NHI and/or MCR in a catalytic slurry oil feed, the insolubility number (IN) for such a feed can be at least about 60, such as at least about 70, at least about 80, or at least about 90. Additionally or alternately, the IN for such a feed can be about 140 or less, such as about 130 or less, about 120 or less, about 110 or less, about 100 or less, about 90 or less, or about 80 or less. Each lower bound noted above for IN can be explicitly contemplated in conjunction with each upper bound noted above for IN. In particular, the IN for a catalytic slurry oil feed can be about 60 to about 140, or about 60 to about 120, or about 80 to about 140.

[0040] An additional favorable feature of hydroprocessing a catalytic slurry oil can be the increase in product volume that can be achieved. Due to the high percentage of aromatic cores in a catalytic slurry oil, hydroprocessing of catalytic slurry oil can result in substantial consumption of hydrogen. The additional hydrogen added to a catalytic slurry oil can result in an increase in volume for the hydroprocessed catalytic slurry oil or volume swell. For example, the amount of C_3+ liquid products generated from hydrotreatment and FCC processing of catalytic slurry oil can be greater than $\sim 100\%$ of the volume of the initial catalytic slurry oil. (A similar proportional increase in volume can be achieved for feeds that include only a portion of deasphalted catalytic slurry oil.) Hydroprocessing within the normal range of commercial hydrotreater operations can enable ~ 2000 - 4000 SCF/bbl (~ 340 Nm^3/m^3 to ~ 680 m^3/m^3) of hydrogen to be added to a feed corresponding to a deasphalted catalytic slurry oil. This can result in substantial conversion of a deasphalted catalytic slurry oil feed to $\sim 700^{\circ}\text{F}$ - ($\sim 371^{\circ}\text{C}$ -) products, such as at least about 40 wt% conversion to $\sim 371^{\circ}\text{C}$ - products, or at least about 50 wt%, or at least about 60 wt%, and up to about 90 wt% or more. In some aspects, the $\sim 371^{\circ}\text{C}$ - product can meet the requirements for a low sulfur diesel fuel blendstock in the U.S. Additionally or alternately, the $\sim 371^{\circ}\text{C}$ - product(s) can be upgraded by further hydroprocessing to a low sulfur diesel fuel or blendstock. The remaining $\sim 700^{\circ}\text{F}+$ ($\sim 371^{\circ}\text{C}+$) product can meet the normal specifications for a $< \sim 0.5$ wt% S bunker fuel or a $< \sim 0.1$ wt% S bunker fuel, and/or may be blended with a distillate range blendstock to produce a finished blend that can meet the specifications for a $< \sim 0.1$ wt% S bunker fuel. Additionally or alternately, a $\sim 343^{\circ}\text{C}+$ product can be formed that can be suitable for use as a $< \sim 0.1$ wt% S bunker fuel without additional

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blending. The additional hydrogen for the hydrotreatment of the catalytic slurry oil can be provided from any convenient source.

[0041] Additionally or alternately, the remaining $\sim 371^{\circ}\text{C}+$ product (and/or portions of the $\sim 371^{\circ}\text{C}+$ product) can be used as feedstock to an FCC unit and cracked to generate additional LPG, gasoline, and diesel fuel, so that the yield of $\sim 371^{\circ}\text{C}-$ products relative to the total liquid product yield can be at least about 60 wt%, or at least about 70 wt%, or at least about 80 wt%. Relative to the feed, the yield of C_3+ liquid products can be at least about 100 vol%, such as at least about 105 vol%, at least about 110 vol%, at least about 115 vol%, or at least about 120 vol%. In particular, the yield of C_3+ liquid products can be about 100 vol% to about 150 vol%, or about 110 vol% to about 150 vol%, or about 120 vol% to about 150 vol%.

[0042] More generally, the systems and methods described herein can be used for processing feedstocks containing one or more types of cracked feeds that have a high density prior to hydroprocessing, such as a density of 1.04 g/cm^3 or more, or 1.06 g/cm^3 or more, or 1.08 g/cm^3 or more, such as up to 1.20 g/cm^3 or possibly still higher. Additionally or alternately, the feedstock including one or more cracked feeds can have an aromatics content of about 40 wt% to about 80 wt%, or about 40 wt% to about 70 wt%, or about 50 wt% to about 80 wt%. Additionally or alternately, the feedstock including one or more cracked feeds can have a S_{BN} of 100 to 250 and an I_{N} of 70 to 180, with the S_{BN} of the feedstock being greater than the I_{N} , the S_{BN} optionally being greater by at least 30, or at least 40, or at least 50. In addition to catalytic slurry oils, other types of cracked stocks include, but are not limited to, heavy coker gas oils (such coker bottoms), steam cracker tars, coal tars, and visbreaker gas oils.

[0043] For example, steam cracker tar (SCT) as used herein is also referred to in the art as "pyrolysis fuel oil". The terms can be used interchangeably herein. The tar will typically be obtained from the first fractionator downstream from a steam cracker (pyrolysis furnace) as the bottoms product of the fractionator, nominally having a boiling point of at least about $550^{\circ}\text{F}+$ ($\sim 288^{\circ}\text{C}+$). Boiling points and/or fractional weight distillation points can be determined by, for example, ASTM D2892. Alternatively, SCT can have a T5 boiling point (temperature at which 5 wt% will boil off) of at least about 550°F ($\sim 288^{\circ}\text{C}$). The final boiling point of SCT can be dependent on the nature of the initial pyrolysis feed and/or the pyrolysis conditions, and typically can be about 1450°F ($\sim 788^{\circ}\text{C}$) or less.

[0044] SCT can have a relatively low hydrogen content compared to heavy oil fractions that are typically processed in a refinery setting. In some aspects, SCT can have a hydrogen content of about 8.0 wt% or less, about 7.5 wt% or less, or about 7.0 wt% or less, or about 6.5 wt% or less. In particular, SCT can have a hydrogen content of about 5.5 wt% to about 8.0 wt%, or

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about 6.0 wt% to about 7.5 wt%. Additionally or alternately, SCT can have a micro carbon residue (or alternatively Conradson Carbon Residue) of at least about 10 wt%, or at least about 15 wt%, or at least about 20 wt%, such as up to about 40 wt% or more.

[0045] SCT can also be highly aromatic in nature. The paraffin content of SCT can be about 2.0 wt% or less, or about 1.0 wt% or less, such as having substantially no paraffin content. The naphthene content of SCT can also be about 2.0 wt% or less or about 1.0 wt% or less, such as having substantially no naphthene content. In some aspects, the combined paraffin and naphthene content of SCT can be about 1.0 wt% or less. With regard to aromatics, at least about 30 wt% of SCT can correspond to 3-ring aromatics, or at least 40 wt%. In particular, the 3-ring aromatics content can be about 30 wt% to about 60 wt%, or about 40 wt% to about 55 wt%, or about 40 wt% to about 50 wt%. Additionally or alternately, at least about 30 wt% of SCT can correspond to 4-ring aromatics, or at least 40 wt%. In particular, the 4-ring aromatics content can be about 30 wt% to about 60 wt%, or about 40 wt% to about 55 wt%, or about 40 wt% to about 50 wt%. Additionally or alternately, the 1-ring aromatic content can be about 15 wt% or less, or about 10 wt% or less, or about 5 wt% or less, such as down to about 0.1 wt%.

[0046] Due to the low hydrogen content and/or highly aromatic nature of SCT, the solubility number (S_{BN}) and insolubility number (I_N) of SCT can be relatively high. SCT can have a S_{BN} of at least about 100, and in particular about 120 to about 230, or about 150 to about 230, or about 180 to about 220. Additionally or alternately, SCT can have an I_N of about 70 to about 180, or about 100 to about 160, or about 80 to about 140. Further additionally or alternately, the difference between S_{BN} and I_N for the SCT can be at least about 30, or at least about 40, or at least about 50, such as up to about 150.

[0047] SCT can also have a higher density than many types of crude or refinery fractions. In various aspects, SCT can have a density at 15°C of about 1.08 g/cm³ to about 1.20 g/cm³, or 1.10 g/cm³ to 1.18 g/cm³. By contrast, many types of vacuum resid fractions can have a density of about 1.05 g/cm³ or less. Additionally or alternately, density (or weight per volume) of the heavy hydrocarbon can be determined according to ASTM D287 - 92 (2006) Standard Test Method for API Gravity of Crude Petroleum and Petroleum Products (Hydrometer Method), which characterizes density in terms of API gravity. In general, the higher the API gravity, the less dense the oil. API gravity can be 5° or less, or 0° or less, such as down to about -10° or lower.

[0048] Contaminants such as nitrogen and sulfur are typically found in SCT, often in organically-bound form. Nitrogen content can range from about 50 wppm to about 10,000 wppm elemental nitrogen or more, based on total weight of the SCT. Sulfur content can range from about 0.1 wt% to about 10 wt%, based on total weight of the SCT.

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[0049] Coker bottoms represent another type of cracked feed suitable for hydroprocessing, optionally in combination with a catalytic slurry oil and/or steam cracker tar and/or other cracked fractions. Coking is a thermal cracking process that is suitable for conversion of heavy feeds into fuels boiling range products. The feedstock to a coker typically also includes 5 wt% to 25 wt% recycled product from the coker, which can be referred to as coker bottoms. This recycle fraction allows metals, asphaltenes, micro-carbon residue, and/or other solids to be returned to the coker, as opposed to being incorporated into a coker gas oil product. This can maintain a desired product quality for the coker gas oil product, but results in a net increase in the amount of light ends and coke that are generated by a coking process. The coker bottoms can correspond to a fraction with a T10 distillation point of at least 550°F (288°C), or at least 300°C, or at least 316°C, and a T90 distillation point of 566°C or less, or 550°C or less, or 538°C or less. The coker recycle fraction can have an aromatic carbon content of about 20 wt% to about 50 wt%, or about 30 wt% to about 45 wt%, and a micro carbon residue content of about 4.0 wt% to about 15 wt%, or about 6.0 wt% to about 15 wt%, or about 4.0 wt% to about 10 wt%, or about 6.0 wt% to about 12 wt%.

[0050] In some aspects, the weight percent of catalytic slurry oil in the feed can be greater than or equal to the amount of coker bottoms. The amount of coker bottoms in the feed can generally be from about 5 wt% to about 50 wt%, or about 10 wt% to about 50 wt%, or about 20 wt% to about 35 wt%. The amount of catalytic slurry oil in the feed can be about 20 wt% to about 95 wt%, or about 20 wt% to about 70 wt%, or about 40 wt% to about 95 wt%, or about 50 wt% to about 95 wt%. In aspects where the feed is deasphalted prior to hydroprocessing, the feed can optionally further include 5 wt% to 40 wt% of a vacuum resid fraction. The vacuum resid fraction can have a T10 distillation point of about 510°C or greater, or about 538°C or greater, or about 566°C or greater.

Feedstock Particle Removal

[0051] In some aspects, a feedstock including one or more cracked feeds can include various types of particles, such as catalyst fines present in a catalytic slurry oil and/or coke fines present in a steam cracker tar. Such particles can optionally be removed (such as partially removed to a desired level) by any convenient method, such as filtration. In some aspects, an improved method of removing particles from a blended feed can correspond to removing a portion of particles from the blended feed by settling, followed by using electrostatic filtration to remove additional particles. Additionally or alternately, particles can be removed from a blended feed using a filter with a relatively uniform porosity. Such a filter can optionally be used in conjunction with settling, so that larger particles can be removed from the feed prior to filtration by the filter.

[0052] Settling can provide a convenient method for removing larger particles from a feed. During a settling process, a feed can be held in a settling tank or other vessel for a period of time. This time period can be referred to as a settling time. The feed can be at a settling temperature during the settling time. While any convenient settling temperature can potentially be used (such as a temperature from about 20°C to about 200°C), a temperature of about 100°C or greater (such as at least 105°C, or at least 110°C) can be beneficial for allowing the viscosity of the blended feed to be low enough to facilitate settling. Additionally or alternately, the settling temperature can be about 200°C or less, or about 150°C or less, or about 140°C or less. In particular, the settling temperature can be about 100°C to about 200°C, or about 105°C to about 150°C, or about 110°C to about 140°C. The upper end of the settling temperature can be less important, and temperatures of still greater than 200°C may also be suitable.

[0053] After the settling time, the particles can be concentrated in a lower portion of the settling tank. The blended feed including a portion of catalytic slurry oil and a portion of steam cracker tar can be removed from the upper portion of the settling tank while leaving the particle enriched bottoms in the tank. The settling process can be suitable for reducing the concentration of particles having a particle size of about 25 µm or greater from the blended feed.

[0054] After removing the larger particles from the blended feed, the blended feed can be passed into an electrostatic separator. An example of a suitable electrostatic separator can be a Gulftronic™ electrostatic separator available from General Atomic. An electrostatic separator can be suitable for removal of particles of a variety of sizes, including both larger particles as well as particles down to a size of about 5 µm or less or even smaller. However, it can be beneficial to remove larger particles using a settling process to reduce or minimize the accumulation of large particles in an electrostatic separator. This can reduce the amount of time required for flush and regeneration of an electrostatic separator.

[0055] In an electrostatic separator, dielectric beads within the separator can be charged to polarize the dielectric beads. A fluid containing particles for removal can then be passed into the electrostatic separator. The particles can be attracted to the dielectric beads, allowing for particle removal. After a period of time, the electrostatic separator can be flushed to allow any accumulated particles in the separator to be removed.

[0056] In various aspects, an electrostatic separator can be used in combination with a settling tank for particle removal. Performing electrostatic separation on an blended feed effluent from a settling tank can allow for reduction of the number of particles in a blended feed to about 500 wppm or less, or about 100 wppm or less, or about 50 wppm or less, such as down to about 20 wppm or possibly lower. In particular, the concentration of particles in the blended feed after

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electrostatic separation can be about 0 wppm to about 500 wppm, or about 0 wppm to about 100 wppm, or about 0 wppm to about 50 wppm, or about 1 wppm to about 20 wppm. In some aspects, a single electrostatic separation stage can be used to reduce the concentration of particles in the blended feed to a desired level. In some aspects, two or more electrostatic separation stages in series can be used to achieve a target particle concentration.

[0057] Another option for removal of particles from a feed, optionally after settling, can be to use a filter with a relatively uniform porosity. Examples of such filters correspond to the HyPulse® LSI filters available from Mott Corporation of Farmington, CT. Such filters can be formed using a sintered metal fabrication technology that can allow for accurate control of porosity. Having uniform porosity can assist with having particulates form a cake on the inside of the filter while reducing or minimizing the amount of particulates that penetrate into the screen and/or reducing or minimizing the amount of cake-bridging between filter elements.

Additional Feedstocks

[0058] In some aspects, at least a portion of a feedstock for processing as described herein can correspond to a vacuum resid fraction or another type 950°F+ (510°C+) or 1000°F+ (538°C+) fraction. Another example of a method for forming a 950°F+ (510°C+) or 1000°F+ (538°C+) fraction is to perform a high temperature flash separation. The 950°F+ (510°C+) or 1000°F+ (538°C+) fraction formed from the high temperature flash can be processed in a manner similar to a vacuum resid.

[0059] A vacuum resid fraction or a 950°F+ (510°C+) fraction formed by another process (such as a flash fractionation bottoms or a bitumen fraction) can be deasphalted at low severity to form a deasphalted oil. Optionally, the feedstock can also include a portion of a conventional feed for lubricant base stock production, such as a vacuum gas oil.

[0060] A vacuum resid (or other 510°C+) fraction can correspond to a fraction with a T5 distillation point (ASTM D2892, or ASTM D7169 if the fraction will not completely elute from a chromatographic system) of at least about 900°F (482°C), or at least 950°F (510°C), or at least 1000°F (538°C). Alternatively, a vacuum resid fraction can be characterized based on a T10 distillation point (ASTM D2892 / D7169) of at least about 900°F (482°C), or at least 950°F (510°C), or at least 1000°F (538°C).

[0061] Resid (or other 510°C+) fractions can be high in metals. For example, a resid fraction can be high in total nickel, vanadium and iron contents. In an aspect, a resid fraction can contain at least 0.00005 grams of Ni/V/Fe (50 wppm) or at least 0.0002 grams of Ni/V/Fe (200 wppm) per gram of resid, on a total elemental basis of nickel, vanadium and iron. In other aspects, the

heavy oil can contain at least 500 wppm of nickel, vanadium, and iron, such as up to 1000 wppm or more.

[0062] Contaminants such as nitrogen and sulfur are typically found in resid (or other 510°C+) fractions, often in organically-bound form. Nitrogen content can range from about 50 wppm to about 10,000 wppm elemental nitrogen or more, based on total weight of the resid fraction. Sulfur content can range from 500 wppm to 100,000 wppm elemental sulfur or more, based on total weight of the resid fraction, or from 1000 wppm to 50,000 wppm, or from 1000 wppm to 30,000 wppm.

[0063] Still another method for characterizing a resid (or other 510°C+) fraction is based on the Conradson carbon residue (CCR) of the feedstock. The Conradson carbon residue of a resid fraction can be at least about 5 wt%, such as at least about 10 wt% or at least about 20 wt%. Additionally or alternately, the Conradson carbon residue of a resid fraction can be about 50 wt% or less, such as about 40 wt% or less or about 30 wt% or less.

Hydroprocessing of Feedstock Including One or More Cracked Fractions

[0064] A feedstock including one or more cracked fractions can be hydroprocessed to form a hydroprocessed effluent. This can include hydrotreatment and/or hydrocracking to remove heteroatoms (such as sulfur and/or nitrogen) to desired levels, reduce Conradson Carbon content, and/or provide viscosity index (VI) uplift. Additionally or alternately, the hydroprocessing can be performed to achieve a desired level of conversion of higher boiling compounds in the feed to fuels boiling range compounds. Depending on the aspect, a feedstock can be hydroprocessed by demetallization, aromatics saturation, hydrotreating, hydrocracking, or a combination thereof.

[0065] In various aspects, the aromatics content of the feedstock can be at least 50 wt%, or at least 55 wt%, or at least 60 wt%, or at least 65 wt%, or at least 70 wt%, or at least 75 wt%, such as up to 90 wt% or more. Additionally or alternately, the saturates content of the feedstock can be 50 wt% or less, or 45 wt% or less, or 40 wt% or less, or 35 wt% or less, or 30 wt% or less, or 25 wt% or less, such as down to 10 wt% or less. In this discussion and the claims below, the aromatics content and/or the saturates content of a fraction can be determined based on ASTM D7419.

[0066] Depending on the aspect, the hydroprocessing can be performed in a configuration including at least one hydroprocessing stage with recycle of an aromatics-enriched stream as part of the feedstock, or in a configuration with multiple hydroprocessing stages. The reaction conditions during demetallization and/or hydrotreatment and/or hydrocracking of the feedstock can be selected to generate a desired level of conversion of a feed. Any convenient type of reactor, such as fixed bed (for example trickle bed) reactors can be used. Conversion of the feed

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can be defined in terms of conversion of molecules that boil above a temperature threshold to molecules below that threshold. The conversion temperature can be any convenient temperature, such as ~700°F (371°C) or 1050°F (566°C). The amount of conversion can correspond to the total conversion of molecules within the combined hydrotreatment and hydrocracking stages. Suitable amounts of conversion of molecules boiling above 1050°F (566°C) to molecules boiling below 566°C include 30 wt% to 100 wt% conversion relative to 566°C, or 30 wt% to 90 wt%, or 30 wt% to 70 wt%, or 40 wt% to 90 wt%, or 40 wt% to 80 wt%, or 40 wt% to 70 wt%, or 50 wt% to 100 wt%, or 50 wt% to 90 wt%, or 50 wt% to 70 wt%. In particular, the amount of conversion relative to 566°C can be 30 wt% to 100 wt%, or 50 wt% to 100 wt%, or 40 wt% to 90 wt%. Additionally or alternately, suitable amounts of conversion of molecules boiling above ~700°F (371°C) to molecules boiling below 371°C include 10 wt% to 70 wt% conversion relative to 371°C, or 10 wt% to 60 wt%, or 10 wt% to 50 wt%, or 20 wt% to 70 wt%, or 20 wt% to 60 wt%, or 20 wt% to 50 wt%, or 30 wt% to 70 wt%, or 30 wt% to 60 wt%, or 30 wt% to 50 wt%. In particular, the amount of conversion relative to 371°C can be 10 wt% to 70 wt%, or 20 wt% to 50 wt%, or 30 wt% to 60 wt%.

[0067] The hydroprocessed effluent can also be characterized based on the product quality. After hydroprocessing (hydrotreating and/or hydrocracking), the liquid (C₃+) portion of the hydroprocessed deasphalted oil / hydroprocessed effluent can have a sulfur content of about 1000 wppm or less, or about 500 wppm or less, or about 100 wppm or less (such as down to ~0 wppm). Additionally or alternately, the hydroprocessed deasphalted oil / hydroprocessed effluent can have a nitrogen content of 200 wppm or less, or 100 wppm or less, or 50 wppm or less (such as down to ~0 wppm). Additionally or alternately, the liquid (C₃+) portion of the hydroprocessed deasphalted oil / hydroprocessed effluent can have a MCR content and/or Conradson Carbon residue content of 2.5 wt% or less, or 1.5 wt% or less, or 1.0 wt% or less, or 0.7 wt% or less, or 0.1 wt% or less, or 0.02 wt% or less (such as down to ~0 wt%). MCR content and/or Conradson Carbon residue content can be determined according to ASTM D4530. Further additionally or alternately, the effective hydroprocessing conditions can be selected to allow for reduction of the n-heptane asphaltene content of the liquid (C₃+) portion of the hydroprocessed deasphalted oil / hydroprocessed effluent to less than about 1.0 wt%, or less than about 0.5 wt%, or less than about 0.1 wt%, and optionally down to substantially no remaining n-heptane asphaltenes. The hydrogen content of the liquid (C₃+) portion of the hydroprocessed deasphalted oil / hydroprocessed effluent can be at least about 10.5 wt%, or at least about 11.0 wt%, or at least about 11.5 wt%, such as up to about 13.5 wt% or more.

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[0068] In aspects where the feedstock includes catalytic slurry oil, coker bottoms, and/or steam cracker tar, the I_N of the hydroprocessed effluent can be at least 5 lower than the I_N of the feedstock prior to hydroprocessing, or at least 10 lower.

[0069] After hydroprocessing, the liquid (C_3+) portion of the hydroprocessed effluent can have a volume of at least about 95% of the volume of the corresponding feed to hydroprocessing, or at least about 100% of the volume of the feed, or at least about 105%, or at least about 110%, such as up to about 150% of the volume. In particular, the yield of C_3+ liquid products can be about 95 vol% to about 150 vol%, or about 110 vol% to about 150 vol%. Optionally, the C_3 and C_4 hydrocarbons can be used, for example, to form liquefied propane or butane gas as a potential liquid product. Therefore, the C_3+ portion of the effluent can be counted as the “liquid” portion of the effluent product, even though a portion of the compounds in the liquid portion of the hydrotreated effluent may exit the hydrotreatment reactor (or stage) as a gas phase at the exit temperature and pressure conditions for the reactor.

[0070] In some aspects, the portion of the hydroprocessed effluent having a boiling range / distillation point of less than about 700°F (~371°C) can be used as a low sulfur fuel oil or blendstock for low sulfur fuel oil. In other aspects, such a portion of the hydroprocessed effluent can be used (optionally with other distillate streams) to form ultra low sulfur naphtha and/or distillate (such as diesel) fuel products, such as ultra low sulfur fuels or blendstocks for ultra low sulfur fuels. The portion having a boiling range / distillation point of at least about 700°F (~371°C) can be used as an ultra low sulfur fuel oil having a sulfur content of about 0.1 wt% or less or optionally blended with other distillate or fuel oil streams to form an ultra low sulfur fuel oil or a low sulfur fuel oil. In some aspects, at least a portion of the liquid hydrotreated effluent having a distillation point of at least about ~371°C can be used as a feed for FCC processing. In still other aspects, the portion having a boiling range / distillation point of at least about 371°C can be used as a feedstock for lubricant base oil production.

[0071] Optionally, a feed can initially be exposed to a demetallization catalyst prior to exposing the feed to a hydrotreating catalyst. Deasphalted oils can have metals concentrations (Ni + V + Fe) on the order of 10 – 100 wppm. A combined catalytic slurry oil / coker bottoms feed can include still higher levels of metals. Exposing a conventional hydrotreating catalyst to a feed having a metals content of 10 wppm or more can lead to catalyst deactivation at a faster rate than may be desirable in a commercial setting. Exposing a metal containing feed to a demetallization catalyst prior to the hydrotreating catalyst can allow at least a portion of the metals to be removed by the demetallization catalyst, which can reduce or minimize the deactivation of the hydrotreating catalyst and/or other subsequent catalysts in the process flow.

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Commercially available demetallization catalysts can be suitable, such as large pore amorphous oxide catalysts that may optionally include Group VI and/or Group VIII non-noble metals to provide some hydrogenation activity.

[0072] In various aspects, the feedstock can be exposed to a hydrotreating catalyst under effective hydrotreating conditions. The catalysts used can include conventional hydroprocessing catalysts, such as those comprising at least one Group VIII non-noble metal (Columns 8 – 10 of IUPAC periodic table), preferably Fe, Co, and/or Ni, such as Co and/or Ni; and at least one Group VI metal (Column 6 of IUPAC periodic table), preferably Mo and/or W. Such hydroprocessing catalysts optionally include transition metal sulfides that are impregnated or dispersed on a refractory support or carrier such as alumina and/or silica. The support or carrier itself typically has no significant/measurable catalytic activity. Substantially carrier- or support-free catalysts, commonly referred to as bulk catalysts, generally have higher volumetric activities than their supported counterparts.

[0073] The catalysts can either be in bulk form or in supported form. In addition to alumina and/or silica, other suitable support/carrier materials can include, but are not limited to, zeolites, titania, silica-titania, and titania-alumina. Suitable aluminas are porous aluminas such as gamma or eta having average pore sizes from 50 to 200 Å, or 75 to 150 Å (as determined by ASTM D4284); a surface area (as measured by the BET method) from 100 to 300 m²/g, or 150 to 250 m²/g; and a pore volume of from 0.25 to 1.0 cm³/g, or 0.35 to 0.8 cm³/g. More generally, any convenient size, shape, and/or pore size distribution for a catalyst suitable for hydrotreatment of a distillate (including lubricant base stock) boiling range feed in a conventional manner may be used. Preferably, the support or carrier material is an amorphous support, such as a refractory oxide. Preferably, the support or carrier material can be free or substantially free of the presence of molecular sieve, where substantially free of molecular sieve is defined as having a content of molecular sieve of less than about 0.01 wt%.

[0074] The at least one Group VIII non-noble metal, in oxide form, can typically be present in an amount ranging from about 2 wt% to about 40 wt%, preferably from about 4 wt% to about 15 wt%. The at least one Group VI metal, in oxide form, can typically be present in an amount ranging from about 2 wt% to about 70 wt%, preferably for supported catalysts from about 6 wt% to about 40 wt% or from about 10 wt% to about 30 wt%. These weight percents are based on the total weight of the catalyst. Suitable metal catalysts include cobalt/molybdenum (1-10% Co as oxide, 10-40% Mo as oxide), nickel/molybdenum (1-10% Ni as oxide, 10-40% Co as oxide), or nickel/tungsten (1-10% Ni as oxide, 10-40% W as oxide) on alumina, silica, silica-alumina, or titania.

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[0075] The hydroprocessing is carried out in the presence of hydrogen. A hydrogen stream is, therefore, fed or injected into a vessel or reaction zone or hydroprocessing zone in which the hydroprocessing catalyst is located. Hydrogen, which is contained in a hydrogen “treat gas,” is provided to the reaction zone. Treat gas, as referred to herein, can be either pure hydrogen or a hydrogen-containing gas, which is a gas stream containing hydrogen in an amount that is sufficient for the intended reaction(s), optionally including one or more other gasses (e.g., nitrogen and light hydrocarbons such as methane). The treat gas stream introduced into a reaction stage will preferably contain at least about 50 vol. % and more preferably at least about 75 vol. % hydrogen. Optionally, the hydrogen treat gas can be substantially free (less than 1 vol%) of impurities such as H₂S and NH₃ and/or such impurities can be substantially removed from a treat gas prior to use.

[0076] Hydrogen can be supplied at a rate of from about 100 SCF/B (standard cubic feet of hydrogen per barrel of feed) (17 Nm³/m³) to about 10000 SCF/B (1700 Nm³/m³). Preferably, the hydrogen is provided in a range of from about 2000 SCF/B (340 Nm³/m³) to about 10000 SCF/B (1700 Nm³/m³). Hydrogen can be supplied co-currently with the input feed to the hydrotreatment reactor and/or reaction zone or separately via a separate gas conduit to the hydrotreatment zone.

[0077] The effective hydrotreating conditions can optionally be suitable for incorporation of a substantial amount of additional hydrogen into the hydrotreated effluent. During hydrotreatment, the consumption of hydrogen by the feed in order to form the hydrotreated effluent can correspond to at least about 1500 SCF/bbl (~260 Nm³/m³) of hydrogen, or at least about 1700 SCF/bbl (~290 Nm³/m³), or at least about 2000 SCF/bbl (~330 Nm³/m³), or at least about 2200 SCF/bbl (~370 Nm³/m³), such as up to about 5000 SCF/bbl (~850 Nm³/m³) or more. In particular, the consumption of hydrogen can be about 1500 SCF/bbl (~260 Nm³/m³) to about 5000 SCF/bbl (~850 Nm³/m³), or about 2000 SCF/bbl (~340 Nm³/m³) to about 5000 SCF/bbl (~850 Nm³/m³), or about 2200 SCF/bbl (~370 Nm³/m³) to about 5000 SCF/bbl (~850 Nm³/m³).

[0078] Hydrotreating conditions can include temperatures of 200°C to 450°C, or 315°C to 425°C; pressures of 250 psig (1.8 MPag) to 5000 psig (34.6 MPag) or 300 psig (2.1 MPag) to 3000 psig (20.8 MPag), or about 2.9 MPag to about 13.9 MPag (~400 to ~2000 psig); liquid hourly space velocities (LHSV) of 0.1 hr⁻¹ to 10 hr⁻¹, or 0.1 hr⁻¹ to 5.0 hr⁻¹; and a hydrogen treat gas rate of from about 430 to about 2600 Nm³/m³ (~2500 to ~15000 SCF/bbl), or about 850 to about 1700 Nm³/m³ (~5000 to ~10000 SCF/bbl).

[0079] In aspects where multiple hydroprocessing stages are used, a second (or subsequent) hydrotreatment stage can be operated with hydrotreating conditions that include a temperature

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that is 20°C – 100°C lower than a temperature associated with a first hydroprocessing stage; a pressure that is 1.5 MPag – 10 MPag lower than a pressure associated with a first hydroprocessing stage (or 1.5 MPag – 5 MPag); and/or a space velocity that is 0.2 hr⁻¹ – 2.0 hr⁻¹ greater than a space velocity associated with a first hydroprocessing stage. Optionally, a hydrotreating catalyst in a second stage can be the same as a hydroprocessing catalyst in a first stage.

[0080] In various aspects, the feedstock can be exposed to a hydrocracking catalyst under effective hydrocracking conditions. Hydrocracking catalysts typically contain sulfided base metals on acidic supports, such as amorphous silica alumina, cracking zeolites such as USY, or acidified alumina. Often these acidic supports are mixed or bound with other metal oxides such as alumina, titania or silica. Examples of suitable acidic supports include acidic molecular sieves, such as zeolites or silicoaluminophosphates. One example of suitable zeolite is USY, such as a USY zeolite with cell size of 24.30 Angstroms or less. Additionally or alternately, the catalyst can be a low acidity molecular sieve, such as a USY zeolite with a Si to Al ratio of at least about 20, and preferably at least about 40 or 50. ZSM-48, such as ZSM-48 with a SiO₂ to Al₂O₃ ratio of about 110 or less, such as about 90 or less, is another example of a potentially suitable hydrocracking catalyst. Still another option is to use a combination of USY and ZSM-48. Still other options include using one or more of zeolite Beta, ZSM-5, ZSM-35, or ZSM-23, either alone or in combination with a USY catalyst. Non-limiting examples of metals for hydrocracking catalysts include metals or combinations of metals that include at least one Group VIII metal, such as nickel, nickel-cobalt-molybdenum, cobalt-molybdenum, nickel-tungsten, nickel-molybdenum, and/or nickel-molybdenum-tungsten. Additionally or alternately, hydrocracking catalysts with noble metals can also be used. Non-limiting examples of noble metal catalysts include those based on platinum and/or palladium. Support materials which may be used for both the noble and non-noble metal catalysts can comprise a refractory oxide material such as alumina, silica, alumina-silica, kieselguhr, diatomaceous earth, magnesia, zirconia, or combinations thereof, with alumina, silica, alumina-silica being the most common (and preferred, in one embodiment).

[0081] When only one hydrogenation metal is present on a hydrocracking catalyst, the amount of that hydrogenation metal can be at least about 0.1 wt% based on the total weight of the catalyst, for example at least about 0.5 wt% or at least about 0.6 wt%. Additionally or alternately when only one hydrogenation metal is present, the amount of that hydrogenation metal can be about 5.0 wt% or less based on the total weight of the catalyst, for example about 3.5 wt% or less, about 2.5 wt% or less, about 1.5 wt% or less, about 1.0 wt% or less, about 0.9 wt% or less, about 0.75 wt% or less, or about 0.6 wt% or less. Further additionally or alternately when more

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than one hydrogenation metal is present, the collective amount of hydrogenation metals can be at least about 0.1 wt% based on the total weight of the catalyst, for example at least about 0.25 wt%, at least about 0.5 wt%, at least about 0.6 wt%, at least about 0.75 wt%, or at least about 1 wt%. Still further additionally or alternately when more than one hydrogenation metal is present, the collective amount of hydrogenation metals can be about 35 wt% or less based on the total weight of the catalyst, for example about 30 wt% or less, about 25 wt% or less, about 20 wt% or less, about 15 wt% or less, about 10 wt% or less, or about 5 wt% or less. In embodiments wherein the supported metal comprises a noble metal, the amount of noble metal(s) is typically less than about 2 wt %, for example less than about 1 wt%, about 0.9 wt % or less, about 0.75 wt % or less, or about 0.6 wt % or less. It is noted that hydrocracking under sour conditions is typically performed using a base metal (or metals) as the hydrogenation metal.

[0082] In various aspects, the conditions selected for hydrocracking can depend on the desired level of conversion, the level of contaminants in the input feed to the hydrocracking stage, and potentially other factors. For example, hydrocracking conditions in a single stage, or in the first stage and/or the second stage of a multi-stage system, can be selected to achieve a desired level of conversion in the reaction system. Hydrocracking conditions can be referred to as sour conditions or sweet conditions, depending on the level of sulfur and/or nitrogen present within a feed. For example, a feed with 100 wppm or less of sulfur and 50 wppm or less of nitrogen, preferably less than 25 wppm sulfur and/or less than 10 wppm of nitrogen, represent a feed for hydrocracking under sweet conditions.

[0083] A hydrocracking process under sour conditions can be carried out at temperatures of about 550°F (288°C) to about 840°F (449°C), hydrogen partial pressures of from about 1500 psig to about 5000 psig (10.3 MPag to 34.6 MPag), liquid hourly space velocities of from 0.05 h⁻¹ to 10 h⁻¹, and hydrogen treat gas rates of from 35.6 m³/m³ to 1781 m³/m³ (200 SCF/B to 10,000 SCF/B). In other embodiments, the conditions can include temperatures in the range of about 600°F (343°C) to about 815°F (435°C), hydrogen partial pressures of from about 1500 psig to about 3000 psig (10.3 MPag-20.9 MPag), and hydrogen treat gas rates of from about 213 m³/m³ to about 1068 m³/m³ (1200 SCF/B to 6000 SCF/B). The LHSV can be from about 0.25 h⁻¹ to about 50 h⁻¹, or from about 0.5 h⁻¹ to about 20 h⁻¹, preferably from about 1.0 h⁻¹ to about 4.0 h⁻¹.

[0084] In some aspects, a portion of the hydrocracking catalyst can be contained in a second reactor stage. In such aspects, a first reaction stage of the hydroprocessing reaction system can include one or more hydrotreating and/or hydrocracking catalysts. The conditions in the first reaction stage can be suitable for reducing the sulfur and/or nitrogen content of the feedstock. A separator can then be used in between the first and second stages of the reaction system to

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remove gas phase sulfur and nitrogen contaminants. One option for the separator is to simply perform a gas-liquid separation to remove contaminant. Another option is to use a separator such as a flash separator that can perform a separation at a higher temperature. Such a high temperature separator can be used, for example, to separate the feed into a portion boiling below a temperature cut point, such as about 350°F (177°C) or about 400°F (204°C), and a portion boiling above the temperature cut point. In this type of separation, the naphtha boiling range portion of the effluent from the first reaction stage can also be removed, thus reducing the volume of effluent that is processed in the second or other subsequent stages. Of course, any low boiling contaminants in the effluent from the first stage would also be separated into the portion boiling below the temperature cut point. If sufficient contaminant removal is performed in the first stage, the second stage can be operated as a “sweet” or low contaminant stage.

[0085] Still another option can be to use a separator between the first and second stages of the hydroprocessing reaction system that can also perform at least a partial fractionation of the effluent from the first stage. In this type of aspect, the effluent from the first hydroprocessing stage can be separated into at least a portion boiling below the distillate (such as diesel) fuel range, a portion boiling in the distillate fuel range, and a portion boiling above the distillate fuel range. The distillate fuel range can be defined based on a conventional diesel boiling range, such as having a lower end cut point temperature of at least about 350°F (177°C) or at least about 400°F (204°C) to having an upper end cut point temperature of about 700°F (371°C) or less or 650°F (343°C) or less. Optionally, the distillate fuel range can be extended to include additional kerosene, such as by selecting a lower end cut point temperature of at least about 300°F (149°C).

[0086] In aspects where the inter-stage separator is also used to produce a distillate fuel fraction, the portion boiling below the distillate fuel fraction includes, naphtha boiling range molecules, light ends, and contaminants such as H₂S. These different products can be separated from each other in any convenient manner. Similarly, one or more distillate fuel fractions can be formed, if desired, from the distillate boiling range fraction. The portion boiling above the distillate fuel range represents the potential lubricant base stocks. In such aspects, the portion boiling above the distillate fuel range is subjected to further hydroprocessing in a second hydroprocessing stage.

[0087] A hydrocracking process under sweet conditions can be performed under conditions similar to those used for a sour hydrocracking process, or the conditions can be different. In an embodiment, the conditions in a sweet hydrocracking stage can have less severe conditions than a hydrocracking process in a sour stage. Suitable hydrocracking conditions for a non-sour stage can include, but are not limited to, conditions similar to a first or sour stage. Suitable

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hydrocracking conditions can include temperatures of about 500°F (260°C) to about 840°F (449°C), hydrogen partial pressures of from about 1500 psig to about 5000 psig (10.3 MPag to 34.6 MPag), liquid hourly space velocities of from 0.05 h⁻¹ to 10 h⁻¹, and hydrogen treat gas rates of from 35.6 m³/m³ to 1781 m³/m³ (200 SCF/B to 10,000 SCF/B). In other embodiments, the conditions can include temperatures in the range of about 600°F (343°C) to about 815°F (435°C), hydrogen partial pressures of from about 1500 psig to about 3000 psig (10.3 MPag-20.9 MPag), and hydrogen treat gas rates of from about 213 m³/m³ to about 1068 m³/m³ (1200 SCF/B to 6000 SCF/B). The LHSV can be from about 0.25 h⁻¹ to about 50 h⁻¹, or from about 0.5 h⁻¹ to about 20 h⁻¹, preferably from about 1.0 h⁻¹ to about 4.0 h⁻¹.

[0088] In still another aspect, the same conditions can be used for hydrotreating and hydrocracking beds or stages, such as using hydrotreating conditions for both or using hydrocracking conditions for both. In yet another embodiment, the pressure for the hydrotreating and hydrocracking beds or stages can be the same.

[0089] In yet another aspect, a hydroprocessing reaction system may include more than one hydrocracking stage. If multiple hydrocracking stages are present, at least one hydrocracking stage can have effective hydrocracking conditions as described above, including a hydrogen partial pressure of at least about 1500 psig (10.3 MPag). In such an aspect, other hydrocracking processes can be performed under conditions that may include lower hydrogen partial pressures. Suitable hydrocracking conditions for an additional hydrocracking stage can include, but are not limited to, temperatures of about 500°F (260°C) to about 840°F (449°C), hydrogen partial pressures of from about 250 psig to about 5000 psig (1.8 MPag to 34.6 MPag), liquid hourly space velocities of from 0.05 h⁻¹ to 10 h⁻¹, and hydrogen treat gas rates of from 35.6 m³/m³ to 1781 m³/m³ (200 SCF/B to 10,000 SCF/B). In other embodiments, the conditions for an additional hydrocracking stage can include temperatures in the range of about 600°F (343°C) to about 815°F (435°C), hydrogen partial pressures of from about 500 psig to about 3000 psig (3.5 MPag-20.9 MPag), and hydrogen treat gas rates of from about 213 m³/m³ to about 1068 m³/m³ (1200 SCF/B to 6000 SCF/B). The LHSV can be from about 0.25 h⁻¹ to about 50 h⁻¹, or from about 0.5 h⁻¹ to about 20 h⁻¹, and preferably from about 1.0 h⁻¹ to about 4.0 h⁻¹.

[0090] In aspects where multiple hydroprocessing stages are used, a second (or subsequent) hydrocracking stage can be operated with hydrocracking conditions that include a temperature that is 20°C – 100°C lower than a temperature associated with a first hydroprocessing stage; a pressure that is 1.5 MPag – 10 MPag lower than a pressure associated with a first hydroprocessing stage (or 1.5 MPag – 5 MPag); and/or a space velocity that is 0.2 hr⁻¹ – 2.0 hr⁻¹ greater than a space velocity associated with a first hydroprocessing stage. Optionally, a

hydrocracking catalyst in a second stage can be the same as a hydroprocessing catalyst in a first stage.

Temperature-Based and Solvent-Based Separations

[0091] In various aspects, the hydroprocessed effluent from the first hydroprocessing stage (or only stage for a single stage reaction system) can be passed into a separation stage. A separation stage can include separations based on boiling point / distillation point, solvent-based separations, or a combination thereof. Both boiling point-based and solvent-based separations can be used to form an aromatics-enriched fraction and an aromatics-depleted fraction from a hydroprocessed effluent. For boiling point-based separations, forming an aromatics-enriched fraction and an aromatics-depleted fraction can be based in part on selecting a separation cut point that allows naphthenes of a certain ring type (such as 4-ring naphthenes) to be separated into an aromatics-depleted fraction while aromatics of that ring type (such as 4-ring aromatics) are separated into the aromatics-enriched fraction. Suitable distillation cut points for this type of separation can be cut points at about 371°C (for 3-rings) or about 454°C (for 4-rings). This can result in, for example, formation of an aromatics-depleted fraction with a T90 distillation point or T95 distillation point of 371°C or less (or 454°C or less) and a corresponding aromatics-enriched fraction with a T10 or T5 distillation point of 371°C or more (or 454°C or more). Such fractional weight distillation points can be determined, for example, according ASTM D2887.

[0092] Two types of solvent processing can be performed on the combined higher boiling portion from vacuum distillation and the deasphalted bottoms. The first type of solvent processing is a solvent extraction to reduce the aromatics content and/or the amount of polar molecules. The solvent extraction process selectively dissolves aromatic components to form an aromatics-rich extract phase while leaving the more paraffinic components in an aromatics-poor raffinate phase. Naphthenes are distributed between the extract and raffinate phases. Typical solvents for solvent extraction include phenol, furfural and N-methyl pyrrolidone. By controlling the solvent to oil ratio, extraction temperature and method of contacting distillate to be extracted with solvent, one can control the degree of separation between the extract and raffinate phases. Any convenient type of liquid-liquid extractor can be used, such as a counter-current liquid-liquid extractor. Depending on the initial concentration of aromatics in the deasphalted bottoms, the raffinate phase can have an aromatics content of about 5 wt% to about 25 wt%. For typical feeds, the aromatics contents will be at least about 10 wt%.

[0093] In some aspects, the deasphalted bottoms and the higher boiling fraction from vacuum distillation can be solvent processed together. Alternatively, the deasphalted bottoms and the higher boiling fraction can be solvent processed separately, to facilitate formation of different

types of lubricant base oils. For example, the higher boiling fraction from vacuum distillation can be solvent extracted and then solvent dewaxed to form a Group I base oil while the deasphalted bottoms are solvent processed to form a brightstock. Of course, multiple higher boiling fractions could also be solvent processed separately if more than one distinct Group I base oil and/or brightstock is desired.

[0094] The raffinate from the solvent extraction is preferably under-extracted. In such preferred aspects, the extraction is carried out under conditions such that the raffinate yield is maximized while still removing most of the lowest quality molecules from the feed. Raffinate yield may be maximized by controlling extraction conditions, for example, by lowering the solvent to oil treat ratio and/or decreasing the extraction temperature. The raffinate from the solvent extraction unit can then be solvent dewaxed under solvent dewaxing conditions to remove hard waxes from the raffinate.

[0095] Solvent deasphalting is another type of solvent extraction process. Instead of using an aromatic solvent to form a high-aromatic content extract that is compatible with the solvent, solvent deasphalting involves using an aliphatic solvent to form a reduced aromatic content deasphalted oil that is compatible with the deasphalting solvent. In some aspects, suitable solvents for high yield deasphalting methods as described herein include alkanes or other hydrocarbons (such as alkenes) containing 4 to 7 carbons per molecule, or 5 to 7 carbons per molecule. Examples of suitable solvents include n-butane, isobutane, n-pentane, C₄₊ alkanes, C₅₊ alkanes, C₄₊ hydrocarbons, and C₅₊ hydrocarbons. In some aspects, suitable solvents for low yield deasphalting can include C₃ hydrocarbons, such as propane, or alternatively C₃ and/or C₄ hydrocarbons. Examples of suitable solvents for low yield deasphalting include propane, n-butane, isobutane, n-pentane, C₃₊ alkanes, C₄₊ alkanes, C₃₊ hydrocarbons, and C₄₊ hydrocarbons.

[0096] In this discussion, a solvent comprising C_n (hydrocarbons) is defined as a solvent composed of at least 80 wt% of alkanes (hydrocarbons) having n carbon atoms, or at least 85 wt%, or at least 90 wt%, or at least 95 wt%, or at least 98 wt%. Similarly, a solvent comprising C_{n+} (hydrocarbons) is defined as a solvent composed of at least 80 wt% of alkanes (hydrocarbons) having n or more carbon atoms, or at least 85 wt%, or at least 90 wt%, or at least 95 wt%, or at least 98 wt%.

[0097] In this discussion, a solvent comprising C_n alkanes (hydrocarbons) is defined to include the situation where the solvent corresponds to a single alkane (hydrocarbon) containing n carbon atoms (for example, n = 3, 4, 5, 6, 7) as well as the situations where the solvent is composed of a mixture of alkanes (hydrocarbons) containing n carbon atoms. Similarly, a solvent comprising C_{n+} alkanes (hydrocarbons) is defined to include the situation where the

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solvent corresponds to a single alkane (hydrocarbon) containing n or more carbon atoms (for example, $n = 3, 4, 5, 6, 7$) as well as the situations where the solvent corresponds to a mixture of alkanes (hydrocarbons) containing n or more carbon atoms. Thus, a solvent comprising C_{4+} alkanes can correspond to a solvent including n -butane; a solvent include n -butane and isobutane; a solvent corresponding to a mixture of one or more butane isomers and one or more pentane isomers; or any other convenient combination of alkanes containing 4 or more carbon atoms. Similarly, a solvent comprising C_{5+} alkanes (hydrocarbons) is defined to include a solvent corresponding to a single alkane (hydrocarbon) or a solvent corresponding to a mixture of alkanes (hydrocarbons) that contain 5 or more carbon atoms. Alternatively, other types of solvents may also be suitable, such as supercritical fluids. In various aspects, the solvent for solvent deasphalting can consist essentially of hydrocarbons, so that at least 98 wt% or at least 99 wt% of the solvent corresponds to compounds containing only carbon and hydrogen. In aspects where the deasphalting solvent corresponds to a C_{4+} deasphalting solvent, the C_{4+} deasphalting solvent can include less than 15 wt% propane and/or other C_3 hydrocarbons, or less than 10 wt%, or less than 5 wt%, or the C_{4+} deasphalting solvent can be substantially free of propane and/or other C_3 hydrocarbons (less than 1 wt%). In aspects where the deasphalting solvent corresponds to a C_{5+} deasphalting solvent, the C_{5+} deasphalting solvent can include less than 15 wt% propane, butane and/or other $C_3 - C_4$ hydrocarbons, or less than 10 wt%, or less than 5 wt%, or the C_{5+} deasphalting solvent can be substantially free of propane, butane, and/or other $C_3 - C_4$ hydrocarbons (less than 1 wt%).

[0098] Deasphalting of heavy hydrocarbons, such as vacuum resids, is known in the art and practiced commercially. A deasphalting process typically corresponds to contacting a heavy hydrocarbon with an alkane solvent (propane, butane, pentane, hexane, heptane etc and their isomers), either in pure form or as mixtures, to produce two types of product streams. One type of product stream can be a deasphalted oil extracted by the alkane, which is further separated to produce deasphalted oil stream. A second type of product stream can be a residual portion of the feed not soluble in the solvent, often referred to as rock or asphaltene fraction. The deasphalted oil fraction can be further processed into make fuels or lubricants. The rock fraction can be further used as blend component to produce asphalt, fuel oil, and/or other products. The rock fraction can also be used as feed to gasification processes such as partial oxidation, fluid bed combustion or coking processes. The rock can be delivered to these processes as a liquid (with or without additional components) or solid (either as pellets or lumps).

[0099] During solvent deasphalting, the input feed to the solvent deasphalting unit can be mixed with a solvent. Portions of the feed that are soluble in the solvent are then extracted,

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leaving behind a residue with little or no solubility in the solvent. The portion of the deasphalted feedstock that is extracted with the solvent is often referred to as deasphalted oil. Typical solvent deasphalting conditions include mixing a feedstock fraction with a solvent in a weight ratio of from about 1 : 2 to about 1 : 10, such as about 1 : 8 or less. Typical solvent deasphalting temperatures range from 40°C to 200°C, or 40°C to 150°C, depending on the nature of the feed and the solvent. The pressure during solvent deasphalting can be from about 50 psig (~345 kPag) to about 1000 psig (~6900 kPag).

Examples of Reaction System Configurations

[00100] FIG. 1 schematically shows an example of a reaction system for processing a feedstock including one or more cracked fractions. In FIG. 1, a feedstock 105 that includes one or more cracked fractions and a hydrogen-containing stream 101 are introduced into first stage hydroprocessing reactor(s) 110. The first stage hydroprocessing reactor(s) 110 can correspond to, for example, fixed bed (such as trickle bed) reactor(s) that include demetallization catalyst, hydrotreating catalyst, and/or hydrocracking catalyst. This results in generation of a hydroprocessed effluent 115. The hydroprocessed effluent 115 can be separated in a separation stage. In FIG. 1, the separation stage corresponds to a separation stage based on performing a boiling point separation. The separation stage in FIG. 1 includes both an atmospheric distillation tower 120 and a vacuum distillation tower 130. Optionally, other separators (such as a flash separator for removing light ends prior to atmospheric distillation tower 120) could also be included as part of a separation stage.

[00101] Atmospheric distillation tower 120 can separate hydroprocessed effluent 115 into various fractions, such as light ends 122, naphtha boiling range fraction(s) 124, distillate fuel boiling range fraction(s) 126, and a bottoms fraction 128. The bottoms fraction 128 can be passed into a vacuum distillation tower 130 for further separation. For example, the vacuum distillation tower 130 can separate bottoms fraction 128 into any remaining light components 132, an intermediate boiling range fraction 135, and a bottoms fraction 138. The boiling range for intermediate boiling range fraction 135 can be dependent on the desired composition. For example, an intermediate boiling range fraction 135 with a T90 distillation point of ~482°C or more can be suitable for removing the heaviest, most aromatic components prior to second stage hydroprocessing unit 140. As another example, an intermediate boiling range fraction 135 with a T90 distillation point of ~371°C or less can be suitable for producing an intermediate boiling range fraction that contains a majority of the three-ring naphthenes from bottoms fraction 128 while the vacuum bottoms 138 contains a majority of the three-ring aromatics from bottoms fraction 128. Based on this split of three-ring naphthenes and three-ring aromatics into different

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fractions, the intermediate boiling range fraction 135 can correspond to an aromatics-depleted fraction while vacuum bottoms fraction 138 can correspond to an aromatics-enriched fraction. In this type of example, instead of using the vacuum bottoms 138 as part of the feed to fluid catalytic cracking unit 190, the vacuum bottoms can be recycled (not shown) to first hydroprocessing stage 110. This type of recycle is illustrated in connection with the single stage processing configuration shown in FIG. 2, as discussed further below. As still another example, an intermediate boiling range fraction 135 with a T90 distillation point of $\sim 454^{\circ}\text{C}$ or less can be suitable for producing an intermediate boiling range fraction that contains a majority of the four-ring naphthenes (and optionally five-ring naphthenes) from bottoms fraction 128 while the vacuum bottoms 138 contains a majority of the four-ring aromatics from bottoms fraction 128. This type of vacuum bottoms 138 can also be suitable for recycle to first hydroprocessing stage 110. Based on this split of four-ring naphthenes and four-ring aromatics into different fractions, the intermediate boiling range fraction 135 can correspond to an aromatics-depleted fraction while vacuum bottoms fraction 138 can correspond to an aromatics-enriched fraction.

[00102] The intermediate boiling range fraction 135 can then be passed into a second hydroprocessing stage 140, along with optional hydrogen-containing stream 141. The resulting second stage hydroprocessed effluent 145 can be passed through a separation stage 150, such as a knock-out drum, to remove lower boiling components. For example, lower boiling component stream 152 can correspond to a C_6 - stream (i.e., a stream including roughly n-hexane, cyclohexane, and lower boiling components). The remaining heavier portion 155 of hydrotreated effluent 145 can then be used for any convenient purpose. For example, the twice-hydroprocessed effluent 145 (or the remaining heavier portion 155 thereof) can be suitable for use as a feed to a fluid catalytic cracking unit 190. Optionally, vacuum bottoms 138 can also be passed into fluid catalytic cracking unit 190. Additionally or alternately, twice-hydroprocessed effluent 145 (or the remaining heavier portion 155 thereof) can be suitable for use as a feed to a hydrocracking unit or as a blend component for a low sulfur fuel oil.

[00103] The flow paths in FIG. 1 can represent fluid communication between the components. Fluid communication can refer to direct fluid communication or indirect fluid communication. Indirect fluid communication refers to fluid communication where one or more intervening process elements are passed through for fluids (and/or solids) that are communicated between the indirectly communicating elements. For example, vacuum distillation tower 130 is in indirect fluid communication with first stage hydroprocessing reactor(s) 110 via atmospheric distillation tower 120.

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[00104] FIG. 2 schematically shows another example of a reaction system for processing a feedstock including one or more cracked fractions. The example of a reaction system in FIG. 2 corresponds to a single stage reaction system that uses recycle to achieve at least some of the benefits of a multi-stage reaction system. In FIG. 2, a feedstock 205 that includes one or more cracked fractions and a hydrogen-containing stream 201 are introduced into hydroprocessing reactor(s) 210. The hydroprocessing reactor(s) 210 can correspond to, for example, fixed bed (such as trickle bed) reactor(s) that include demetallization catalyst, hydrotreating catalyst, and/or hydrocracking catalyst. This results in generation of a hydroprocessed effluent 215. The hydroprocessed effluent 215 can be separated in a separation stage. In FIG. 2, the separation stage corresponds to a separation stage based on performing a boiling point separation. The separation stage in FIG. 2 includes both an atmospheric distillation tower 220 and a vacuum distillation tower 230. Optionally, other separators (such as a flash separator for removing light ends prior to atmospheric distillation tower 220) could also be included as part of a separation stage.

[00105] Atmospheric distillation tower 220 can separate hydroprocessed effluent 215 into various fractions, such as light ends 222, naphtha boiling range fraction(s) 224, distillate fuel boiling range fraction(s) 226, and a bottoms fraction 228. The bottoms fraction 228 can be passed into a vacuum distillation tower 230 for further separation. For example, the vacuum distillation tower 230 can separate bottoms fraction 228 into any remaining light components 232 and a plurality of heavier fractions. The heavier fractions can include an intermediate boiling range fraction 235 with a T90 distillation point of $\sim 454^{\circ}\text{C}$ or less that contains a majority of the four-ring naphthenes (and optionally five-ring naphthenes) from bottoms fraction 228. The heavier fractions can also include a vacuum bottoms 238 that contains a majority of the four-ring aromatics from bottoms fraction 228.

[00106] The intermediate boiling range fraction 235 can optionally be passed into a separation stage 250, such as a knock-out drum, to remove lower boiling components. For example, lower boiling component stream 252 can correspond to a C₆- stream (i.e., a stream including roughly n-hexane, cyclohexane, and lower boiling components). The remaining heavier portion 255 of intermediate boiling range fraction 235 can then be used for any convenient purpose. For example, the intermediate boiling range fraction 235 (or the remaining heavier portion 255 thereof) can be suitable for use as a feed to a fluid catalytic cracking unit 290. Additionally or alternately, the intermediate boiling range fraction 235 (or the remaining heavier portion 255 thereof) can be suitable for use as a feed to a hydrocracking unit or as a blend component for a low sulfur fuel oil.

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[00107] FIG. 3 shows still another example of a configuration for hydroprocessing a feedstock that includes one or more cracked fractions. In FIG. 3, the configuration includes a separation stage that involves both boiling point separation and solvent-based separation. In FIG. 3, a feedstock 305 that includes one or more cracked fractions and a hydrogen-containing stream 301 are introduced into first stage hydroprocessing reactor(s) 310. The first stage hydroprocessing reactor(s) 310 can correspond to, for example, fixed bed (such as trickle bed) reactor(s) that include demetallization catalyst, hydrotreating catalyst, and/or hydrocracking catalyst. This results in generation of a hydroprocessed effluent 315. The hydroprocessed effluent 315 can be separated in a separation stage. In FIG. 3, the separation stage includes an optional flash separator 350 and a solvent extraction unit 360, or another type of solvent-based separator. Another option for a solvent-based separator can be a solvent deasphalting unit. The hydroprocessed effluent 315 can be passed into optional flash separator 350 to remove a lower boiling portion of the effluent, such as a C₆- stream 352. The remaining portion 355 of hydroprocessed effluent 315 can then be solvent extracted 360. Solvent extraction unit 360 can generate an extract fraction 368 that is enriched in aromatics relative to the remaining portion 355 of hydroprocessed effluent 315. After (optional) removal of solvent, the extract 368 can be recycled as part of feedstock 305, which can increase the aromatics content of feedstock 305 prior to entering first stage hydroprocessing reactor(s) 310. Solvent extraction unit 360 can also generate a raffinate fraction 365 that is passed into second hydroprocessing stage reactor(s) 340, along with a (optional) hydrogen-containing stream (not shown). The second hydroprocessing stage reactor(s) 340 can generate a second hydroprocessed effluent 345. In some aspects, such as the configuration shown in FIG. 3, the second hydroprocessed effluent 345 can be fractionated in a separation stage 370 (such as an atmospheric distillation tower) to form, for example, naphtha boiling range fraction(s) 374, distillate fuel boiling range fraction(s) 376, and a bottoms fraction 378. The bottoms fraction 378 can be used as a feed for fluid catalytic cracking, or alternatively the bottoms fraction 378 can be recycled as part of the input flow into second stage hydroprocessing reactor(s) 340. As an alternative, separation stage 370 could correspond to a knock-out drum or other flash separator for removing lower boiling components (such as a C₆-stream) from second hydroprocessing effluent 345, and a remaining portion of second hydroprocessing effluent 345 could be used as a low sulfur fuel oil and/or passed into a fluid catalytic cracking unit. This type of configuration is further illustrated in FIG. 4, in conjunction with illustrating a single stage hydroprocessing configuration that involves a solvent-based separation stage.

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[00108] In FIG. 4, a feedstock 405 that includes one or more cracked fractions and a hydrogen-containing stream 401 are introduced into hydroprocessing reactor(s) 410. The hydroprocessing reactor(s) 410 can correspond to, for example, fixed bed (such as trickle bed) reactor(s) that include demetallization catalyst, hydrotreating catalyst, and/or hydrocracking catalyst. This results in generation of a hydroprocessed effluent 415. The hydroprocessed effluent 415 can be separated in a separation stage. In FIG. 4, the separation stage corresponds to an atmospheric distillation tower 420 followed by a solvent extraction unit 460. Atmospheric distillation tower 420 can produce a light ends fraction 422, naphtha boiling range fraction(s) 424, distillate boiling range fraction(s) 426, and a bottoms fraction 428. Bottoms fraction 428 can then be passed into solvent extraction unit 460 (or alternatively a solvent deasphalting unit) to produce an extract 468 and a raffinate 465. Extract 468 can be recycled for use as part of feedstock 405. Raffinate 465 can be passed into fluid catalytic cracking unit 490. Alternatively, raffinate 465 could be used as a low sulfur fuel oil and/or as a feed for a hydrocracking unit (not shown).

Example 1 – Properties of Catalytic Slurry Oils and Hydroprocessed Catalytic Slurry Oils

[00109] Catalytic slurry oils were obtained from fluid catalytic cracking (FCC) processes operating on various feeds. Table 1 shows results from characterization of the catalytic slurry oils. Additionally, a blend of catalytic slurry oils from several FCC process sources was also formed and characterized.

Table 1 – Characterization of Catalytic Slurry Oils

	CSO 1	CSO 2	CSO 3	CSO 4	CSO X (Blend)
API Gravity (15°C)	-7.5	-9.0	1.2	-5.0	-3.0
S (wt%)	4.31	4.27	1.11	1.82	3.07
N (wppm)	1940	2010	1390	1560	1750
H (wt%)	6.6	6.5	8.4	7.0	7.3
MCR (wt%)	11.5	14.6	4.7	13.4	12.5
n-heptane insolubles (wt%)	4.0	8.7	0.4	5.0	0.7
GCD (ASTM D2887) (wt%)					
< 316°C	2		4		3
316°C – 371°C	11		13		12
371°C – 427°C	43		40		36
427°C – 482°C	27		26		28
482°C – 538°C	7		10		10

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538°C – 566°C	2		2		2
566°C+	8		5		9

[00110] As shown in Table 1, typical catalytic slurry oils (or blends of such slurry oils) can represent a low value and/or challenged feed. The catalytic slurry oils have an API Gravity at 15°C of less than 1.5, and often less than 0. The catalytic slurry oils can have sulfur contents of greater than 1.0 wt%, nitrogen contents of at least 1000 wppm, and hydrogen contents of less than 8.5 wt%, or less than 7.5 wt%, or less than 7.0 wt%. The catalytic slurry oils can also be relatively high in micro carbon residue (MCR), with values of at least 4.5 wt%, or at least 6.5 wt%, and in some cases greater than 10 wt%. The catalytic slurry oils can also contain a substantial n-heptane insolubles (asphaltene) content, for example at least 0.3 wt%, or at least 1.0 wt%, or at least 4.0 wt%. It is noted that the boiling range of the catalytic slurry oils has more in common with a vacuum gas oil than a vacuum resid, as less than 10 wt% of the catalytic slurry oils corresponds to 566°C+ compounds, and less than 15 wt% corresponds to 538°C+ compounds.

[00111] The blend of catalytic slurry oils (CSO X) from Table 1 was used as a feedstock for a pilot scale processing plant. The blend of catalytic slurry oils had a density of 1.12 g/cm³, a T10 distillation point of 354°C, a T50 of 427°C, and a T90 of 538°C. The blend contained roughly 12 wt% MCR, had a sulfur content of ~3 wt%, a nitrogen content of ~2500 wppm, and a hydrogen content of ~7.4 wt%. A compositional analysis of the blend determined that the blend included 10 wt% saturates, 70 wt% aromatics with 4 or more rings, and 20 wt% aromatics with 1 – 3 rings.

[00112] The blend was used as a feedstock for hydroprocessing in a reaction system with a single hydrotreating stage. The feedstock was exposed to a commercially available medium pore NiMo supported hydrotreating catalyst. The start of cycle conditions were a total pressure of ~2600 psig, ~0.25 LHSV, ~370°C, and ~10,000 SCF/B of hydrogen treat gas. The conditions resulted in total product with an organic sulfur content of about 125 wppm. The total product from hydroprocessing was analyzed. The total product at start of run included 3 wt% H₂S; 1 wt% of C₄- (i.e., light ends); 5 wt% naphtha boiling range compounds; 47 wt% of 177°C – 371°C (diesel boiling range) compounds, which had a sulfur content of less than 15 wppm; and 45 wt% of 371°C+ compounds. The 371°C+ compounds had a specific gravity of ~1.0 g/cm³. The 371°C+ fraction was suitable for use as a hydrocracker feed, a FCC feed, and/or sale as a fuel oil. The yield of 566°C+ compounds was 2.5 wt%. Hydrogen consumption at the start of hydroprocessing was ~3400 SCF/B. The feed was processed in the pilot reactor for 300 days,

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with adjustments to the conditions to maintain the organic sulfur content in the total product at roughly 125 wppm. The end of cycle conditions were ~2600 psig, ~0.25 LHSV, ~410°C, and ~10,000 SCF/B of hydrogen treat gas. The total product at end of run included 3 wt% H₂S; 3 wt% of C₄- (i.e., light ends); 8 wt% naphtha boiling range compounds; 45 wt% of 177°C – 371°C (diesel boiling range) compounds, which had a sulfur content of less than 15 wppm; and 41 wt% of 371°C+ compounds with a specific gravity of 1.0 g/cm³. Hydrogen consumption at the end of hydroprocessing was ~3300 SCF/B. By the end of the run, greater than 90 wt% of the 566°C+ compounds were being converted. There was no build up in pressure during the course of the run. This lack of pressure build up and the general stability of the run, particularly at the end of run conditions which included a temperature of 410°C, was surprising.

[00113] Without being bound by any particular theory, it is believed that the surprising stability of the process is explained in part by the S_{BN} and I_N values of the hydrotreated effluent during the course of the processing run, and the corresponding difference between those values. FIG. 5 shows measured values for the S_{BN} and I_N of the liquid portion (C₅+) of the hydroprocessed effluent in relation to the amount of 566°C+ conversion. The amount of 566°C+ conversion roughly corresponds to the length of processing time, as the amount of conversion roughly correlates with the temperature increases required to maintain the organic sulfur content of the hydroprocessed effluent at the desired target level of ~125 wppm. As shown in FIG. 5, both the S_{BN} and the I_N of the hydroprocessed effluent decrease with increasing conversion, but the difference between S_{BN} and I_N in the hydroprocessed effluent remains relatively constant at roughly 40 to 50. This unexpectedly large difference in S_{BN} and I_N even at 90+ wt% conversion relative to 566°C indicates that the hydroprocessed effluent should have a low tendency to cause coke formation in the reactor and/or otherwise deposit solids that can cause plugging.

Example 2 – Solvent Separation for Processing of Cracked Fractions

[00114] The blend of catalytic slurry oils from Example 1 was processed in a configuration similar to the first hydroprocessing stage and separation stage shown in FIG. 3. The first stage hydroprocessing (hydrotreating) conditions included exposing the feed to a catalyst similar to the NiMo supported catalyst in Example 1 at a total pressure of ~2400 psig, ~1.0 LHSV, ~370°C, and ~10,000 SCF/B of hydrogen treat gas in order to generate a hydroprocessed effluent. The total liquid product of the hydroprocessed effluent had a density of 1.04 g/cm³ and an organic sulfur content of about 0.5 wt%. A gas-liquid separator was then used to remove light ends (C₆-) and H₂S from the effluent. The remaining portion of the effluent, corresponding to the total liquid product, had an initial boiling point of about 227°C, which was high enough to allow for solvent extraction of the total liquid product without further distillation. The total liquid product was

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solvent extracted using N-methylpyrrolidone at a 0.25 : 1 (v/v) treat rate. This resulted in a raffinate phase corresponding to 60 wt% of the total liquid product and an extract phase corresponding to 40 wt% of the total liquid product. The raffinate product had a density of 0.99 g/cm³ and a sulfur content of ~500 wppm. The raffinate stream was suitable for use as a high value LSFO (low sulfur fuel oil) blending component for blending off of vacuum resid streams with greater than 0.5 wt% sulfur content. Alternatively, the raffinate could be used as a high quality feed for a distillate hydrocracker, which could correspond to the second stage hydroprocessing unit shown in FIG. 3. Based on the product quality into the second stage hydroprocessing unit, it is expected that the second stage hydroprocessing unit could be operated at a LHSV of 0.5 hr⁻¹ or greater while producing a diesel boiling range product with a sulfur content of 15 wppm or less. If gasoline production is more desirable, still another option could be to hydrotreat the raffinate prior to passing the twice hydrotreated effluent into a fluid catalytic cracking process.

[00115] In another processing run, the blend of catalytic slurry oils from Example 1 was processed in a configuration similar to FIG. 4. The hydroprocessing (hydrotreating) conditions included exposing the feed to a catalyst similar to the NiMo supported catalyst in Example 1 at a total pressure of ~2400 psig, ~0.25 LHSV, ~370°C, and ~10,000 SCF/B of hydrogen treat gas in order to generate a hydroprocessed effluent. The total liquid product had a density of 0.98 g/cm³ and a sulfur content of ~150 wppm. The total liquid product was distilled to form ~ 5 wt% of a naphtha boiling range fraction (C₆ – 177°C), ~50 wt% of a diesel fuel boiling range fraction (177°C – 371°C), and ~45 wt% of 371°C+ bottoms. The 371°C+ product was extracted with N-methylpyrrolidone at a 0.25 : 1 (v/v) treat rate, which split the 371°C+ product into 70 wt% of raffinate with a density of 0.94 g/cm³ and 30 wt% of an extract with a density of 1.07 g/cm³. The raffinate can be suitable for further processing in, for example, a fluid catalytic cracker, while the extract can be, for example, blended off as low sulfur fuel oil and/or recycled back as part of the feedstock to the hydroprocessing stage.

Additional Embodiments

[00116] Embodiment 1. A method for processing a heavy cracked feedstock, comprising: exposing a feedstock comprising a density at 15°C of 1.06 g/cm³ or more and at least 50 wt% of one or more 343°C+ cracked fractions (or at least 60 wt%, or at least 70 wt%) to a hydroprocessing catalyst under fixed bed hydroprocessing conditions to form a hydroprocessed effluent, the one or more 343°C+ cracked fractions having an aromatics content of 40 wt% or more relative to a weight of the one or more 343°C+ cracked fractions, a 343°C+ portion of the hydroprocessed effluent having a density at 15°C of 1.04 g/cm³ or less; separating the

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hydroprocessed effluent in one or more separation stages to form an aromatics-enriched fraction and an aromatics-depleted fraction; and exposing at least a portion of the aromatics-enriched fraction to a second hydroprocessing catalyst under second fixed bed hydroprocessing conditions to form a second hydroprocessed effluent.

[00117] Embodiment 2. The method of Embodiment 1, wherein exposing the feedstock to the hydroprocessing catalyst further comprises exposing the at least a portion of the aromatics-enriched fraction to the hydroprocessing catalyst, wherein the hydroprocessing conditions comprise the second hydroprocessing conditions, and wherein the hydroprocessed effluent comprises the second hydroprocessed effluent.

[00118] Embodiment 3. The method of any of the above embodiments, wherein the separating the hydroprocessed effluent in one or more separation stages comprises performing a separation based on boiling point to form an aromatics-enriched fraction and an aromatics-depleted fraction.

[00119] Embodiment 4. The method of Embodiment 3, wherein the aromatics-enriched fraction has a T10 distillation point of 371°C or more, and the aromatics-depleted fraction has a T90 distillation point of 371°C or less; or wherein the aromatics-enriched fraction has a T10 distillation point of 454°C or more, and the aromatics-depleted fraction has a T90 distillation point of 454°C or less.

[00120] Embodiment 5. The method of Embodiment 1 or 2, wherein the separating the hydroprocessed effluent in one or more separation stages comprises performing a solvent-based separation to form an aromatics-enriched fraction and an aromatics-depleted fraction, the solvent-based separation optionally comprising solvent extraction using an aromatic solvent, the aromatic solvent optionally comprising N-methylpyrrolidone.

[00121] Embodiment 6. The method of Embodiment 5, wherein the separating the hydroprocessed effluent in one or more separation stages further comprises performing a separation based on boiling point prior to performing the solvent-based separation to form the aromatics-enriched fraction and the aromatics-depleted fraction.

[00122] Embodiment 7. The method of any of the above embodiments, the method further comprising exposing at least a portion of the aromatics-depleted fraction to a distillate hydroprocessing catalyst under distillate fixed bed hydroprocessing conditions to form a distillate hydroprocessing effluent, a 177°C – 371°C portion of the distillate hydroprocessing effluent optionally having a sulfur content of 50 wppm or less (or 15 wppm or less).

[00123] Embodiment 8. A method for processing a heavy cracked feedstock, comprising: exposing a feedstock comprising a density at 15°C of 1.06 g/cm³ or more and at least 50 wt% of

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one or more 343°C+ cracked fractions (or at least 60 wt%, or at least 70 wt%) to a hydroprocessing catalyst under fixed bed hydroprocessing conditions to form a hydroprocessed effluent, the one or more 343°C+ cracked fractions having an aromatics content of 40 wt% or more relative to a weight of the one or more 343°C+ cracked fractions, a 343°C+ portion of the hydroprocessed effluent having a density at 15°C of 1.04 g/cm³ or less; separating, from the hydroprocessed effluent, a first fraction comprising a T10 distillation point of at least 260°C (or at least 300°C, or at least 340°C) and a T90 distillation point of 454°C or less and a second fraction comprising a T10 distillation point of at least 427°C; and exposing at least a portion of the first fraction to a distillate hydroprocessing catalyst under distillate fixed bed hydroprocessing conditions to form a distillate hydroprocessing effluent, a 177°C – 371°C portion of the distillate hydroprocessing effluent optionally having a sulfur content of 50 wppm or less (or 15 wppm or less).

[00124] Embodiment 9. The method of any of the above embodiments, wherein the one or more 343°C+ cracked fractions comprise a catalytic slurry oil, a coker bottoms fraction, a steam cracker tar fraction, a coal tar, a visbreaker gas oil, or a combination thereof; or wherein the one or more 343°C+ cracked fractions consist essentially of a catalytic slurry oil.

[00125] Embodiment 10. The method of Embodiment 9, further comprising settling the catalytic slurry oil prior to exposing the feed to the hydroprocessing catalyst, the settled catalytic slurry oil having a catalyst fines content of 1 wppm or less.

[00126] Embodiment 11. The method of any of the above embodiments, wherein the one or more 343°C+ cracked fractions comprise about 2 wt% or more n-heptane insolubles and the hydroprocessed effluent comprises about 1 wt% or less n-heptane insolubles; or wherein the one or more 343°C+ cracked fractions comprise at least a first amount of micro carbon residue, and the hydroprocessed effluent comprises less than about half of the first amount of micro carbon residue; or wherein the one or more 343°C+ cracked fractions comprise at least 3 wt% of a 566°C+ portion, the effective hydroprocessing conditions being effective for 55 wt% or more conversion of the feedstock relative to 566°C (or 65 wt% or more, or 75 wt% or more); or a combination thereof.

[00127] Embodiment 12. The method of any of the above embodiments, wherein an I_N of at least one of the first hydroprocessed effluent and the second hydroprocessed effluent is 10 or more lower than an I_N of the feedstock (or 20 or more lower, or 30 or more lower); or wherein a difference between an S_{BN} of the hydroprocessed effluent and the I_N of the hydroprocessed effluent is at least 30, or at least 40; or a combination thereof.

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[00128] Embodiment 13. The method of any of the above embodiments, wherein the feedstock comprises 4.0 wt% or more of micro carbon residue (or 6.0 wt% or more); or wherein the catalytic slurry oil comprises 5.0 wt% or more of micro carbon residue (or 7.0 wt% or more, or 10 wt% or more); or wherein the hydroprocessed effluent comprises 4.0 wt% or less of micro carbon residue (or 3.0 wt% or less, or 2.0 wt% or less); or wherein the feedstock comprises at least 1.0 wt% of organic sulfur, the hydroprocessed effluent comprising 1000 wppm or less of organic sulfur (or about 500 wppm or less, or about 200 wppm or less); or a combination thereof.

[00129] Embodiment 14. A system for processing a cracked feedstock, comprising: a first hydroprocessing reactor comprising a first hydroprocessing inlet, a first hydroprocessing outlet, and a fixed bed comprising a first hydroprocessing catalyst, the first hydroprocessing inlet comprising a feedstock comprising a density at 15°C of 1.06 g/cm³ or more and at least 50 wt% of one or more 343°C+ cracked fractions, the one or more 343°C+ cracked fractions having an aromatics content of 40 wt% or more relative to a weight of the one or more cracked fractions, the first hydroprocessing outlet comprising a hydroprocessed effluent; a separation stage comprising a separation inlet, a first separation outlet, and a second separation outlet, the first separation inlet being in fluid communication with the first hydroprocessing outlet, a first separation outlet comprising a hydroprocessed effluent fraction having a T90 distillation point of 454°C or less, a second separation outlet comprising a hydroprocessed effluent fraction having a T10 distillation point of at least 427°C; and a second hydroprocessing reactor comprising a second hydroprocessing inlet, a second hydroprocessing outlet, and a fixed bed comprising a second hydroprocessing catalyst, the second hydroprocessing inlet being in fluid communication with the first separation outlet, the first hydroprocessing inlet optionally being in fluid communication with the second separation outlet, the system optionally further comprising a fluid catalytic cracking reactor in indirect fluid communication with the second hydroprocessing outlet.

[00130] Embodiment 15. A hydroprocessed effluent, a second hydroprocessed effluent, or a distillate hydroprocessing effluent made according to any of Embodiments 1 – 13.

[00131] When numerical lower limits and numerical upper limits are listed herein, ranges from any lower limit to any upper limit are contemplated. While the illustrative embodiments of the invention have been described with particularity, it will be understood that various other modifications will be apparent to and can be readily made by those skilled in the art without departing from the spirit and scope of the invention. Accordingly, it is not intended that the scope of the claims appended hereto be limited to the examples and descriptions set forth herein but rather that the claims be construed as encompassing all the features of patentable novelty

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which reside in the present invention, including all features which would be treated as equivalents thereof by those skilled in the art to which the invention pertains.

[00132] The present invention has been described above with reference to numerous embodiments and specific examples. Many variations will suggest themselves to those skilled in this art in light of the above detailed description. All such obvious variations are within the full intended scope of the appended claims.

CLAIMS:

1. A method for processing a heavy cracked feedstock, comprising:
exposing a feedstock comprising a density at 15°C of 1.06 g/cm³ or more and at least 50 wt% of one or more 343°C+ cracked fractions (or at least 60 wt%, or at least 70 wt%) to a hydroprocessing catalyst under fixed bed hydroprocessing conditions to form a hydroprocessed effluent, the one or more 343°C+ cracked fractions having an aromatics content of 40 wt% or more relative to a weight of the one or more 343°C+ cracked fractions, a 343°C+ portion of the hydroprocessed effluent having a density at 15°C of 1.04 g/cm³ or less;
separating the hydroprocessed effluent in one or more separation stages to form an aromatics-enriched fraction and an aromatics-depleted fraction; and
exposing at least a portion of the aromatics-enriched fraction to a second hydroprocessing catalyst under second fixed bed hydroprocessing conditions to form a second hydroprocessed effluent.
2. The method of Claim 1, wherein exposing the feedstock to the hydroprocessing catalyst further comprises exposing the at least a portion of the aromatics-enriched fraction to the hydroprocessing catalyst, wherein the hydroprocessing conditions comprise the second hydroprocessing conditions, and wherein the hydroprocessed effluent comprises the second hydroprocessed effluent.
3. The method of any of the above claims, wherein the separating the hydroprocessed effluent in one or more separation stages comprises performing a separation based on boiling point to form an aromatics-enriched fraction and an aromatics-depleted fraction.
4. The method of Claim 3, wherein the aromatics-enriched fraction has a T10 distillation point of 371°C or more, and the aromatics-depleted fraction has a T90 distillation point of 371°C or less; or
wherein the aromatics-enriched fraction has a T10 distillation point of 454°C or more, and the aromatics-depleted fraction has a T90 distillation point of 454°C or less.
5. The method of Claim 1 or 2, wherein the separating the hydroprocessed effluent in one or more separation stages comprises performing a solvent-based separation to form an aromatics-enriched fraction and an aromatics-depleted fraction, the solvent-based separation optionally comprising solvent extraction using an aromatic solvent, the aromatic solvent optionally comprising N-methylpyrrolidone.
6. The method of Claim 5, wherein the separating the hydroprocessed effluent in one or more separation stages further comprises performing a separation based on boiling point prior to

performing the solvent-based separation to form the aromatics-enriched fraction and the aromatics-depleted fraction.

7. The method of any of the above claims, the method further comprising exposing at least a portion of the aromatics-depleted fraction to a distillate hydroprocessing catalyst under distillate fixed bed hydroprocessing conditions to form a distillate hydroprocessing effluent, a 177°C – 371°C portion of the distillate hydroprocessing effluent optionally having a sulfur content of 50 wppm or less (or 15 wppm or less).

8. A method for processing a heavy cracked feedstock, comprising:

exposing a feedstock comprising a density at 15°C of 1.06 g/cm³ or more and at least 50 wt% of one or more 343°C+ cracked fractions (or at least 60 wt%, or at least 70 wt%) to a hydroprocessing catalyst under fixed bed hydroprocessing conditions to form a hydroprocessed effluent, the one or more 343°C+ cracked fractions having an aromatics content of 40 wt% or more relative to a weight of the one or more 343°C+ cracked fractions, a 343°C+ portion of the hydroprocessed effluent having a density at 15°C of 1.04 g/cm³ or less;

separating, from the hydroprocessed effluent, a first fraction comprising a T10 distillation point of at least 260°C (or at least 300°C, or at least 340°C) and a T90 distillation point of 454°C or less and a second fraction comprising a T10 distillation point of at least 427°C; and

exposing at least a portion of the first fraction to a distillate hydroprocessing catalyst under distillate fixed bed hydroprocessing conditions to form a distillate hydroprocessing effluent, a 177°C – 371°C portion of the distillate hydroprocessing effluent optionally having a sulfur content of 50 wppm or less (or 15 wppm or less).

9. The method of any of the above claims, wherein the one or more 343°C+ cracked fractions comprise a catalytic slurry oil, a coker bottoms fraction, a steam cracker tar fraction, a coal tar, a visbreaker gas oil, or a combination thereof; or

wherein the one or more 343°C+ cracked fractions consist essentially of a catalytic slurry oil.

10. The method of Claim 9, further comprising settling the catalytic slurry oil prior to exposing the feed to the hydroprocessing catalyst, the settled catalytic slurry oil having a catalyst fines content of 1 wppm or less.

11. The method of any of the above claims, wherein the one or more 343°C+ cracked fractions comprise about 2 wt% or more n-heptane insolubles and the hydroprocessed effluent comprises about 1 wt% or less n-heptane insolubles; or wherein the one or more 343°C+ cracked fractions comprise at least a first amount of micro carbon residue, and the hydroprocessed effluent comprises less than about half of the first amount of micro carbon residue; or

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wherein the one or more 343°C+ cracked fractions comprise at least 3 wt% of a 566°C+ portion, the effective hydroprocessing conditions being effective for 55 wt% or more conversion of the feedstock relative to 566°C (or 65 wt% or more, or 75 wt% or more); or

a combination thereof.

12. The method of any of the above claims, wherein an I_N of at least one of the first hydroprocessed effluent and the second hydroprocessed effluent is 10 or more lower than an I_N of the feedstock (or 20 or more lower, or 30 or more lower); or wherein a difference between an S_{BN} of the hydroprocessed effluent and the I_N of the hydroprocessed effluent is at least 30, or at least 40; or a combination thereof.

13. The method of any of the above claims, wherein the feedstock comprises 4.0 wt% or more of micro carbon residue (or 6.0 wt% or more); or

wherein the catalytic slurry oil comprises 5.0 wt% or more of micro carbon residue (or 7.0 wt% or more, or 10 wt% or more); or

wherein the hydroprocessed effluent comprises 4.0 wt% or less of micro carbon residue (or 3.0 wt% or less, or 2.0 wt% or less); or

wherein the feedstock comprises at least 1.0 wt% of organic sulfur, the hydroprocessed effluent comprising 1000 wppm or less of organic sulfur (or about 500 wppm or less, or about 200 wppm or less); or

a combination thereof.

14. A system for processing a cracked feedstock, comprising:

a first hydroprocessing reactor comprising a first hydroprocessing inlet, a first hydroprocessing outlet, and a fixed bed comprising a first hydroprocessing catalyst, the first hydroprocessing inlet comprising a feedstock comprising a density at 15°C of 1.06 g/cm³ or more and at least 50 wt% of one or more 343°C+ cracked fractions, the one or more 343°C+ cracked fractions having an aromatics content of 40 wt% or more relative to a weight of the one or more cracked fractions, the first hydroprocessing outlet comprising a hydroprocessed effluent;

a separation stage comprising a separation inlet, a first separation outlet, and a second separation outlet, the first separation inlet being in fluid communication with the first hydroprocessing outlet, a first separation outlet comprising a hydroprocessed effluent fraction having a T90 distillation point of 454°C or less, a second separation outlet comprising a hydroprocessed effluent fraction having a T10 distillation point of at least 427°C; and

a second hydroprocessing reactor comprising a second hydroprocessing inlet, a second hydroprocessing outlet, and a fixed bed comprising a second hydroprocessing catalyst, the second hydroprocessing inlet being in fluid communication with the first separation outlet, the

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first hydroprocessing inlet optionally being in fluid communication with the second separation outlet, the system optionally further comprising a fluid catalytic cracking reactor in indirect fluid communication with the second hydroprocessing outlet.

15. A hydroprocessed effluent, a second hydroprocessed effluent, or a distillate hydroprocessing effluent made according to any of claimss 1 – 13.

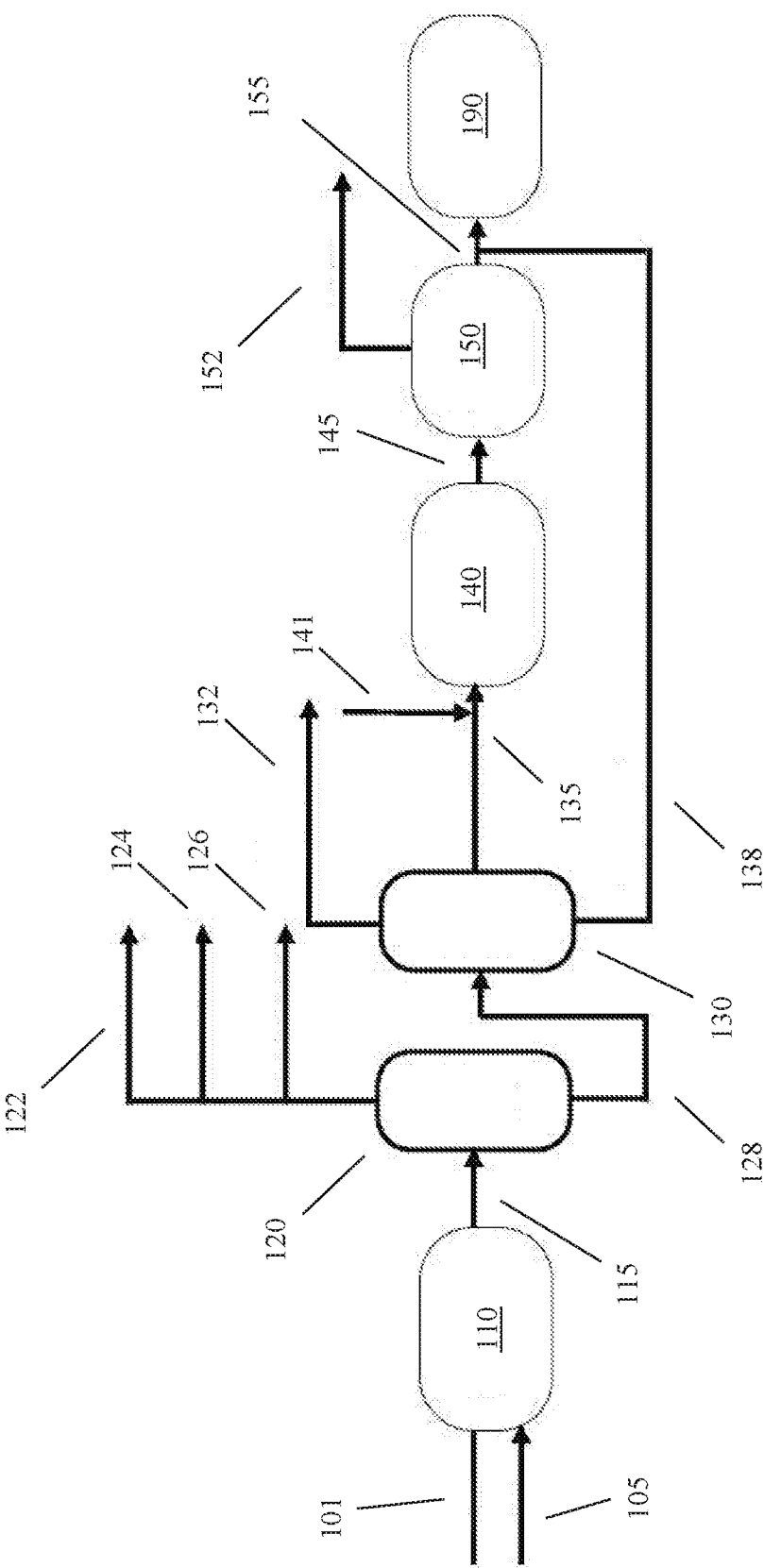


FIG. 1

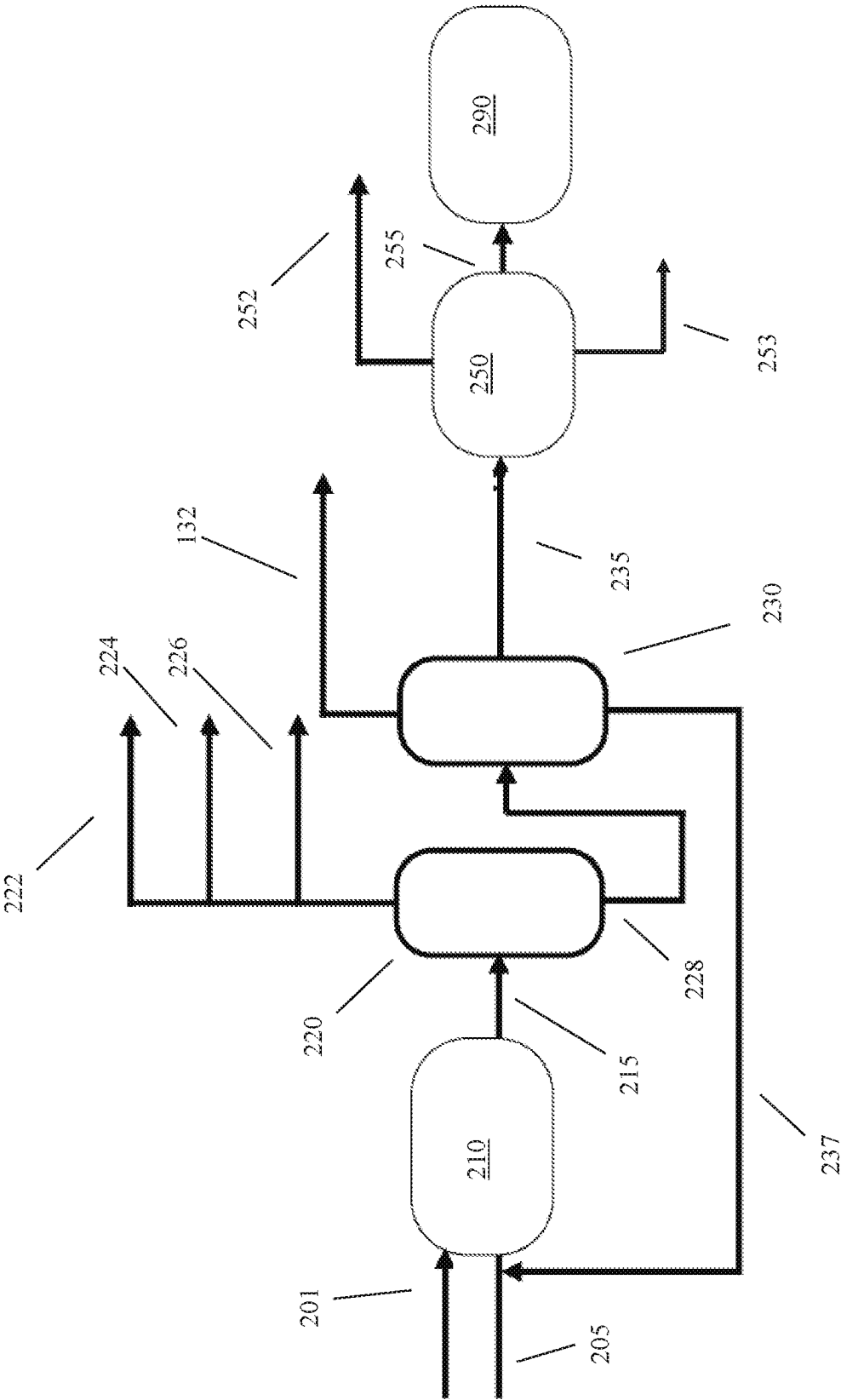


FIG. 2

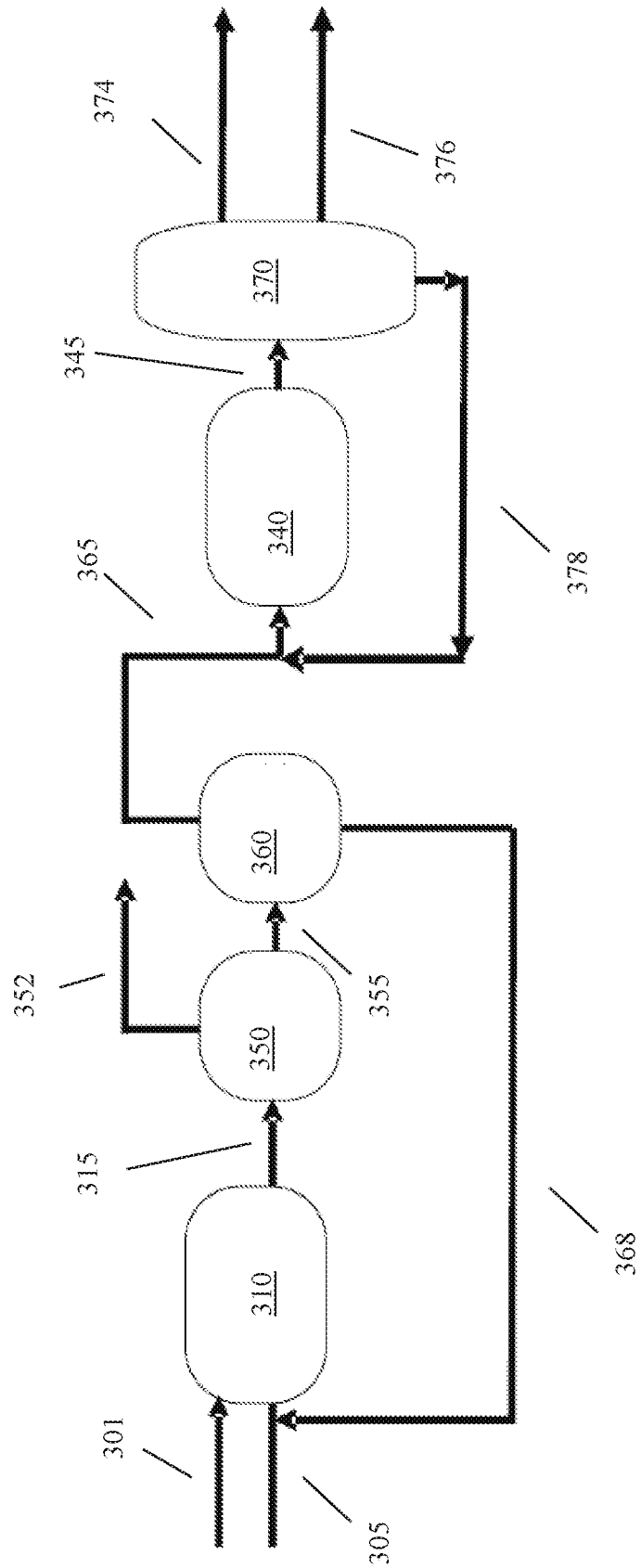


FIG. 3

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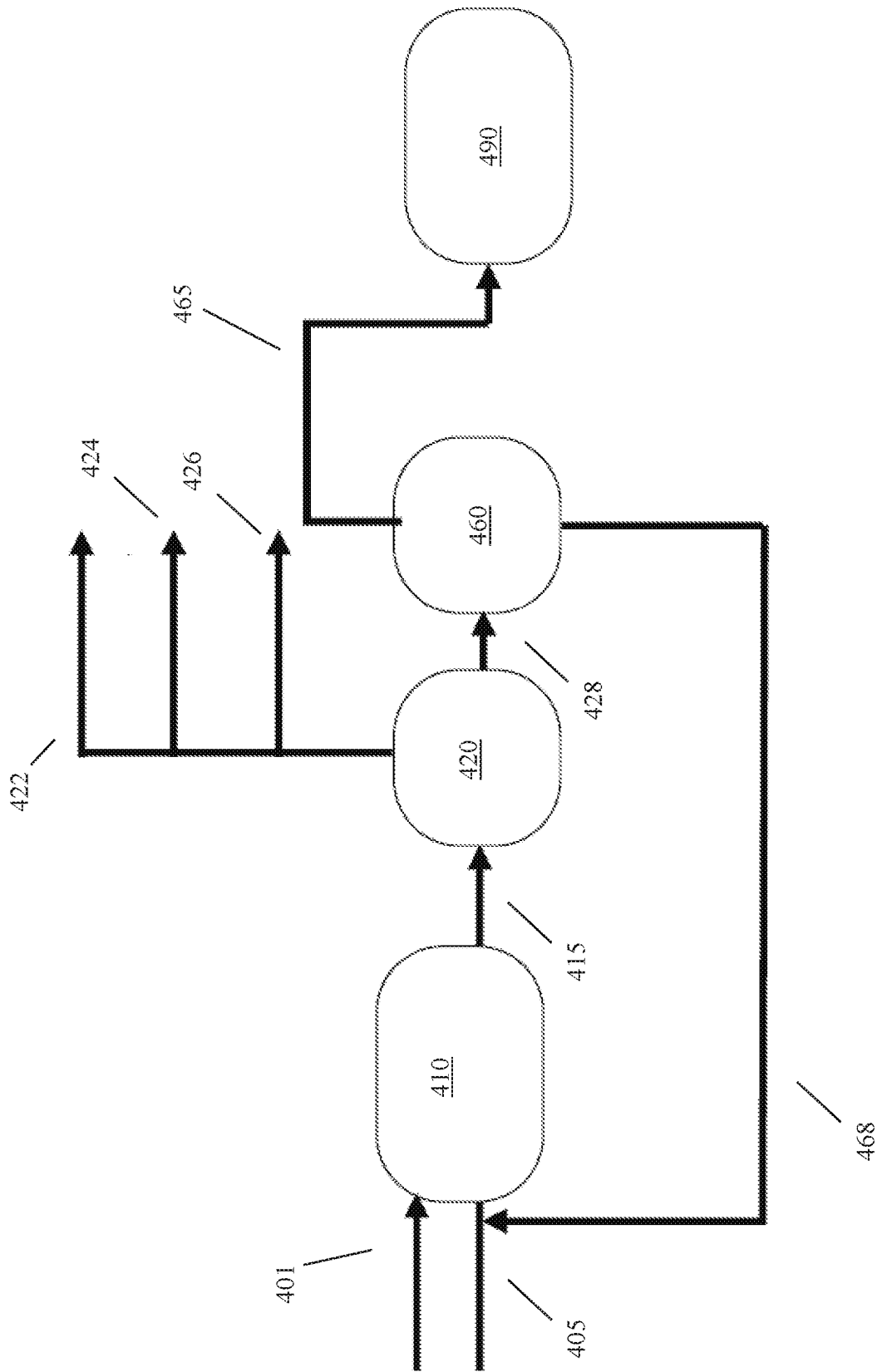


FIG. 4

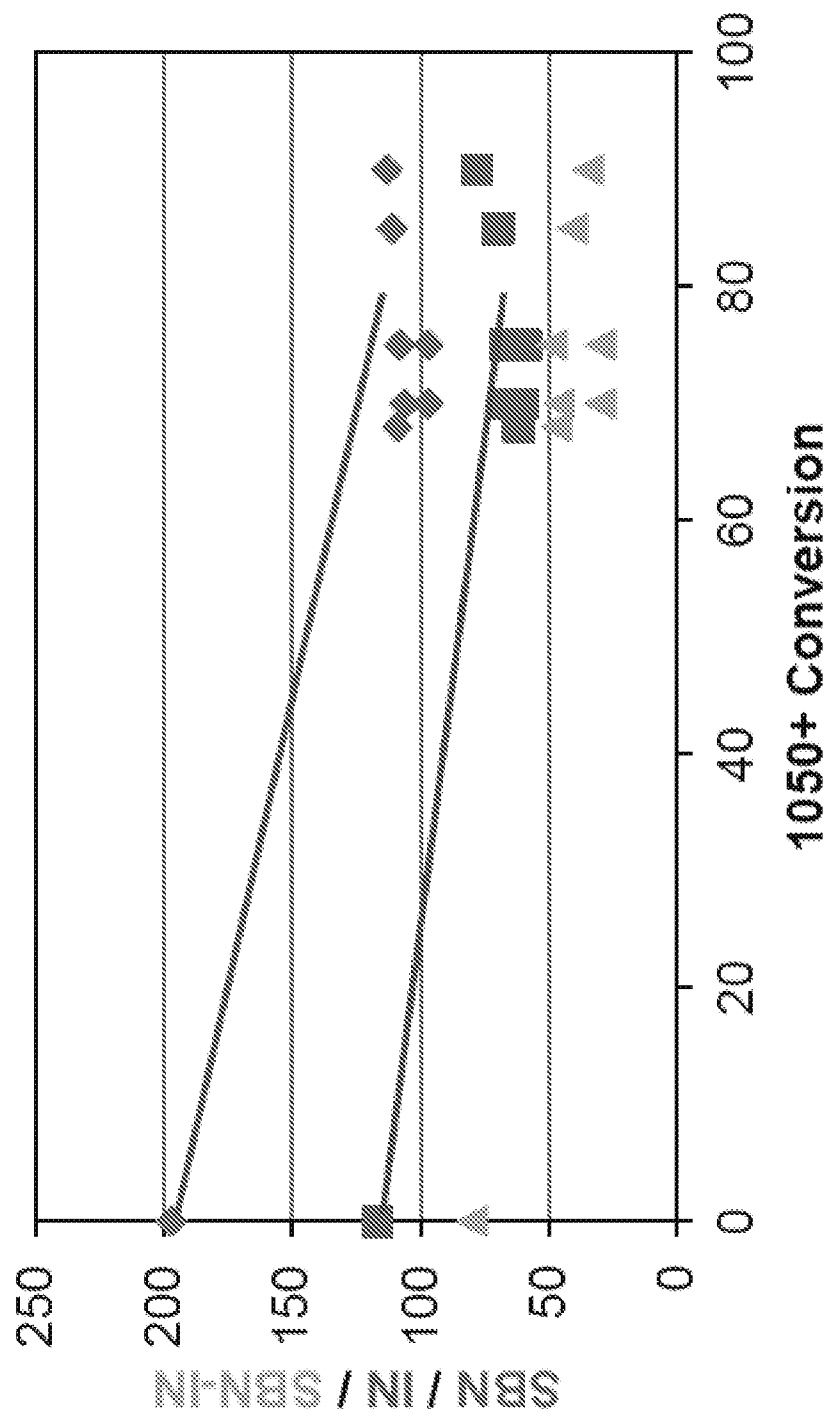


FIG. 5