

1 **NEW HYDROCRACKING PROCESS FOR THE PRODUCTION OF**
2 **HIGH QUALITY DISTILLATES FROM HEAVY GAS OILS**

3

4

5 **FIELD OF THE INVENTION**

6

7 This invention is directed to processes for the conversion of material boiling in
8 the Vacuum Gas Oil boiling range to high quality middle distillates and/or
9 naphtha and lighter products, and more particularly to a multiple stage
10 process using a single hydrogen loop.

11

12

13 **BACKGROUND OF THE INVENTION**

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15 In the refining of crude oil, gas oil hydrocrackers are used to convert heavy
16 gas oils to lighter products using a single reaction stage or multiple reaction
17 stages. In most instances, the various reaction stages operate at similar
18 pressure levels. Where pressure levels are different, separate hydrogen
19 loops are employed. Multiple reaction stages are used to achieve the
20 following:

21

- 22 • high conversion employing minimum reactor volume and catalyst
23 volume overall
- 24 • better product qualities
- 25 • lower hydrogen consumption

26

27 The use of multiple reaction systems however involves more equipment
28 including multiple expensive high-pressure pumps and compressors.

29

30 U.S. Patent No. 5,980,729 discloses a configuration with multiple reaction
31 zones in a single hydrogen loop. The process uses a hot stripper
32 downstream of the denitrification/desulfurization zone. Liquid from the hot
33 stripper is pumped to the hydrocracking reactor upstream of the hydrotreating

1 reactor. Recycle oil from the fractionation section is also pumped back to the
2 hydrocracking reactor.

3

4 In conventional hydroprocessing, it is necessary to transfer hydrogen from a
5 vapor phase into the liquid phase where it will be available to react with a
6 petroleum molecule at the surface of the catalyst. This is accomplished by
7 circulating very large volumes of hydrogen gas and the oil through a catalyst
8 bed. The oil and the hydrogen flow through the bed and the hydrogen is
9 absorbed into a thin film of oil that is distributed over the catalyst. Because
10 the amount of hydrogen required can be large, 1000 to 5000 SCF/bbl of liquid,
11 and the amount of catalyst required can also be large, the reactors are very
12 large and can operate at severe conditions, from a few hundred psi to as
13 much as 5000 psi and temperatures from around 400°F to 900°F.

14

15 U.S. Pat. No. 6,224,747 teaches hydrocracking a VGO stream in a
16 hydrocracking reaction zone within an integrated hydroconversion process.
17 Effluent from the hydrocracking reaction zone is combined with a light
18 aromatic-containing feed stream, and the blended stream hydrotreated in a
19 hydrotreating reaction zone. The hydrocracked effluent serves as a heat sink
20 for the hydrotreating reaction zone. The integrated reaction system provides
21 a single hydrogen supply and recirculation system for use in two reaction
22 systems. There is no temperature control between the hydrocracking reaction
23 zone and the hydrotreating reaction zone, however.

24

25 U.S. Pat. No. 3,592,757 (Baral) illustrates temperature control between zones
26 by means of heat exchangers, as in the instant invention. Baral does not
27 employ a single hydrogen loop, as does the instant invention. Baral discloses
28 a hydrofiner (similar to a hydrotreater) operating in series with a hydrocracker,
29 with a fraction of the product fed to a hydrogenator. A gas oil feed is fed with
30 both make-up and recycle hydrogen to a hydrofiner. A recycle stream and
31 additional recycle hydrogen are added to the hydrofiner product stream, and

1 the mixture is fed to a hydrocracker. The hydrocracker product stream is
2 cooled and separated into a vapor and a liquid stream. The vapor stream is
3 passed to a recycle hydrogen compressor recycle to the hydrofiner. The
4 liquid stream is fractionated into a top, middle, and bottom stream. The
5 bottom stream is recycled to the hydrocracker. The middle stream is mixed
6 with hydrogen from a make-up hydrogen compressor and directed to a
7 hydrogenator. Hydrogen recovered from the hydrogenator is compressed in a
8 stage of the make-up hydrogen compressor and directed to the hydrofiner.
9
10 U.S. Pat. No. 5,114,562 (Haun et al.) teaches a two-stage
11 hydrodesulfurization (similar to hydrotreating) and hydrogenation process for
12 distillate hydrocarbons. There is heat exchange between the two stages, but
13 a single hydrogen loop is not employed. Two separate reaction zones are
14 employed in series, the first zone for hydrodesulfurization and a second zone
15 for hydrogenation. A feed is mixed with recycled hydrogen and fed to a
16 desulfurization reactor. Hydrogen sulfide is stripped from the desulfurization
17 reactor product by a countercurrent flow of hydrogen. The liquid product
18 stream from this stripping operation is mixed with relatively clean recycled
19 hydrogen and the mixture is fed to a hydrogenation reaction zone. Hydrogen
20 is recovered from the hydrogenation reactor and recycled as a split stream to
21 both the desulfurization reactor and the hydrogenation reactor. The hydrogen
22 from the stripping operation is passed through a separator, mixed with the
23 portion of the recycled hydrogen directed to the hydrogenation reactor,
24 compressed, passed through a treating step and recycled to the
25 hydrogenation reactor. Thus, the hydrocarbon feed stream passes in series
26 through the desulfurization and hydrogenation reactors, while relatively low
27 pressure hydrogen is provided for the desulfurization step and relatively high
28 pressure hydrogen is provided for the hydrogenation step.
29

SUMMARY OF THE INVENTION

In accordance with a first aspect of the present invention, there is provided a method for hydroprocessing a hydrocarbon feedstock, said method employing multiple reaction zones within a single reaction loop, comprising the following steps: (a) passing a hydrocarbonaceous feedstock to a first hydroprocessing zone having one or more beds containing hydroprocessing catalyst, the hydroprocessing zone being maintained at hydroprocessing conditions, wherein the feedstock is contacted with catalyst and hydrogen; (b) passing the effluent of step (a) directly to a hot high pressure separator, wherein the effluent is contacted with a hot, hydrogen-rich stripping gas to produce a vapor stream comprising hydrogen, hydrocarbonaceous compounds boiling at a temperature below the boiling range of the hydrocarbonaceous feedstock, hydrogen sulfide, ammonia, and a bottoms stream comprising hydrocarbonaceous compounds boiling in approximately the same range of said hydrocarbonaceous feedstock along with a portion of the hydrocarbonaceous compounds boiling in the diesel boiling range; (c) passing the vapor stream from step (b) after cooling and partial condensation to a hot hydrogen stripper containing at least one bed of hydrotreating catalyst, where it is contacted countercurrently with hydrogen, while the liquid stream of step (b) is passed to a second stage reactor; (d) passing the overhead vapor stream from the hot hydrogen stripper of step (c), after cooling and contacting with water, to a first cold high pressure separator where hydrogen, hydrogen sulfide and light hydrocarbonaceous gases are removed overhead and a liquid stream comprising naphtha and middle distillates is passed to fractionation, thereby removing most of the ammonia and some of the hydrogen sulfide (as ammonium bi-sulfide in the sour water stream as it leaves the cold high-pressure separator); (e) passing the liquid stream from the hot hydrogen stripper of step (c) to a bed of hydroprocessing catalyst in the second reactor stage wherein the liquid is contacted under hydroprocessing conditions with the catalyst, in the presence of hydrogen; (f) passing the overhead from the cold high pressure separator of step (d) to an amine absorber, where hydrogen sulfide is removed before hydrogen is compressed and recycled to hydroprocessing vessels within the loop; (g) passing the separator bottoms of step (b) to a second reaction stage where it

is contacted with at least one bed of hydrocracking catalyst in the presence of hydrogen to produce a vapor stream and liquid effluent; (h) passing the vapor stream of step (g) after cooling to a second cold high-pressure separator where a vapor stream is removed comprising primarily hydrogen and light hydrocarbonaceous gases; (i) passing the liquid effluent of step (g) after cooling to the cold high-pressure separator of step (h) to separate hydrogen and light hydrocarbonaceous gases from the liquid effluent; (j) passing the vapor stream from steps (h) and (i) after further cooling and separation of condensate, to the make-up hydrogen compressor; (k) passing the compressed hydrogen from the make-up hydrogen compressor to the primary reactor loop; and (l) passing the liquid effluent from steps (h) and (i) to the fractionation system.

In accordance with a second aspect of the present invention, there is provided an integrated hydroconversion process having at least two stages, each stage possessing at least one reaction zone, comprising: (a) combining a first refinery stream with a first hydrogen-rich gaseous stream to form a first feedstock; (b) passing the first feedstock to a reaction zone of the first stage, which is maintained at conditions sufficient to effect a boiling range conversion, to form a first reaction zone effluent comprising normally liquid phase components and normally gaseous phase components; (c) passing the first reaction zone effluent of step (b) to a heat exchanger or series of exchangers, where it exchanges heat with a second refinery stream; (d) combining the first reaction zone effluent of step (b) with the second refinery stream of step (c) to form a second feedstock; (e) passing the second feedstock of step (d) to a reaction zone of the second stage, which is maintained at conditions sufficient for converting at least a portion of the aromatics present in the second refinery stream, to form a second reaction zone effluent; (f) separating the second reaction zone effluent of step (e) into a liquid stream comprising products and a second hydrogen-rich gaseous stream; (g) recycling at least a portion of the second hydrogen-rich gaseous stream of step (f) to a reaction zone of the first stage; and (h) passing the liquid stream comprising products of step (f) to a fractionation column, wherein product streams comprise gas or naphtha stream removed overhead one or more middle distillate streams, and a bottoms stream suitable for further processing.

In accordance with a third aspect of the present invention, there is provided an integrated hydroconversion process having; at least two stages, each stage possessing at least one reaction zone, comprising: (a) combining a first refinery stream with a first hydrogen-rich gaseous stream to form a first feedstock; (b) passing the first feedstock to a reaction zone of the first stage, which is maintained at conditions sufficient to effect a boiling range conversion, to form a first reaction zone effluent comprising normally liquid phase components and normally gaseous phase components; (c) passing the first reaction zone effluent of step (b) to a heat exchanger or series of exchangers, where it exchanges heat with other refinery streams; (d) passing the effluent of step (c) to a hot high pressure separator, where it is separated into a liquid stream which is passed to fractionation, and a gaseous stream, which is combined with a second refinery stream which comprises light cycle oil, light gas oil, atmospheric gas oil or mixtures of all three; (e) passing the combined gaseous stream of step (d) to a reaction zone of the second stage, which is maintained at conditions sufficient for converting at least a portion of the aromatics present in the second refinery stream, to form a second reaction zone effluent; (f) separating the second reaction zone effluent of step (e) into a liquid stream comprising products and a second hydrogen-rich gaseous stream; (g) recycling at least a portion of the second hydrogen-rich gaseous stream of step (f) to a reaction zone of the first stage; and (h) passing the liquid stream comprising products of step (f) to a fractionation column, wherein product streams comprise a gas or naphtha stream removed overhead, one or more middle distillate streams, and a bottoms stream suitable for further processing.

The first embodiment is disclosed in Figure 1. The process configuration for the first embodiment is different from U.S. Pat. No. 5,980,729 in many aspects. The primary reactor can be a combination hydrotreating- hydrocracking reactor that uses no recycle liquid.

Liquid from the hot stripper downstream of the reactor can be reduced in pressure to a subsequent reaction stage where hydrocracking reactions are completed. Pump may not be involved in the transfer of liquid. Also, the second hydrocracking stage may operate at lower pressure than the primary reaction stage.

Moderate to high conversion can be achieved using a single hydrogen loop. Product quality can be modulated to just meet specifications, eliminating product loss and saving hydrogen. The pressures of the reaction stages can be maintained at levels suited for particular types of feed characteristics, i.e., only the first stage reactor that processes the most difficult feed must operate at the highest pressure level. High-temperature, high-pressure pumps may not be involved in the process. The second hydrocracking reactor stage can operate in either co-current or counter-current mode with respect to the reaction gas, which may be primarily make-up hydrogen. The second hydrocracking reaction stage can be fed with high purity make-up hydrogen to maximize hydrogen partial pressure. The second stage may be loaded with very high activity catalyst that can be used for hydrocracking at relatively low pressures.

The second embodiment is disclosed in Figures 2 and 3. A VGO stream can be initially hydrocracked in a first-stage hydrocracking reaction zone within an integrated hydroconversion process. The integrated hydroconversion process may possess at least one hydrocracking stage and at least one hydrotreating stage. Effluent from the first-stage hydrocracking

reaction zone can be combined with a light aromatic-containing feed stream, and the blended stream can be hydrotreated in a second stage, which comprises a hydrotreating reaction zone. Heat exchange can occur between the first-stage hydrocracking reaction zone and the second-stage hydrotreating reaction zone, permitting the temperature control of the first-stage hydrotreating zone. The temperature of the first-stage hydrotreater is lower than that of the first-stage hydrocracker. This improves the aromatic saturation of the converted hydrocarbons and also allows the catalyst of the first-stage hydrotreating zone to be different from the catalyst in subsequent hydrocracking zones that may be present. In one embodiment, the effluent from the first-stage hydrotreater is heated in an exchanger, then passed to a hot high pressure separator, where overhead light ends are removed and passed to a cold high pressure separator. In the cold high pressure separator, hydrogen and hydrogen sulfide gas can be removed overhead and materials boiling in the gasoline and diesel range may be passed to a fractionator. Hydrogen sulfide can be subsequently removed in an absorber and hydrogen can be compressed and recirculated to be used as interbed quench, as well as mixed with vacuum gas oil feed.

The liquid effluent of the hot high pressure separator, which may contain materials boiling in the diesel range, is also passed to the fractionator. The fractionator bottoms may be subsequently hydrocracked and products may be subsequently hydrotreated in units not depicted.

The second embodiment offers several notable benefits. A method for hydroprocessing two refinery streams using a single hydrogen supply and a single hydrogen recovery system is disclosed. Furthermore, there can be provided a method for hydrocracking a refinery stream and hydrotreating a second refinery stream with a common hydrogen feed supply. The feed to the hydrocracking reaction zone is not poisoned with contaminants present in the feed to the hydrotreating reaction

zone. Disclosed, further herein is hydroprocessing two or more dissimilar refinery streams in an integrated hydroconversion process while maintaining good catalyst life and high yields of the desired products, particularly distillate range refinery products. Such dissimilar refinery streams may originate from different refinery processes, such as a VGO, derived from the effluent of a VGO hydrotreater, which can contain relatively few catalyst contaminants and/or aromatics, and an FCC cycle oil or straight run diesel, which may contain substantial amounts of aromatic compounds.

BRIEF DESCRIPTION OF THE DRAWINGS

The Figures illustrate multiple reaction stages employing a single hydroprocessing loop. Figure 1 depicts the use of an interstage hot stripper and an interstage hot separator.

Figure 2 illustrates a hydrocracker and hydrotreater, in series, in a single hydrogen loop separated by a heat exchanger. Light and heavy materials are separated from each other. Hydrogen and hydrogen sulfide might be removed from the light products. Hydrogen is compressed and recirculated. Products are sent to a fractionator.

Figure 3 illustrates a hydrocracking step followed by separation and fractionation. Material removed overhead is combined with a light aromatic stream and hydrotreated. Hydrogen is separated from the hydrotreated effluent and recirculated. Products are sent to a fractionator.

DETAILED DESCRIPTION OF THE INVENTION

Description of Figure 1

Preheated Oil feed in stream 1 is mixed with hydrogen in stream 40, which is preheated recycle and make-up hydrogen gas (reactor feed gas). The feed has been preheated in a process heat exchanger pumped up to the reactor pressure by the feed pump. The mixture of feed and reactor feed gas, now in stream 2, gets further preheated by heat exchange (in exchanger 41) and a final furnace (42) before it enters the first stage, downflow fixed bed primary reactor (3). The primary or first stage reactor contains multiple beds of hydroprocessing catalysts which may be beds of either hydrotreating or hydrocracking catalysts. Cold hydrogen from the recycle gas compressor is used as interbed quench (4,5,6).

The effluent 7 of the first stage reactor, which has been hydrotreated and partially hydrocracked, contains hydrogen sulfide, ammonia, light gases, naphtha, middle distillate, and hydrotreated vacuum gas oil. The effluent enters the hot high pressure separator (8) at slightly lower pressure and at slightly lower temperature, where most of the diesel and lighter material is separated from the unconverted oil. The hot high pressure separator has disc and doughnut type trays. Hydrogen rich gas, heated in exchanger 38, is introduced at the bottom for stripping through stream 9.

Stream 11 contains the overhead from the hot high-pressure separator. At this point, external feeds boiling in the middle-distillates boiling range such as Light Cycle Oil (LCO), Light Coker Gas Oil (LCGO), Atmospheric Gas Oil (AGO), Light Visbreaker Gas Oil (LVBGO), etc., can be introduced (10). Stream 11 is cooled by process heat exchange or by steam generation before entering the high-pressure hydrogen stripper-hydrotreater (14). Liquid in stream 11 flows downward through a bed of packing containing hydrotreating

1 catalyst, while being contacted with countercurrent flowing hydrogen from
2 stream 25.

3

4 The overhead stream 15 contains primarily hydrogen, ammonia and hydrogen
5 sulfide, along with some light gases and naphtha. It is cooled by process heat
6 exchange (44), contacted with water (45), and further cooled by air cooling
7 (46) before being fed to the Cold High Pressure Separator No. 1 (17). The
8 water injection allows the removal of most of the ammonia from the hydrogen
9 gas as ammonium bisulfide solution. Hydrogen, hydrogen sulfide and light
10 hydrocarbonaceous gases are removed overhead as stream 18. Stream 20 is
11 a sour water stream containing ammonium bisulfide. Stream 19 is a
12 hydrocarbonaceous stream containing naphtha, kerosene and diesel range
13 products. Stream 18 is sent to an amine absorber (21) where almost the
14 entire quantity of hydrogen sulfide is removed from the hydrogen-rich stream
15 by contacting with amine (47). After removal of the hydrogen sulfide, the gas
16 is sent for compression to the recycle gas compressor (23). The compressed
17 recycle gas (24) is split into streams 25 and 26. Stream 26 is further split into
18 the first stage recycle gas feed (27) and stream 28 that supplies the quench to
19 the first stage. Risk amine leaves the amine absorber as stream 48.

20

21 Bottoms from the hot high-pressure separator, stream 12, can be reduced in
22 pressure and cooled down by process heat exchange before being fed to the
23 second stage reactor (30) where hydrocracking reactions are completed and
24 unconverted material in stream 12 is further converted to diesel and lighter
25 products. The second stage reactor is fed with high purity make-up hydrogen
26 (31) from an intermediate stage of the make-up hydrogen compressor (49).
27 The hydrogen, in the preferred mode, flows up the reactor in countercurrent
28 fashion for maximizing the benefits of hydrogen partial pressure. The
29 invention will also work with co-current introduction of make-up hydrogen.
30 The second stage reactor feed gas requirements in terms of adequate
31 gas-to-oil ratio can be met by introducing all of the make-up hydrogen

1 required in all reaction stages to the front of second stage reactor. The
2 invention has the provision, however, to introduce recycle hydrogen from the
3 recycle gas compressor through stream 35.

4

5 The second reaction stage operates under a clean, ammonia and hydrogen
6 sulfide free environment and thus hydrocracking rate constants are much
7 higher. Catalyst deactivation is much reduced. These factors enable the
8 operation at lower hydrogen partial pressures and with reduced catalyst
9 requirements.

10

11 The lower bed or beds of the second stage reactor (30) can be loaded with
12 hydrotreating catalyst where diesel range material (16) from the hydrogen
13 stripper (14) can be introduced for completion of aromatic saturation and other
14 hydroprocessing reactions. Alternately, stream 16 can be diverted directly to
15 the fractionation section if the diesel quality is adequate.

16

17 There are at least two, preferably three to four, beds of hydroprocessing
18 catalyst in reactor 30. The catalyst can be either base metal or noble metal
19 hydroprocessing catalyst.

20

21 Stream 33, which comes from the top of the reactor, contains primarily
22 hydrogen, although some H₂S and ammonia may be present. It is cooled by
23 process heat exchange (50) before being sent to Cold High Pressure
24 Separator No. 2 (17.5). The overhead vapor of Cold High Pressure Separator
25 No. 2 passes to the make-up hydrogen compressor (49), to the final stage of
26 compression.

27

28 The liquid effluent from reactor 30, Stream 34, which contains light gases,
29 naphtha, middle distillate and hydrotreated gas oil, is cooled by process heat
30 exchange (51) and sent to Cold High Pressure Separator No. 2 (17.5).

31

1 Bottoms (line 37) from the Cold High Pressure Separator No. 2 is sent to
2 fractionation.

3
4 The Make-up hydrogen compressor (49) is a multi-stage machine with
5 typically three to four compression stages. After each stage of compression,
6 the gas is cooled and any condensate knocked out in a knock-out drum
7 (KOD). For this invention, the gas to the second reaction stage is withdrawn
8 after an intermediate stage of compression. The gas stream (31) is sent to
9 the second reaction stage (30) and is returned via the Cold High Pressure
10 Separator No. 2 (stream 36) to the final stage of compression of the make-up
11 hydrogen compressor.

12
13 After the final stage of compression, the high-pressure make-up hydrogen is
14 sent to the first reaction stage, stream 39 and to the hot separator.

15
16 Reference is now made to Figure 2, which discloses preferred embodiments
17 of the invention. Not included in the figures are various pieces of auxiliary
18 equipment such as heat exchangers, condensers, pumps and compressors,
19 which are not essential to the invention.

20
21 In Figure 2, two downflow reactor vessels, 5 and 15 are depicted. Between
22 them is heat exchanger 20. Each vessel contains at least one reaction zone.
23 The first-stage reaction, hydrocracking, occurs in vessel 5. The second-stage
24 reaction, hydrotreating, occurs in vessel 15. Each vessel is depicted as
25 having three catalyst beds. The first reaction vessel 5 is for cracking a first
26 refinery stream 1. The second reaction vessel 15 is for removing
27 nitrogen-containing and aromatic molecules from a second refinery stream 17.
28 A suitable volumetric ratio of the catalyst volume in the first reaction vessel to
29 the catalyst volume in the second reaction vessel encompasses a broad
30 range, depending on the ratio of the first refinery stream to the second refinery
31 stream. Typical ratios generally lie between 20:1 and 1:20. A preferred

1 volumetric range is between 10:1 and 1:10. A more preferred volumetric ratio
2 is between 5:1 and 1:2.

3

4 In the integrated process, a first refinery stream 1 is combined with a
5 hydrogen-rich gaseous stream 4 to form a first feedstock 12. The stream
6 exiting furnace 30, stream 13, is passed to first reaction vessel 5.

7 Hydrogen-rich gaseous stream 4 contains greater than 50% hydrogen, the
8 remainder being varying amounts of light gases, including hydrocarbon gases.

9 The hydrogen-rich gaseous stream 4 shown in the drawing is a blend of
10 make-up hydrogen 3 and recycle hydrogen 26. While the use of a recycle
11 hydrogen stream is generally preferred for economic reasons, it is not
12 required. First feedstock 1 may be heated in one or more exchangers, such
13 as exchanger 10, emerging as stream 12, and in one or more heaters, such
14 as heater 30, (emerging as stream 13) before being introduced to first
15 reaction vessel 5 in which hydrocracking preferably occurs. Hydrotreating
16 preferably occurs in vessel 15.

17

18 Hydrogen may also be added as a quench stream through lines 6 and 7, and
19 9 and 11, (which also come from hydrogen stream 4) for cooling the first and
20 the second reaction stages, respectively. The effluent from the hydrocracking
21 stage, stream 14 is cooled in heat exchanger 20 by stream 2. Stream 2 boils
22 in the diesel range and may be light cycle oil, light gas oil, atmospheric gas
23 oil, or a mixture of the three. Stream 2 emerges from exchanger 20 as
24 stream 16 and combines stream 14 as it emerges from exchanger 20 to form
25 combined feedstock 17. Hydrogen in stream 8 joins the combined feedstock
26 17 before it enters vessel 15. Stream 17 enters vessel 15 for hydrotreatment,
27 and exits as stream 18.

28

29 The second reaction stage, found in vessel 15, contains at least one bed of
30 catalyst, such as hydrotreating catalyst, which is maintained at conditions

1 sufficient for converting at least a portion of the nitrogen compounds and at
2 least a portion of the aromatic compounds in the second feedstock.

3

4 Hydrogen stream 4 may be recycle hydrogen from compressor 40.

5 Alternately, stream 4 may be a fresh hydrogen stream, originating from
6 hydrogen sources external to the present process.

7

8 Stream 18, the second reaction zone effluent, contains thermal energy which
9 may be recovered by heat exchange, such as in heat exchanger 10. Second
10 stage effluent 18 emerges from exchanger 10 as stream 19 and is passed to
11 hot high pressure separator 25. The liquid effluent of the hot high pressure
12 separator 25, stream 22 is passed to fractionation. The overhead gaseous
13 stream from separator 25, stream 21, is combined with water from stream 23
14 for cooling. The now cooled stream 21 enters the cold high pressure
15 separator 35. Light liquids are passed to fractionation in stream 27 (which
16 joins stream 22) and sour water is removed through stream 34. Gaseous
17 overhead stream 24 passes to amine absorber 45, for the removal of
18 hydrogen sulfide gas. Purified hydrogen then passes, through stream 26, to
19 the compressor 40, where it is recompressed and passed as recycle to one or
20 more of the reaction vessels and as a quench stream for cooling the reaction
21 zones. Such uses of hydrogen are well known in the art.

22

23 An example separation scheme for a hydroconversion process is taught in
24 U.S. Patent No. 5,082,551, the entire disclosure of which is incorporated
25 herein by reference for all purposes.

26

27 The absorber 45 may include means for contacting a gaseous component of
28 the reaction effluent 19 with a solution, such as an alkaline aqueous solution,
29 for removing contaminants such as hydrogen sulfide and ammonia which may
30 be generated in the reaction stages and may be present in reaction effluent
31 19. The hydrogen-rich gaseous stream 24 is preferably recovered from the

1 separation zone at a temperature in the range of 100°F-300°F or
2 100°F-200°F.

3
4 Liquid stream 22 is further separated in fractionator 50 to produce overhead
5 gasoline stream 28, naphtha stream 29, kerosene fraction 31, diesel stream
6 32 and fractionator bottoms 33. A preferred distillate product has a boiling
7 point range within the temperature range 250°F-700°F. A gasoline or naphtha
8 fraction having a boiling point range within the temperature range C₅-400°F is
9 also desirable.

10
11 In Figure 3, two downflow reactor vessels, 5 and 15, are depicted. The first
12 stage reaction, hydrocracking, occurs in vessel 5. The second stage,
13 hydrotreating, occurs in vessel 15. Each vessel contains at least one reaction
14 zone. Each vessel is depicted as having three catalyst beds. The first
15 reaction vessel 5 is for cracking a first refinery stream 1. The second reaction
16 vessel 15 is for removing nitrogen-containing and aromatic molecules from a
17 second refinery stream 34. A suitable volumetric ratio of the catalyst volume
18 in the first reaction vessel to the catalyst volume in the second reaction vessel
19 encompasses a broad range, depending on the ratio of the first refinery
20 stream to the second refinery stream. Typical ratios generally lie between
21 20:1 and 1:20. A preferred volumetric range is between 10:1 and 1:10. A
22 more preferred volumetric ratio is between 5:1 and 1:2.

23
24 In the integrated process, a first refinery stream 1 is combined with a
25 hydrogen-rich gaseous stream 4 to form a first feedstock 12 which is passed
26 to first reaction vessel 5. Hydrogen-rich gaseous stream 4 contains greater
27 than 50% hydrogen, the remainder being varying amounts of light gases,
28 including hydrocarbon gases. The hydrogen-rich gaseous stream 4 shown in
29 the drawing is a blend of make-up hydrogen 3 and recycle hydrogen 26.
30 While the use of a recycle hydrogen stream is generally preferred for
31 economic reasons, it is not required. First feedstock 1 may be heated in one

1 or more exchangers or in one or more heaters before being combined with
2 hydrogen-rich stream 4 to create stream 12. Stream 12 is then introduced to
3 first reaction vessel 5, where the first stage, in which hydrocracking preferably
4 occurs, is located. The second stage is located in vessel 15, where
5 hydrotreating preferably occurs.

6
7 The effluent from the first stage, stream 14 is heated in heat exchanger 20.
8 Stream 14 emerges from exchanger 20 as stream 17 and passes to the
9 "hot/hot" high pressure separator 55. The liquid stream 36 emerges from the
10 "hot/hot" high pressure separator 55 and proceeds to fractionator 60. Stream
11 37 represents products streams for gasoline and naphtha, stream 38
12 represents distillate recycled back to the inlet of hydrotreater 15, and stream
13 39 represents clean bottoms material.

14
15 The gaseous stream 34 emerges from the "hot/hot" high pressure separator
16 55, and joins with stream 2, which boils in the diesel range and may be light
17 cycle oil, light gas oil, atmospheric gas oil, or a mixture of the three. It further
18 combines with hydrogen-rich stream 4 prior to entering vessel 15 for
19 hydrotreatment, and exits as stream 18.

20
21 The second reaction zone, found in vessel 15, contains at least one bed of
22 catalyst, such as hydrotreating catalyst, which is maintained at conditions
23 sufficient for converting at least a portion of the nitrogen compounds and at
24 least a portion of the aromatic compounds in the second feedstock.

25
26 Hydrogen stream 4 may be recycle hydrogen from compressor 40.
27 Alternately, stream 4 may be a fresh hydrogen stream, originating from
28 hydrogen sources external to the present process.

29
30 Stream 18, the second stage effluent, contains thermal energy which may be
31 recovered by heat exchange, such as in heat exchanger 10. Second stage

1 effluent 18 emerges from exchanger 10 as stream 19 and is passed to hot
2 high pressure separator 25. The liquid effluent of the hot high pressure
3 separator 25, stream 22 is passed to fractionation. The overhead gaseous
4 stream from separator 25, stream 21, is combined with water from stream 23
5 for cooling. The now cooled stream 21 enters the cold high pressure
6 separator 35. Light liquids are passed to fractionation in stream 27 (which
7 joins stream 22) and sour water is removed through stream 41. Gaseous
8 overhead stream 24 passes to amine absorber 45, for the removal of
9 hydrogen sulfide gas. Purified hydrogen then passes, through stream 26, to
10 the compressor 40, where it is recompressed and passed as recycle to one or
11 more of the reaction vessels and as a quench stream for cooling the reaction
12 zones. Such uses of hydrogen are well known in the art.

13
14 The absorber 45 may include means for contacting a gaseous component of
15 the reaction effluent 19 (stream 24) with a solution, such as an alkaline
16 aqueous solution, for removing contaminants such as hydrogen sulfide and
17 ammonia which may be generated in the reaction zones and may be present
18 in reaction effluent 19. The hydrogen-rich gaseous stream 24 is preferably
19 recovered from the separation zone at a temperature in the range of
20 100°F-300°F or 100°F-200°F.

21
22 Liquid stream 22 is further separated in fractionator 50 to produce overhead
23 gasoline stream 28, naphtha stream 29, kerosene fraction 31, diesel stream
24 32 and fractionator bottoms 33. A preferred distillate product has a boiling
25 point range within the temperature range 250°F-700°F. A gasoline or naphtha
26 fraction having a boiling point range within the temperature range C₅-400°F is
27 also desirable.

28

1 Feeds

2

3 A wide variety of hydrocarbon feeds may be used in first embodiment of this
4 invention. Typical feedstocks include any heavy or synthetic oil fraction or
5 process stream having a boiling point above 392°F (200°C). Such feedstocks
6 include vacuum gas oils, heavy atmospheric gas oil, delayed coker gas oil,
7 visbreaker gas oil demetallized oils, vacuum residua, atmospheric residua,
8 deasphalted oil, Fischer-Tropsch streams, and FCC streams.

9

10 In the case of the second embodiment one suitable first refinery feed stream
11 is a VGO having a boiling point range starting at a temperature above 500°F
12 (260°C), usually within the temperature range of 500°F-1100°F
13 (260°C-593°C). A refinery stream wherein 75 vol% of the refinery stream
14 boils within the temperature range 650°F-1050°F is an example feedstock for
15 the first reaction zone. The first refinery stream may contain nitrogen, usually
16 present as organonitrogen compounds. VGO feed streams for the first
17 reaction zone contain less than about 200 ppm nitrogen and less than
18 0.25 wt. % sulfur, though feeds with higher levels of nitrogen and sulfur,
19 including those containing up to 0.5 wt. % and higher nitrogen and up to
20 5 wt. % sulfur and higher may be treated in the present process. The first
21 refinery stream is also preferably a low asphaltene stream. Suitable first
22 refinery streams contain less than about 500 ppm asphaltenes, preferably
23 less than about 200 ppm asphaltenes, and more preferably less than about
24 100 ppm asphaltenes. Example streams include light gas oil, heavy gas oil,
25 straight run gas oil, deasphalted oil, and the like. The first refinery stream
26 may have been processed, e.g., by hydrotreating, prior to the present process
27 to reduce or substantially eliminate its heteroatom content. The first refinery
28 stream may comprise recycle components.

29

30 The hydrocracking reaction step removes nitrogen and sulfur from the first
31 refinery feed stream in the first hydrocracking reaction zone and effects a

1 boiling range conversion, so that the liquid portion of the first hydrocracking
2 reaction zone effluent has a normal boiling range below the normal boiling
3 point range of the first refinery feedstock. By "normal" is meant a boiling point
4 or boiling range based on a distillation at one atmosphere pressure, such as
5 that determined in a D1160 distillation. Unless otherwise specified, all
6 distillation temperatures listed herein refer to normal boiling point and normal
7 boiling range temperatures. The process in the first hydrocracking reaction
8 zone may be controlled to a certain cracking conversion or to a desired
9 product sulfur level or nitrogen level or both. Conversion is generally related
10 to a reference temperature, such as, for example, the minimum boiling point
11 temperature of the hydrocracker feedstock. The extent of conversion relates
12 to the percentage of feed boiling above the reference temperature which is
13 converted to products boiling below the reference temperature.

14
15 The hydrocracking reaction zone effluent includes normally liquid phase
16 components, e.g., reaction products and unreacted components of the first
17 refinery stream, and normally gaseous phase components, e.g., gaseous
18 reaction products and unreacted hydrogen. In the process, the hydrocracking
19 reaction zone is maintained at conditions sufficient to effect a boiling range
20 conversion of the first refinery stream of at least about 25%, based on a 650°F
21 reference temperature. Thus, at least 25% by volume of the components in
22 the first refinery stream which boil above about 650°F are converted in the
23 first hydrocracking reaction zone to components which boil below about
24 650°F. Operating at conversion levels as high as 100% is also within the
25 scope of the invention. Example boiling range conversions are in the range of
26 from about 30% to 90% or of from about 40% to 80%. The hydrocracking
27 reaction zone effluent is further decreased in nitrogen and sulfur content, with
28 at least about 50% of the nitrogen containing molecules in the first refinery
29 stream being converted in the hydrocracking reaction zone. Preferably, the
30 normally liquid products present in the hydrocracking reaction zone effluent
31 contain less than about 1000 ppm sulfur and less than about 200 ppm

1 nitrogen, more preferably less than about 250 ppm sulfur and about 100 ppm
2 nitrogen.

3

4 Catalyst

5

6 Each hydroprocessing zone in either embodiment may contain only one
7 catalyst, or several catalysts in combination. In the preferred embodiment,
8 hydrocracking is occurring in the first zone and hydrotreating is occurring in
9 the second zone.

10

11 The hydrocracking catalyst generally comprises a cracking component, a
12 hydrogenation component, and a binder. Such catalysts are well known in the
13 art. The cracking component may include an amorphous silica/alumina phase
14 and/or a zeolite, such as a Y-type or USY zeolite. Catalysts having high
15 cracking activity often employ REX, REY and USY zeolites. The binder is
16 generally silica or alumina. The hydrogenation component will be a Group VI,
17 Group VII, or Group VIII metal or oxides or sulfides thereof, preferably one or
18 more of iron, chromium, molybdenum, tungsten, cobalt, or nickel, or the
19 sulfides or oxides thereof. If present in the catalyst, these hydrogenation
20 components generally make up from about 5% to about 40% by weight of the
21 catalyst. Alternatively, noble metals, especially platinum and/or palladium,
22 may be present as the hydrogenation component, either alone or in
23 combination with the base metal hydrogenation components: iron, chromium
24 molybdenum, tungsten, cobalt, or nickel. If present, the platinum group
25 metals will generally make up from about 0.1% to about 2% by weight of the
26 catalyst.

27

28 Hydrotreating catalyst usually is designed to remove sulfur and nitrogen and
29 provide a degree of aromatic saturation. It will typically be a composite of a
30 Group VI metal or compound thereof, and a Group VIII metal or compound
31 thereof supported on a porous refractory base such as alumina. Examples of

1 hydrotreating catalysts are alumina supported cobalt-molybdenum, nickel
2 sulfide, nickel-tungsten, cobalt-tungsten and nickel-molybdenum. Typically,
3 such hydrotreating catalysts are presulfided.

4
5 Catalyst selection is dictated by process needs and product specifications. In
6 particular, a noble catalyst may be used in the second stage when there is a
7 low amount of H_2S present. A low acidity catalyst may be used in the bottom
8 of the second stage hydrocracker in order to avoid overcracking distillate to
9 gas and naphtha.

10
11 Conditions - Hydrocracking Stage

12
13 Reaction conditions in the hydrocracking reaction zone include a reaction
14 temperature between about 250°C and about 500°C (482°F-932°F),
15 pressures from about 3.5 MPa to about 24.2 MPa (500-3,500 psi), and a feed
16 rate (vol oil/vol cat h) from about 0.1 to about 20 hr^{-1} . Hydrogen circulation
17 rates are generally in the range from about 350 std liters H_2 /kg oil to 1780 std
18 liters H_2 /kg oil (2,310-11,750 standard cubic feet per barrel). Preferred
19 reaction temperatures range from about 340°C to about 455°C (644°F-851°F).
20 Preferred total reaction pressures range from about 7.0 MPa to about
21 20.7 MPa (1,000-3,000 psi). With the preferred catalyst system, it has been
22 found that preferred process conditions include contacting a petroleum
23 feedstock with hydrogen under hydrocracking conditions comprising a
24 pressure of about 13.8 MPa to about 20.7 MPa (2,000-3000 psi), a gas to oil
25 ratio between about 379-909 std liters H_2 /kg oil (2,500-6,000 scf/bbl), a LHSV
26 of between about 0.5-1.5 hr^{-1} , and a temperature in the range of 360°C to
27 427°C (680°F-800°F).

28

1 Feed and Effluent Characteristics - Hydrotreater Stage

2

3 The second refinery feedstream has a boiling point range generally lower than
4 the first refinery feedstream. Indeed, it is a feature of the present process that
5 a substantial portion of the second refinery feedstream has a normal boiling
6 point in the middle distillate range, so that cracking to achieve boiling point
7 reduction is not necessary. Thus, at least about 75 vol% of a suitable second
8 refinery stream has a normal boiling point temperature of less than about
9 1000°F. A refinery stream with at least about 75% v/v of its components
10 having a normal boiling point temperature within the range of 250°F-700°F is
11 an example of a preferred second refinery feedstream.

12

13 The process of this invention is particularly suited for treating middle distillate
14 streams which are not suitable for high quality fuels. For example, the
15 process is suitable for treating a second refinery stream which contains high
16 amounts of nitrogen and/or high amounts of aromatics, including streams
17 which contain up to 90% aromatics and higher. Example second refinery
18 feedstreams which are suitable for treating in the present process include
19 straight run vacuum gas oils, including straight run diesel fractions, from crude
20 distillation, atmospheric tower bottoms, or synthetic cracked materials such as
21 coker gas oil, light cycle oil or heavy cycle oil.

22

23 After the first refinery feedstream is treated in the hydrocracking stage, the
24 first hydrocracking reaction zone effluent is combined with the second
25 feedstock, and the combination passed together with hydrogen over the
26 catalyst in the hydrotreating stage. Since the hydrocracked effluent is already
27 relatively free of the contaminants to be removed by hydrotreating, the
28 hydrocracker effluent passes largely unchanged through the hydrotreater.
29 And unreacted or incompletely reacted feed remaining in the effluent from the
30 hydrotreater is effectively isolated from the hydrocracker zone to prevent
31 contamination of the catalyst contained therein.

1 However, the presence of the hydrocracker effluent plays an important and
2 unexpected economic benefit in the integrated process. Leaving the
3 hydrocracker, the effluent carries with it substantial thermal energy. This
4 energy may be used to heat the second reactor feedstream in a heat
5 exchanger before the second feedstream enters the hydrotreater. This
6 permits adding a cooler second feed stream to the integrated system than
7 would otherwise be required, and saves on furnace capacity and heating
8 costs.

9
10 As the second feedstock passes through the hydrotreater, the temperature
11 again tends to increase due to exothermic reaction heating in the second
12 zone. The hydrocracker effluent in the second feedstock serves as a heat
13 sink, which moderates the temperature increase through the hydrotreater.
14 The heat energy contained in the liquid reaction products leaving the
15 hydrotreater is further available for exchange with other streams requiring
16 heating. Generally, the outlet temperature of the hydrotreater will be higher
17 than the outlet temperature of the hydrocracker. In this case, the instant
18 invention will afford the added heat transfer advantage of elevating the
19 temperature of the first hydrocracker feed for more effective heat transfer.
20 The effluent from the hydrocracker also carries the unreacted hydrogen for
21 use in the first-stage hydrotreater without any heating or pumping requirement
22 to increase pressure.

23 24 Conditions - Hydrotreater Stage

25
26 The hydrotreater is maintained at conditions sufficient to remove at least a
27 portion of the nitrogen compounds and at least a portion of the aromatic
28 compounds from the second refinery stream. The hydrotreater will operate at
29 a lower temperature than the hydrocracker, except for possible temperature
30 gradients resulting from exothermic heating within the reaction zones,
31 moderated by the addition of relatively cooler streams into the one or more

1 reaction zones. Feed rate of the reactant liquid stream through the reaction
2 zones will be in the region of 0.1 to 20 hr⁻¹ liquid hourly space velocity. Feed
3 rate through the hydrotreater will be increased relative to the feed rate through
4 the hydrocracker by the amount of liquid feed in the second refinery
5 feedstream and will also be in the region of 0.1 to 20 hr⁻¹ liquid hourly space
6 velocity. These process conditions selected for the first reaction zone may be
7 considered to be more severe than those conditions normally selected for a
8 hydrotreating process.

9
10 At any rate, hydrotreating conditions typically used in the hydrotreater will
11 include a reaction temperature between about 250°C and about 500°C
12 (482°F-932°F), pressures from about 3.5 MPa to about 24.2 MPa
13 (500-3,500 psi), and a feed rate (vol oil/vol cat h) from about 0.1 to about
14 20 hr⁻¹. Hydrogen circulation rates are generally in the range from about
15 350 std liters H₂/kg oil to 1780 std liters H₂/kg oil (2,310-11,750 standard cubic
16 feet per barrel). Preferred reaction temperatures range from about 340°C to
17 about 455°C (644°F-851°F). Preferred total reaction pressures range from
18 about 7.0 MPa to about 20.7 MPa (1,000-3,000 psi). With the preferred
19 catalyst system, it has been found that preferred process conditions include
20 contacting a petroleum feedstock with hydrogen in the presence of the
21 layered catalyst system under hydrocracking conditions comprising a
22 pressure of about 16.0 MPa (2,300 psi), a gas to oil ratio at from about
23 379-909 std liters H₂/kg oil (2,500 scf/bbl to about 6,000 scf/bbl), a LHSV of
24 between about 0.5-1.5 hr⁻¹, and a temperature in the range of 360°C to 427°C
25 (680°F-800°F). Under these conditions, at least about 50% of the aromatics
26 are removed from the second refinery stream in the hydrotreater. It is
27 expected that as much as 30-70% or more of the nitrogen present in the
28 second refinery stream would also be removed in the process. However,
29 cracking conversion in the hydrotreater would be generally low, typically less
30 than 20%. Standard methods for determining the aromatic content and the
31 nitrogen content of refinery streams are available. These include ASTM

1 D5291 for determining the nitrogen content of a stream containing more than
2 about 1500 ppm nitrogen. ASTM D5762 may be used for determining the
3 nitrogen content of a stream containing less than about 1500 ppm nitrogen.
4 ASTM D2007 may be used to determine the aromatic content of a refinery
5 stream.

6
7 Products
8

9 The embodiments of this invention are especially useful in the production of
10 middle distillate fractions boiling in the range of about 250-700°F (121-371°C).
11 A middle distillate fraction is defined as having an approximate boiling range
12 from about 250 to 700°F. At least 75 vol%, preferably 85 vol%, of the
13 components of the middle distillate have a normal boiling point of greater than
14 250°F. At least about 75 vol%, preferably 85 vol%, of the components of the
15 middle distillate have a normal boiling point of less than 700°F. The term
16 "middle distillate" includes the diesel, jet fuel and kerosene boiling range
17 fractions. The kerosene or jet fuel boiling point range refers to the range
18 between 280 and 525°F (38-274°C). The term "diesel boiling range" refers to
19 hydrocarbons boiling in the range from 250 to 700°F (121-371°C).

20
21 Gasoline or naphtha may also be produced in the process of this invention.
22 Gasoline or naphtha normally boils in the range below 400°F (204°C), or C₅-.
23 Boiling ranges of various product fractions recovered in any particular refinery
24 will vary with such factors as the characteristics of the crude oil source, local
25 refinery markets and product prices.

26
27 Heavy hydrotreated gas oil, another product of this invention, usually boils in
28 the range from 550 to 700°F.

29

1 Example

2

3 These are the conditions and results obtained using the process depicted in

4 Figure 1:

5

	Stage 1	Hydrogen Stripper/Hydrotreater	Stage 2
Catalyst Hydrogenation Component	Base metal	Base metal	Base metal or Noble Metal
LHSV, hr ⁻¹	0.3-3.5	0.3-2.0	0.5-7.0
Operating Temperature, °C	300-440	250-400	250-440
Reactor Inlet Pressure, kg/cm ² g	100-230	90-220	80-170
Gas/Oil Ratio Nm ³ /m ³	160-1500	160-1500	160-850
Conversion, %	20-70		25-75
Kerosene Smoke Point, mm	13-25		20-40
Diesel Cetane Number	30-55		50-75

6

7 Generally, cetane uplift is 20 to 45 and improvement in kerosene smoke point

8 is 7-27 mm.

1 WHAT IS CLAIMED IS:

2

3 1. A method for hydroprocessing a hydrocarbon feedstock, said method
4 employing multiple reaction zones within a single reaction loop,
5 comprising the following steps:

6

7 (a) passing a hydrocarbonaceous feedstock to a first hydroprocessing
8 zone having one or more beds containing hydroprocessing
9 catalyst, the hydroprocessing zone being maintained at
10 hydroprocessing conditions, wherein the feedstock is contacted
11 with catalyst and hydrogen;

12

13 (b) passing the effluent of step (a) directly to a hot high pressure
14 separator, wherein the effluent is contacted with a hot,
15 hydrogen-rich stripping gas to produce a vapor stream comprising
16 hydrogen, hydrocarbonaceous compounds boiling at a temperature
17 below the boiling range of the hydrocarbonaceous feedstock,
18 hydrogen sulfide, ammonia, and a bottoms stream comprising
19 hydrocarbonaceous compounds boiling in approximately the same
20 range of said hydrocarbonaceous feedstock along with a portion of
21 the hydrocarbonaceous compounds boiling in the diesel boiling
22 range;

23

24 (c) passing the vapor stream from step (b) after cooling and partial
25 condensation to a hot hydrogen stripper containing at least one
26 bed of hydrotreating catalyst, where it is contacted countercurrently
27 with hydrogen, while the liquid stream of step (b) is passed to a
28 second stage reactor;

29

30 (d) passing the overhead vapor stream from the hot hydrogen stripper
31 of step (c), after cooling and contacting with water, to a first cold

- 1 high pressure separator where hydrogen, hydrogen sulfide and
2 light hydrocarbonaceous gases are removed overhead and a liquid
3 stream comprising naphtha and middle distillates is passed to
4 fractionation, thereby removing most of the ammonia and some of
5 the hydrogen sulfide (as ammonium bi-sulfide in the sour water
6 stream as it leaves the cold high-pressure separator);
7
8 (e) passing the liquid stream from the hot hydrogen stripper of step (c)
9 to a bed of hydroprocessing catalyst in the second reactor stage
10 wherein the liquid is contacted under hydroprocessing conditions
11 with the catalyst, in the presence of hydrogen;
12
13 (f) passing the overhead from the cold high pressure separator of
14 step (d) to an amine absorber, where hydrogen sulfide is removed
15 before hydrogen is compressed and recycled to hydroprocessing
16 vessels within the loop;
17
18 (g) passing the separator bottoms of step (b) to a second reaction
19 stage where it is contacted with at least one bed of hydrocracking
20 catalyst in the presence of hydrogen to produce a vapor stream
21 and liquid effluent;
22
23 (h) passing the vapor stream of step (g) after cooling to a second cold
24 high-pressure separator where a vapor stream is removed
25 comprising primarily hydrogen and light hydrocarbonaceous gases;
26
27 (i) passing the liquid effluent of step (g) after cooling to the cold
28 high-pressure separator of step (h) to separate hydrogen and light
29 hydrocarbonaceous gases from the liquid effluent;
30

- (j) passing the vapor stream from steps (h) and (i) after further cooling and separation of condensate, to the make-up hydrogen compressor;
 - (k) passing the compressed hydrogen from the make-up hydrogen compressor to the primary reactor loop; and
 - (l) passing the liquid effluent from steps (h) and (i) to the fractionation system.
- 2. The method of claim 1, step (g), in which the hydrogen flows in a countercurrent direction to the liquid effluent of claim 1, step (b).
 - 3. The method of claim 1, wherein the feed is selected from the group consisting of vacuum gas oil, heavy atmospheric gas oil, delayed coker gas oil, visbreaker gas oil, demetallized oils, FCC light cycle oil, vacuum residua deasphalted oil, Fischer-Tropsch streams, and FCC streams.
 - 4. The method of claim 1, wherein the cetane number improvement occurring in step (e) ranges from 20 to 45.
 - 5. The method of claim 1, wherein the kerosene smoke point improvement occurring in step (e) ranges from 7 to 27.
 - 6. The method of claim 1, wherein the hydroprocessing catalyst of both stage 1 and stage 2 comprises both a cracking component and a hydrogenation component.
 - 7. An integrated hydroconversion process having at least two stages, each stage possessing at least one reaction zone, comprising:

- 1 (a) combining a first refinery stream with a first hydrogen-rich gaseous
2 stream to form a first feedstock;
3
- 4 (b) passing the first feedstock to a reaction zone of the first stage,
5 which is maintained at conditions sufficient to effect a boiling range
6 conversion, to form a first reaction zone effluent comprising
7 normally liquid phase components and normally gaseous phase
8 components;
9
- 10 (c) passing the first reaction zone effluent of step (b) to a heat
11 exchanger or series of exchangers, where it exchanges heat with
12 a second refinery stream;
13
- 14 (d) combining the first reaction zone effluent of step (b) with the
15 second refinery stream of step (c) to form a second feedstock;
16
- 17 (e) passing the second feedstock of step (d) to a reaction zone of the
18 second stage, which is maintained at conditions sufficient for
19 converting at least a portion of the aromatics present in the second
20 refinery stream, to form a second reaction zone effluent;
21
- 22 (f) separating the second reaction zone effluent of step (e) into a liquid
23 stream comprising products and a second hydrogen-rich gaseous
24 stream;
25
- 26 (g) recycling at least a portion of the second hydrogen-rich gaseous
27 stream of step (f) to a reaction zone of the first stage; and
28
- 29 (h) passing the liquid stream comprising products of step (f) to a
30 fractionation column, wherein product streams comprise gas or

- 1 naphtha stream removed overhead, one or more middle distillate
2 streams, and a bottoms stream suitable for further processing.
3
- 4 8. The process according to Claim 7 wherein the reaction zone of step 1(b)
5 stage is maintained at hydrocracking reaction conditions, including a
6 reaction temperature in the range of from about 340°C to about 455°C
7 (644°F-851°F), a reaction pressure in the range of about 3.5-24.2 MPa
8 (500-3500 pounds per square inch), a feed rate (vol oil/vol cat h) from
9 about 0.1 to about 10 hr⁻¹ and a hydrogen circulation rate ranging from
10 about 350 std liters H₂/kg oil to 1780 std liters H₂/kg oil (2,310-11,750
11 standard cubic feet per barrel).
12
- 13 9. The process according to Claim 7 wherein the reaction zone of step 1(e)
14 is maintained at hydrotreating reaction conditions, including a reaction
15 temperature in the range of from about 250°C to about 500°C
16 (482°F-932°F), a reaction pressure in the range of from about 3.5 MPa
17 to 24.2 MPa (500-3,500 psi), a feed rate (vol oil/vol cat h) from about 0.1
18 to about 20 hr⁻¹, and a hydrogen circulation rate in the range from about
19 350 std liters H₂/kg oil to 1780 std liters H₂/kg oil (2,310-11,750 standard
20 cubic feet per barrel).
21
- 22 10. An integrated hydroconversion process having at least two stages, each
23 stage possessing at least one reaction zone, comprising:
24
- 25 (a) combining a first refinery stream with a first hydrogen-rich gaseous
26 stream to form a first feedstock;
27
- 28 (b) passing the first feedstock to a reaction zone of the first stage,
29 which is maintained at conditions sufficient to effect a boiling range
30 conversion, to form a first reaction zone effluent comprising

- 1 normally liquid phase components and normally gaseous phase
2 components;
3
- 4 (c) passing the first reaction zone effluent of step (b) to a heat
5 exchanger or series of exchangers, where it exchanges heat with
6 other refinery streams;
7
- 8 (d) passing the effluent of step (c) to a hot high pressure separator,
9 where it is separated into a liquid stream which is passed to
10 fractionation, and a gaseous stream, which is combined with a
11 second refinery stream which comprises light cycle oil, light gas oil,
12 atmospheric gas oil or mixtures of all three;
13
- 14 (e) passing the combined gaseous stream of step (d) to a reaction
15 zone of the second stage, which is maintained at conditions
16 sufficient for converting at least a portion of the aromatics present
17 in the second refinery stream, to form a second reaction zone
18 effluent;
19
- 20 (f) separating the second reaction zone effluent of step (e) into a liquid
21 stream comprising products and a second hydrogen-rich gaseous
22 stream;
23
- 24 (g) recycling at least a portion of the second hydrogen-rich gaseous
25 stream of step (f) to a reaction zone of the first stage; and
26
- 27 (h) passing the liquid stream comprising products of step (f) to a
28 fractionation column, wherein product streams comprise a gas or
29 naphtha stream removed overhead, one or more middle distillate
30 streams, and a bottoms stream suitable for further processing.