

TARGETED HYDROGENATION HYDROCRACKING**FIELD OF THE INVENTION**

This invention is directed to a partial conversion hydrocracking (HCR) unit, in which unconverted oil is fed to a Fluid Catalytic Cracking (FCC) unit.

BACKGROUND OF THE INVENTION

In the refining of crude oil, vacuum gas oil hydrotreaters and hydrocrackers are employed to remove impurities such as sulfur, nitrogen and metals from the feed. Typically, the middle distillate boiling material (boiling in the range from 250° F.-735°F.) from VGO hydrotreating or moderate severity hydrocrackers does not meet the smoke point, the cetane number or the aromatic specification required.

Removal of these impurities in subsequent hydroprocessing stages (often known as upgrading), creates more valuable middle distillate products. Hydroprocessing technology (which encompasses hydrotreating, hydrocracking and hydrodewaxing processes) aims to increase the value of the crude oil by fundamentally rearranging molecules. The end products are also made more environmentally friendly.

In most cases, this middle distillate is separately upgraded by a middle distillate hydrotreater or, alternatively, the middle distillate is blended into the general fuel oil pool or used as home heating oil. Recently hydroprocessing schemes have been developed which permit the middle distillate to be hydrotreated in the same high pressure loop as the vacuum gas oil hydrotreating reactor or the moderate severity hydrocracking reactor. The investment cost saving and/or utilities saving are significant since a separate middle distillate hydrotreater is not required.

There are U.S. patents which are directed to multistage hydroprocessing

1 within a single high pressure hydrogen loop. In U.S. Patent No. 6,797,154,
2 high conversion of heavy gas oils and the production of high quality middle
3 distillate products are possible in a single high-pressure loop with reaction
4 stages operating at different pressure and conversion levels. The flexibility
5 offered is great and allows the refiner to avoid decrease in product quality
6 while at the same time minimizing capital cost. Feeds with varying boiling
7 ranges are introduced at different sections of the process, thereby minimizing
8 the consumption of hydrogen and reducing capital investment.

9
10 U.S. Pat. No. 6,787,025 also discloses multi-stage hydroprocessing for the
11 production of middle distillates. A major benefit of this invention is the
12 potential for simultaneously upgrading difficult cracked stocks such as Light
13 Cycle Oil, Light Coker Gas Oil and Visbroken Gas Oil or Straight-Run
14 Atmospheric Gas Oils utilizing the high-pressure environment required for
15 mild hydrocracking.

16
17 U.S. Pat. No. 7,238,277 provides very high to total conversion of heavy oils to
18 products in a single high-pressure loop, using multiple reaction stages. The
19 second stage or subsequent stages may be a combination of co-current and
20 counter-current operation. The benefits of this invention include conversion of
21 feed to useful products at reduced operating pressures using lower catalyst
22 volumes. Lower hydrogen consumption also results. A minimal amount of
23 equipment is employed. Utility consumption is also minimized.

24
25 U.S. Publication 20050103682 relates to a multi-stage process for
26 hydroprocessing gas oils. Preferably, each stage possesses at least one
27 hydrocracking zone. The second stage and any subsequent stages possess
28 an environment having a low heteroatom content. Light products, such as
29 naphtha, kerosene and diesel, may be recycled from fractionation (along with
30 light products from other sources) to the second stage (or a subsequent
31 stage) in order to produce a larger yield of lighter products, such as gas and
32 naphtha. Pressure in the zone or zones subsequent to the initial zone is from

1 500 to 1000 psig lower than the pressure in the initial zone, in order to provide
2 cost savings and minimize overcracking.

3

4 Most refineries run the FCC unit at full capacity for optimal asset utilization.
5 During planned and/or unplanned shutdown of Residue Desulfurization unit(s)
6 feeding FCC unit, it is desirable to reduce the conversion in the FCC feed
7 hydrocracker in order to maximize the feed to FCC unit. The patents
8 disclosed above do not address the following issues:

- 9 1. Production of on-specification Jet and Diesel products during low
10 conversion HCR operation.
- 11 2. Avoidance of undesirable over-saturation of the unconverted oil
12 (UCO) from the HCR unit feeding FCC unit and reduce hydrogen
13 consumption. Normally, further aromatic saturation of the middle
14 distillate products from a low conversion HCR is achieved in a
15 separate Post Treatment unit.

16

17 SUMMARY OF THE INVENTION

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19 A new process scheme has been developed to design a partial conversion
20 hydrocracking (HCR) unit, feeding the unconverted oil to a FCC unit. The
21 steps of this invention are summarized as follows:

22

23 A method for hydroprocessing a hydrocarbon feedstock, said method
24 employing multiple hydroprocessing zones within a single reaction loop,
25 each zone having one or more catalyst beds, comprising the following
26 steps:

27

- 28 (a) passing a hydrocarbonaceous feedstock to a first
29 hydroprocessing zone having one or more beds containing
30 hydroprocessing catalyst, the hydroprocessing zone being
31 maintained at hydroprocessing conditions, wherein the
32 feedstock is contacted with catalyst and hydrogen;

33

- 1 (b) passing the effluent of step (a) directly to a hot high pressure
2 separator, wherein the effluent is separated to produce a vapor
3 stream comprising hydrogen, hydrocarbonaceous compounds
4 boiling at a temperature below the boiling range of the
5 hydrocarbonaceous feedstock, hydrogen sulfide and ammonia
6 and a liquid stream comprising hydrocarbonaceous
7 compounds boiling approximately in the range of said
8 hydrocarbonaceous feedstock;
9
- 10 (c) passing the vapor stream of step (b) after cooling and partial
11 condensation, to a hot high pressure separator where it is
12 flashed, thereby producing an overhead vapor stream and a
13 liquid stream, wherein the liquid stream, which comprises
14 hydrotreated hydrocarbons in the middle distillate range, is
15 passed to a second hydroprocessing zone;
16
- 17 (d) passing the overhead vapor stream from the hot high
18 pressure separator of step (c), after cooling and contact with
19 water, said vapor stream comprising hydrogen, ammonia,
20 hydrogen sulfide, light gases and naphtha, to a cold high
21 pressure separator, where hydrogen, hydrogen sulfide, and
22 light hydrocarbonaceous gases are removed overhead,
23 ammonia is removed from the cold high pressure separator as
24 ammonium bisulfide in the sour water stripper, and naphtha
25 and middle distillates are passed to fractionation
26
- 27 (e) passing the liquid stream from the hot high pressure separator
28 of step (b) to a hot low pressure separator, where it is flashed
29 to produce an overhead stream comprising gases and a liquid
30 stream comprising unconverted oil;
31
- 32 (f) passing the liquid stream of step (e) which comprises
33 unconverted oil, to a steam stripper, where lighter material is

1 removed overhead as a vapor stream, and a liquid stream,
2 which comprises stripped unconverted oil, is recovered.

3

4 BRIEF DESCRIPTION OF THE FIGURE

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6 The Figure illustrates the flow scheme of the current invention.

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9 DETAILED DESCRIPTION OF THE INVENTION

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11 Feeds

12 A wide variety of hydrocarbon feeds may be used in the instant invention.

13 Typical feedstocks include any heavy or synthetic oil fraction or process

14 stream having a boiling point above 392° F. (200° C.). Such feedstocks

15 include vacuum gas oils (VGO), heavy coker gas oil (HCGO), heavy

16 atmospheric gas oil (AGO), light coker gas oil (LCGO), visbreaker gas oil

17 (VBGO), demetallized oils (DMO), vacuum residua, atmospheric residua,

18 deasphalted oil (DAO), Fischer-Tropsch streams, Light Cycle Oil, Light Cycle

19 Gas Oil and other FCC product streams.

20

21 Products

22 The process of this invention is especially useful in the production of middle

23 distillate fractions boiling in the range of about 250-700°F (121-371°C). A

24 middle distillate fraction is defined as having an approximate boiling range

25 from about 250 to 700°F. At least 75 vol.%, preferably 85 vol.% of the

26 components of the middle distillate have a normal boiling point of greater than

27 250°F. At least about 75 vol.%, preferably 85 vol.% of the components of the

28 middle distillate have a normal boiling point of less than 700°F. The term

29 "middle distillate" includes the diesel, jet fuel and kerosene boiling range

30 fractions. The kerosene or jet fuel boiling point range refers to the range

31 between 280 and 525°F (138-274°C). The term "diesel boiling range" refers to

32 hydrocarbons boiling in the range from 250 to 700°F (121-371°C).

33

1 Gasoline or naphtha may also be produced in the process of this invention.
2 Gasoline or naphtha normally boils in the range below 400°F. (204°C), or C₅ to
3 400°F. Boiling ranges of various product fractions recovered in any particular
4 refinery will vary with such factors as the characteristics of the crude oil
5 source, local refinery markets and product prices.

6

7 Conditions

8 "Hydroprocessing conditions" is a general term which refers primarily in this
9 application to hydrocracking or hydrotreating.

10

11 Hydrotreating conditions include a reaction temperature between
12 400°F.-950°F. (204°C.-482°C.), preferably 600°F.-850°F. (315°C.-464°C.); a
13 pressure between 500 to 5000 psig (pounds per square inch gauge)
14 (3.5-34.6 MPa), preferably 1000 to 3000 psig (7.0-20.8 MPa); a feed rate
15 (LHSV) of 0.3 hr⁻¹ to 20 hr⁻¹ (v/v) preferably from 0.5 to 4.0; and overall
16 hydrogen consumption 300 to 2000 SCF per barrel of liquid hydrocarbon feed
17 (63.4-356 m³/m³ feed).

18

19 Typical hydrocracking conditions include a reaction temperature of from
20 400°F.-950°F. (204°C.-510°C.), preferably 650°F.-850°F. (315°C.-454°C.).
21 Reaction pressure ranges from 500 to 5000 psig (3.5-4.5 MPa), preferably
22 1000-3000 psig (7.0-20.8 MPa). LHSV ranges from 0.1 to 15 hr⁻¹ (v/v),
23 preferably 0.5 to 5.0 hr⁻¹. Hydrogen consumption ranges from 500 to 2500
24 SCF per barrel of liquid hydrocarbon feed (89.1-445 m³H₂/m³ feed).

25

26 Catalyst

27 A hydroprocessing zone may contain only one catalyst, or several catalysts in
28 combination.

29

30 The hydrocracking catalyst generally comprises a cracking component, a
31 hydrogenation component and a binder. Such catalysts are well known in the
32 art. The cracking component may include an amorphous silica/alumina phase
33 and/or a zeolite, such as a Y-type or USY zeolite. Catalysts having high

1 cracking activity often employ REX, REY and USY zeolites. The binder is
2 generally silica or alumina. The hydrogenation component will be a Group VI,
3 Group VII, or Group VIII metal or oxides or sulfides thereof, preferably one or
4 more of molybdenum, tungsten, cobalt, or nickel, or the sulfides or oxides
5 thereof. If present in the catalyst, these hydrogenation components generally
6 make up from about 5% to about 40% by weight of the catalyst. Alternatively,
7 platinum group metals, especially platinum and/or palladium, may be present
8 as the hydrogenation component, either alone or in combination with the base
9 metal hydrogenation components molybdenum, tungsten, cobalt, or nickel. If
10 present, the platinum group metals will generally make up from about 0.1% to
11 about 2% by weight of the catalyst.

12

13 Hydrotreating catalyst is typically a composite of a Group VI metal or
14 compound thereof, and a Group VIII metal or compound thereof supported on
15 a porous refractory base such as alumina. Examples of hydrotreating
16 catalysts are alumina supported cobalt-molybdenum, nickel sulfide, nickel-
17 tungsten, cobalt-tungsten and nickel-molybdenum. Typically, such
18 hydrotreating catalysts are presulfided.

19

20 In some cases, high activity hydrotreating catalyst suitable for high levels of
21 hydrogenation, is employed. Such catalysts have high surface areas (greater
22 than 140 m²/gm) and high densities (0.7-0.95 gm/cc). The high surface
23 area increases reaction rates due to generally increased dispersion of the
24 active components. Higher density catalysts allow one to load a larger
25 amount of active metals and promoter per reactor volume, a factor which is
26 commercially important. Since deposits of coke are thought to cause the
27 majority of the catalyst deactivation, the catalyst pore volume should be
28 maintained at a modest level (0.4-0.6). A high activity catalyst is at times
29 desired in order to reduce the required operating temperatures. High
30 temperatures lead to increased coking.

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1 Description of the preferred embodiment

2 Please refer to the Figure:

3 In this process scheme, fresh feed (Stream 9) is passed to the top of fixed
4 bed hydrotreater reactor 10. Hydrogen passes through stream 1. Stream 29
5 is a sidestream from stream 1. From stream 29, streams 3 and 4 add
6 hydrogen in between the first and second beds, and second and third beds of
7 reactor 10 respectively. Hydrotreater 10 is loaded with a high activity
8 hydrotreating catalyst, where most of the feed impurities (heteroatoms) such
9 as nitrogen, sulfur, etc. are removed and some degree of aromatic saturation
10 is achieved.

11

12 The hydrotreated reactor effluent (stream 12) exchanges heat in exchanger 5
13 with the reactor feed (stream 2 prior to entering the exchanger 5 and stream 9
14 upon leaving the exchanger 5). Stream 12 is flashed in hot high pressure
15 separator 40 at high temperature and pressure conditions to recover most of
16 the unconverted oil (UCO) components in the liquid phase (stream 13). Vapor
17 leaves separator 40 overhead in line 22, and heat is exchanged with
18 hydrogen stream 31 in exchanger 25. Stream 22, which is made up of more
19 than 85wt% diesel and lighter material, preheats the fractionator feed (not
20 shown in the Figure) and generates high pressure steam. Stream 22 is finally
21 cooled to about 200°C in the hot high pressure separator vapor/recycle gas
22 exchanger 25. Stream 22 is then flashed in hot high pressure separator 50.
23 At these relatively high pressure and low temperature conditions, most of the
24 hydrotreated jet and diesel range material is recovered as liquid stream 27 at
25 high pressure, which is pumped(pump 35) to the feedstream (stream 11) ,
26 which passes to hydrocracking reactor 20, for further processing. The
27 overhead vapor from the hot high pressure separator 50, stream 23, is then
28 cooled in an air cooler (not shown) before entering a cold high pressure
29 separator (not shown). The overhead vapor stream, stream 23, comprises
30 hydrogen, ammonia, and hydrogen sulfide, along with light gases and
31 naphtha. In the cold high pressure separator (not shown) hydrogen, hydrogen
32 sulfide, and light hydrocarbonaceous gases are removed overhead, ammonia
33 is removed from the cold high pressure separator as ammonium bisulfide in

1 the sour water stripper. Naphtha and middle distillates are passed to
2 fractionation.

3

4 Stream 13 passes to hot low pressure separator 60, where it is flashed.
5 Vapor is removed as stream 28. The hot low pressure separator bottoms are
6 removed as stream 73 and passed to UCO (unconverted oil) stripper 30. The
7 material of stream 73 is stream stripped in stripper 30 to recover any lighter
8 material in the UCO stream. Lighter material is removed as stream 26. Jet
9 and diesel range material is withdrawn as a side draw 17 from the column.
10 Side draw 17 combines with stream 19, stripper bottoms 16 (UCO) to become
11 stream 19. A side stream 18 may be taken from bottoms stream 16. Stream
12 19, recycle oil, is pumped, via pump 45, to storage drum 70. The recycle oil
13 exits storage drum 70 through stream 21 and is pumped, by means of pump
14 55, to stream 11. Stream 11 is heated in exchanger 15 prior to entering
15 hydrocracking reactor 20 for further aromatic saturation. The overhead liquid
16 stream 26 from the UCO stripper 30 is sent to the main product stripper, and
17 the offgas is sent to fuel gas (not shown).

18

19 The hydrotreated, stripped UCO (stream 16) from the bottom of the UCO
20 stripper, is an excellent quality FCC feed. At this point, a part of stripped
21 unconverted oil (stream 18) is sent out as FCC feed. Further saturation of the
22 FCC feed is thus avoided. Only a limited portion of the UCO (mixed with
23 stream 19, is passed to hydrocracker 20 for further saturation of aromatic
24 components and conversion to distillate products. The amount recycled back
25 is based on the desired overall conversion level.

26

27 The second stage hydrocracking reactor 20 is loaded with hydrocracking
28 catalyst and operates under a clean environment (no heteroatoms), ideally
29 selectively converting the UCO to desired products and further saturating the
30 aromatic components to achieve required jet and diesel properties at different
31 conversion levels. Stream 32 is a sidestream from stream 1. From stream 32,
32 streams 7 and 8 add hydrogen in between the first and second beds and
33 second and third beds of reactor 20 respectively.

Both the hydrotreating reactor 10 and hydrocracking reactor 20 are designed for the maximum conversion desired. During lower conversion operation, the hydrotreating reaction is maintained at the same temperature as the highest conversion case in order to achieve target denitrification and desulfurization. The temperature of the hydrocracking reactor is reduced at lower conversions.

The effluent (stream 72) of second stage hydrocracking reactor 20 is cooled (in exchanger 15) to preheat second stage reactor feed (stream 11) fractionator feed and cold low pressure separator liquid stream. Stream 72, now renumbered stream 74, combines with hot high pressure separator vapor stream 23 for further cooling and the removal of high pressure recycle gas. The hydrocarbon liquid from the cold high pressure separator (not shown) is sent to the fractionation section (not shown) for product recovery.

EXAMPLE

The following table highlights the advantages of the process scheme of this invention over a conventional process scheme for a 65,000 BPOD (barrel per operating day) hydrocracking unit: The table indicates that there is no need in the current invention for post treatment in order to reach desired product specifications. Furthermore, less hydrogen is consumed in the scheme of the current invention than in the conventional case.

	Conventional Process Scheme	New Process Scheme
Process Scheme	Single Stage Once-Through	Targeted-Hydrogenation Hydrocracking
Fresh Feed Rate, BPSD	65,000	65,000
Overall LHSV, 1 hr	0.7 – 0.9	0.7 – 0.9
Product Yield	Base	Similar to Base case
Jet & Diesel Quality	Needs post treatment for aromatic saturation at low conversion	On specification product at all desired conversions: Jet : S<10 ppm; Smoke Point>24mm Diesel : S<10 ppm; Cetane Index>50
Chemical H ₂ Consumption	Base	Base – 160 SCFB at 60% conv. Base – 100 SCFB at 40% conv. Base – 50 SCFB at 30% conv.

CLAIMS

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3 What is claimed is:

4 1.) A method for hydroprocessing a hydrocarbon feedstock, said method
5 employing multiple hydroprocessing zones within a single reaction
6 loop, each zone having one or more catalyst beds, comprising the
7 following steps:

8

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

14

15 (b) passing the effluent of step (a) directly to a hot high pressure
16 separator, wherein the effluent is separated to produce a vapor
17 stream comprising hydrogen, hydrocarbonaceous compounds
18 boiling at a temperature below the boiling range of the
19 hydrocarbonaceous feedstock, hydrogen sulfide and ammonia and
20 a liquid stream comprising hydrocarbonaceous compounds boiling
21 approximately in the range of said hydrocarbonaceous feedstock;

22

23 (c) passing the vapor stream of step (b) after cooling and partial
24 condensation, to a second hot high pressure separator where it is
25 flashed, thereby producing an overhead vapor stream and a liquid
26 stream, wherein the liquid stream, which comprises hydrotreated
27 hydrocarbons in the middle distillate range, is passed to a second
28 hydroprocessing zone;

29

30 (d) passing the overhead vapor stream from the hot high pressure
31 separator of step (c), after cooling and contact with water, said
32 vapor stream comprising hydrogen, ammonia, hydrogen sulfide,
33 light gases and naphtha, to a cold high pressure separator, where

1 hydrogen, hydrogen sulfide, and light hydrocarbonaceous gases
2 are removed overhead, ammonia is removed from the cold high
3 pressure separator as ammonium bisulfide in the sour water
4 stripper, and naphtha and middle distillates are passed to
5 fractionation;

6

7 (e) passing the liquid stream from the hot high pressure separator
8 of step (b) to a hot low pressure separator, where it is flashed to
9 produce an overhead stream comprising gases and a liquid stream
10 comprising unconverted oil;

11

12 (f) passing the liquid stream of step (e) which comprises
13 unconverted oil, to a steam stripper, where a vapor stream is
14 removed overhead and a liquid stream, which comprises stripped
15 unconverted oil, is recovered.

16

17 2.) The process of claim 1, wherein at least a portion of the stripped
18 unconverted oil of step (f) is passed to a fluid catalytic cracking unit as
19 feed.

20

21 3.) The process of claim 1, wherein at least a portion of the stripped
22 unconverted oil of step (f) is combined with the liquid effluent of step(c)
23 to form a liquid stream which is passed to the second
24 hydroprocessing zone.

25

26 4.) The process of claim 1, wherein the second hydroprocessing zone
27 contains at least one bed of hydroprocessing catalyst suitable for
28 aromatic saturation and ring opening.

29

30 5.) The process of claim 4, wherein the liquid stream is contacted under
31 hydroprocessing conditions with the hydroprocessing catalyst, in the
32 presence of hydrogen to produce middle distillate products.

33

- 1 6.) The process of claim 1, wherein the hydroprocessing conditions of
2 step (a) comprise a reaction temperature of from 400°F.-950°F.
3 (204°C.-510°C.), a reaction pressure in the range from 500 to 5000 psig
4 (3.5-34.5 MPa), an LHSV in the range from 0.1 to 15 hr⁻¹ (v/v), and
5 hydrogen consumption in the range from 500 to 2500 scf per barrel of
6 liquid hydrocarbon feed (89.1-445 m³ H₂/m³ feed).
7
- 8 7.) The process of claim 6, wherein the hydroprocessing conditions of
9 step 1(a) preferably comprise a temperature in the range from
10 650°F.-850°F. (343°C.-454°C.), reaction pressure in the range from
11 1500-3500 psig (10.4-24.2 MPa), LHSV in the range from 0.25 to 2.5
12 hr⁻¹, and hydrogen consumption in the range from 500 to 2500 scf per
13 barrel of liquid hydrocarbon feed (89.1-445 m³ H₂/m³ feed).
14
- 15 8.) The process of claim 1, wherein the hydroprocessing conditions of
16 step 1(e) comprise a reaction temperature of from 400°F.-950°F.
17 (204°C.-510°C.), a reaction pressure in the range from 500 to 5000 psig
18 (3.5-34.5 MPa), an LHSV in the range from 0.1 to 15 hr⁻¹ (v/v), and
19 hydrogen consumption in the range from 500 to 2500 scf per barrel of
20 liquid hydrocarbon feed (89.1-445 m³ H₂/m³ feed).
21
- 22 9.) The process of claim 9, wherein the hydroprocessing conditions of
23 step 1(e) preferably comprise a temperature in the range from
24 650 m³ H₂/m³ feed F.-850 m³ H₂/m³ feed F. (343°C.-454°C.), reaction
25 pressure in the range from 1500-3500 psig (10.4-24.2 MPa), LHSV in
26 the range from 0.25 to 2.5 hr⁻¹, and hydrogen consumption in the
27 range from 500 to 2500 scf per barrel of liquid hydrocarbon feed
28 (89.1-445 m³ H₂/m³ feed).
29
- 30 10.) The process of claim 1, wherein the feed to step 1(a) comprises
31 hydrocarbons boiling in the range from 500°F. to 1500°F.
32

- 1 11.) The process of claim 1, wherein the feed is selected from the group
2 consisting of vacuum gas oil, heavy atmospheric gas oil, delayed coker
3 gas oil, visbreaker gas oil, FCC light cycle oil, and deasphalted oil.
4
- 5 12.) The process of claim 1, wherein the hydroprocessing catalyst
6 comprises both a cracking component and a hydrogenation
7 component.
8
- 9 13.) The process of claim 12, wherein the hydrogenation component is
10 selected from the group consisting of Ni, Mo, W, Pt and Pd or
11 combinations thereof.
12
- 13 14.) The process of claim 3, wherein the cracking component may be
14 amorphous or zeolitic.
15
- 16 15.) The process of claim 11, wherein the zeolitic component is selected
17 from the group consisting of Y, USY, REX, and REY zeolites.
18
- 19 16.) The process of claim 1, wherein the middle distillate products produced
20 do not require additional treatment to meet product specifications.
21
- 22 17.) The process of claim 16, wherein the sulfur content of jet fuel is less
23 than 10 ppm, the smoke point is greater than 24mm, the sulfur content
24 of diesel is less than 10 ppm, and the cetane index is greater than 50.
25
- 26 18.) The process of claim 1, in which smaller amounts of hydrogen are used
27 than in single stage once-through hydrocracking.
28
- 29 19.) The process of claim 18, wherein the amount of hydrogen used is 160
30 SCF per barrel lower than the amount used in single stage once-
31 through hydrocracking at 60% conversion, 100 SCF per barrel lower
32 than the amount used in single stage once-through hydrocracking at

1 40% conversion, and 50 SCF per barrel lower than the amount used in
2 single stage once-through hydrocracking at 30% conversion.

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4 20.) The process of claim 1, wherein the hydrotreating occurs in the first
5 reaction zone and hydrocracking occurs in the second reaction zone.

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