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(72) **Inventeurs/Inventors**:
ZERPA REQUES, NESTOR GREGORIO, CA;
DE CLERK, ARNO, CA;
XIA, YUHAN, CA;
OMER, AYYUB ABDULJAWAD, CA

(73) **Propriétaire/Owner**:
NEXEN ENERGY ULC, CA

(74) **Agent**: NORTON ROSE FULBRIGHT CANADA LLP/S.E.N.C.R.L., S.R.L.

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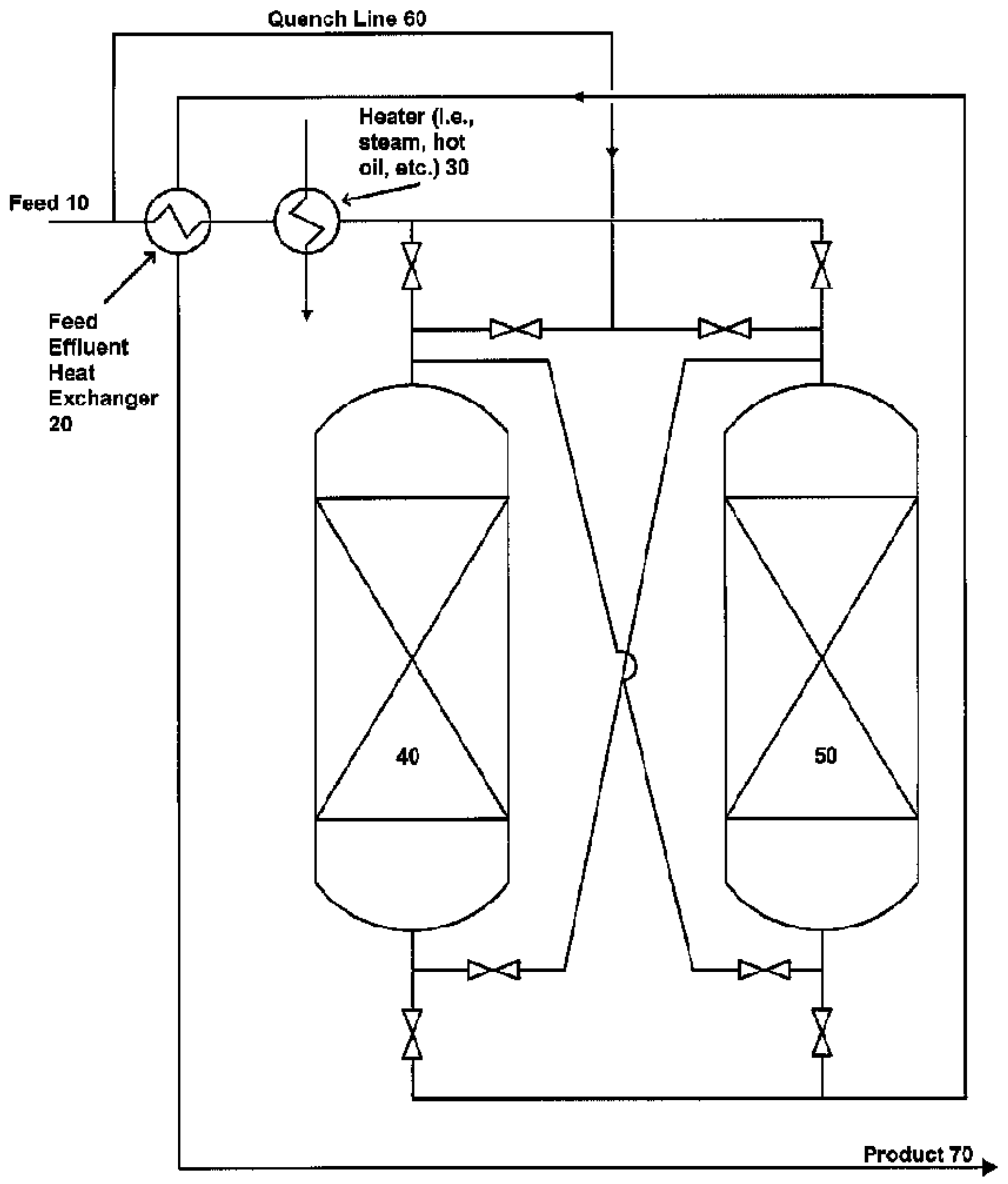
(54) Title: OLEFINS REDUCTION OF A HYDROCARBON FEED USING OLEFINS-AROMATICS ALKYLATION



(57) Abrégé/Abstract:
A process for reducing the olefins and/or di-olefins content of a hydrocarbon-containing feed, such as that resulting from the thermal cracking of heavy oil, thermally cracked bitumen, or thermally cracked petroleum are disclosed. The process involves alkylation of di-olefins and olefins with aromatics already present in the hydrocarbon feed and without removal of nitrogen-, sulfur- or oxygen-containing compounds present in the feed. The process can take place without the need to add an external source of hydrogen, olefins or aromatics.

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(71) **Applicant**: **NEXEN ENERGY ULC** [CA/CA]; 801 7th Ave SW, Calgary, Alberta T2P 3P7 (CA).

(72) **Inventors**: **ZERPA REQUES, Nestor, Gregorio**; 59 Arbour Ridge Way NW, Calgary, Alberta T3G 3T1 (CA). **DE CLERK, Arno**; 686 Romaniuk Road, Edmonton, Alberta T6R 1A5 (CA). **XIA, Yuhan**; 59 Arbour Ridge Way NW, Calgary, Alberta T3G 3T1 (CA). **OMER, Ayyub, Abduljawad**; 136 Martinwood Place NE, Calgary, Alberta T3J 3H5 (CA).

(74) **Agents**: **THERIAULT, Andre** et al.; Norton Rose Fulbright Canada LLP / S.E.N.CR.L., s.r.l., 1 Place Ville Marie, Suite 2500, Montreal, Quebec H3B 1R1 (CA).

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(54) **Title**: OLEFINS REDUCTION OF A HYDROCARBON FEED USING OLEFINS- AROMATICS ALKYLATION

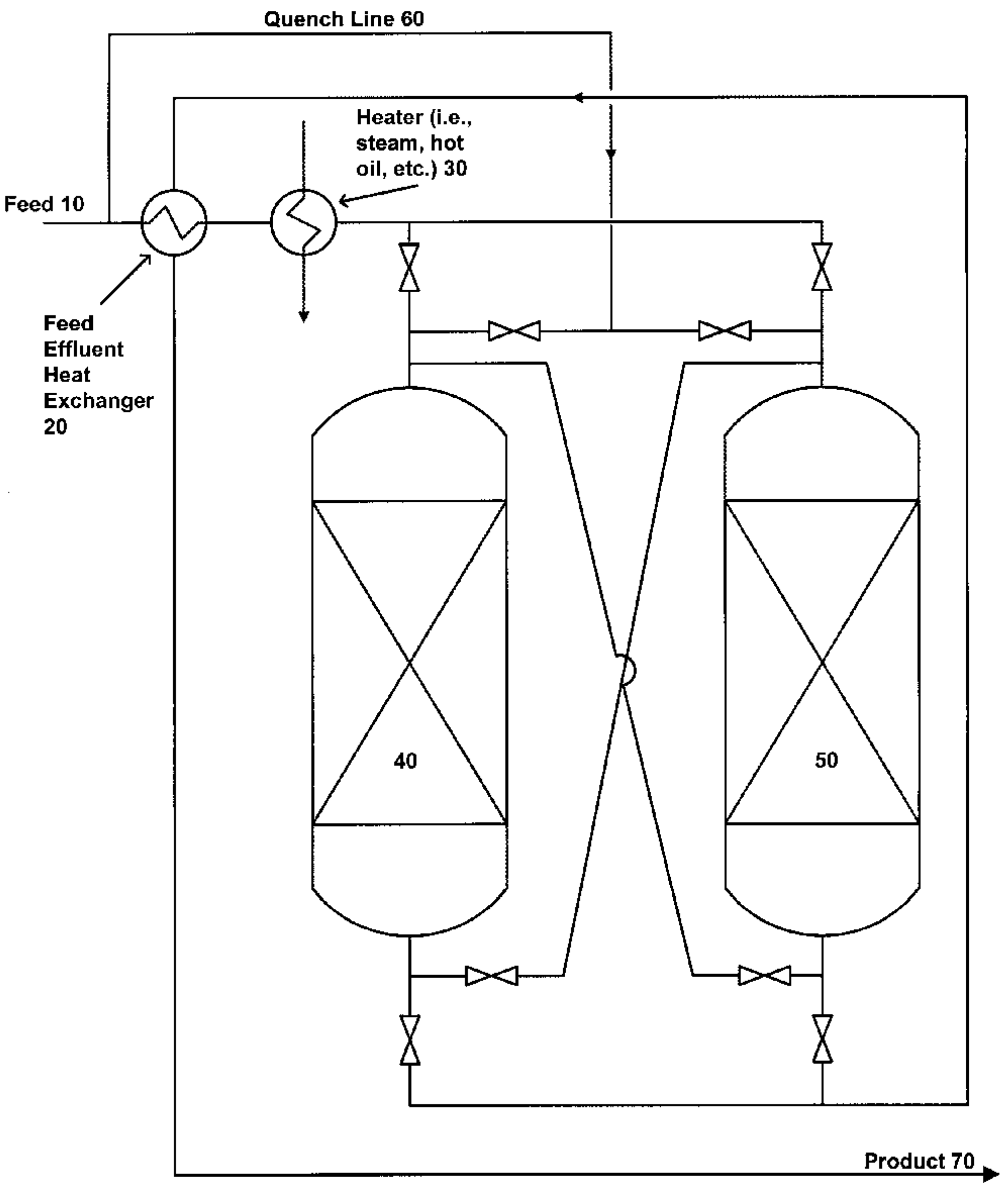

FIG. 1

(57) **Abstract**: A process for reducing the olefins and/or di-olefins content of a hydrocarbon-containing feed, such as that resulting from the thermal cracking of heavy oil, thermally cracked bitumen, or thermally cracked petroleum are disclosed. The process involves alkylation of di-olefins and olefins with aromatics already present in the hydrocarbon feed and without removal of nitrogen-, sulfur- or oxygen-containing compounds present in the feed. The process can take place without the need to add an external source of hydrogen, olefins or aromatics.

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OLEFINS REDUCTION OF A HYDROCARBON FEED USING OLEFINS-AROMATICS ALKYLATION

FIELD

The present invention relates to olefins and/or di-olefins reduction via alkylation of olefins and di-olefins present in a hydrocarbon-containing product, such as heavy oil and/or thermally cracked bitumen or other thermally cracked hydrocarbon.

BACKGROUND

Commonly, heavy oils and/or bitumen are difficult to transport from their production areas due to their high viscosities at typical handling temperatures. For example in Canada, there are typically three specifications that must be met by any oil to be acceptable for pipeline transport. The viscosity should be below the maximum viscosity limit (e.g. < 350 cSt at 7.5 °C), the density should be below the maximum density limit (e.g. < 940 $kg{\cdot}m^{-3}$ at 15.6 °C, or >19 °API), and the olefins (including di-olefins) content should be below the maximum limit (e.g. < 1 wt% as 1-decene equivalent).

As is well known, light oils generally have much lower viscosity values and therefore flow easier through pipelines than heavy oils. Regardless of the recovery method used for their extraction, including costly thermal enhanced oil recovery (EOR) methods, heavy oils and bitumen generally need to be diluted by blending the oil with low density and low viscosity diluents, typically gas condensate, naphtha and/or lighter oil to make the heavy oils and/or bitumen transportable over long distances. For example in Canada, when adding diluent to bitumen to produce transportable oil (also known as “DilBit”), the volume of diluent is typically 30 to 35% of the total product. There are several disadvantages of adding diluent to bitumen to produce transportable oil, including:

- Well remoteness makes the construction of pipelines for sending or returning the diluents to the heavy hydrocarbon (bitumen) production zone considerably expensive; and
- Availability of diluents, typically light hydrocarbons, is steadily decreasing worldwide, making them more and more expensive to procure.

The upgrading of heavy oil and/or bitumen to a product that meets the specifications for pipeline transport is known in the industry. Upgrading has become an attractive alternative for converting heavy oil and/or bitumen into transportable oil or oil that requires less diluent to be transportable, and in some cases upgrading is the only viable alternative in order to transport heavy oil to refineries and market places.

One upgrading approach involves the chemical processing of the heavy oil and/or bitumen by a suitable combination of conversion and separation steps. Most chemical processing for converting heavy oil and/or bitumen into transportable oil are cracking based systems and usually include at least one form of cracking, e.g. thermal cracking, and at least one form of hydro-processing. The cracking step is employed to reduce the viscosity and density of the heavy oil and/or bitumen. The hydro-processing step is employed to reduce the olefin and di-olefin content of the heavy oil and/or bitumen.

Moderate thermal cracking such as visbreaking or more severe thermal processes such as coking systems have been proposed in the prior art to reduce the viscosity and density of heavy oils and/or bitumen.

A disadvantage of these processes is the production of cracked material comprising olefins and di-olefins. If left untreated, olefins, and more particularly the more reactive conjugated C_4 and C_5 di-olefins (i.e. butadiene, 1,3-pentadiene, 2-methyl-1,3-butadiene) may react with each other, with oxygen (such as oxygen in the air) or other reactive compounds (e.g. organic acids, carbonyls, amines, etc.), to form long chain polymers (polymerization reaction) commonly referred to as gums. Gums of this nature are known to foul process equipment. It has been reported in U.S. Patent 6,210,560 assigned to Exxon Research and Engineering Company, that these polymerization reactions occur at a significant rate for petroleum oils having a di-olefins value per UOP-326 of 4 grams iodine/100 grams oil or greater in the temperature range of 232°C to 324°C; particularly 260°C to 304°C. Below this temperature range, the reaction rate is too slow for significant polymerization formation to occur and above this temperature range the chemical bonds are broken thermally faster than they are formed. It has also been reported in in U.S. Patent Application Publication No. 2012/0273394 A1 assigned to UOP LLC, that the fouling tendencies of oils having a di-olefins value per UOP-326 of less than about 2 grams iodine/100 grams oil are deemed acceptable.

As excessive olefins content in hydrocarbon streams can lead to fouling of refining equipment and pipelines, methods to reduce the amount of olefins in a hydrocarbon stream are sought in the industry. A particular challenge is to reduce the amount of olefins in facilities where it is impractical to supply or generate hydrogen to make use of hydro-processing.

SUMMARY

According to one aspect, there is provided a process to reduce olefins content in a hydrocarbon feed, such as a cracked hydrocarbon feed, said process comprising: alkylating said olefins in said hydrocarbon with aromatic compounds for a time sufficient to reduce the olefins content under the reaction temperature at issue, wherein the aromatics and olefins content of the feed is such that there is no need to add extraneous aromatic compounds, hydrogen, or extraneous olefins for the reaction to occur. The hydrocarbon feed is derived from a thermal cracking process or any other process that involves cracking of a hydrocarbon. When an appropriate hydrocarbon feed is used, there is also no need to add an external source of hydrogen in the process, or to perform hydro-processing prior to carrying out the alkylation. Also, the olefins content in the resulting hydrocarbon product may be reduced sufficiently to meet a pipeline and/or operational specification.

According to one aspect, there is a process to treat a hydrocarbon feed, the hydrocarbon feed being produced as a result of treatment of a hydrocarbon, the treatment being an upgrading procedure that includes a cracking step. The hydrocarbon feed contains olefinic material, the amount of olefins being detectable by various methods. The hydrocarbon feed also contains nitrogen bases. If the feed has olefinic material above a certain undesirable level, such as above the level that would be suitable to meet pipeline specifications, the hydrocarbon feed may be treated by reacting the feed with an alkyl acceptor. The alkyl acceptor may be an aromatic compound, and the aromatic compound may be present in the hydrocarbon feed.

According to one aspect, there is a process to treat a hydrocarbon feed that contains nitrogen-, oxygen- and sulfur-containing compounds, which includes bases (alkaline materials) The process teaches a way in which an olefin- and aromatic- containing hydrocarbon feed with bases can be treated without extraneous hydrogen to reduce the olefins content.

In accordance with one aspect there is provided a process to reduce the olefins content in a hydrocarbon feed, said process comprising: alkylating said hydrocarbon feed with an aromatic component within the feed without the requirement to add external sources of aromatics, olefins, or hydrogen and without the requirement to adjust or predetermine the ratio of olefins content to aromatics content in the hydrocarbon feed prior to the alkylation; wherein the alkylation is carried out for a duration, and at a temperature sufficient to reduce the olefins content of the hydrocarbon feed, wherein said alkylating said hydrocarbon feed uses at least one catalyst, wherein the hydrocarbon feed includes basic compounds, and wherein said alkylating is carried out at a temperature of greater than 150 °C.

In accordance with another aspect there is provided a process for producing a hydrocarbon product meeting pipeline specification, comprising: a. selecting a hydrocarbon feed produced by thermal cracking of a heavy oil feed, the hydrocarbon feed containing olefins and di-olefins, and basic compounds; b. alkylating the olefins and di-olefins in the presence of a catalyst under conditions such that the olefins and di-olefins within the feed react with aromatic compounds in the feed, without the requirement to import additional olefins, aromatics or hydrogen into the reaction, wherein the conditions comprise a temperature of greater than 150 °C; and c. recovering hydrocarbon product from the reaction, the product having substantially the same volume as the hydrocarbon feed, the product having at least one of the following characteristics: i. an olefins content of less than 1 wt% as 1-decene equivalent as determined using HNMR; ii. a bromine number of less than 1.1 g Br/100 g; and iii. a di-olefins value per UOP-326 of about 4 grams iodine/100 grams oil or less.

BRIEF DESCRIPTION OF THE FIGURES

Figure 1 is a depiction of a top flow down lead-lag configuration of one embodiment.

Figure 2 is a depiction of a bottom flow up lead-lag configuration of one embodiment.

DETAILED DESCRIPTION

As used herein, the following terms have the following meanings:

"Cracking" includes any process for breaking down heavier molecules into smaller ones. "Thermal cracking" refers to an example of a cracking process that uses heat to perform such breaking of heavier molecules into smaller ones.

"Heavy oil" includes any oil that does not flow easily and more particularly, heavy oil generally includes any oil having an API of less than 22.2. Most heavy oils have a viscosity range from 10,000 cP to 10,000,000 cP (cP = centipoise). Heavy oil includes bitumen, which is found in oil sands.

"Hydrocarbon" is an organic compound consisting primarily of hydrogen and carbon, and in some instances, the hydrocarbon may also contain heteroatoms such as sulfur, nitrogen and oxygen.

"Hydrocarbon feed" refers to a hydrocarbon that may be processed by the methods of this invention. Generally, the hydrocarbon feed would have at least some olefins and/or di-olefins, and would be produced as a result of a cracking operation. Examples of hydrocarbon feeds include the whole or distillate fractions of any of the following: cracked oil, cracked heavy oil, cracked bitumen, cracked shale oil, cracked coal derived liquids, thermally cracked oil, thermally cracked heavy oil, thermally cracked bitumen, thermally cracked shale oil, thermally cracked coal derived liquids among other hydrocarbons. The term "hydrocarbon feed" as used throughout may include any fractions or distillates that arise from an upgrading process.

"Hydrocarbon product" as used herein refers to the hydrocarbon produced following a method to reduce the olefins content.

"Olefin" means any of a class of unsaturated open-chain hydrocarbons such as ethylene, having the general formula C_nH_{2n}, or unsaturated cyclic hydrocarbons such as cyclohexene, having the general formula $C_nH_{(2n-2)}$. As used herein, the term "olefin" is meant to encompass a mixture having olefins and/or di-olefins. "Di-olefins" are olefins that contain at least two double bonds.

"Pipeline specification" refers to the composition of an oil that allow it to be transported by pipeline. For example, bitumen will not flow through a pipeline efficiently and must be mixed with diluents and/or upgraded to be readied for pipeline transportation.

Conversion technologies that accomplish cracking (e.g. visbreaking, thermal cracking, coking, steam cracking, etc.) generate olefins and di-olefins. The presence of excessive quantities of olefins and di-olefins in a hydrocarbon stream or hydrocarbon feed can cause problems such as coking and/or fouling of catalysts and/or heat exchanger surfaces and other equipment failures. Also, a hydrocarbon product having a high olefins content is not suitable for pipeline specifications, as olefins tends to be reactive and increase the risk of fouling in the refineries or other facilities that process the cracked hydrocarbon stream, rendering the cracked hydrocarbon stream unmarketable. Accordingly, the olefins and di-olefins content of hydrocarbon feed should be reduced to achieve appropriate operational and pipeline specifications.

Most methods to reduce di-olefins and olefins content currently involve expensive hydro-processing infrastructures such as hydrotreaters and associated equipment, which require an extraneous source of hydrogen. Also, many methods to reduce the olefins and di-olefins content of a hydrocarbon stream require the addition of extraneous components such as hydrogen (H_2) and/or another hydrocarbon. For example, when upgrading is employed to treat a feed material, separate equipment is needed for hydrogen (H_2) generation to enable hydro-processing. In addition to the required hydro-processing unit, units needed to produce hydrogen (H_2) (e.g. Steam Methane Reforming) increase the complexity and cost of the oil upgrading facility and increase the carbon footprint of the overall process.

The alkylation of olefins and aromatics described in the prior art usually involves an aromatic-containing hydrocarbon material which has been subjected to a hydro-processing step at least once, before being employed as aromatic feed material to the alkylation process. The aromatic feed material supplied to the alkylation process in these instances is generally substantially free of basic compounds. Many methods of alkylation of olefins and aromatics usually involve a separate olefin-containing feed material, which is substantially free of nitrogen-, oxygen- and sulfur-containing compounds. Also, the processes for alkylating olefins and aromatics that are described in the art usually involve the deliberate addition of extraneous olefins to the aromatics-containing feed.

U.S. Patent Nos. 8,395,006 and 7,476,774 by ExxonMobil Research and Engineering Co. propose the use of a technology referred to as BenzOUT™ Reformate-alkylation to reduce the benzene content in gasoline. The technology uses two essential feed streams, a benzene-containing aromatic stream and a light olefin (such as ethylene or propylene) stream imported into the process. U.S. Patent No. 5,120,890, U.S. Patent Application Publication No. 2006/0194998 and PCT Patent Application Publication No. WO 2007/068008 describe other means to remove benzene using an olefin. In all of these cases, extraneous olefins are brought into the process with the specific purpose of decreasing the benzene levels in naphtha destined to make finished gasoline. Further, the benzene-containing aromatic stream was subjected to at least one hydro-processing step, usually a naphtha hydrotreater, before catalytic naphtha reforming. The hydro-processing step is the source of the reformate.

Other technologies to reduce the levels of benzene using alkylation with an olefin are described in Oil Gas J. 1993, 91(13):63-69 and Energy Fuels 2008, 22:1449-1455. In all instances the specific objective is to reduce the amount of benzene in the mixture, because the benzene content in gasoline is limited by fuel specifications. In all of these cases extraneous olefins are introduced into the process for the specific purpose of performing alkylation of benzene to reduce the amount of benzene in the treated product.

Alkylation of aromatics with olefins using aluminum chloride, solid phosphoric acid and more recently silica-alumina based zeolites as catalysts is described in Appl. Catal. A 2001, 221:283-294, Catal. Today 2002, 73:3-22 and, Appl. Catal. A 2005, 280, 89-103. These processes involve the production of alkyl aromatics as higher value chemical commodities from the feed material. The purpose of the alkylation processes describe therein is to specifically produce alkyl aromatics, which can be sold at a higher price than the feed materials. In these processes, the aromatic feed and the olefinic materials are separately purified before being added together for the reaction in order to limit any deleterious materials for either the process or the catalyst.

Extraneous olefins are also added in processes for the alkylation of sulfur-containing aromatics (thiophenic-compounds) to facilitate separation of the sulfur-containing aromatics, as described in Catal. Today 2008, 130:190-194 and Appl. Catal. B 2006, 64:254-261. Processes can also be found for alkylating oxygen-containing aromatics, such as phenol, with extraneous olefins as described in Microporous Mesoporous Mater. 2000, 39:457-463 and J. Mol. Catal. A 2005, 193:251-257. Such processes usually have the objective of producing a specific alkylated aromatic product.

As excessive olefins and di-olefins content is undesirable, disclosed herein is a method to reduce the olefins and di-olefins content of a hydrocarbon feed such that the resulting hydrocarbon product meets a pipeline and/or an operational specification. By "operational specification", it is meant that the resulting hydrocarbon product has a suitable composition to limit the risk of fouling during transport and limit the risk of fouling during operation of a refinery. For example, a hydrocarbon product with a suitable "operational specification" would have reduced fouling tendencies when used in equipment such as pipelines, heat exchangers or reactor beds.

While the aforementioned prior patents and literature describe methods to reduce the aromatics content of a refinery product, or the on-purpose manufacturing of an alkyl aromatic compound as product, the present disclosure is instead directed at a method to reduce the olefins and di-olefins content in a hydrocarbon feed. The present invention differs from the prior art in a number of respects:

(a) Olefins are not added extraneously.

(b) The process removes olefins and di-olefins without extraneous hydrogen (e.g. H_2 is not added or present, and there is no gas phase. The alkylation can be performed in a single liquid phase).

(c) The olefins and the aromatics involved in the alkylation reaction are not supplied as separate feed streams, nor required to be available as separate streams. The hydrocarbon feed

contains the olefins and the aromatics, and as such there is no need for two distallate feeds to the alkylation process. The necessary components can come from a single feed.
(d) The feed to the process does not require any part of the feed material to be subjected to a hydro-processing step, either before, or after the alkylation process.
(e) The feed material contains bases that would normally be thought to render such the feed material incompatible with the processes and acid catalysts employed for the alkylation of aromatics with olefins. For example, the hydrocarbon feed contains nitrogen bases.
(f) The feed material contains di-olefins, which are generally excluded from processes for alkylation of aromatics with olefins.

The feed that is treated by the process described herein is a cracked hydrocarbon stream and the olefins and di-olefins in t the feed are produced as a result of a cracking process, such as thermal cracking. The olefins are present as a consequence of the upstream processing and the olefins were neither intentionally produced, added to the feed or considered desirable components of the feed. The resulting "upgraded" hydrocarbon product meets at least of an operating specification or a pipeline specification. Using the presently described method, there is no need to import any of the reactants into the process. This is in contrast to the BenzOUT™ and other processes, since in those processes, the necessary components for the reaction to occur must be added to the feed and at least part of the feed material was subjected to a prior hydro-processing step. In the present method, the hydrocarbon feed (e.g. produced as a result of a cracking process) comprises the necessary reactants to facilitate the reduction of olefins and di-olefins content in the hydrocarbon feed, without the need to "import" reactants into the feed, such as olefins, di-olefins, aromatics and/or hydrogen.

Thus, the presently described method allows for:

(a) the application of olefins-aromatics alkylation to a feed material that due to its boiling point distribution and presence of basic nitrogen containing compounds, has not previously been considered suitable for olefins-aromatics alkylation. In particular, alkylation catalysts are acid catalysts that can be deactivated (e.g. neutralized) when there are basic compounds (e.g. compounds with alkaline properties) present in the feed. In the prior art examples of which Applicant is aware, the aromatic feed was extensively hydro-processed, or purified by separation before subjected to alkylation. Also, in all of the previous prior art examples the Applicant is aware of, the olefinic feed was limited to short-chain, usually C_2, C_3 or C_4 normally gaseous olefins, lacked substantial levels of di-olefins and the olefinic feed was purified by separation beforealkylation.

(b) an industrially acceptable conversion rate for olefins-aromatics alkylation using acid catalysis of a feed material that contains compounds that are normally considered poisons for typical olefins-aromatics alkylation catalysts, such as basic nitrogen-containing or other compounds with alkaline properties, as well as strongly adsorbing heteroatom containing compounds;

(c) the application of olefins-aromatics alkylation as a process to reduce the olefins content of a thermally cracked hydrocarbon containing product; and

(d) the application of olefins-aromatics alkylation as an alternative to hydro-processing to make a hydrocarbon containing product suitable for operation and/or pipeline transport.

(e) the application of olefins-aromatics alkylation to a hydrocarbon feed that has not been hydro-processed, or subjected to separation for the removal of material in the stream that may be deleterious to typical olefins-aromatics alkylation catalysts.

The alkylation reaction according to the present disclosure is a reaction comprising the addition of an olefinic group to an aromatic group (e.g. the aromatic group is essentially the alkyl acceptor and would be available within the hydrocarbon feed). The olefins-aromatics alkylation reaction produces an alkylated aromatic compound with reduced olefins content. The reaction can occur without the use of an external source olefins, di-olefins, aromatics and/or hydrogen and, without substantial loss of the volume of material as compared to the hydrocarbon feed. The reduction in volume in the hydrocarbon product is only around 0.1%-10% v/v as compared to the original hydrocarbon feed. Thus, the volume of the product is substantially similar to the original feed. The exact volume decrease is dependent on the original olefins/di-olefins content of starting hydrocarbon feed and other reaction conditions. As would be appreciated, the olefinic group that takes part in the reaction may be a sole olefin, or it may be attached to and/or form part of a molecule containing at least one other functional group. Similarly, the aromatic group may be a sole aromatic hydrocarbon or may be attached to, and/or form part of a molecule that contains at least one other functional group.

Using the method disclosed herein, the olefins content of the hydrocarbon feed is reduced to a di-olefins value per UOP-326 of about 4 grams iodine/100 grams oil or less. In some embodiments, the di-olefins value per UOP-326 is about 2 grams iodine/100 grams oil, and in yet other embodiments, the di-olefins value per UOP-326 is about 0 grams iodine/100 grams oil. That being said, other methods for representing olefins and di-olefins content in a hydrocarbon, include but are not limited to Proton Nuclear Magnetic Resonance (^{1}H NMR) spectroscopy, Bromine Number per ASTM-D1159, etc., and any of these methods may be used as well to quantify the olefinicity of the resulting hydrocarbon product. The analytical method used to determine the olefins and, more specifically, the di-olefins content is the choice of the person of ordinary skill in the art.

The process is conducted at a pressure and temperature which facilitates olefins and di-olefins alkylation with aromatics (olefins-aromatics alkylation). Generally, suitable temperatures are below about 380°C. In one embodiment, the temperature range is from about 50°C to about 380°C. In yet another embodiment, the temperature range is from about 150°C to about 350°C.

The process is conducted at a pressure wherein the reactants and resultant hydrocarbon product are substantially liquid. While the transition phase from liquid to vapour is pressure and temperature dependent, the methods disclosed herein can be carried out from about 0 to about 8 MPa. In one embodiment, a suitable pressure range is from about 2 to about 5 MPa.

The process has a weight hourly space velocity of from about 0.01 to about 20 h^{-1}, In one embodiment, the weight hourly space velocity is from about 0.02 to about 2.0 h^{-1}.

According to one or more embodiments the method for olefins alkylation with aromatics includes contacting the feed material (e.g. hydrocarbon feed or hydrocarbon stream that is usually produced from an upgrading process such as thermal cracking) with at least one acid catalyst, at a temperature below around 380 °C and at a pressure sufficient for the reactants and resultant hydrocarbon product to be substantially liquid. The acid catalyst may be a heterogeneous catalyst selected from the group consisting of supported liquid phase catalysts, solid acid catalysts and homogeneous catalysts immobilized on a solid support.

Because the hydrocarbon feed would generally be produced as a result of an upgrading process, such as an upgrading process for the processing of heavy oils, the hydrocarbon feed may have species that include heteroatoms such as sulfur, nitrogen and oxygen. These types of heteroatoms can sometimes be problematic for acid catalysis. However, the present method can operate despite the presence of heteroatoms in the feed.

Preferably, said liquid phase catalyst comprises Brønsted acids (e.g. H_2SO_4, HF) and Lewis acids (e.g. BF_3). In one embodiment, the heterogeneous catalyst is selected from the group consisting of silica-alumina, structured silica-alumina molecular sieves, such as but not limited to MCM-41, and crystalline silica-alumina zeolites, such as but not limited to zeolites of the families MWW, BEA, MOR, MFI and FAU, solid phosphoric acid (SPA), aluminophosphase and silico-aluminophosphates, such as zeolites of the AEL family, heteropolyacids, acidic resins and acidified metal oxides and mixtures thereof. The acid catalyst may be a heterogeneous catalyst, facilitating the separation of the hydrocarbon product from the catalyst.

In certain embodiments, said at least one acid catalyst has sufficient acid strength to catalyze the olefins-aromatics alkylation reaction, as well as an acid strength distribution to retain sufficient activity in contact with the basic compounds that are present in the hydrocarbon feed. The temperature and acid catalyst are selected such that an optimal combination of olefins-aromatics alkylation activity and smallest amount of catalyst inhibition by compounds that are strongly adsorbing, or are basic in nature in the hydrocarbon product (with reduced olefins levels) is achieved.

In one embodiment, the process comprises a reactor containing an acid catalyst through which the olefins, di-olefins and aromatics contained in the feed material are treated at appropriate operating conditions, resulting in a hydrocarbon product which meets at least one operating and/or pipeline specification.

In one embodiment, the hydrocarbon feed may be derived from heavy oil or bitumen upgrading, the feed comprising at least one olefinic, di-olefinic and aromatic component, and optionally further comprising at least one heteroatom component. In one or more specific embodiments, the hydrocarbon feed of the present invention may be a cracked distillate or a blend of virgin and cracked distillates. The properties of a typical blend of 28°C to 343°C boiling range virgin and cracked distillates from Athabasca bitumen are presented below.

Conversion of the feed in the reactor is achieved through the selective parameters of the reactor, the catalyst, combinations of catalysts and the operating conditions within the reactor. In one embodiment, different combinations of the reactor, catalyst and operating conditions of the reactor convert the olefins in the hydrocarbon feed to a sufficient degree to meet the conversion objective. In certain embodiments, the conversion objectives include processing the feed to achieve at least one operation and/or transport specification.

The process may comprise one or more post-conditioning steps. These steps will be determined based on the design of the conversion step (e.g. catalytic step) and the downstream process requirements (e.g. the intended use of the resulting hydrocarbon product). Post-conditioning steps may include phase separation, product fractionation and/or heat exchange.

Referring now to Figures 1 and 2, there is depicted a feedstock 10 preheated in feed-effluent heat exchanger 20 then further heated in heater 30. The feedstock then enters the lead reactor 40 followed by the lag reactor 50 (depending on the position of the valves the lead and lag reactor may be reversed in positioning in the process depending on process conditions and/or catalysts aging) and exits the reactor. Before entering the lag reactor 50 and being that the reaction is exothermic, the effluent of the lead reactor 40 is quenched using the inter reactor quench feed stream 60. The effluent from the lag reactor product is then cooled in the feed-effluent heat exchanger 20 before exiting the system as the product 70. In Figure 1, the configuration is top flow down, lead-lag. In Figure 2, the configuration is bottom flow up, lead lag.

In all of the above configurations, the following should be considered:

The first step may be feed conditioning, although this step is not always needed. The process may require feed conditioning to adjust the temperature and pressure before the feed enters the reactor. The process may also require certain additions to the hydrocarbon feed. For example, it may be useful to reduce the content of contaminants that would reduce the efficacy of the later catalytic steps, although the method disclosed herein is designed to function despite the presence of these contaminants. There may also be a need for pH adjustment for example. The need for feed conditioning is determined by the condition and composition of the hydrocarbon feed, which is in turn a function of the process by which the feed is generated. For example, certain hydrocarbon feeds produced by suitable upgrading processes may not require pre-conditioning, while other feeds may require some pre-conditioning. Other examples of pre-conditioning may include passing the hydrocarbon feed through a first conditioning catalyst and/or absorbent material to improve the operation and/or performance of the acid catalyst. Another example of pre-conditioning involves altering the temperature of the hydrocarbon feed prior to carrying out the alkylation process. As a person skilled in the art would appreciate, the pre-conditioning steps may be carried out in the same vessel as the olefins-aromatics alkylation reaction or alternatively, the pre-conditioning steps may be carried out in different vessels.

Also, in respect of the selection of the hydrocarbon feed to the reaction, any suitable feed may be used, provided that at least some olefins and aromatics are present. Thus, the feed may comprises one or more various fractions or distillates from an upgrading process. As a person skilled in the art would appreciate, an upgrading process may produce different fractions, ranging from light oils to heavy bottoms. Combinations of these fractions may be combined or used alone as the hydrocarbon feed. Suitable feed compositions would generally have between 5 and 40% olefins content on a per weight basis and have between 5 and 40% aromatics content on a per weight basis. An example of a suitable upgrading process that may be used to treat a heavy hydrocarbon feed to produce an upgraded product with olefinic hydrocarbon material is Applicant's U.S. Patent Application Publication No. 20140138287. The upgraded oil produced from the upgrading process may comprise several distillate streams of hydrocarbons of different boiling points, and each distillate stream or combinations of the distillate streams may be used as the hydrocarbon feed for the processes of this invention. As an example, the minimum and maximum ranges of various components that may be in the feed are as follows (note that olefins in this context includes di-olefins):

Maximum Ranges:

Aromatics, %w	40
Olefins, %w	45
Diolefins, %w	10
Sulfur, %w	3.6
Nitrogen, %w	0.2
Basic Nitrogen, ppmw	600

Minimum Ranges:

The minimum ranges could be:

Aromatics, %w	5
Olefins, %w	1
Diolefins, %w	0.2
Sulfur, %w	0
Nitrogen, %w	0
Basic Nitrogen, ppmw	0

The second step in the process is conversion of the feed in the reactor. The reactor, the catalyst and the operating conditions are chosen to allow this process to achieve desirable olefins content reduction. Using the teachings of the present invention, it is possible to select different combinations of the reactor, catalyst and operating conditions that will convert the olefins in the feed to a sufficient degree to meet the desired operating and/or pipeline specifications objective(s). The catalyst is chosen so that the catalyst can perform the reaction without being poisoned or otherwise inhibited to an extent that the reaction cannot occur. In practice, this means that if an acid catalyst is used, the reaction conditions are chosen such that the acid catalyst would not become irreversibly poisoned with basic compounds present in the feed. The temperature can be chosen to prevent the acid catalyst from reacting with basic compounds, or at least, from becoming irreversibly bound to the basic compounds.

The hydrocarbon product produced from the reactor may still have some olefins and di-olefins. Therefore, it is possible to carry out a second conversion step, either in the same reactor or in a second reactor. The second conversion step can be carried out with the same catalyst or with a different catalyst. As an example, it may be appropriate to use a first catalyst to remove the di-olefins, and then, use a second catalyst to remove the olefins. The first catalyst and second catalyst may differ, and the reaction conditions can be chosen to be appropriate for each catalyst.

The third step is the hydrocarbon product conditioning (e.g. post-conditioning steps). The types of product conditioning may include addition of diluents, etc. This step is not always needed, as it is contemplated that the hydrocarbon product produced from the reaction will meet a pipeline specification or an operational specification. However, particular post-conditioning or post-treating steps may be needed depending on the intended use of the hydrocarbon product. Post-conditioning may include but is not limited to operations such as cooling steps to cool the hydrocarbon product exiting the reactor, gas removal and/or phase separation, product fractionation, various heat exchange steps and other post-conditioning steps that may be required depending on the intended use of the hydrocarbon product.

Catalyst

The type of catalyst affects the selection of the reactor and operating conditions. Two aspects to consider are the catalyst form and the catalytically active material within the catalyst responsible for catalyzing the olefins-aromatics alkylation reaction:

(a) A heterogeneous catalyst having a particle size and particle morphology suitable for use in a packed bed may be used. Catalysts that are suitable for use in packed bed reactors are known in the art. Such catalysts have a smaller chance of contaminating the hydrocarbon product as the catalyst-product separation tends to be easier, allowing for simpler reactor and operating configurations. This may be advantageous when used in a field upgrading application (e.g. when the upgrading occurs on site), as such field upgrading applications are most economical when there is less complex equipment set-up.

(b) The catalytically active material has acidic properties. The catalytically active material may be incorporated into a heterogeneous catalyst. The catalytically active material may be in the form of a solid catalyst, a supported liquid phase catalyst, or a homogeneous catalyst immobilized on a solid support, for example. The catalytically active materials with acidic properties may be promoted with metals, even though metal promoters are not specifically required by the processes disclosed herein.

It is known that among others, the following heterogeneous catalysts are catalytically active materials for liquid phase aromatic alkylation:

(b.1) Zeolites of the framework type FAU, like Y-zeolite (e.g. J. Mol. Catal. A 2007, 277:1-14 and Appl. Catal. A 1999, 182:407-411);

(b.2) Zeolites of the framework type BEA, like Beta-zeolite (e.g. Appl. Catal. A 1997, 153:233-241);

(b.3) Zeolites of the framework type MOR, like mordenite (e.g. J. Mol. Catal. A 2004, 223:305-311);

(b.4) Zeolites of the framework type MFI, like ZSM-5 (e.g. Energy Fuels 2008, 22:1449-1455);

(b.5) Zeolites of the framework type MWW, like MCM-22 (e.g. Appl. Catal. A 2005, 292:68-75 and J. Catal. 2005, 236:45-54);

(b.6) Zeolites of the framework type MTW, like ZSM-12 (e.g. Catal. Rev.-Sci. Eng. 2002, 44:375-421);

(b.7) Amorphous silica-alumina based catalysts (e.g. Ind. Eng. Chem. Res. 2005, 44:5535-5541);

(b.8) Natural clays, such as montmorillonite (e.g. Helv. Chim. Acta 1987, 70:577-586);

(b.9) Solid phosphoric acid (e.g. Ind. Eng. Chem. Res. 2006, 45:7399-7408 and J. Am. Chem. Soc. 1945, 67:1060-1062);

(b.10) Acidic resins, such as sulfonated styrene-divinylbenzene copolymers (e.g. React. Func. Polym. 2000, 44:1-7).

The aforementioned list is by no means exhaustive. However, this process can be performed using appropriate catalysts despite the presence of contaminants typically found in industrial feed materials and which may be deleterious to acid catalysts in general. Thus, in contrast to some known methods of alkylation , the process of the present invention can be used at temperatures below around 380 °C with a hydrocarbon feed that contains potential catalyst poisons in the feed.

An amorphous silica-alumina catalyst or a crystalline silica-alumina catalyst may be used in this process. The silica-alumina catalyst may have a SiO_2 to Al_2O_3 ratio of 0-99 wt% for example, but in some circumstances, it may be appropriate to use a catalyst having a SiO_2 to Al_2O_3 ratio of 5-75 wt%. The silica-alumina catalyst is generally activated by calcination at a temperature in the range 500-600 °C.

The choice of catalyst type is based on accessibility and performance in the presence of basic compounds, such as pyridine, which may be acid catalyst poisons. Basic nitrogen compounds are typically present in most hydrocarbon feed materials that have not been hydro-processed.

Operating conditions

The operating conditions are selected to match the catalyst employed. There are a number of guiding principles in selecting appropriate operating conditions. These are:

(a) The temperature range depends on the catalyst selected and the contaminants present in the hydrocarbon feed. Olefin-aromatic alkylation reactions are favored thermodynamically by low temperature. However, the contaminants present in the hydrocarbon feed, in particular basic nitrogen species, tend to deactivate the acid sites in the catalyst and render it ineffective after a short exposure time. It has been found however, that operating the catalyst at higher temperatures the catalyst can perform the reaction without being poisoned or otherwise inhibited to an extent that the reaction cannot occur. If too high a temperature is used however the rate of olefins polymerization reactions that promote fouling and catalyst deactivation will prevail. Furthermore, the operating temperature needs to consider the rate of catalytic or thermal cracking such that the amount of olefins produced during the cracking process does not exceed the target olefins concentration at the outlet of the reactor. The maximum operating temperature is in the range of about 320-380°C for silica-alumina catalysts. The lower operating temperature limit is determined by the activity of the catalyst in the presence of basic heteroatom feed contaminants, where below this point olefin-aromatic alkylation reactions are too slow for significant olefins conversion to occur.

In order to perform acid catalysis at an industrially meaningful rate, the minimum operating temperature must be sufficiently high to avoid excessive catalyst poisoning by irreversible adsorption of compounds. In the presence of nitrogen bases in the feed, the minimum operating temperature is in the range of from about 200-300 °C for silica-alumina catalysts. The temperature conditions appropriate for acidic resin and SPA catalysts are from about 50-380°C, and more particularly, from about 150-350°C.

(b) The pressure should be sufficient to keep most of the hydrocarbon feed material in the liquid phase at the operating temperature. This limits the amount of light olefins that may be present in the vapour phase and that may pass through the reactor unconverted. A typical operating pressure is in the range 0-8 MPa. For operating temperatures in the range of about 300-380°C, which is typical for silica-alumina catalysts, the pressure range is about 2-5 MPa.

(c) The flow rate is determined by the olefins conversion requirements for the selected combination of catalyst and operating conditions. The range of weight hourly space velocity (WHSV) is 0.01 to 20 h^{-1}. The WHSV range is generally in the range of 0.02 to 2 h^{-1}.

Optimum conditions for olefins conversion are adjusted to meet final product specifications and may be determined empirically, depending on changes in feed composition, selected catalyst types (i.e. one or more catalyst types used in conjunction) and aging, number of reactor beds (i.e. single vs. multiple with interstage cooling) and other unit constraints. Process operating conditions can be optimized using strategies known in the art.

(d) Depending on the catalyst, it may be beneficial to add water, or compounds that may produce water, such as alcohols, to the feed.

Reactor

The process may be conducted in a conventional packed bed reactor. The catalyst is contained and retained in a process vessel that is designed according to principles known in

the art. In one embodiment, a single adiabatic packed bed (fixed bed) reactor is employed. The use of multiple catalyst beds within the reactor, the use of inter-bed quench feed stream, the use of more than one reactor and product recycling may all be considered.
The adiabatic temperature increase should be controlled. Aromatic alkylation with olefins and olefins dimerization (a possible side-reaction) are both exothermic. Implementation of heat management strategies in the reactor design is known in the art.

As depicted in the figures, the reactor may be operated either in down flow or up flow configuration. The operation of the reactor in an up flow configuration with a liquid filled catalyst bed improves heat transfer and catalyst wetting, and maximizes liquid holdup. This configuration also facilitates removal of heavy products (gums) typical of di-olefins reactions from the catalyst by dissolving it in the liquid product, as described in patents related to the buildup of fouling agents (e.g. U.S. Patent No. 4,137,274). The operation of the reactor in a down flow configuration facilitates maintenance and catalyst replacement, since the contaminated portion of the catalyst will be concentrated at the top of the reactor where it is easier to access and replace.

From an operational point of view, the reactor may further be designed for easy maintenance and catalyst replacement in the field.

Hydrocarbon Feed

The hydrocarbon feed for this process is generally derived from a cracking process. It is known that cracking processes generate olefins, through the cracking of high molecular weight feeds. The hydrocarbon feed is thus generally olefins-rich. This process has particular applicability in the upgrading and subsequent removal of olefins from heavy oils. Heavy oils tend to have high aromatics content. The aromatics are retained during the upgrading process, such that the feed has both suitable amounts of olefinic and aromatic components to be used as reactants in this process. Also, the resulting lighter oils derived from the heavy oils through upgrading still contain heteroatoms and nitrogen bases normally thought to be deleterious for acid catalysis. Thus, the method disclosed herein can operate despite the presence of these components.

An example of an upgrading process that produces a suitable feed is described in co-pending PCT Patent Application No. PCT/CA2013/001066. This upgrading process can potentially produce hydrocarbon streams having different molecular weight distributions. For example, following distillation, there may be multiple distillate streams produced, each of a different weight distribution, and each containing different proportions of aromatic and olefinic components. An appropriate fraction of distillate can be selected for use in the methods of this invention, or alternatively, distillate streams can be combined to provide for a suitable hydrocarbon feed.

As discussed throughout, one aspect of the present process is that the olefins-aromatics alkylation can take place irrespective of the ratio of olefins to aromatics in the hydrocarbon feed. Thus, unlike previously described olefins-aromatics alkylations, the present process does not require strict control of the olefins-aromatics ratio. In practice, the ratio of olefins to aromatics should not exceed around a 3:1 molar ratio. This ratio is based on the physical constraints of each aromatic molecule, which will seldom have space on the aromatic rings (due to steric constraints) to allow addition of more than three olefinic molecules per aromatic molecule. There is, however, no lower limit on the ratio of olefins to aromatics. The optimal ratio of catalyst to olefins is determined by the weight hourly space velocity. The process allow for the ability to perform olefins-aromatics alkylation in a feed material that contains other species that are normally considered catalyst poisons for acid catalysts. Catalysts and operating conditions are chosen so that combinations that would normally cause the catalyst to be poisoned are essentially neutralized in their effect. Finally, it is noted that the hydrocarbon feed may contain any size (e.g. carbon length) of olefins. Many hydrocarbon feeds would contain olefins are in the naphtha range, i.e. C_5-C_{11} olefins, but more generally, the olefins in the feed are expected to be in the approximate range of C_2-C_{15}. This would include both the naphtha and kerosene range, as well as C_2's, C_3's and C_4's, which may be dissolved in the naphtha.

The method will be more readily understood by reference to the following examples, which are included merely for the purposes of illustration of certain aspects of the embodiments. The examples are not intended to limit the invention, as one of skill in the art would recognize from the above teachings and the following examples that other techniques and methods can satisfy the claims and can be employed without departing from the scope of the method.

EXAMPLES

Example 1: Modeling of Olefins/Di-olefins Conversion

Based on proprietary modeling, the following chart depicts three Examples (A, B and C) of a feed blend of 28 °C to 343°C boiling range virgin and cracked distillates with typical properties from Athabasca Bitumen after treating over amorphous silica alumina catalyst and the resulting products:

	Feed	Amorphous Silica Alumina Catalyst		
	343°C- Virgin and Cracked Distillates from Bitumen	Example A	Example B	Example C
Olefins Conversion %w	-	5	50	100
Di-olefins Conversion %w	-	52	71	100
Weight %	100	100	100	100
Volume %	100	99.86	98.62	97.22
Volume Loss %v	-	0.14	1.38	2.78
Oil Properties				
Average Boiling Point, °C	252			
Specific Gravity	0.8799	0.8812	0.8922	0.9050
API	29.32	29.08	27.10	24.85
Average Molecular Weight	189			
UOP K	11.16			
Bromine # per ASTM-D1159, g Br_2/100 g	18	17	8	0
Diene # per UOP-326, g I_2/100 g	4.3	2.0	1.0	0.0
Aromatics %w	28.4	29.2	35.6	42.9
Olefins %w	14.5	13.8	7.3	0.0
Di-olefins %w	2.1	1.0	0.6	0.0
Viscosity, cSt @ 7.5°C	5.8	6.0	8.4	13.1
Viscosity, cSt @ 20°C	4.2	4.4	5.9	8.8
Composition, %w				
Carbon	85.57	85.57	85.57	85.57
Hydrogen	11.66	11.66	11.66	11.66
Sulfur	2.63	2.63	2.63	2.63
Nitrogen	0.06	0.06	0.06	0.06
Oxygen	0.08	0.08	0.08	0.08
Ni + V, ppmw	0	0	0	0

Example 2: Experimental Zeolite Catalyst on thermally cracked oil sands derived bitumen at 249°C

A zeolite catalyst of the framework type MOR was employed for the conversion of a distillate cut containing straight run and thermally cracked oil sands derived bitumen. A batch reactor was loaded with 4-5 g catalyst and the distillate cut, which had a API gravity of 19.5 and an olefins content of 4.6 mol C=C per gallon distillate. The catalyst and feed were brought into contact at different temperatures and at an equivalent weight hourly space velocity of 1.4 1/h. The olefins conversions obtained at different temperatures are listed in the following table:

Temperature (°C)	249	304	316	346
Bromine number (g Br/100g)	11.9	8.0	10.6	7.4
Olefins conversion (%)	35	56	42	60

This example demonstrates the conversion at 249 °C, showing that the reaction can be meaningfully performed in an olefins-aromatics alkylation unit that does not require a fired heater for feed preheating, which is particularly advantageous in field upgrading of oil and bitumen.

Example 3: Amorphous silica alumina catalyst on oil sands bitumen and thermally cracked deasphalted oil

This example evaluated olefins conversion of an atmospheric distillate was evaluated by laboratory experimentation using batch reactors. The experimental feed was an atmospheric distillate containing virgin oil sands bitumen and thermally cracked deasphalted oil. In a typical experiment the reactor was loaded with 4.5 g of amorphous silica alumina catalyst and 6.3 g of distillate. The reactor was operated under autogenous pressure and was exposed to the temperatures indicated in the following table for 1 hour by heating in a fluidized sand bath heater. The decrease in olefins content of the distillate was determined by bromine number titration and was found to be as follows:

Temperature (°C)	315	330	350
Olefins conversion (%)	47	57	68

Example 4: Solid phosphoric acid catalyst on thermally cracked oil sands bitumen distillate

An oil sands bitumen distillate containing thermally cracked distillate was contacted with a solid phosphoric acid catalyst at an equivalent weight hourly space velocity of 2.8 1/h (equivalent residence time of 22 minutes) at 159-162 °C. The hydration level of the catalyst was not controlled during the test. While these conditions were not optimized, a 14 % olefins conversion was still achieved.

Example 5: Silica and alumina catalyst on mixture of virgin oil sands bitumen and thermally cracked deasphalted oil

A pilot-scale reactor was loaded with 140g Siralox 30 catalyst. This catalyst contained silica and alumina in a 30:70 ratio and is classified as an amorphous silica alumina catalyst. The catalyst was loaded as extrudates with a particle diameter of 3 mm. The reactor internal diameter was 44.5 mm. The reactor was continuously operated with an industrial feed material containing a mixture of virgin oil sands bitumen and thermally cracked deasphalted oil. The feed had an olefins content of 18.3 ± 0.7 g Br/100 g, which included a di-olefins content of 1.67 ± 0.07 determined as per the UOP 326-08 standard test method. The specific gravity of the distillate feed was 0.9374 determined using an Anton Paar Abbemat 200 density meter. The operating pressure was maintained in the range 5-6 MPa.

The conversion data for the following temperature (T) and weight hourly space velocity (WHSV) conditions were obtained with freshly loaded catalyst:

T (°C)	WHSV (1/h)	Bromine number (g Br/100 g)	Olefins conversion (%)	Di-olefins conversion (%)
315	1.0	12.9 ± 0.2	30	46
350	1.0	12.7 ± 0.2	31	31
315	0.5	5.5 ± 0.2	70	not determined
350	0.5	8.0 ± 0.1	56	not determined

Because the industrial feed contained nitrogenous bases and di-olefins that could deactivate and foul the catalyst, it was of interest to evaluate the stability and performance of the catalyst. The average stable olefins conversion for a run length of over 2000 minutes was 20 ± 3 % at 350 °C, 5.2 MPa and WHSV of 0.5 1/h. This steady state conversion level was maintained over a period of more than 1300 minutes before the conditions were changed to perform the next test.

Based on this performance, near complete olefins conversion is anticipated with SIRALOX™ 30 operated at 350 °C and WHSV of 0.11/h. This demonstrates that the North American pipeline specification limit for olefins can be met under these operating conditions.

Example 6: Experimental catalyst-absent conversion for comparative analysis

A blank experiment was performed using a batch reactor operated at 350 °C for 1 hour to determine whether the observed results could be explained in terms of thermal alkylation and oligomerization of the olefins (commensurate with results from 2005 Energy Fuels 19:1462-1467). It was found that the extent of olefins conversion due to thermal reactions, in the absence of a catalyst as taught in this invention, was 4 % or less.

Example 7: Material balance analysis using silica-alumina catalyst

In order to determine the potential loss of material during the process described by this invention, a test run on a pilot-reactor was performed. The duration of the test run was 44 hours. A silica-alumina catalyst was employed at 315 °C, 5 MPa and a WHSV of 1 h^{-1}. At the end of the test run, a material balance was performed. The difference was determined between the total amount of oil sands bitumen derived material fed to the process and liquid product collected after process, which was cooled down to near ambient temperature. The liquid recovery was 98%, demonstrating that there was no substantial loss of liquid hydrocarbons due to the practice of the processes of this disclosure. Thus, this method allows for the production of a hydrocarbon product that is substantially similar in volume to the starting feed. By "substantially similar", it is meant that the volume of the hydrocarbon product is around 90% or more of the starting feed.

As many changes may be made to the preferred embodiment of the invention without departing from the scope thereof; it is intended that all matter contained herein be considered illustrative of the invention and not in a limiting sense.

What is claimed is:

1. A process to reduce the olefins content in a hydrocarbon feed, said process comprising:

alkylating said hydrocarbon feed with an aromatic component within the feed without the requirement to add external sources of aromatics, olefins, or hydrogen and without the requirement to adjust or predetermine the ratio of olefins content to aromatics content in the hydrocarbon feed prior to the alkylation;

wherein the alkylation is carried out for a duration, and at a temperature sufficient to reduce the olefins content of the hydrocarbon feed,

wherein said alkylating said hydrocarbon feed uses at least one catalyst,

wherein the hydrocarbon feed includes basic compounds, and

wherein said alkylating is carried out at a temperature of greater than 150 °C.

2. The process of claim 1, wherein the hydrocarbon feed has not been pre-treated to remove basic compounds prior to the alkylation.

3. The process of claim 1, further comprising recovering a hydrocarbon product following the alkylation, and alkylating the hydrocarbon product using a second catalyst, the second catalyst being different from the catalyst used in the first alkylation step.

4. The process of claim 1, wherein the hydrocarbon feed comprises oxygen-containing, nitrogen-containing and/or sulfur-containing compounds that could cause catalyst poisoning.

5. The process of claim 1, wherein the at least one catalyst is an acid catalyst.

6. The process of claim 5, wherein said acid catalyst is a heterogeneous catalyst.

7. The process of claim 6, wherein said heterogeneous catalyst is selected from the group consisting of silica-alumina, structured silica-alumina molecular sieves, MCM-41, crystalline silica-alumina zeolites, zeolites of the families MWW, BEA, MOR, MFI and FAU, solid phosphoric acid (SPA), aluminophosphase and silico-aluminophosphates, zeolites of the AEL family, heteropolyacids, acidic resins, acidified metals and mixtures thereof.

8. The process of claim 6, wherein said heterogeneous catalyst is a Bronsted acid or a Lewis acid.

9. The process of claim 1, wherein said hydrocarbon feed is produced as a result of a cracking process.

10. The process of claim 9, wherein the hydrocarbon feed is the result of cracking a light oil, medium oil, heavy oil, extra heavy oil, bitumen, oil sands, shale oil, oil shales, coal and coal derived liquids, biomass and biomass derived liquids, organic waste and waste derived liquids, or a mixture thereof.

11. The process of claim 1, further comprising controlling the water content of the alkylating step.

12. The process of claim 1, wherein the process is carried out for sufficient time that the alkylated hydrocarbon feed meets at least one operating specification.

13. The process of claim 1, wherein the process is carried out for sufficient time that the alkylated hydrocarbon feed meets at least one pipeline specification.

14. The process of claim 1, wherein said olefinic content in said hydrocarbon product has:

a di-olefins value per UOP-326 of about 4 grams iodine/100 grams oil or less;

an olefins content of less than 1 wt% as 1-decene equivalent as determined using HNMR; or

a bromine number of less than 1.1 g Br/100 g.

15. The process of claim 1, wherein the process is carried out at a temperature below 380°C.

16. The process of claim 1, wherein the process is conducted at a pressure wherein the hydrocarbon feed and the hydrocarbon product are substantially liquid.

17. The process of claim 1, wherein said the process is carried out at a weight hourly space velocity of from about 0.01 to about 20 h^{-1}.

18. The process of claim 1, further comprising pre-conditioning the hydrocarbon feed prior to alkylating.

19. A process for producing a hydrocarbon product meeting pipeline specification, comprising:

a. selecting a hydrocarbon feed produced by thermal cracking of a heavy oil feed, the hydrocarbon feed containing olefins and di-olefins, and basic compounds;

b. alkylating the olefins and di-olefins in the presence of a catalyst under conditions such that the olefins and di-olefins within the feed react with aromatic compounds in the feed, without the requirement to import additional olefins, aromatics or hydrogen into the reaction, wherein the conditions comprise a temperature of greater than 150 °C; and

c. recovering hydrocarbon product from the reaction, the product having substantially the same volume as the hydrocarbon feed, the product having at least one of the following characteristics:
    i. an olefins content of less than 1 wt% as 1-decene equivalent as determined using HNMR;
    ii. a bromine number of less than 1.1 g Br/100 g; and
    iii. a di-olefins value per UOP-326 of about 4 grams iodine/100 grams oil or less.

20. The process of claim 1, wherein the aromatics content in the hydrocarbon feed ranges from 5 to 40 % weight and the olefins content ranges from 1 to 45 % weight.

21. The process of claim 1, wherein the sulfur content of the hydrocarbon feed ranges from 0 to 4 % weight and the nitrogen content ranges from 0 to 0.2 % weight.

22. The process of clam 1, wherein the alkylation is carried out in a single liquid phase.

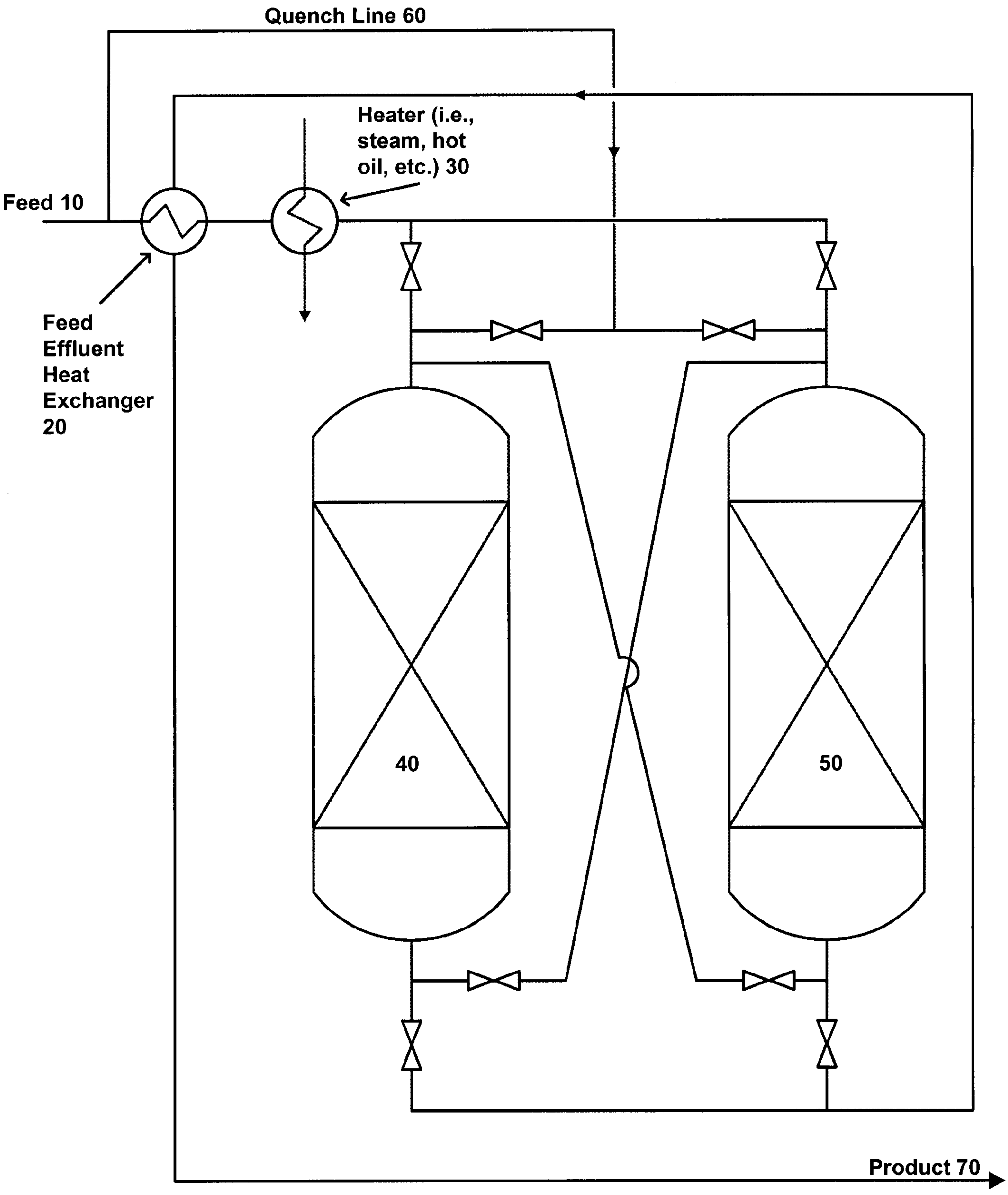

Quench Line 60
Heater (i.e., steam, hot oil, etc.) 30
Feed 10
Feed Effluent Heat Exchanger 20
40
50
Product 70

FIG. 1

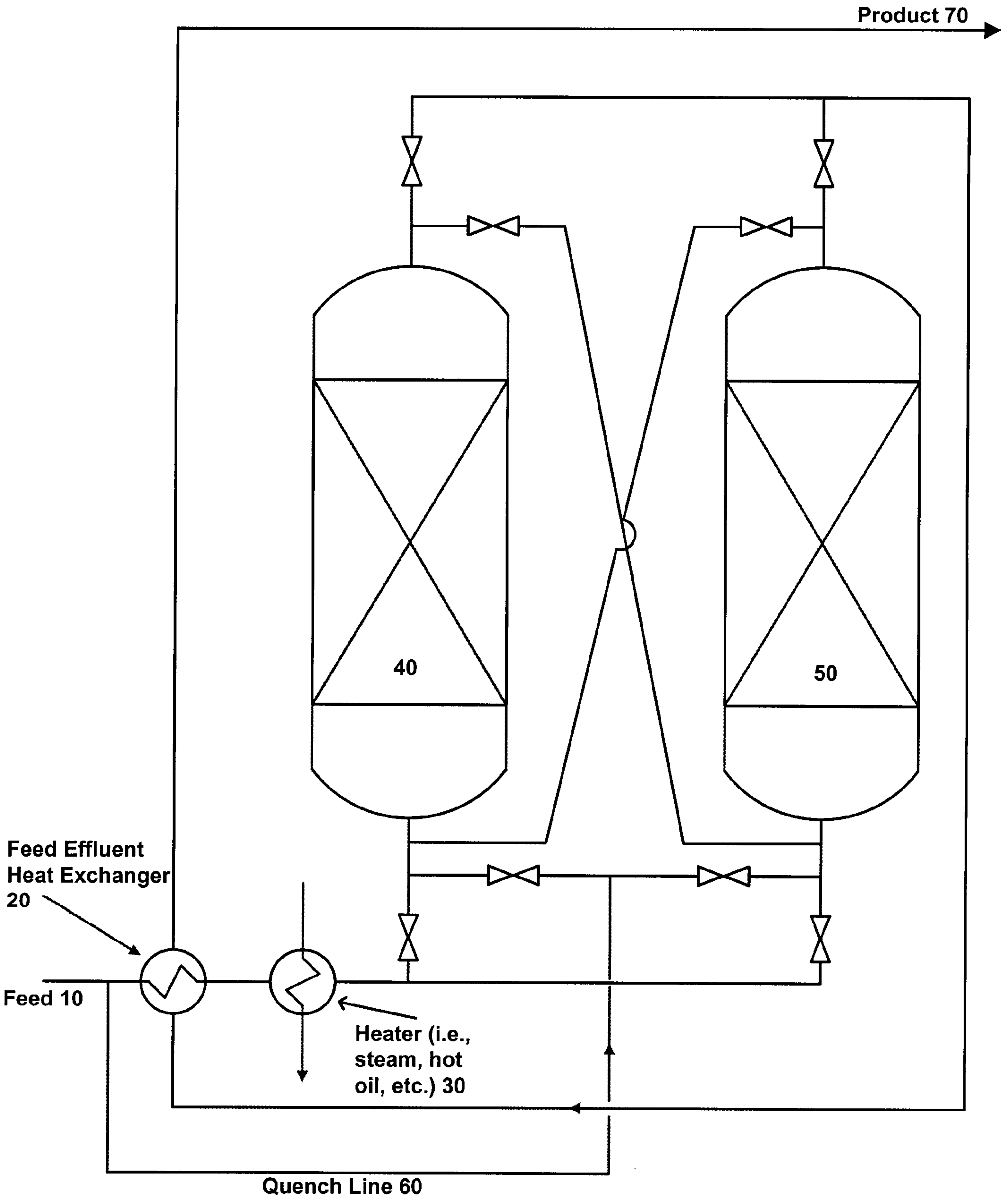
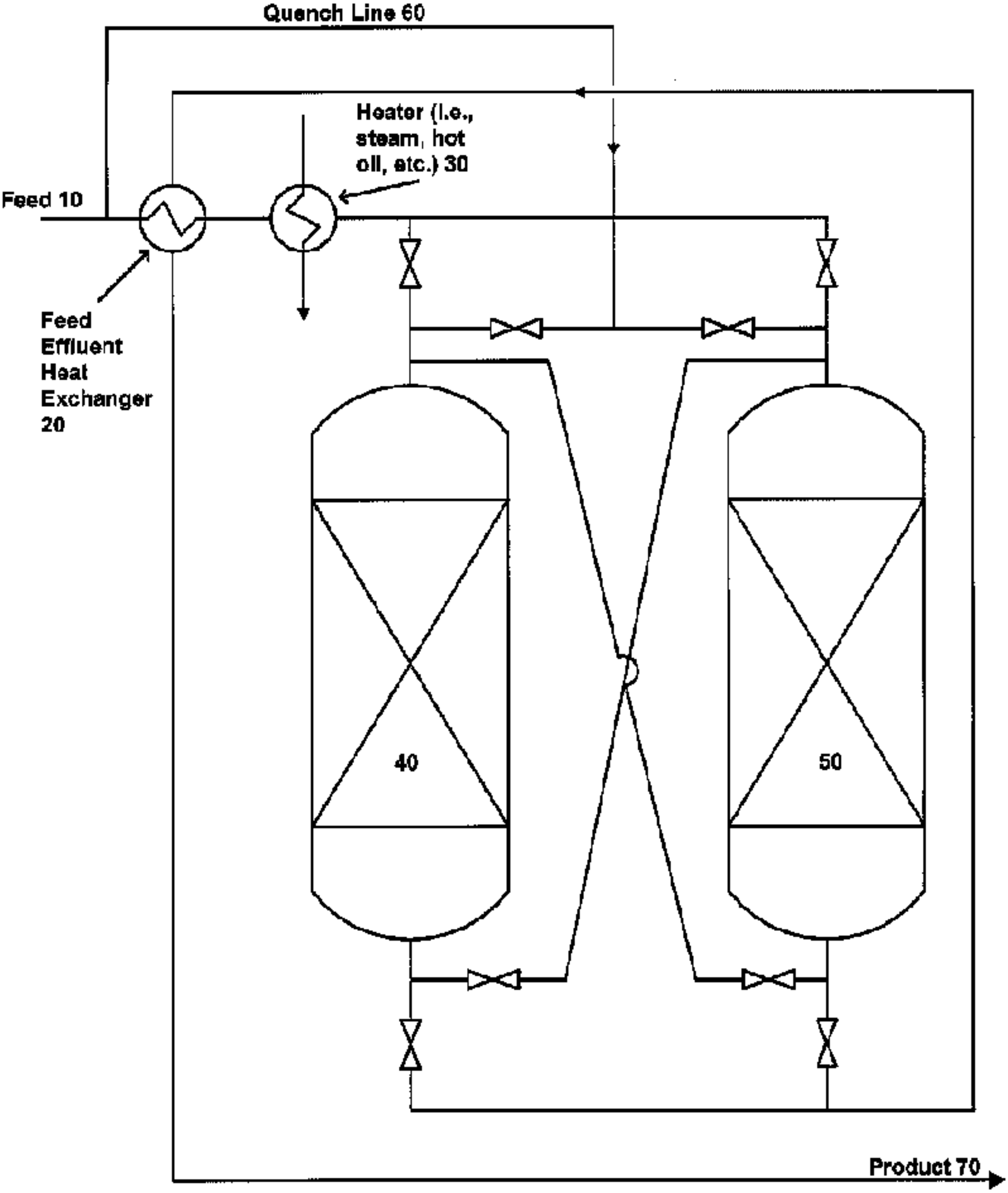

Product 70
40
50
Feed Effluent
Heat Exchanger
20
Feed 10
Heater (i.e.,
steam, hot
oil, etc.) 30
Quench Line 60

FIG. 2


Quench Line 60
Heater (i.e., steam, hot oil, etc.) 30
Feed 10
Feed Effluent Heat Exchanger 20
40
50
Product 70