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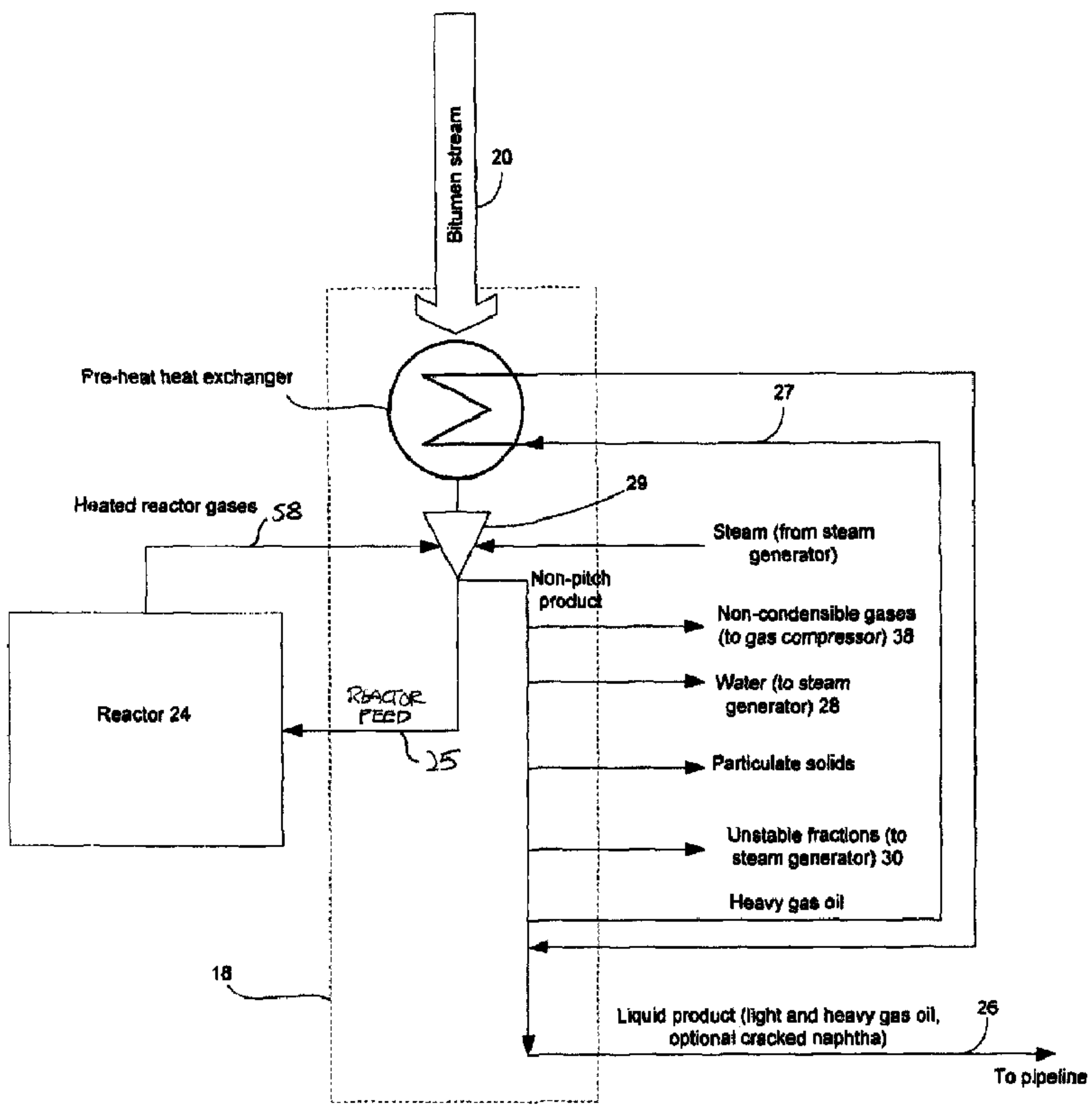
(54) **Titre : PROCEDE POUR LA VALORISATION DE PRODUITS DE PETROLE LOURD ET DE BITUME**
(54) **Title: PROCESS FOR UPGRADING HEAVY OIL AND BITUMEN PRODUCTS**



(57) **Abrégé/Abstract:**
A process for upgrading bitumen recovered from an oil reservoir without hydrogen production is par-ticularly useful in field upgrading applications. In this pro-cess, recovered bitumen enters a fractionator and is con-tacted with heated gases from a

(57) **Abrégé(suite)/Abstract(continued):**

fluidized bed reactor. The bitumen and heated gases are fractionated into segregated products including at least a liquid pitch, unstable frac-tions, and an upgraded liquid product. The liquid pitch is introduced into the reactor to produce a vapor phase liquid product; the reactor comprises solid particles moving through the reactor and a fluidizing gas fluidizing the solid particles at a conversion temperature which is suitable for facilitating the conversion of at least some of the liquid pitch into the vapor phase liquid product. The heated gases comprising the vapor phase liquid product and fluidizing gas are directed from the reactor to the fractionator to con-tact the bitumen stream. In this process, enough of the seg-regated unstable fractions are burned that the liquid prod-uct and any remaining unstable fractions meets pipeline specifications without hydrogen treatment of any of the re-maining unstable fractions.

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(54) **Title**: PROCESS FOR UPGRADING HEAVY OIL AND BITUMEN PRODUCTS

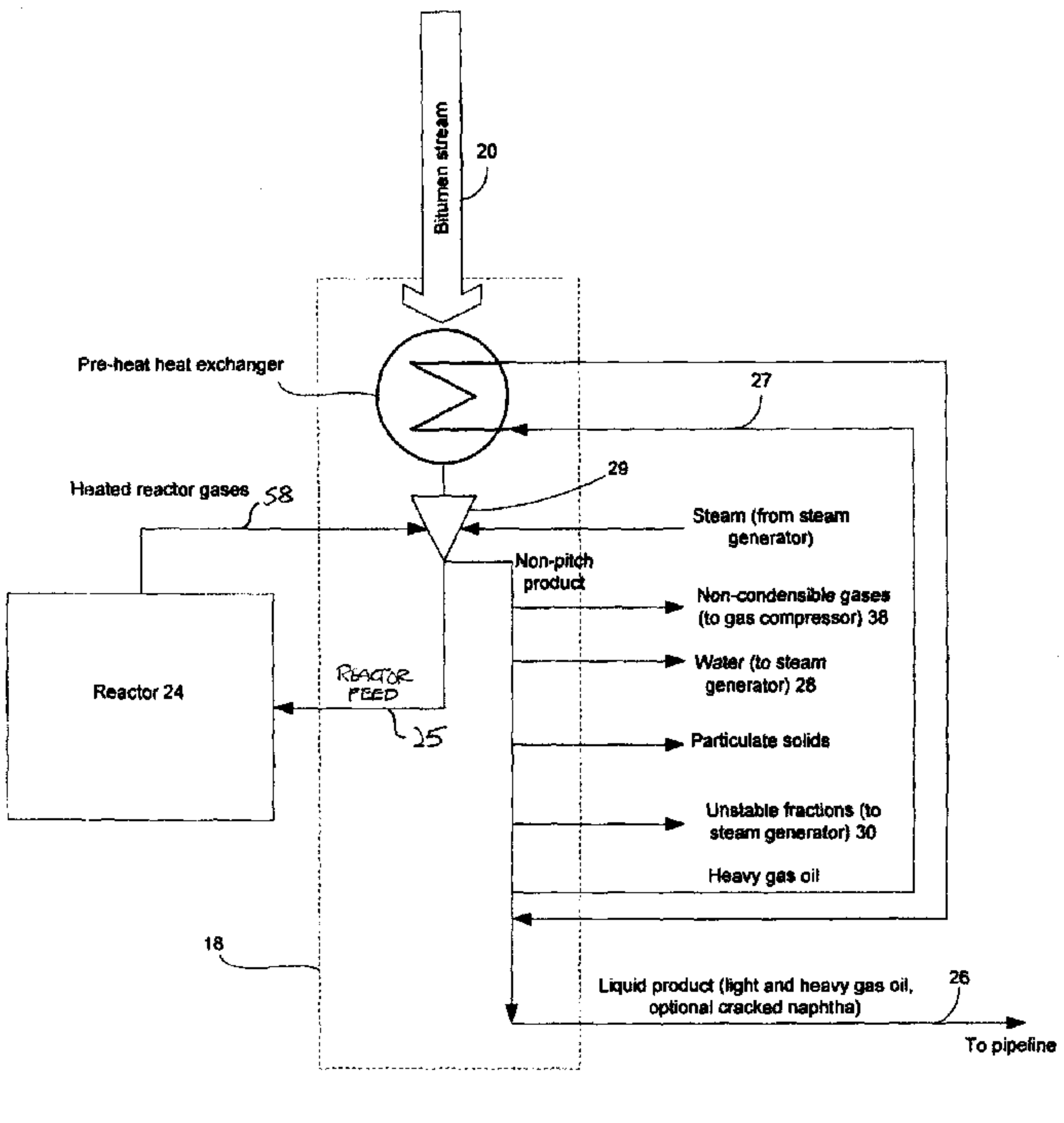

FIGURE 3

(57) **Abstract**: A process for upgrading bitumen recovered from an oil reservoir without hydrogen production is particularly useful in field upgrading applications. In this process, recovered bitumen enters a fractionator and is contacted with heated gases from a fluidized bed reactor. The bitumen and heated gases are fractionated into segregated products including at least a liquid pitch, unstable fractions, and an upgraded liquid product. The liquid pitch is introduced into the reactor to produce a vapor phase liquid product; the reactor comprises solid particles moving through the reactor and a fluidizing gas fluidizing the solid particles at a conversion temperature which is suitable for facilitating the conversion of at least some of the liquid pitch into the vapor phase liquid product. The heated gases comprising the vapor phase liquid product and fluidizing gas are directed from the reactor to the fractionator to contact the bitumen stream. In this process, enough of the segregated unstable fractions are burned that the liquid product and any remaining unstable fractions meets pipeline specifications without hydrogen treatment of any of the remaining unstable fractions.

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Process for Upgrading Heavy Oil and Bitumen Products

Field of the Invention

This invention relates generally to oil processing, and in particular, to a system for upgrading heavy oil and bitumen products.

Background

As conventional access to crude oil declines, more emphasis has been placed upon devising ways to economically exploit the abundant ultra-heavy oil (also known as “bitumen”) reserves present most notably in Venezuela, Canada and the United States. Depending upon the reserve, some of this oil is not recoverable by conventional means, as the oil will not flow at the ambient temperature. In Canada, and specifically in Alberta the majority of bitumen is present as a semi-homogenous mixture of solid hydrocarbon and inorganic sand and clay, referred to as “oil sand”. The recovery of bitumen in Canada introduces an additional challenge, as the heavy oil / bitumen must first be recovered from the sand aggregate prior to introduction into upgrading units. One popular method of oil recovery from oil sands is thermal recovery, which involves in situ heating of the oil/sand aggregate, often using steam as the heating medium. The thermal energy in the steam liquefies the heavy oil / bitumen, which can then be collected and pumped to the surface. Examples of thermal recovery processes include steam assisted gravity drainage (SAGD) and cyclic steam stimulation (CSS).

Additional processing is required for bitumen and heavy oils before they can be introduced into refining infrastructure for light crude oil; such processing is known as “upgrading”. The degree of upgrading depends upon how far the oil to be processed (“feedstock”) deviates from light oil, when compared using standard refining metrics. In a conventional upgrading system for converting a heavy, low quality material into a conventional light oil analog, the feedstock is usually introduced into a plant where it is usually separated into a pitch and non-pitch fraction. "Pitch" as generally understood in the industry means the fraction of the oil boiling above approximately 975°F, as measured by the standard ASTM method. This physical separation does not introduce any chemical changes to the molecules in the oil, but rather separates any higher quality oil from the heavier, low quality fraction. The heavier pitch fraction, representing typically 30-50% by weight of the feed mixture is then introduced into a primary upgrading (PUG) facility where it is subjected to conditions where the large molecules "crack" into smaller ones, resulting in a liquid with a lower boiling point than the starting material. Typically, a boiling point of less than 975°F is targeted for the product liquid, based on producing an acceptable feedstock for conventional downstream refining equipment. Significant sulfur is released from the oil as part of this processing step. Depending upon the technology employed in the PUG facility, elemental hydrogen may also be introduced to the oil to remove nitrogen and any remaining sulfur, and to increase the hydrogen content of the oil. In addition to producing a hydrocarbon stream that is liquid at ambient conditions (the "liquid products"), the PUG facility will also produce a non-condensable, sour, gaseous hydrocarbon stream (the "gas"), and a hydrogen-deficient solid byproduct which is often coke. If a catalytic process is employed in the PUG facility, a purge of catalyst will also be required, upon which some of the coke resides.

The liquid products produced by the PUG facility are then subjected to secondary upgrading (SUG) in a SUG facility. In this facility, elemental hydrogen is added catalytically to the liquid products to increase the hydrogen content of the hydrocarbon therein, and sulfur, nitrogen and metals are removed from the liquid products. Typically, the SUG facility utilizes fixed-bed catalytic reactors.

A significant amount of infrastructure is required to support the PUG and SUG facilities in prior art systems. For example, steam methane reforming (SMR), gasification or other hydrogen generation means must be provided to generate the hydrogen required by both the SUG and possibly the PUG facilities.

The conventional upgrading system is energy and resource intensive, complex, and expensive to set up and maintain. Reasons include: the remote and localized nature of oil sand reservoirs result in expensive labour costs; the use of expensive diluent (a low molecular weight hydrocarbon) to assist in the separation of the heavy oil from the mixture of bitumen and water that is initially recovered at surface by the thermal recovery process; and, SAGD, CSS and other thermal recovery processes are very energy intensive, requiring high pressure steam that is typically produced by combusting natural gas.

Both hydrogen addition and coking processes have been incorporated into known systems for the processing of heavy oil and bitumen. The benefits derived from the PUG processes based on hydrogen addition are derived at the expense of significant incremental capital and operating expense. Many of these costs are the result of incorporating hydrogen production equipment and processes, expensive and complex bitumen/heavy oil conversion reactors in hydrogen service, as well as costs associated with catalyst and incremental feedstock for hydrogen production. The economic penalties applied to realize the benefits of SUG are similar to those associated with PUG.

Because of these high expenses, conventional upgrading (whether full or partial upgrading) systems are economic only at higher production outputs, typically in the order of above ~60,000 bbl/d. This limits the degree of integration that can be achieved with thermal production, which typically boasts ~20,000- 30,000 bbl/d of production. “Field upgrading” is a concept used to refer to upgrading on a relatively small scale, usually constructed adjacent to the SAGD or other production facility. This is of particular interest to the many small scale bitumen/heavy oil producers operating a single SAGD facility, or “pod”. To date there have been no commercial applications of the field upgrading concept, as no system utilizing PUG technology has proven economic at such smaller scales.

Summary Of Invention

It is an objective of the invention to provide a solution to at least some of the deficiencies in the prior art.

One particular objective of the invention is to provide an improved system for upgrading heavy oil and bitumen.

According to one aspect of the invention, there is provided a process for upgrading bitumen recovered from an oil reservoir without hydrogen treatment. This process comprises the following steps:

(a) in a fractionator, contact the recovered bitumen with heated gases from a fluidized bed reactor and fractionate the bitumen and heated gases into segregated products including at least a liquid pitch, unstable fractions, and an upgraded liquid product;

(b) introduce the liquid pitch into the reactor to produce a vapor phase liquid product, the reactor comprising solid particles moving through the reactor and a fluidizing gas fluidizing the solid particles at a conversion temperature which is suitable for facilitating the conversion of at least some of the liquid pitch into the vapor phase liquid product;

(c) direct the heated gases comprising the vapor phase liquid product and fluidizing gas from the reactor to the fractionator to contact the bitumen stream; and

(d) burning enough of the segregated unstable fractions that the liquid product and any remaining unstable fractions meets pipeline specifications without hydrogen treatment of any of the remaining unstable fractions.

During the step of fractionating the bitumen and heated gases, there can also produced a non-condensable gas at least some of which is used as fluidization gas in the reactor. The reactor can also produce coke when the liquid pitch is converted into the vapor phase liquid product, in which case the process further comprises using at least some of the coke to generate steam for use in recovering the bitumen.

During the step of fractionating the bitumen and heated gases, there can also be produced cracked naphtha, in which case the process further comprises using at least some of the cracked naphtha to generate steam for use in recovering the bitumen. At least some of the cracked naphtha can be recovered as liquid naphtha for use as an upgraded product. The amount of cracked naphtha included in the liquid product depends upon the fraction of this liquid that is unstable, and the ability of the resulting mixture to meet pipeline specifications.

The bitumen can be recovered by steam assisted gravity drainage or cyclic steam stimulation or other known steam recovery techniques. The steam used for such recovery techniques can be generated by a circulating fluidized bed steam generator and clean up facility.

In this process, at least some of the coke can be partially oxidizing to generate heat; in such case, flue gas generated as a result of the partial oxidation is directed to the circulating fluidized bed steam generator and clean up facility. Sulfur is removed from the combustion gas generated in the process through contact with lime.

Also, at least some of the coke partially oxidized to generate heat can be used to heat one or more of the solid particles, fluidizing gas, and low grade steam.

The reactor output and bitumen input can be selected such that the amount of coke and naphtha produced is sufficient to meet all the energy requirements of the circulating fluidized bed steam generator and clean up facility. The reactor output can be selected such that the amount of coke and naphtha produced is also or additionally sufficient to meet all the energy requirements for sufficiently heating the solid particles and fluidizing gas for use in the reactor.

In the fractionator, the heated gases can be contacted with the bitumen such that the boiling temperature of volatile material in the bitumen is reduced, thereby enabling fractionation without use of atmospheric and vacuum columns, which are elements of a traditional upgrading flowsheet.

The process and system according to the above aspects of the invention provides advantages over prior art systems and processes by reducing the capital scope beyond eliminating the need for hydrogen generation, and providing additional benefits as will be described.

Brief Description of Drawings

Figure 1 is a flowsheet of a system for producing an upgraded oil product from heavy oil or bitumen according to one embodiment of the invention.

Figure 2 is a schematic view of a cross-flow fluid bed reactor used in the system of Figure 1.

Figure 3 is a schematic view of the fractionation process used in the system in Figure 1.

Detailed Description of Embodiments of the Invention

Any terms not directly defined in this description shall be understood to have the meanings commonly associated with them as understood within the art of the invention.

According to one embodiment of the invention and as shown in Figures 1 to 3, a system 10 is provided which upgrades heavy oil and bitumen without using hydrogen injection. As this system 10 does not incorporate the equipment, processes, and materials associated with hydrogen injection, this system 10 can be economically deployed for smaller scale "field upgrading" applications, wherein the upgrader feed rate is approximately equal to the production rate of a single pod oil sands thermal recovery facility such as SAGD or CSS, which is typically in the order of 20,000 - 30,000 bbl/day of a heavy oil/ bitumen feedstock. The system is located at an oil sands reservoir and is used to extract and upgrade bitumen into an intermediate product which meets pipeline specifications, and may also meet refinery specifications for refining by a light crude oil refinery (not shown). In this embodiment, the system 10 is designed for operation at an oil sands reservoir in Alberta, Canada in which bitumen is recovered using SAGD techniques, and details of the operating parameters, inputs and outputs of components in the system 10 are provided for this specific application; however, it is to be understood that such disclosed operating parameters, inputs and outputs are provided merely to illustrate one specific application of the system 10 and that different operating parameters, inputs and outputs can be specified depending on the particular application of the system 10.

Referring now to Figure 1, bitumen is produced by SAGD in a bitumen production facility 12. Bitumen is comprised of a mixture of virgin heavy gas oil and pitch. For brevity the term "bitumen" is used in this description to conveniently refer to both bitumen and heavy oil. Steam used by the SAGD process is generated by a steam generator and gas clean-up facility ("steam generator") 14 which is fluidly coupled to the bitumen production facility 12 by steam line 13. In this embodiment, bitumen production is rated at 20,000 bbl/day which can be met by well known SAGD techniques. However, other bitumen recovery techniques such as CSS can be employed within the scope of this invention. In this SAGD application, 874,000 lb/hr steam, saturated at 1,300 psig, is injected into the ground through a vertical injection wellbore (not shown). The injection wellbore changes direction and continues horizontally through the reservoir, where holes in the wellbore permit the steam to escape, creating a heated "steam chamber" around the injector wellbore. The steam provides the energy required to melt the bitumen contained in the oil sand within the steam chamber; the melted bitumen drains by gravity into a collection wellbore (not shown) that runs parallel to the injector wellbore. The bitumen and condensed steam, contaminated with some particulate matter are pumped to the surface. The water and solids are separated from the oil in a gravity settler (not shown). An oil-soluble diluent is added to the mixture at a rate of 5,000 bbl/d prior to entry into the separation vessel, in order to assist in the separation. "Diluent" refers to a light, virgin oil that is used to dilute heavy oil in order to reduce its density and viscosity.

The diluent / bitumen mixture ("diluted bitumen") is then fed to a diluent splitter 16 at a rate of 25,000 bbl/d, where it is heated through indirect heat exchange with light and heavy gas oil from a fractionation apparatus 18, steam from the steam generator 14, and diluent from the diluent splitter 16 to a temperature of 235°C (conduits for these fluids to the diluent splitter 16 are not shown in Figure 1). The diluent splitter 16 includes a fractionator column (not shown) designed to separate the diluent from the bitumen in the diluted bitumen stream in a manner that is known in the art. Steam (9,000 lb/hr at 55 psig) from the steam generator 14 is introduced into the bottom of the fractionator column to assist in the separation (steam supply conduit not shown). The column contains components as known in the art to effect contacting between the vapour and liquid streams within the column. The liquid streams consist almost entirely of hydrocarbon, while the vapour is comprised of water and hydrocarbon. The column's components include an overhead receiver in which condensed steam is separated from the condensed diluent by gravity. The separated liquid diluent (about 5,000 bbl/d) is recycled via a return diluent stream 22 to the bitumen production facility 12 for reuse. The separated bitumen (about 20,000 bbl/d) is fed to the fractionation apparatus 18 as a liquid bitumen stream 20. The condensed steam is returned to the bitumen production facility 12 to be de-oiled (water return line not shown). After de-oiling, the water is returned to a water treatment facility 32 for purification before being transported to the steam generator 14 for conversion to steam.

The fractionation apparatus 18 is comprised of two primary vessels: a scrubber and a fractionator column (both not shown in Figure 1). The two vessels can alternatively be combined into one, as is often done industrially, but are kept separate in this embodiment for convenience and layout considerations. The incoming bitumen stream 20 is preheated to 300°C through indirect contact in conventional heat transfer equipment with a heavy gas oil pump-around loop 27 (see Figure 3) drawn from the bottom of the fractionator column; as will be described below, heavy gas oil is a liquid product from a primary upgrader reactor 24 that is condensed in the fractionation apparatus 18 The warmed bitumen stream 20 is then introduced into the scrubber vessel, where it is distributed onto the top of internal components that are designed to operate in a fouling service, which may be, for instance shed decks (not shown). These internal components are designed to effect contacting of the relatively heavy bitumen stream 20 with heated gases from the primary upgrader reactor 24 that is introduced into the scrubber below the internal components. The contacting will remove particulate solids that are entrained and carried over from the reactor 24. The heated reactor gases (77 MMSCFD) substantially consists of all of the fluidization gas (56 MMSCFD), unreacted vaporized pitch, vaporized cracked naphtha, cracked light gas oil, cracked heavy gas oil, non-condensable gas, water vapour and some suspended coke fines from the reactor 24.

The heated reactor gases are hot and act as a stripping medium, assisting in the separation of pitch from non-pitch content in the bitumen stream 20. The separated pitch materials, in liquid form, along with some gas oil exit the bottom of the scrubber and are introduced as a reactor feed (pitch) stream 25 into the primary upgrader reactor 24 at 350°C. The remaining components of the bitumen stream, combined with the heated gaseous reactor products, form the non-pitch materials and comprise potential liquid products (which are gases in the scrubber). The potential liquid product, along with the non-condensable gas and fluidization gas (81.5 MMSCFD), exit a wash grid (not shown) at the top of the scrubber, and are introduced near the bottom of the fractionator column at 370°C . Figure 3 illustrates the flow of fluids into and out of the fractionator 18.

For the non-pitch material the fractionator column condenses the liquid products and separates them into a number of subfractions based on boiling point. The fractionator column is equipped with standard internal components known in the art for this purpose. Steam (3,300 lb/hr at 55 psig) is fed from the steam generator 14 to the fractionator column to assist in the separation (shown in Figure 3 but not Figure 1), as is common practice. A side stream stripper (not shown) is also incorporated as a means of sharpening the cut point between the gas oil and naphtha cuts. A number of pump-around loops are included in an effort to capture as much of the energy as possible, and to achieve the desired separation, as per conventional fractionator design practices.

The fractionation apparatus 18 produces 7,650 bbl/d of heavy gas oil and 9,650 bbl/d of light gas oil as liquid products, collected as a single liquid product stream 26. The combined product meets pipeline specifications on both density and viscosity metrics, and is discharged from the fractionation apparatus 18 as a liquid products stream 26 for refining. Vapour off the top of the fractionator column is cooled, condensing the steam contained therein into water, which is then recycled via a water conduit 28 to the bitumen production facility 12 for de-oiling. A small portion of naphtha (100 bbl/d) in the pitch free vapour is also condensed, although the majority (>98%) of the cracked naphtha remains vaporized due to the large amount of non-condensable gas in the system 10. Both the vaporized and condensed naphtha streams are routed via cracked naphtha conduit 30 to the steam generator where it is combusted for energy. Since there is no capacity in the system 10 to add hydrogen, the unstable cracked liquid naphtha is not stabilized by hydrogen injection to meet pipeline specifications and is instead combusted on site to produce energy for the upgrading process. The heat contained in the gas oil product leaving the fractionation apparatus 18 is used to preheat the diluted bitumen feed to the diluent splitter 16 by means of conventional heat transfer equipment, and the water to a water treatment apparatus 32. The water treatment apparatus 32 serves to purify water for use by the steam generator 14 (via purified water conduit 34), and receives water for this purpose from the bitumen production facility 12 in the form of deoiled water via line 59, and from a make up water source 36.

The non-condensable gases exiting the fractionation apparatus 18 is routed to a gas compressor 38 via a non-condensable gas conduit 40. The gas compressor 38 operates to increase the pressure of the gas from 5 psig to 50 psig by means of a centrifugal single stage compressor. Non-condensable gas not required for fluidization (7.3 MMSCFD) in the primary upgrader reactor 24 is routed to the steam generator 14 via pressurized gas conduit 42 where it is combusted to produce steam. Gas required for fluidization is supplied to the upgrader reactor 24 via fluidization gas conduit 43. As would be known to one skilled in the art, vaporized naphtha could be recovered from the non-condensable gas with the use of suitable equipment. However, due to the instability of this fraction and the absence of hydrogen addition in the system 10, the capital cost is not justified in the system 10 of this embodiment. Therefore, all of the non-condensable gas and most of the vaporized cracked naphtha in excess of that required for fluidization is combusted to generate steam by the steam generator 14. The balance (56 MMSCFD) is routed to a heater 46 where it is heated to 500°C in tubes inserted into a partial oxidizer (POX) vessel (not shown).

In this embodiment, all of the cracked naphtha is separated from the pipeline-bound liquid product stream 26; in other words, the liquid product is substantially free of unstable fractions. As noted above, the separated unstable fractions (cracked naphtha) can be burned to produce energy for generating steam for the system 10; an additional benefit for separating the unstable fractions from the liquid product is to ensure that the liquid product has sufficient stability to meet pipeline specifications. However, some present pipeline specifications can tolerate liquid product having some amount of unstable fractions; therefore, a lesser amount of liquid naphtha and other unstable fractions can be separated from the product liquid, with the remaining unstable fractions being left in the liquid products for pipeline transport, provided that the liquid product meets pipeline specification. Should the liquid product be transported directly to a refinery, the liquid product may also have to meet refinery specifications. Operation of the fractionation apparatus 18 can be adjusted to change the percentage of unstable fractions that is separated from the liquid product; a bromine test, or equivalent detection methods as known in the art can be used to measure the stability of the liquid product and calculate the minimum amount of unstable fraction that must be fractionated and removed from the liquid product.

The heavy pitch stream 25 from the fractionator apparatus 18, along with some gas oil is fed into the primary upgrading reactor 24 at a rate of 13,700 bbl/d. A primary upgrading reactor suitable for use with the system 10 is disclosed in Applicant's Canadian patent 2,505,632. Referring to Figure 2, the reactor 24 comprises a cross-flow fluidized bed 50 which receives the liquid pitch stream 25. The fluidized bed 50 comprises moving hot solid particles 51 fluidized by the fluidization gas from the fluidization gas conduit 43; the solid particles 51 in the fluidized bed 50 can be coke particles or sand particles and have a bulk horizontal velocity which is generally perpendicular to the vertical upward flow of fluidization gas. The fluidization gas is introduced into the bottom of the reactor 24 at a rate of 56 MMSCFD such that bubbling conditions are achieved in the fluidized bed 50. As noted above, the fluidization gas is comprised of a mixture of non-condensable gas and cracked naphtha, although there may be also small concentrations of vaporized light gas oil and water.

The liquid pitch stream 25 is introduced into the fluidized bed 50 by means of nozzles (not shown). The liquid pitch engulfs the solid particles 51 which move horizontally through the reactor 24. The energy contained in the fluidized solids support the chemical conversion of the pitch into lower boiling hydrocarbon products that continue until all of the feed material has been exhausted. The solid particles 51 drop in temperature as the feed liquid reacts. The cooled solid particles 51 exit the reactor 24 and are transported through cooled solids transfer lines 56 to the heater 46. The cooled solids are heated in the heater 46 and are returned to the reactor 24 via heated solids transfer line 57 to maintain a mean operating temperature of 500°C. The heated reactor gases, containing fluidization gas, unconverted pitch, non-condensable gas and the liquid products that are gaseous at reactor conditions, are passed through a series of cyclones to remove any entrained solids. The mixture of heated reactor gases is then routed to the fractionation column of the fractionator apparatus 18 via conduit 58.

The primary function of the heater 46 is to heat the cooled solid particles 51 back up from 490°C to the temperature required at the reactor inlet conditions to create a mean operating temperature of 500°C. In this embodiment, the heater 46 is a partial oxidizer (POX) vessel (not shown) that partially oxidizes a portion of the coke; alternatively, other heaters known to those skilled in the art that are suitable for heating the solid particles can also be used. The POX vessel is a fluidized vessel in which the coke is partially combusted under oxygen limiting conditions, at a temperature of 650°C. The POX vessel is also used to preheat the fluidization gas to the reactor 24, and to partially meet the site demand for superheating low grade steam (8,750 lb/hr at 55 psig). The POX vessel is equipped with two different sets of heat exchange coils through which fluidization gas and steam are circulated and heated. The heated solid particles 51 are returned from the POX vessel to the reactor 24 via heated solids transfer line 57, while flue gas (66 MMSCFD) resulting from the partial combustion process of the coke is directed via flue gas line 59 into the steam generator 14 for gas cleanup before discharge by flue gas lines 61 through a flare. The coke generated in the reactor that is not consumed in the POX vessel 46 is introduced into an Elutriator vessel (not shown), which separates the solids below a critical size from the larger particles, and returns them to the POX vessel. The balance of the coke (12,000 lb/hr) is routed to the steam generator 14 via coke line 63.

The CFB steam generator 14 has two primary purposes: to produce high quality pressurized steam for multiple applications in the system 10 and to remove sulfur released from the flue gas, coke, naphtha and fuel gas that are combusted in the process. In this embodiment the steam generator 14 produces 901,000 lb/hr at 1300 psig of which 875,000 lb/hr is routed to the SAGD facility and 27,000 lb/hr are routed to the PUG. Of course, the steam generator 14 output can be varied depending on the needs of the system 10. The steam generator 14 comprises a circulating fluid bed boiler (CFB) of a type that as can be obtained from a number of vendors. The CFB is a fluidized bed unit designed to combust a number of fuels in liquid, gaseous or solid form supplied to the steam generator via fuel lines 30, 42, 63 and 65 and the flue gas line (not shown). One particularly suitable fuel is natural gas.

The primary features of the steam generator 14 are:

- limestone (14,500 lb/hr) is introduced into the steam generator 14 via limestone supply line 62 to convert oxidized sulfur into calcium sulfate;

- high pressure steam is produced by circulating treated water through coils located in the fluid bed.;

- heat is produced by combustion of the fuel gas and small amount of condensed naphtha from the fractionation apparatus 18, and the flue gas (66 MMSCFD) and coke from the heater 46. The balance of the energy requirements are met with imported natural gas (19,000 lb/hr) via natural gas supply line 65 which is much less than a conventional steam generator which operates entirely on natural gas;

- particulates are separated from the fuel gas and retained in the system 10 using a series of conventional separation steps, which, depending upon the vendor may include U- beams, cyclones, and bag filters; and

- ash is removed periodically from the CFB.

Most of the steam produced from the steam generator 14 is directed to the bitumen production facility 12 via steam supply line 13, with a small portion used to heat the diluent splitter 16 and preheat the diluted bitumen entering the fractionation apparatus 18 (steam supply line not shown).

A water treatment facility 32 is provided to process water from the bitumen production facility 12, making the water suitable for steam production in the steam generator 14. In this embodiment, falling film evaporator technology is provided for this purpose; however other water treatment technologies suitable for this purpose can be used as is known to those skilled in the art. Falling film evaporator technology is available from a number of vendors and has found utility in SAGD service. In particular, a three-effect evaporation system is used in this embodiment, with a single vapour recompression stage, which provides the energy required for evaporation. The water treatment facility 32 accepts de-oiled water (787,000 lb/hr) from bitumen production facility 12, along with an amount of fresh water makeup (142,000 lb/hr) through line 36. Caustic and scale inhibitor are added in a mixing tank. Air is removed in a de-aerator vessel (not shown), after which the water is introduced into the cascading three effect evaporator system. The purified water generated by the water treatment facility 32 is introduced to the steam generator 14 via purified water supply line 34, while the evaporator condensate is disposed of by deep well injection.

Some notable features of the system 10 include:

- A cross-flow fluid bed primary upgrading (PUG) reactor 24 is used instead of a conventional furnace type delayed coking, or well-mixed fluid bed PUG units. The cross-flow fluid bed PUG reactor 24 generates more liquid products, produces less coke, and retains more of the native hydrogen than conventional coking technologies.

- The coke produced in the cross-flow fluid bed PUG reactor 24 is in a readily-consumable form and can thus be used as fuel to generate steam.

- The volume of fluidization gas required by the cross-flow fluid bed PUG reactor 24 is larger than other fluid bed technologies. When an inert gas is put into contact with a volatile material its boiling point temperature is reduced, a process known as "stripping". The principle of stripping is applied in the system by contacting the fluidization gas with a whole barrel of bitumen, separating the pitch fraction from the non-pitch liquids. This configuration eliminates the need for the traditional atmospheric and vacuum columns, and the associated furnaces, that collectively perform this function;

- A circulating fluidized bed (CFB) steam generator and gas clean up facility 14 is used in the system 10 which is capable of consuming a number of fuels, including coke, hydrocarbon liquids, and fuel gas. This unit 14 is used to produce the high pressure steam required for the thermal production of bitumen.

- The unstable fraction of the naphtha can be consumed in the CFB steam generator and gas clean up facility 14 for energy, increasing the stability of the remaining product liquids while harnessing residual value in the unstable liquids as steam, and reducing the amount of natural gas or other imported fuel required to generate steam.

- Lime as limestone is fed to the CFB steam generator and gas clean up facility 14 to capture the sulfur released by system operation. This enables the capture of sulfur in a single dual purpose unit. Unlike conventional gas cleanup systems, no hydrogen addition is necessary. In the prior art process separate units would be required for each of steam generation, hydrogen sulfide recovery, and two sequential units for sulfur recovery.

- The CFB technology has all of the elements to be considered carbon capture and storage (CCS) ready.

- Water treatment is provided by falling film evaporators, a technology that enables the use of packaged boilers, keeping steam quality high while minimizing costs.

Compared to the conventional upgrader systems, the system 10 shown in Figures 1 to 3 provides the following advantages:

- A significant reduction in the amount of natural gas imported, which reduces one of the largest operating cost drivers. This is achieved by: using combustible byproducts from other components in the system 10 as fuel for the steam generator 14, eliminating the furnaces used in the conventional PUG by using a cross-flow fluid bed PUG 24, eliminating the furnaces in the feed topping facility by using the fractionator apparatus 18, eliminating a separate gas cleanup facility by using a combined steam generator and gas cleanup facility 14, and eliminating a hydrogen production facility.
- A significant reduction in capital scope, which reduces the capital costs. This is realized through elimination of the feed topping facility, the hydrogen generation facility (as no hydrogen treatment is required to stabilize the unstable fractions), the SUG, and integration of gas cleanup with the steam generation facility.

The dramatic reduction in capital and operating costs realized by the present system 10 can be translated into superior project economics, as the savings more than offset any potential discount associated with the resulting upgrader products compared with native light oil.

The advantages introduced by the system 10 justify economic deployment at a much smaller scale, and potentially down to at least 20,000 bbl/d of whole bitumen feed. The application of the current system 10 for processing bitumen from the Peace River region of Alberta, Canada has been described. The bitumen feed consists of 49% (volume basis) pitch. The pitch has an MCR content of 23%. The reservoir basis using SAGD technology is a steam to oil ratio (SOR) of three. The system 10 generates three liquid streams: naphtha (boiling range to 177°C), light gas oil (boiling range from 177°C to 343°C), and heavy gas oil (boiling range from 343°C to 524°C).

The above embodiments have been described by way of example. It will be apparent to persons skilled in the art that a number of variations and modifications can be made to these embodiments.

For example:

- In the embodiment of the invention described above and shown in Figures 1 to 3 a portion of the naphtha fraction is consumed for its energy content, since this fraction is unstable without adding hydrogen, and there is no hydrogen production provided in the system 10. In certain instances the naphtha may have more value as a liquid. This may be the case if, for instance there is a SUG facility in close proximity that can accept the liquid naphtha. Another embodiment of the invention involves installation of additional processing units that will allow for recovery of liquid naphtha from the gas. This equipment is well known in the prior art, and includes such units as Light Ends Recovery (LER), and others.

- In the embodiment of the invention described above and shown in Figures 1 to 3, the alternative fuels produced from the bitumen feed to the reactor 24 may be insufficient to completely meet the energy requirements of the SAGD facility, and in such case natural gas is still required. In another embodiment of the invention the upgrading reactor 24 is increased in size to the point where the alternative fuels generated by the reactor 24 are sufficient to completely meet all of the energy requirements of the system 10, thereby eliminating the need to externally supply natural gas to the system 10. The incremental bitumen required to enable this alternative embodiment is imported into the process, purchased on the open market. The economic benefits of completely eliminating all natural gas requirements, and the incremental liquid products produced from the imported bitumen are achieved with very little incremental capital, since the majority of the equipment does not change in size. This simple change dramatically increases the economics of the process.

- In the embodiment of the invention described above and shown in Figures 1 to 3, alternative fuels are used to offset import natural gas. While the use of alternative fuels is preferred for this embodiment, such use is not necessary. The decision not to consume alternative fuels may be made for environmental reasons. In another embodiment of the invention the energy requirements are met using conventional natural gas fired heating equipment. The solid coke byproduct is stockpiled.

What is clamed is:

1. A process for upgrading bitumen recovered from an oil reservoir without hydrogen treatment comprising:

(a) in a fractionator, contacting the recovered bitumen with heated gases from a fluidized bed reactor and fractionating the bitumen and heated gases into segregated products including at least a liquid pitch, unstable naphtha fraction, and an upgraded liquid product;

(b) introducing the liquid pitch into the reactor to produce a vapor phase liquid product, the reactor comprising solid particles moving through the reactor and a fluidizing gas fluidizing the solid particles at a conversion temperature which is suitable for facilitating the conversion of at least some of the liquid pitch into the vapor phase liquid product;

(c) directing the heated gases comprising the vapor phase liquid product and fluidizing gas from the reactor to the fractionator to contact the bitumen stream; and

(d) burning enough of the segregated unstable naphtha fraction that the upgraded liquid product and any remaining unstable naphtha fraction meets pipeline specifications without hydrogen treatment of any of the remaining unstable naphtha fraction.

2. A process as claimed in claim 1 wherein during the step of fractionating the bitumen and heated gases, there is also produced a non-condensable gas at least some of which is used as the fluidizing gas in the reactor.

3. A process as claimed in claim 1 wherein the reactor produces coke when the liquid pitch is converted into the vapor phase liquid product, and the

process further comprises using at least some of the coke to generate steam for use in recovering the bitumen.

4. A process as claimed in claim 1 wherein in the step of burning unstable fractions, steam is generated for use in recovering the bitumen.

5. A process as claimed in claim 4 wherein the unstable naphtha fraction include cracked naphtha at least some which is recovered as liquid naphtha for use as an upgraded liquid product.

6. A process as claimed in claims 3 or 4 wherein the bitumen is recovered by steam assisted gravity drainage or cyclic steam stimulation.

7. A process as claimed in claim 3 wherein the steam is generated by a circulating fluidized bed steam generator and clean up facility.

8. A process as claimed in claim 7 further comprising partially oxidizing at least some of the coke to generate heat, directing flue gas generated as a result of the partial oxidation to the circulating fluidized bed steam generator and clean up facility, and contacting the flue gas with lime to clean the flue gas.

9. A process as claimed in claim 8 further comprising partially oxidizing at least some of the coke to generate heat, and wherein the heat is used to heat one or more of the solid particles, fluidizing gas, and low grade steam.

10. A process as claimed in claim 9 wherein the reactor output is selected such that the amount of coke or naphtha produced is sufficient to meet all the energy requirements of the circulating fluidized bed steam generator and clean up facility.

11. A process as claimed in claim 10 wherein the reactor output is selected such that the amount of coke or naphtha produced is sufficient to meet all

the energy requirements for sufficiently heating the solid particles and fluidizing gas for use in the reactor.

12. A process as claimed in claim 1 wherein in the fractionator, the heated gases are contacted with the bitumen such that the boiling temperature of volatile material in the bitumen is reduced, thereby enabling fractionation without use of atmospheric and vacuum columns.

13. A process as claimed in claim 1 wherein the oil reservoir is an oil sands reservoir.

14. A process as claimed in claim 1 wherein the reactor is a cross-flow fluidized bed reactor.

15. A process as claimed in claim 1 wherein all of the unstable naphtha fraction is burned that only liquid product remains which meet pipeline specifications.

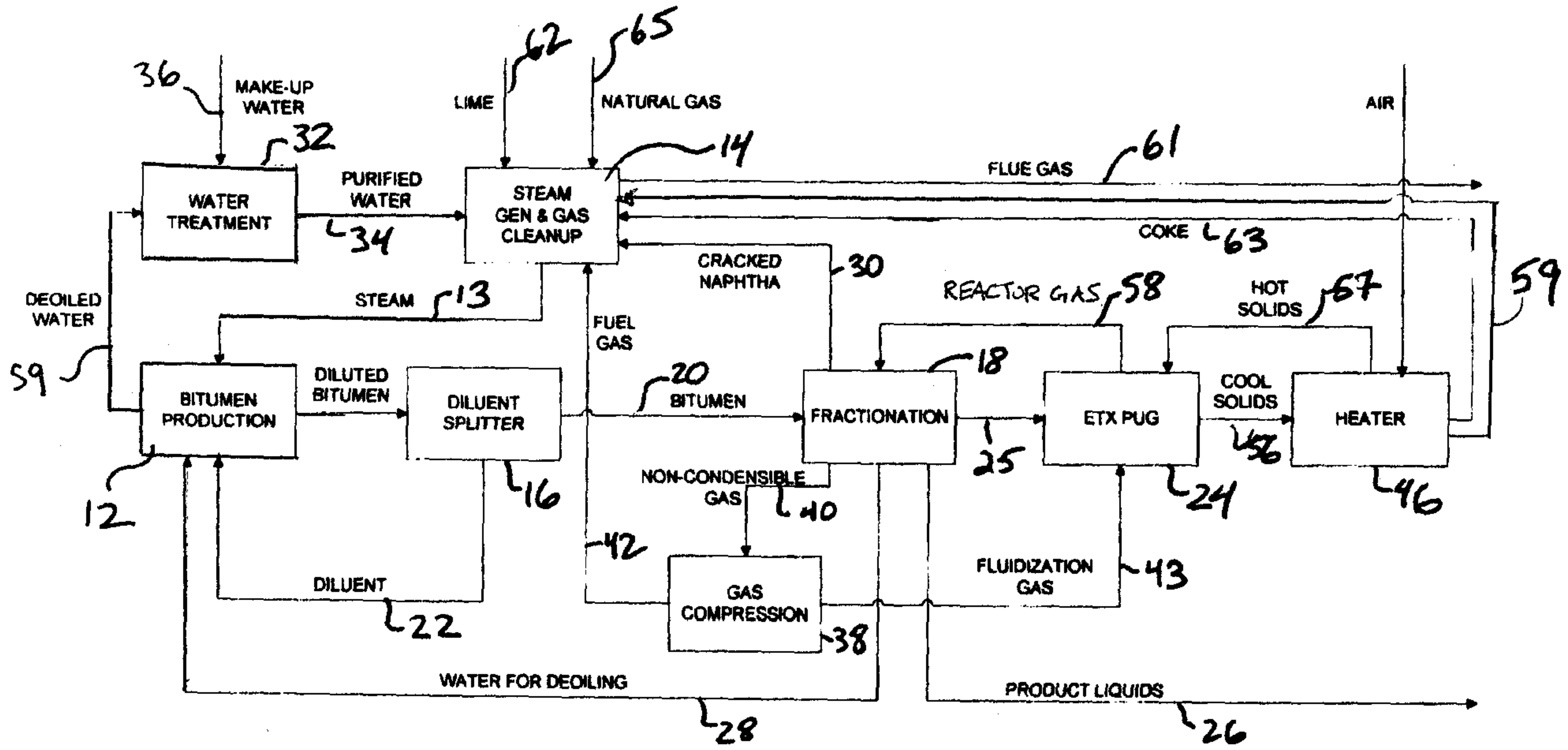


FIGURE 1

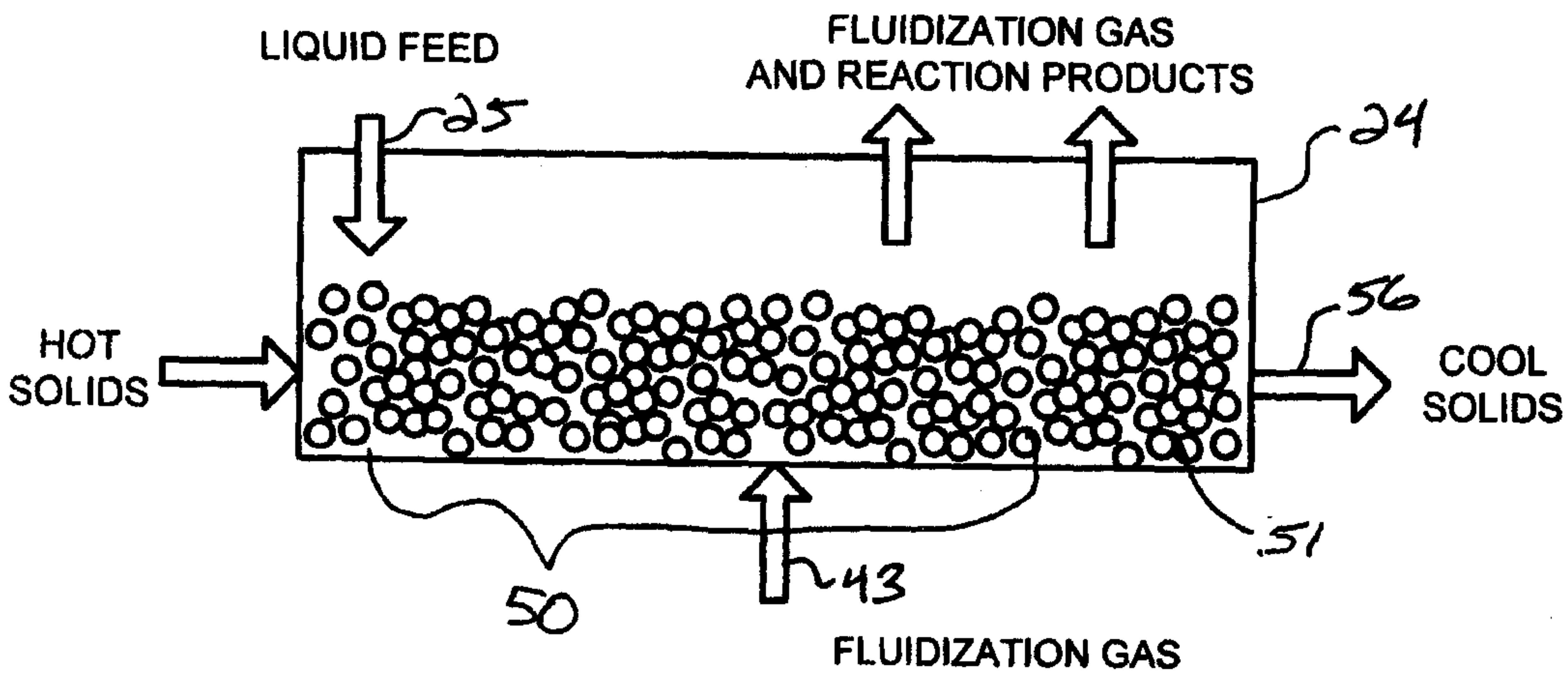


FIGURE 2

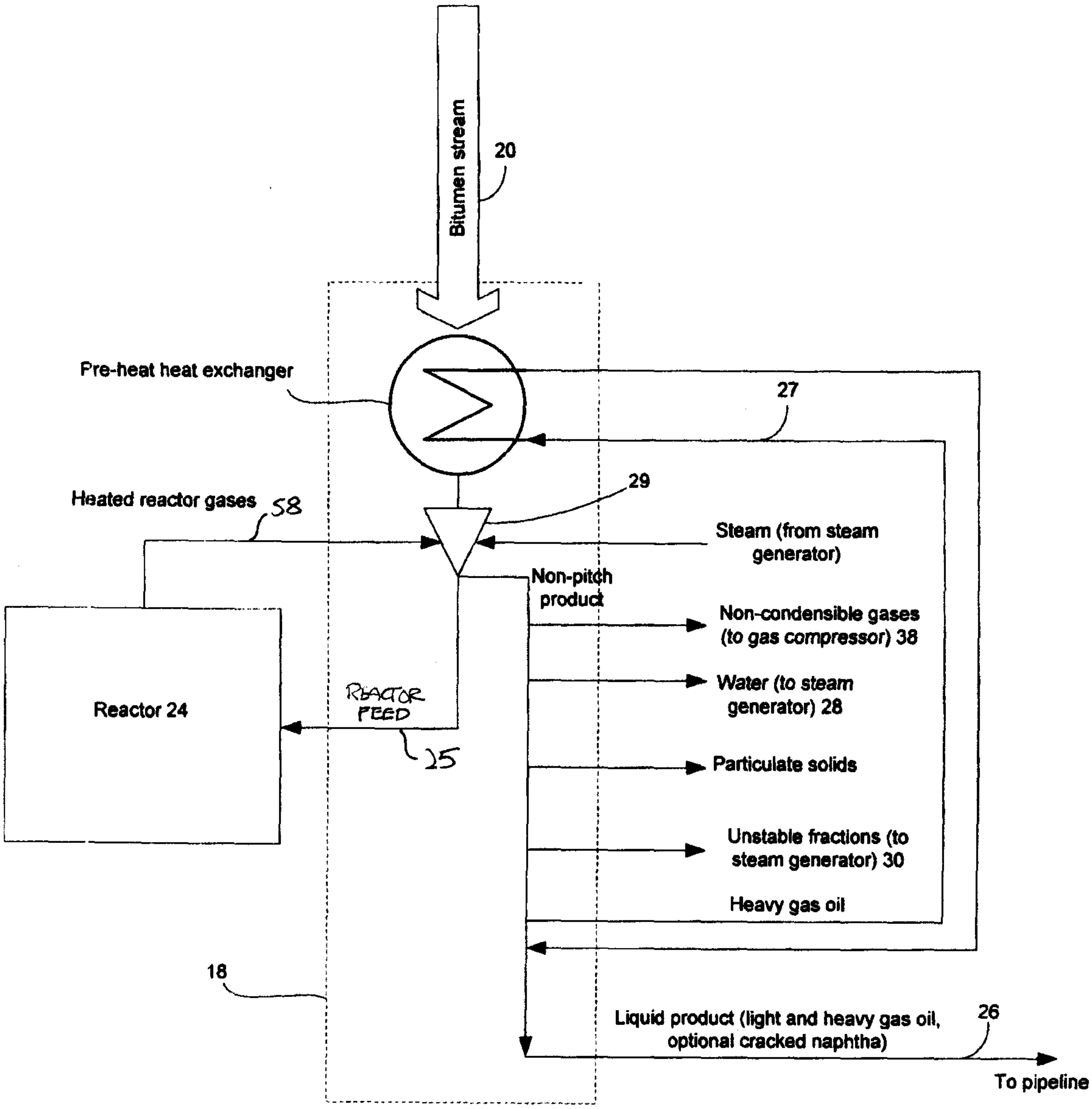
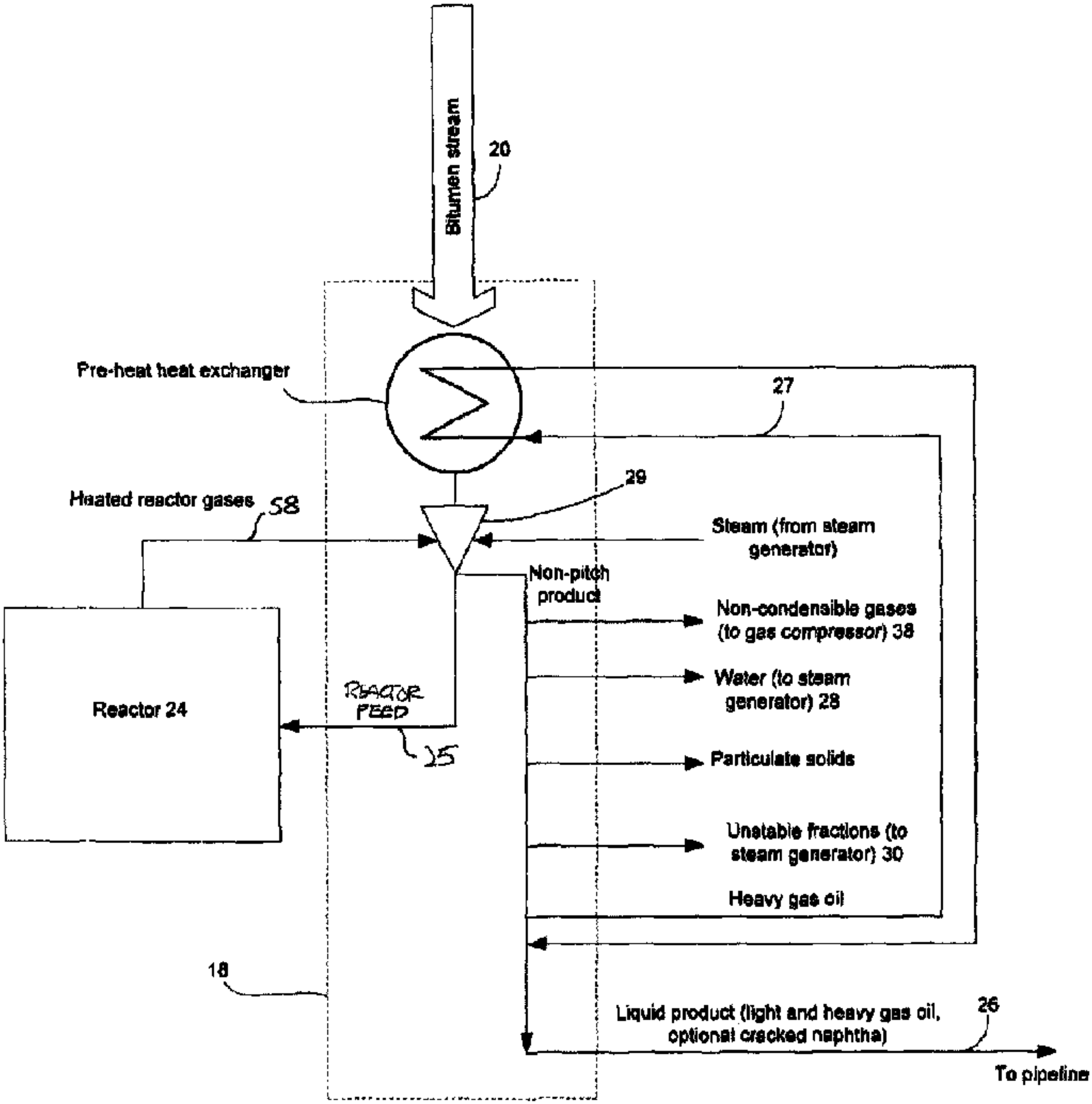

Bitumen stream
20
Pre-heat heat exchanger
27
29
Heated reactor gases
58
Steam (from steam generator)
Non-pitch product
Non-condensible gases (to gas compressor) 38
Water (to steam generator) 28
Reactor 24
REACTOR FEED
25
Particulate solids
Unstable fractions (to steam generator) 30
Heavy gas oil
18
Liquid product (light and heavy gas oil, optional cracked naphtha)
26
To pipeline

FIGURE 3



Bitumen stream
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Steam (from steam generator)
Non-pitch product
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Reactor 24
REACTOR FEED
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Particulate solids
Unstable fractions (to steam generator) 30
Heavy gas oil
18
Liquid product (light and heavy gas oil, optional cracked naphtha)
26
To pipeline