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(54) **PROCESS AND APPARATUS FOR CONVERTING OIL SHALE OR OIL SAND (TAR SAND) TO OIL**

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(57) **ABSTRACT**

Three new embodiments to the Chattanooga Process that convert or upgrade oil bitumen, a combination of oil sands

and bitumen, a combination of sand and bitumen, and oil shale to high grade low sulfur (about 0.1 to 0.5 wt. % sulfur, or less) crude oil. The invention relates to a continuous process for producing synthetic crude oil (SCO) from oil sand bitumen which has been extracted from under ground via in situ processes, or strip mined and extracted via hot water extraction processes before upgrading. It can also apply to kerogen extracted in situ from shale underground. The process involves treating the hot bitumen with sand from an extraction process or the hot oil from kerogen with ground shale containing kerogen in a fluid bed reactor where the reactant and fluidizing medium is only hydrogen. The invention also relates to a continuous process for producing synthetic crude oil (SCO) from oil shale kerogen.

The invention relates to a continuous process for producing synthetic crude oil from oil bearing material, e.g., oil shale or oil sand (tar sand), through continuous process for producing synthetic crude oil from bituminous oil sand (tar sand) or shale. The process includes treating the oil sand (tar sand) or shale to produce a fluidizable feed, feeding the fluidizable feed to a fluidized bed reactor, and fluidizing and reacting the fluidizable feed in the fluidized bed reactor with a feed of hydrogen provided by a feed stream having a stream containing hydrogen in a concentration greater than 90 vol % (90 vol %-100 vol % H<sub>2</sub>).

In one embodiment, the invention relates to a continuous process which can recover methane and ethane from a recycle hydrogen stream. In one embodiment, the process can recover PSA tail gas as feed to a hydrogen plant. In one embodiment, the process can be operated to reduce or eliminate the requirement for externally provided methane feed to the hydrogen plant.



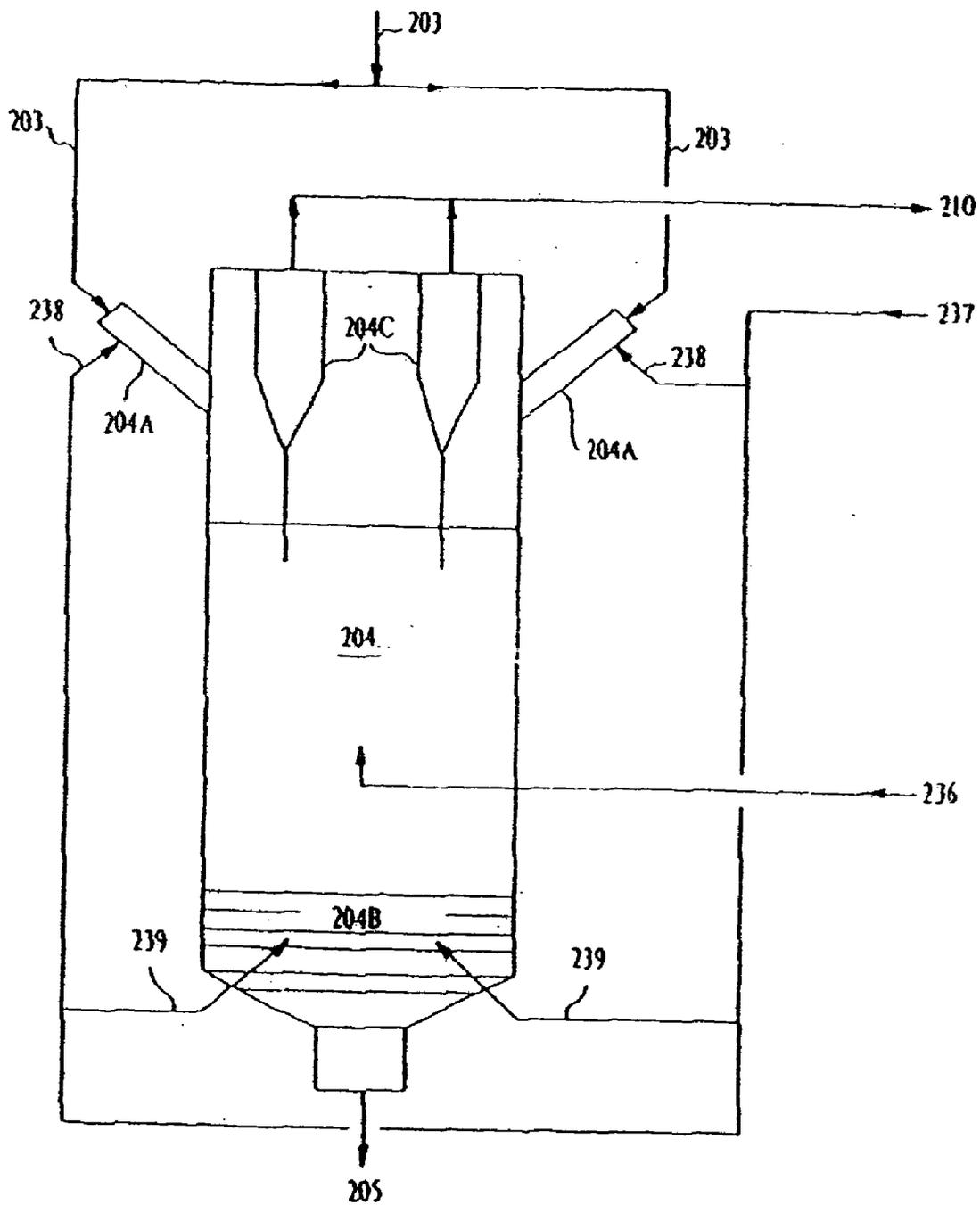


FIG. 2

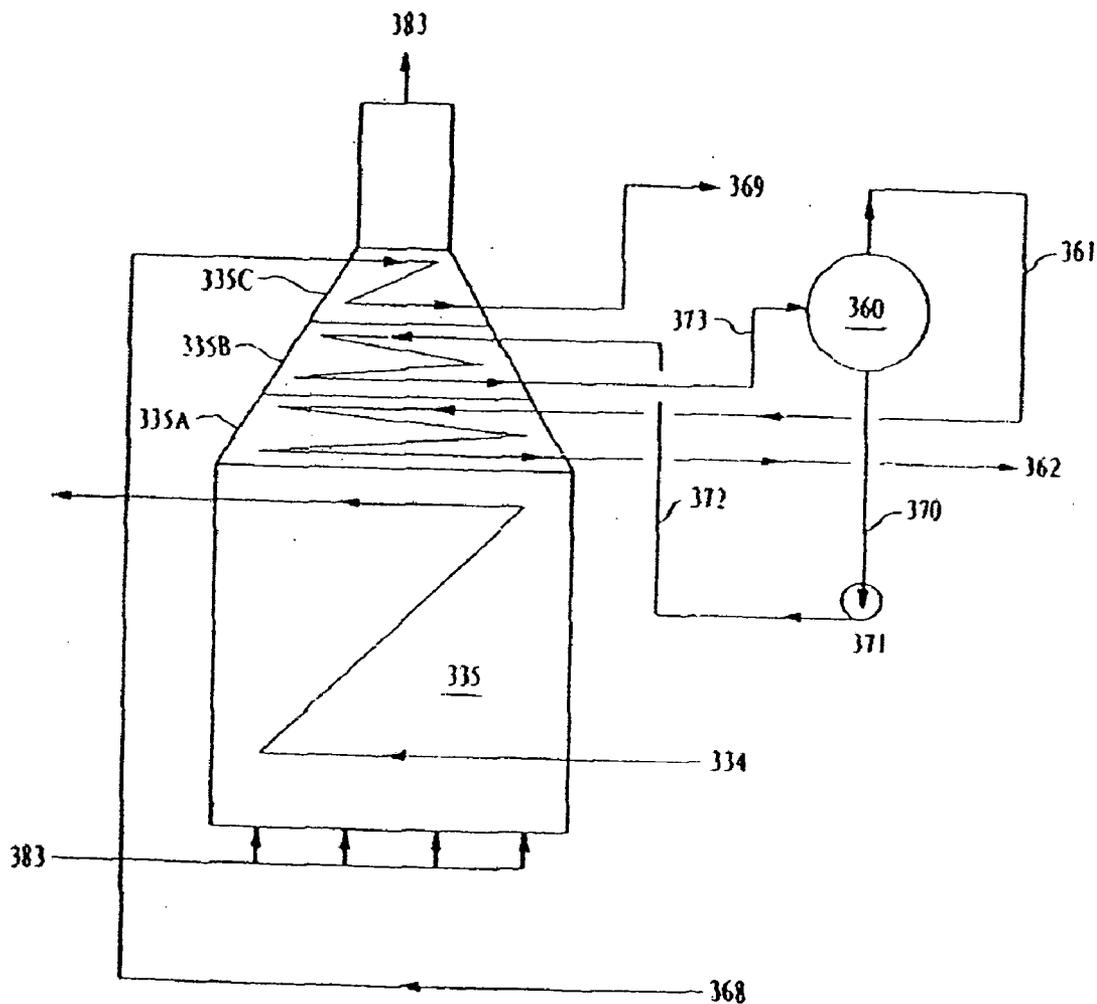


FIG. 3

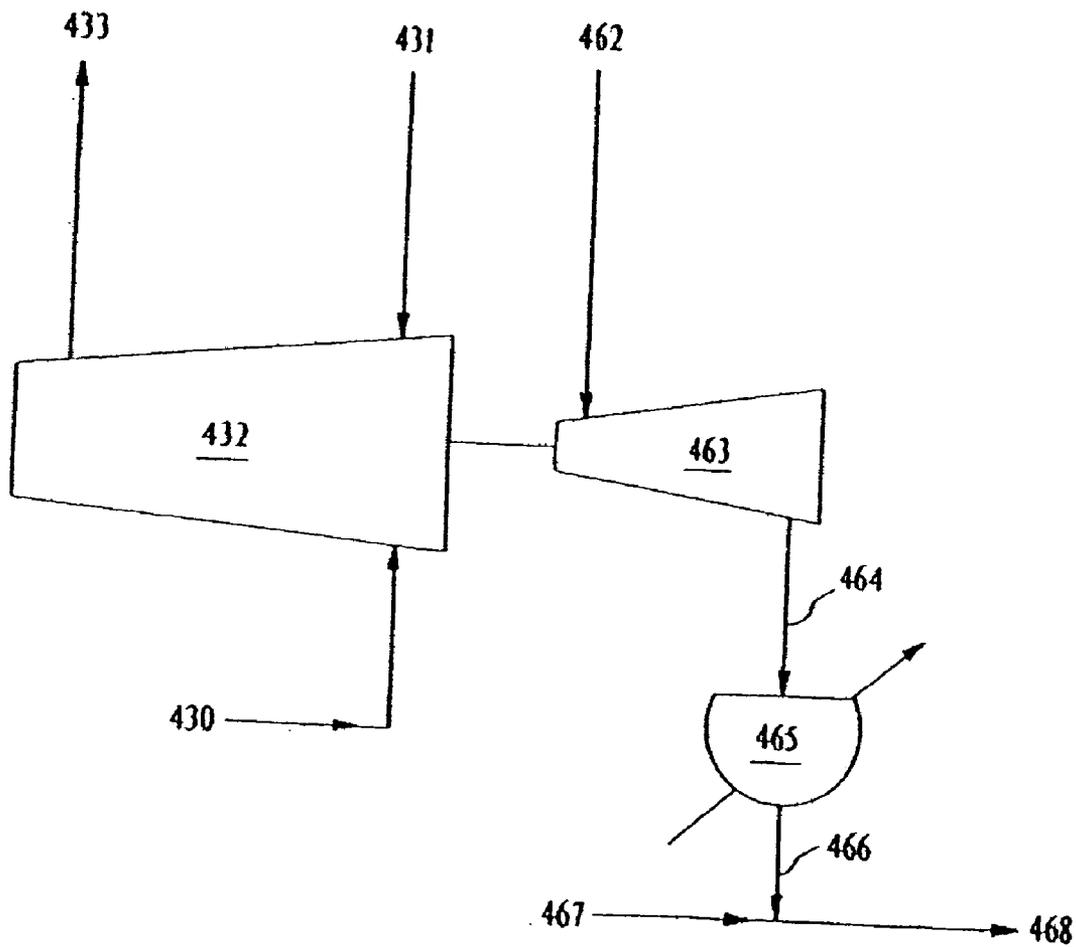


FIG. 4

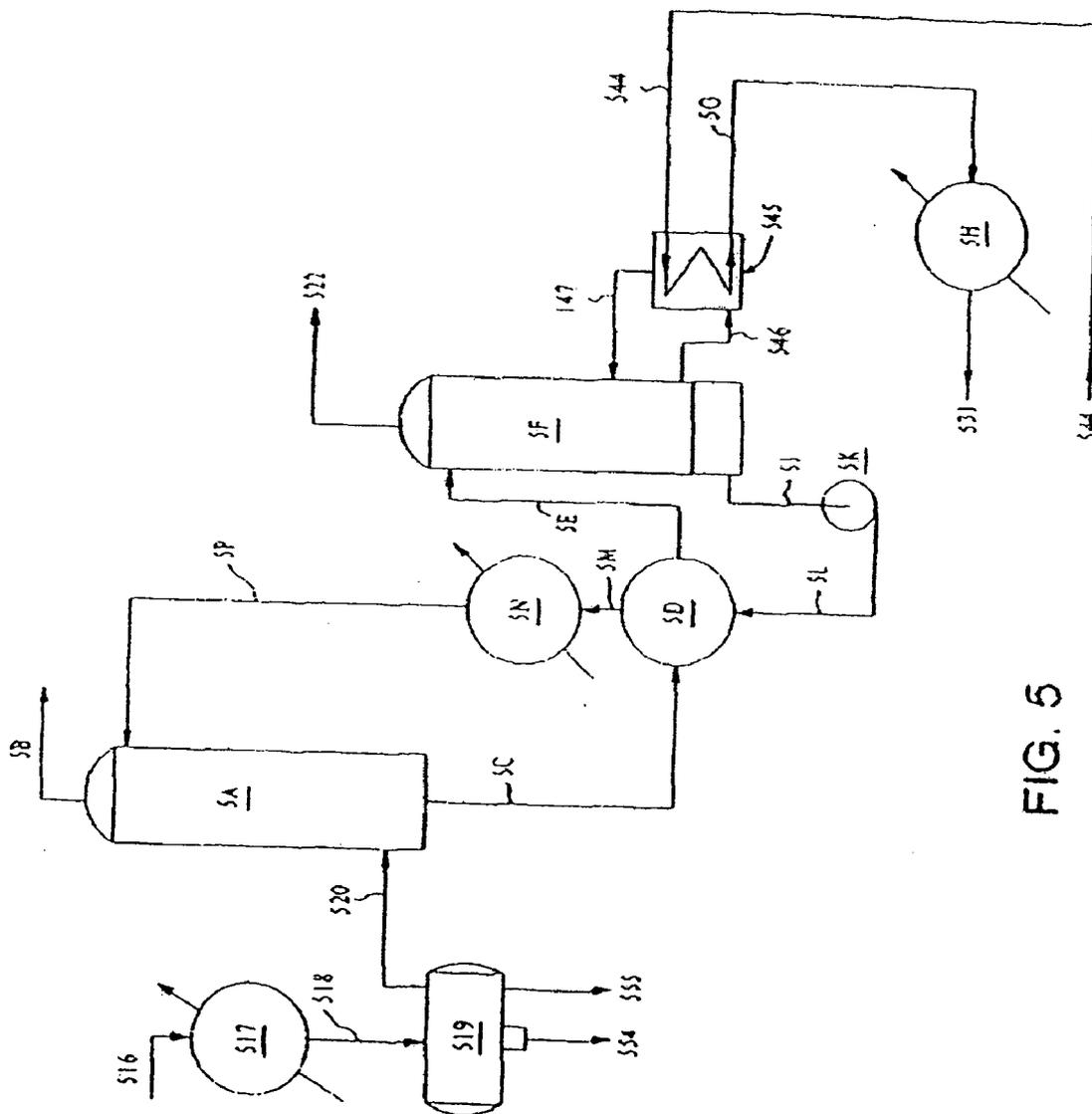


FIG. 5

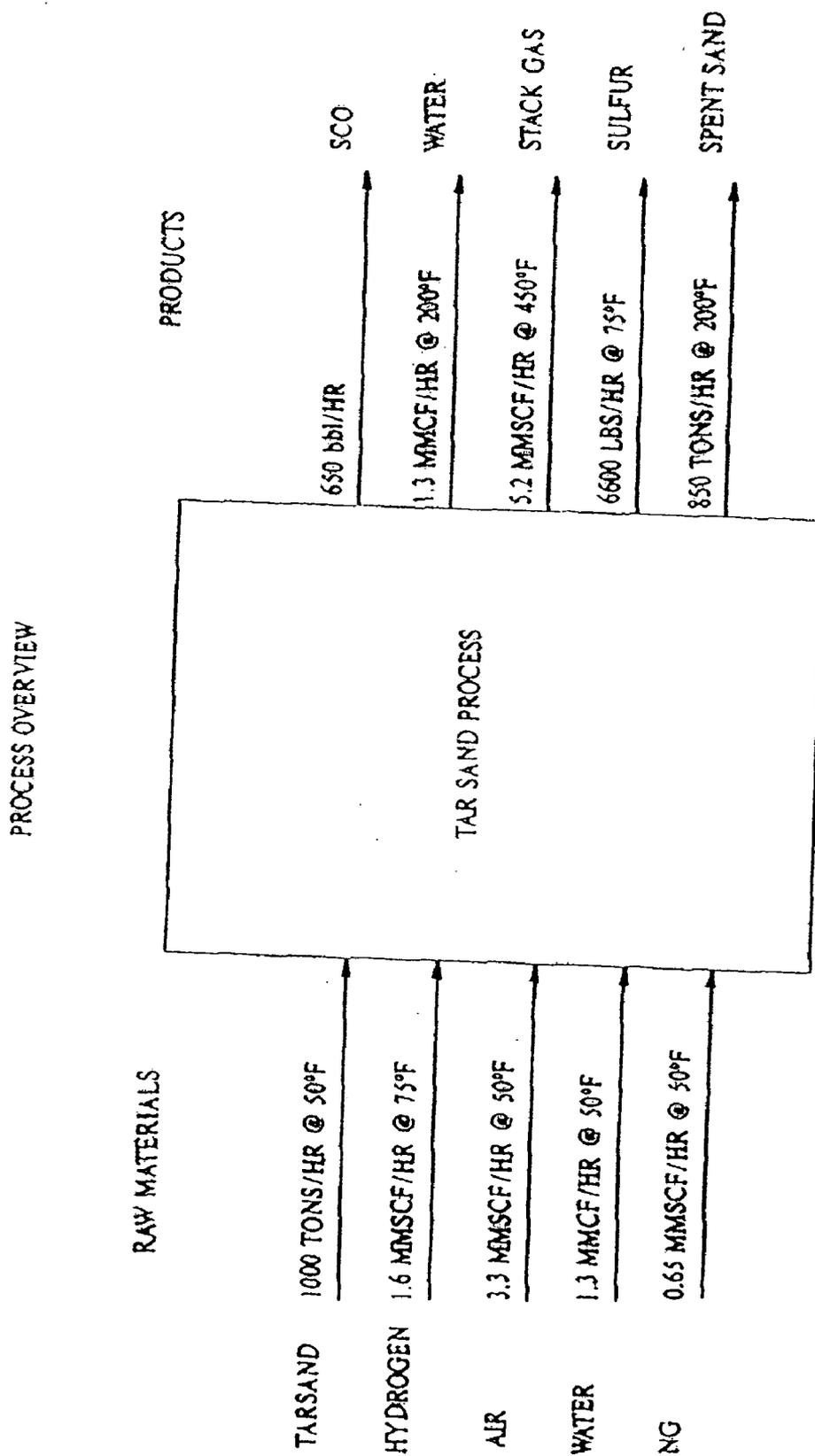


FIG. 6



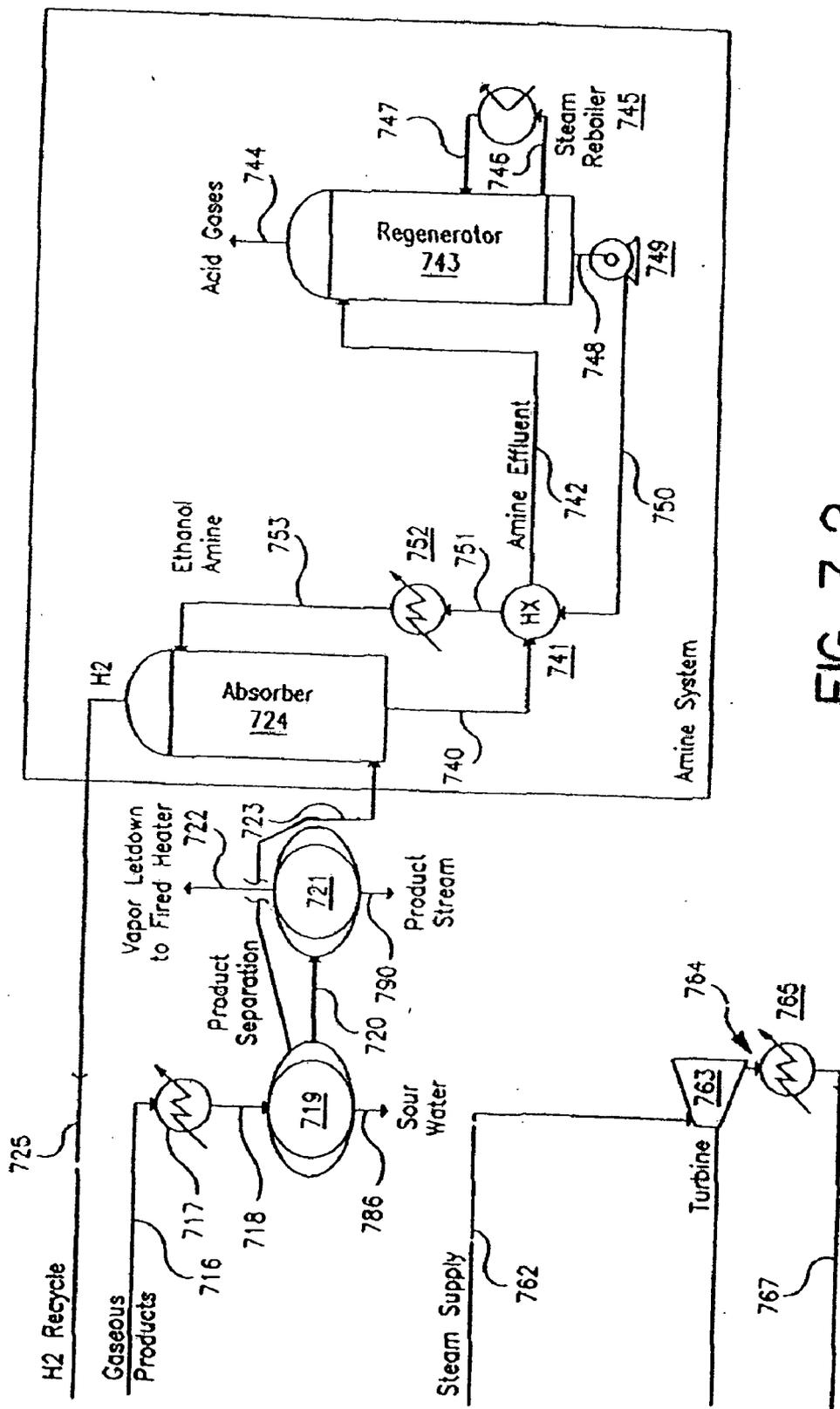


FIG. 7-2

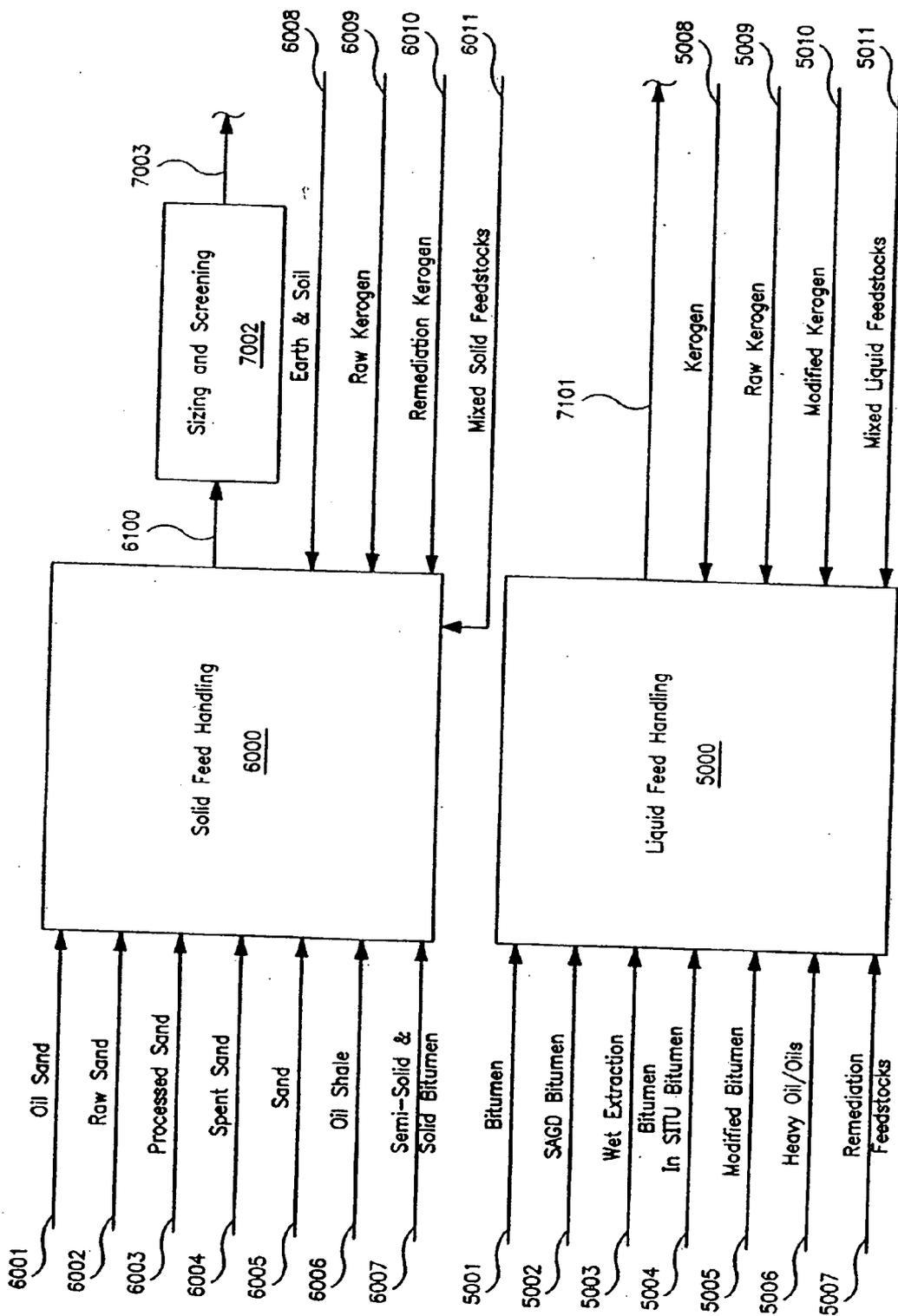


FIG. 7-3

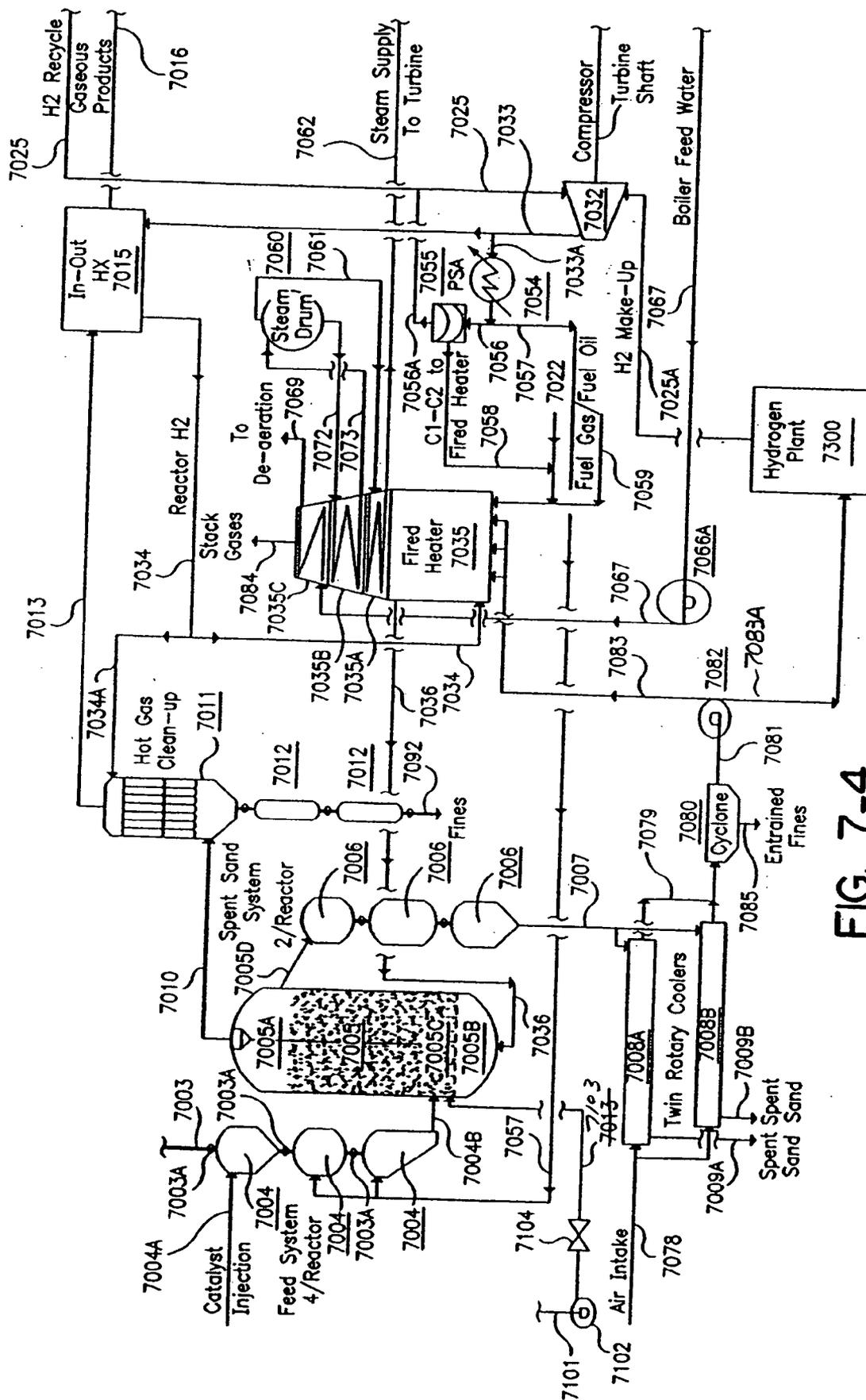


FIG. 7-4



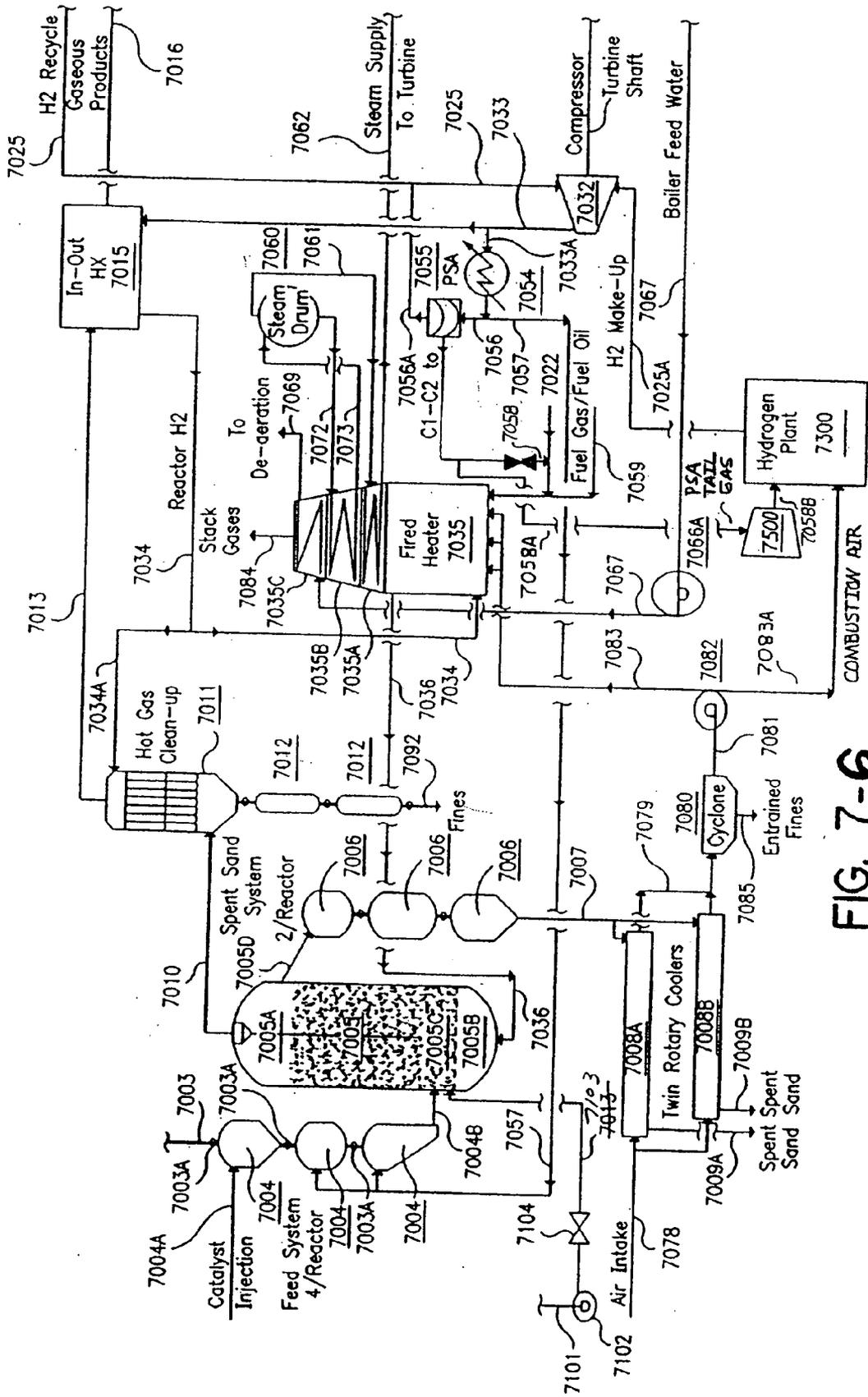


FIG. 7-6

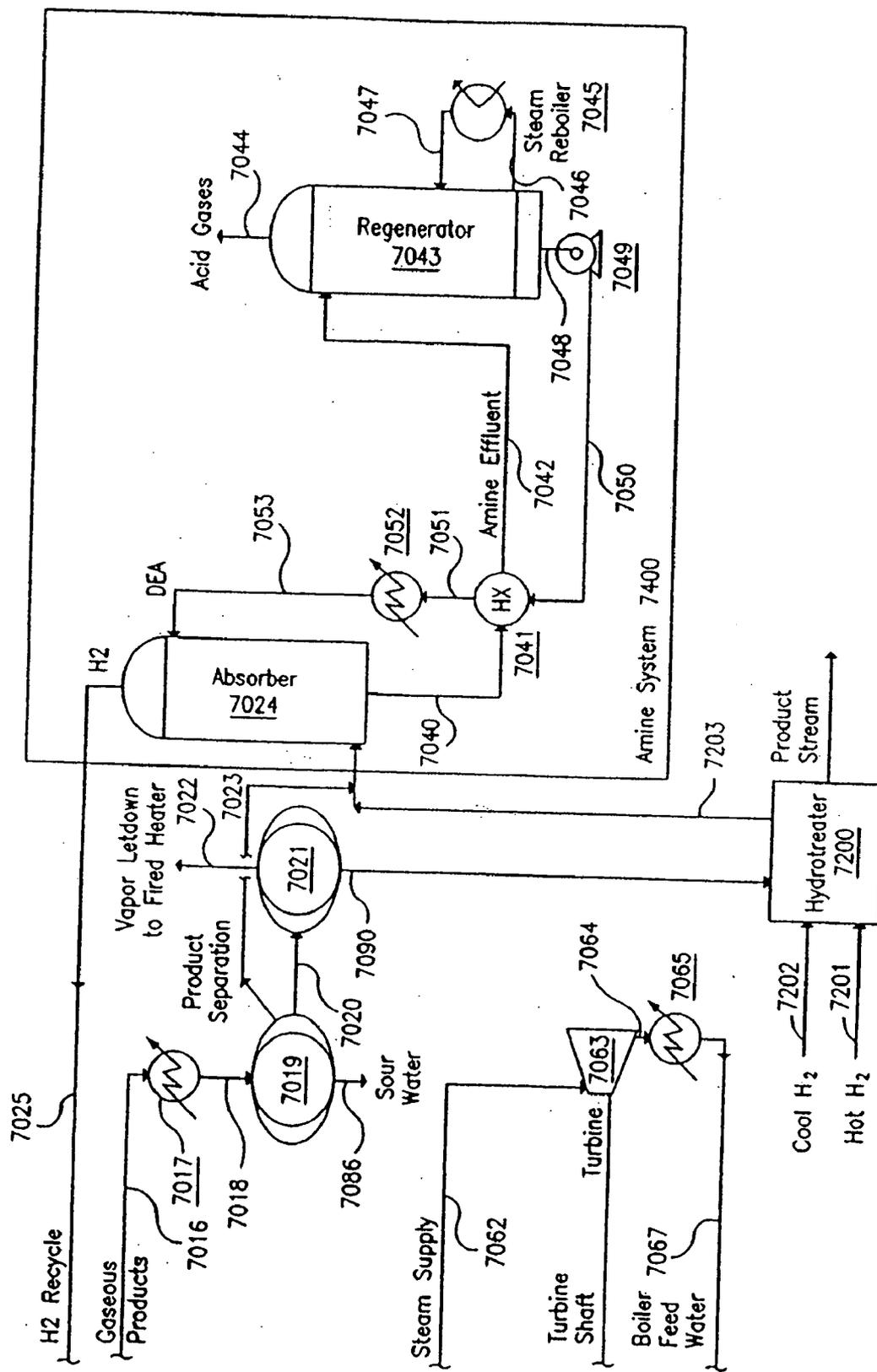


FIG. 7-7

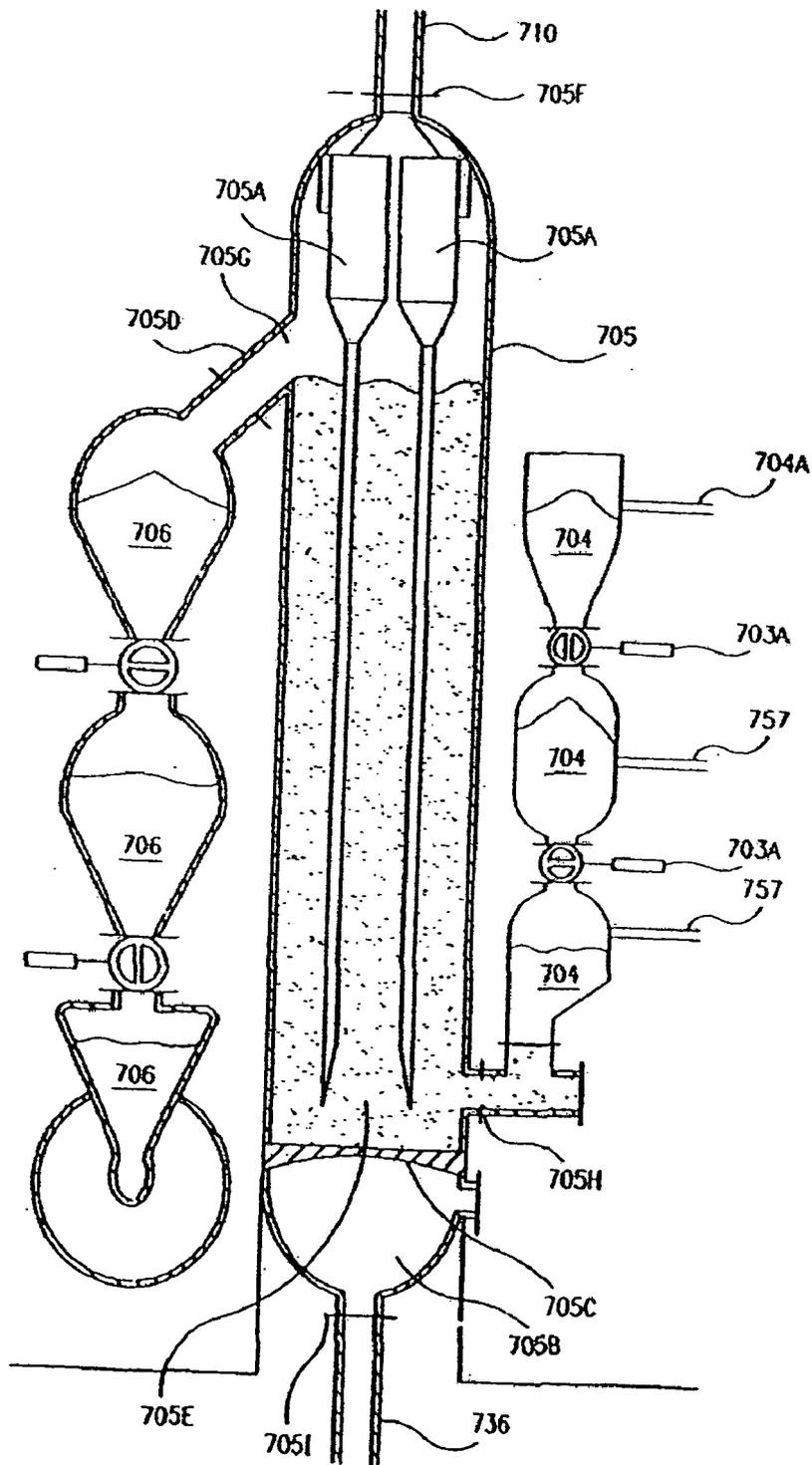
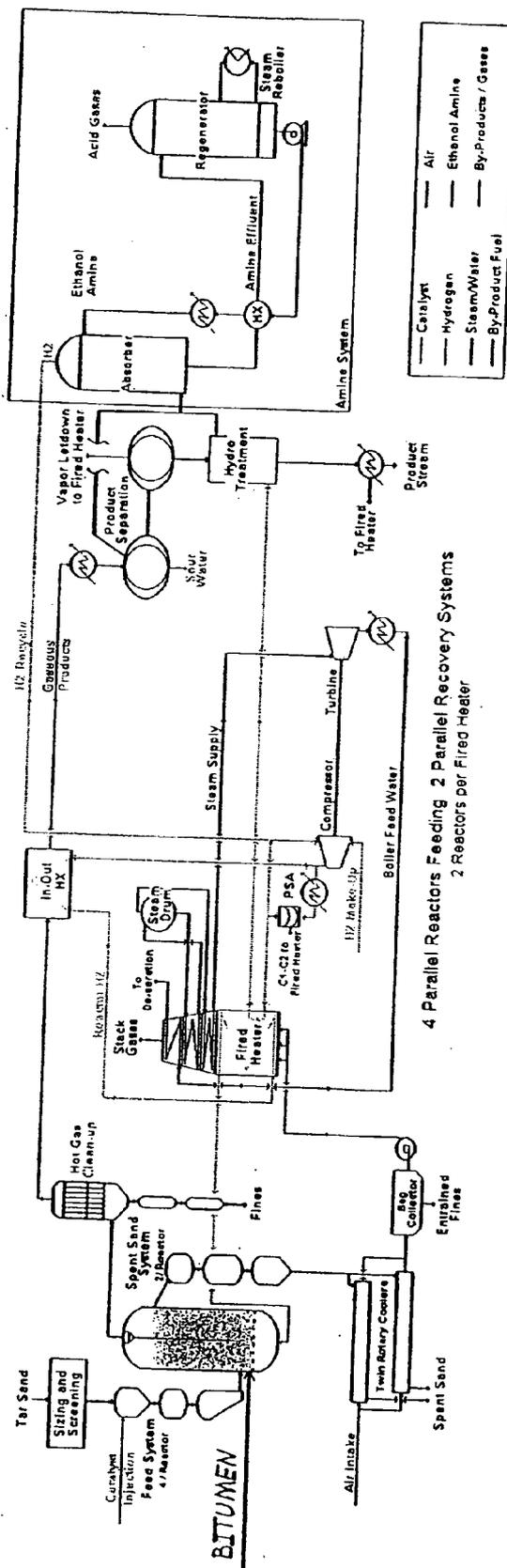


FIG. 8

Figure 9



4 Parallel Reactors Feeding 2 Parallel Recovery Systems  
2 Reactors per Fired Heater

## PROCESS AND APPARATUS FOR CONVERTING OIL SHALE OR OIL SAND (TAR SAND) TO OIL

### CROSS-REFERENCE TO A RELATED APPLICATION

[0001] This is a continuation-in-part application of previously filed copending U.S. application Ser. No. 11/126374 entitled "Process and apparatus for converting oil shale or oil sand (tar sand) to oil" filed May 11, 2005, which claims the benefit of earlier filed provisional application Ser. No. 60/570,815 entitled "Process and apparatus for converting oil shale or oil sand (tar sand) to oil" filed May 14, 2004. This application claims the benefit of U.S. application Ser. No. 11/126374 and this application claims the benefit of provisional application Ser. No. 60/570,815.

[0002] U.S. application Ser. No. 11/126374 entitled "Process and apparatus for converting oil shale or oil sand (tar sand) to oil" filed May 11, 2005, is incorporated by reference herein in its entirety. U.S. provisional application Ser. No. 60/570,815 filed May 14, 2004, is incorporated by reference herein in its entirety.

### FIELD OF THE INVENTION

[0003] The present invention relates to a continuous process for producing synthetic crude oil (SCO) from oil shale or oil sand (also known as "tar sand") and an apparatus for its practice. More specifically, the present invention provides a process for treating dry oil sand (tar sand) or shale without prior beneficiation, in a reactor operating at elevated pressure and temperature conditions, in the presence of hydrogen gas. The invention also relates to a continuous process for producing SCO from oil sand bitumen. The processed bitumen can be extracted from under ground via in situ processes, or strip mined and extracted via hot water extraction processes. The present invention, can also produce SCO from oil shale. Raw kerogen oil can be extracted in situ from shale underground.

[0004] The invention can be used to convert, process or upgrade one or more of bitumen, oil sand, oil shale, kerogen oil and heavy hydrocarbon feedstocks to synthetic crude oil.

### BACKGROUND OF THE INVENTION

[0005] Oil sand (tar sand) systems are known for making synthetic crude oil (SCO), such as those that surface mine and process the oil sand (tar sand), where they first separate sand (85 wt. %) from bitumen (11 wt. %) to avoid processing the sand in the reaction systems. The separated bitumen is converted to sweet, light crude oil by an upgrading refinery operation. Separation of the sand from the bitumen requires beneficiating operations such as floatation cells and secondary separation equipment and processing and equipment to prepare the oil sand (tar sand) for floatation. In these systems, tailing oil recovery is necessary to clear the sand for disposal, however the sand is not completely cleared of bitumen.

[0006] Existing technology uses a large number of physical and chemical processing units for the treatment of wet oil sand (tar sand), e.g., tumblers (being phased out by hydro-pumping), beneficiators including: primary separation vessels with floatation cells and secondary separation systems necessary to recover the bitumen from the oil sand (tar

sand); tailing oil recovery systems which result from the sand not being completely cleared of bitumen; tailing settling ponds which are necessary to settle and separate fine clays and other undesirable solids from the water for floatation since the water must be reused to maximize clean-up to reduce environmental problems. These systems can be large facilities along with the maintenance and systems for reclamation.

[0007] Raw oil sand (tar sand) can be treated in a fluidized bed reactor in the presence of a reducing environment, steam, recycle gases and combustion gases.

### BRIEF SUMMARY OF THE INVENTION

[0008] The process and apparatus of the present invention avoid the use of the large number of physical and chemical processing units used in the processing of wet oil sand (tar sand) by using a single continuous reactor system to hydrocrack and hydrogenate the dry oil sand (tar sand). Moreover, because the present invention directly hydrogenates dry oil sand (tar sand), larger quantities of valuable sweet, light crude oil are obtained.

[0009] The present invention relates to a continuous process for converting oil bearing material, e.g., oil shale or oil sand (tar sand), and an apparatus for its practice.

[0010] One embodiment of the present invention provides a continuous process and an apparatus for its practice where oil bearing material such as the kerogen in oil shale or the bitumen in oil sand (tar sand) is continuously treated.

[0011] In one embodiment, the present invention relates to the treatment of dry oil sand (tar sand).

[0012] In one embodiment, the present invention provides a method and apparatus for converting a oil sand (tar sand) or shale feed to oil which can be conducted in the absence of a beneficiation processes such as, for example, a hot-water extraction process to separate sand or other non reacting solids from bituminous or oil-bearing material in the feed.

[0013] In one embodiment, the present invention provides a process for converting oil sand (tar sand) to oil through the use of a stream containing hydrogen in a concentration greater than 90 vol % (90 vol %-100 vol % H<sub>2</sub>).

[0014] In one embodiment, the present invention provides a heat recovery process whereby hydrogen provides the heat necessary to bring the raw oil sand (tar sand) up to reactor temperature.

[0015] In one embodiment, the present invention provides a process where hydrogen is used for hydrocracking and hydrogenating the bitumen in the oil sand (tar sand) or oil shale.

[0016] In one embodiment, the present invention provides a process for using recycle and make-up hydrogen as a heat transfer vehicle.

[0017] In one embodiment, the present invention provides an improved process for producing oil from oil sand (tar sand) or shale by reacting the oil sand (tar sand) or shale with a stream containing hydrogen in a concentration greater than 90 vol % (90 vol %-100 vol % H<sub>2</sub>) in a fluidized bed reactor, wherein the fluidizing medium is a feed stream including hydrogen gas.

[0018] In one embodiment, the present invention provides a fluidized bed process where one inch or less size oil sand (tar sand) or shale pieces are fed into a fluidized bed reactor near the bottom of the reactor and spent sand and reaction products are removed from near the top of the reactor.

[0019] In each of the following three embodiments:

[0020] Extraction produced bitumen and raw sand or tailings pond sand; or

[0021] in situ produced bitumen and raw sand or tailings pond sand; or

[0022] in situ produced raw kerogen oil and shale, are fed to the reactor via the lock-hopper system for solids and injection lines for liquids.

[0023] In one embodiment, the present invention provides a method of recycling unreacted hydrogen that exits a reactor in which oil sand (tar sand) or oil shale is converted to oil. The method includes purging impurities in the exiting recycle hydrogen stream by acid gas removal through an amine system and by pressure swing adsorption, maintaining the hydrogen at more than about 450 psig throughout the recycle process, admixing fresh hydrogen to the recycle hydrogen stream to form a mixture, and feeding the mixture into the reactor.

[0024] In one embodiment, the invention provides a process for producing oil from an oil bearing feed such as oil sand (tar sand) or oil shale. The process comprises introducing the feed in a fluidizable form into a fluidized bed reactor. A fluidizing medium enters the fluidized bed reactor where it contacts and fluidizes the fluidizable feed. The fluidizing medium includes at least hydrogen. The fluidized feed forms a fluidized bed where the feed reacts with the hydrogen provided by a feed stream having hydrogen in a concentration of from 90-100% at a temperature within the equipment design range of about 800° F. to about 1500° F., with one embodiment having a reaction temperature of 900° F.-1200° F., and another embodiment having a reaction temperature of about 915° F. The reaction products include synthetic crude oil and spent solids which are discharged from the fluidized bed reactor.

[0025] In one embodiment, the invention provides flexibility in the processing of feedstocks. It comprises a process for upgrading oil bearing material e.g. bitumen or kerogen, and an apparatus for its practice.

[0026] In one embodiment, the present invention provides a continuous process and an apparatus for its practice where oil bearing material such as the kerogen from oil shale or the bitumen in oil sand is continuously treated.

[0027] In one embodiment, the present invention provides a process for upgrading extracted bitumen, or in situ produced bitumen or kerogen, through the use of a stream containing hydrogen in a concentration greater than 90 vol % (90 vol %-100 vol % H<sub>2</sub>).

[0028] In one embodiment, the present invention provides the heat necessary to bring the bitumen or kerogen up to reactor temperature of about 800° F.-1500° F.

[0029] In one embodiment, the present invention provides the heat necessary to bring tailings or raw sand up to a temperature in the range of 840° F.-1000° F., with one

embodiment having a temperature of about 915° F. for the fluid bed reactor for processing the bitumen or kerogen.

[0030] In one embodiment, present invention provides a process where hydrogen is used for hydrocracking and hydrogenating the bitumen of an oil sand or a kerogen of oil shale.

[0031] In one embodiment, the present invention provides a process for using recycle and makeup hydrogen as a heat transfer medium.

[0032] In one embodiment, the present invention provides an improved process for producing upgraded oil from extracted bitumen or in situ produced bitumen or kerogen, wherein the fluidizing medium is substantially hydrogen.

[0033] In one embodiment, the present invention provides an improved process for producing upgraded oil from extraction produced bitumen or in situ produced bitumen, or in situ produced kerogen from shale where raw sand or tailings sand or shale respectively is fed into the bottom of the reactor for fluidization with hydrogen and the reaction products are discharged from near the top of the reactor.

[0034] In one embodiment, the present invention provides a method for recycling the excess unreacted hydrogen from the reactor in which bitumen or kerogen is upgraded to high grade low sulfur oil (about 0.1 to 0.5 wt. % sulfur, or less). In the processing of oil sands and bitumen, this can produce an SCO having an API gravity of 28° to 34° and a sulfur content of about 0.1 to 0.5 wt. %. In the case of processing oil shale a lighter product is produced ranging from 28° API to as high as 40° API or more.

[0035] In one embodiment, the present invention provides the dry fluidized bed of sand in a reactor at a temperature of about 900° F. and 600 psig to receive hot bitumen at about 300° F. in a steam-traced line from a conventional, or existing, extraction facility, such as a hot bitumen extraction plant. Sand from existing sand tailings deposits or operations can be transported to the reactor by conventional, or existing, materials handling means. In one embodiment, sand from tailings deposits can be transported to the plant by trucks or hydrotransport as a slurry, dewatered and handled by conveyors, bucket elevators, screened and fed to the reactor.

[0036] In one embodiment, the present invention provides a process for processing the reclaimed sand and extracted bitumen together in the previously discussed reactor to produce a higher grade, about 34° API, low sulfur oil of about 0.1 to 0.5 wt. % sulfur, or less, by the use of an appropriate catalyst.

[0037] In one embodiment, the invention processes in situ produced hot bitumen in a fluid bed reactor located in close proximity to the producing well head. Reclaimed sand is fed in a continuous stream to the reactor to provide the fluid bed medium in addition to the hot bitumen which is fed separately. Recycle and make-up hydrogen pre-heated to about 1500° F. in a central fired heater provide the reactant, fluidizing medium and heat conveyor for the reactor which is operated at about 900° F. and 600 psig. The reclaimed sand can be transported to the site as a water slurry and be dewatered. Spent sand from the reactor containing some coke from the reaction process can then be slurried with the

original incoming water and be transported back to its original or an alternate area for disposal.

[0038] In one embodiment, the present invention provides a process for producing synthetic crude oil, comprising the steps of providing a reactor, a first reactor feed stream comprising a bitumen, and a gas reactor feed comprising hydrogen, contacting the bitumen with the hydrogen in the reactor thereby producing a reactor gas stream, providing a pressure swing adsorption system producing a PSA tail gas, and feeding the PSA tail gas to an integrated hydrogen plant.

[0039] In one embodiment, the first feed stream can comprise an oil sand. The oil sand can comprise a Canadian Athabasca oil sand. In another embodiment the feed can comprise a heavy oil. In one embodiment, the reactor has a fluidized bed formed of an oil sand.

[0040] In one embodiment, the present invention, provides for a process for producing synthetic crude oil, comprising the steps of providing a reactor, a first reactor feed stream comprising a kerogen, and a gas reactor feed comprising hydrogen, contacting the kerogen with the hydrogen in the reactor thereby producing a reactor gas stream, providing a pressure swing adsorption system producing a PSA tail gas, and feeding the PSA tail gas to an integrated hydrogen plant. In one embodiment, the first feed stream comprises an oil shale.

[0041] In one embodiment, the present invention has a reactor with a fluidized bed formed of an oil shale.

[0042] In one embodiment, this process for producing synthetic crude oil further comprises the steps of producing a synthetic crude oil product and burning an amount of the synthetic crude oil product in a fired heater. In another embodiment, the present invention is self-sufficient in production of gas feed to the integrated hydrogen plant. In yet another embodiment, the present invention is self-sufficient in production of fuel to the fired heater.

[0043] The present invention provides for an apparatus for producing a synthetic crude oil.

[0044] The present invention provides for a system for producing a synthetic crude oil.

[0045] The present invention provides for a control system for producing a synthetic crude oil.

#### BRIEF DESCRIPTION OF THE DRAWINGS

[0046] FIG. 1 shows the flow diagram of one embodiment according to the present invention.

[0047] FIG. 2 shows a fluidized bed reactor for converting bitumen in oil sand (tar sand) to viable products in accordance with the present invention.

[0048] FIG. 3 shows a stand-alone fired heater used in the process according to the present invention.

[0049] FIG. 4 shows a compressor for supplying the hydrogen for use in the present invention.

[0050] FIG. 5 shows the flow chart of an acid gas recovery system for use in the present invention.

[0051] FIG. 6 shows the mass balance for one embodiment of the present invention.

[0052] FIG. 7-1 shows a flow diagram of a second embodiment according to the present invention.

[0053] FIG. 7-2 shows a flow diagram of turbine, product separation and amine system of the second embodiment.

[0054] FIG. 7-3 shows a flow diagram of a variety of optional feed streams.

[0055] FIG. 7-4 shows a flow diagram of a reactor operations embodiment.

[0056] FIG. 7-5 shows a flow diagram of and embodiment of turbine, product separation and amine system.

[0057] FIG. 7-6 shows a flow diagram of a third embodiment according to the present invention.

[0058] FIG. 7-7 shows a flow diagram of and embodiment of turbine, product separation and amine system for the third embodiment.

[0059] FIG. 8 shows a fluidized bed reactor embodiment with lock hoppers.

[0060] FIG. 9 shows a process flow diagram for Example 3

[0061] In FIGS. 1-6, common elements are similarly identified except for the "figure number" designation. Thus, all elements depicted in FIG. 1, start off with the number 1, e.g., the reactor in FIG. 1 is identified as "104" and in FIG. 2 the same reactor is identified as "204."

#### DETAILED DESCRIPTION OF THE INVENTION

[0062] In one embodiment of the invention hydrocarbon bearing solids, e.g., dry oil sand (tar sand) or oil shale is processed in a fluidized bed reactor. Reaction occurs in the reactor as the feed constituents undergo reactions, including reactions with hydrogen. The process is operated in some embodiments to avoid decompression of the hydrogen. A first portion of a stream containing hydrogen in a concentration greater than 90 vol % (90 vol %-100 vol % H<sub>2</sub>) is used to feed the oil shale or oil sand (tar sand), which has been comminuted and reduced in size to form particles that are capable of being fluidized, e.g., fluidizable, into the reactor. A second portion of the hydrogen stream is used as the fluidizing medium.

[0063] The concentration of a hydrogen feed can vary in this invention. One embodiment utilizes pure hydrogen feed (100% H<sub>2</sub>). Most embodiments use lower purity hydrogen because light hydrocarbons are formed in the reactor, i.e. methane and ethane, the accumulation of these light gases adversely affects the reaction between bitumen and hydrogen by suppressing the partial vapor pressure of the hydrogen. A recycle hydrogen stream is therefore operated in another embodiment to maintain the hydrogen concentration at about 94 vol. % by purging methane and ethane via the PSA system. The invention can use hydrogen feed concentration as low as 25 vol. %, or lower.

[0064] While the reactor can react with any concentration of hydrogen in the feed stream (1 vol %-100 vol % H<sub>2</sub>). Some embodiments use hydrogen feed streams with H<sub>2</sub> concentrations of 75 vol %-94 vol % and methane and ethane concentrations of 6 vol % to 25 vol %. One embodi-

ment utilizes a hydrogen feed stream having 94 vol % H<sub>2</sub> and 6 vol % methane and ethane mixture.

[0065] In one embodiment, the hydrogen stream that feeds the reactor is formed from fresh make-up hydrogen and recycle hydrogen generated during the process, or obtained from other hydrogen producing processes. A mixed stream of fresh-make-up and recycle hydrogen stream is discharged from a compressor at a first temperature and pressure (e.g., 187° F. and 670 psig), and a portion is diverted for admixture with the fluidizable particles of oil sand (tar sand) or oil shale which are injected into the fluidized bed reactor in a fan like flow, at an acute angle relative to the vertical axis of the reactor or a horizontal plane.

[0066] The reference to fan like flow supra is further described in conjunction with an embodiment where the solids (e.g., oil sands, oil shale, ordinary sand or tailings pond sand) are fed to the reactor in a flow by a combination of gravity and assistance from the high pressure hydrogen motivation stream into the reactor. The inlet line from the lock-hopper system to the reactor can be horizontal or can be inclined relative to the vertical axis of the reactor. Mechanical means such as a screw can be used in certain applications to assist the solids flow into the reactor. The highly agitated conditions in the reactor quickly disperse the solids feed into the fluidized bed. One embodiment utilizes four lock-hopper feed systems which can be equally spaced around the circumference of each reactor.

[0067] In one embodiment, the solids fed to the reactor, either oil sands, oil shale, ordinary sand or tailings pond sand, flow by a combination of gravity and assistance from the high pressure hydrogen motivation stream into the reactor. The actual inlet line from the lock-hopper system to the reactor can be horizontal or can be inclined relative to the vertical axis of the reactor. Mechanical means such as a screw can be used in certain applications to assist the solids flow into the reactor. The highly agitated conditions in the reactor quickly disperse the solids feed into the fluidized bed. The present design provides four lock-hopper feed systems equally spaced around the circumference of each reactor.

[0068] The remainder of the hydrogen stream at said first temperature is indirectly heated to a second higher temperature by indirect heat exchange with overhead products from the fluidized bed reactor. The second higher temperature is the temperature of the hydrogen stream from the discharge of the compressor that has passed through the in-out heat exchanger in order to cool the reactor over head stream. It can be about 875° F.; having been heated from 187° F. by indirect heat exchange with hot reactor over head stream. Indirect simply means the heat transfer from a hotter gas to a cooler gas takes place through a solid metallic retaining wall so that mixing of the two gas streams cannot occur.

[0069] The hydrogen stream at the second temperature (e.g., about 875° F.) is conveyed to a direct fired heater where the hydrogen stream is heated to a third temperature higher than said second temperature (e.g., about 1200° F.-1500° F.) and then used as the fluidizing medium in the reactor to fluidize the oil sand (tar sand) or oil shale fluidizable particles that have been injected with the first portion of the hydrogen stream.

[0070] In the fluidized bed reactor the bitumen in the oil sand (tar sand), or the kerogen in the oil shale change phase

and can be reacted via endothermic and exothermic reactions, including reactions with hydrogen, to produce spent oil sand (tar sand) or oil shale and an overhead product stream that contains hydrogen, hydrogen sulfide, sulfur gases, C<sub>1</sub>+C<sub>2</sub> hydrocarbons, ammonia, fines (sand particles and clay) and vaporous products. The overhead product stream is first separated in cyclone separators within the reactor which help maintain the bed level and separate solids.

[0071] The reaction products formed in the reactor that upon cooling, condensing and separation from the other stream components form the liquid product SCO. In the liquid phase these products are grouped as naphtha, distillate and gas oil.

[0072] In the reactor, contact between the feeds and hydrogen takes place within the fluidized bed of raw sand, or tailings pond sand or shale. The particular solid being fed, either raw sand, tailings pond sand or shale, establishes the fluid bed for the particular combination of feed stocks being processed. The reaction between hydrogen and bitumen or between hydrogen and raw kerogen oil produces reaction products in gaseous or vapor form. Also, the water contained in oil sands, raw sands or tailings pond sand fed to the reactor vaporizes to form steam. The product and water vapors exit the reactor via the normal flow pattern through the internal cyclone separators at the top of the reactor and flow downstream to the next stage in the process circuit.

[0073] The solids fed to the reactor, either oil sands, oil shale, ordinary sand or tailings pond sand, flow by a combination of gravity and assistance from the high pressure hydrogen motivation stream into the reactor. The actual inlet line from the lock-hopper system to the reactor can be horizontal or can be inclined relative to the vertical axis of the reactor. Mechanical means such as a screw can be used in certain applications to assist the solids flow into the reactor. The highly agitated conditions in the reactor quickly disperse the solids feed into the fluidized bed. The present design provides four lock-hopper feed systems equally spaced around the circumference of each reactor.

[0074] In one embodiment, the overhead product stream is first separated from the fluid bed solids by the reactor space above the level of the solids by the reactor space above the level of the solids overflow line and the internal cyclone separators. (The bed level can be established by the by the height of the solids overflow lines above the reactor support grid).

[0075] The reactor effluent stream of gases and product vapors separated from most of the reactor sand and shale can be considered the "First separated overhead" product.

[0076] The first separated overhead product is conveyed to a hot gas filter(s) to provide a cleaned product stream. The hot gas filter can remove unstrained sand, shale or fines that escape from the reactor and cleanses that hot overhead stream. In one embodiment, the temperature of the product overhead stream after the hot gas filter is about 915° F. but can range 840° to 1000° F. The cleaned product stream at a first temperature (e.g., 840° to 1000° F.) is conveyed to a first heat exchange unit where heat is transferred to a second portion of the hydrogen stream and results in a product stream at a second temperature lower than said first product stream temperature (e.g., about 450° F.). The second portion

hydrogen stream can constitute a recycle and make-up hydrogen stream which can be fed to the fired heater and to the reactor. The second temperature lower than said first product stream temperature of the recycle stream (second portion) is in one embodiment the temperature of the overhead product stream after the in-out heat exchanger, which for one embodiment is about 450° F. which is lower than 9150 F, the normal temperature before the in-out heat exchanger. This second temperature can however vary (300° F. to 900° F.).

[0077] The product stream at said second temperature is conveyed to a condenser to further reduce its temperature to a third temperature lower than the second product stream temperature which in one embodiment can be cooled in a range of about 100° F. to about 4500. The overhead gas stream can be cooled to 100° F. from about 4500 F to condense the product vapors. In some embodiments additional heat recovery can take place between the in-out heat exchanger and the condenser. This can result in a lower temperature of the gas stream going to the condenser, perhaps as low as 200° F.

[0078] The product stream at said third temperature contains liquid and gas fractions and is conveyed to a separator where the gas fraction is removed, sent to an amine scrubber, and recycled as a scrubbed recycle hydrogen stream, and the liquid fraction is removed as oil product (SCO).

[0079] The recycle hydrogen is conveyed to a compressor where it is combined with fresh make-up hydrogen for use in the fluidized bed reactor as the first and second hydrogen stream portions.

[0080] The following definitions are employed in this disclosure.

[0081] "Bitumen" Bitumen is a naturally occurring petroleum-based tar-like material. It is the hydrocarbon component found in tar or oil sands. Bitumen, the hydrocarbon component found in oil sands, varies in composition depending upon the degree of loss of the more volatile components. It can vary from a very viscous, tar-like, semi-solid material to solid forms. The hydrocarbon types found in bitumen comprise aliphatics, aromatics, resins, and asphaltenes. A typical bitumen hydrocarbon might be composed of:

[0082] 19 wt. % aliphatics (which can range from 5 wt. %-30 wt. %, or higher)

[0083] 30 wt. % aromatics (which can range from 15 wt. %-50 wt. %, or higher)

[0084] 32 wt. % resins (which can range from 15 wt. %-50 wt. %, or higher)

[0085] 19 wt. % asphaltenes (which can range from 5 wt. %-30 wt. %, or higher)

[0086] Sulfur content of bitumen can range in excess of 7 wt. %.

[0087] In addition bitumen can contain water and nitrogen compounds ranging from less than 0.4 wt. % to in excess of 0.7 wt. %. The metals content, while small, can be removed to avoid contamination of the product SCO. In one embodiment metals are removed with the coke which forms on the spent sand. Nickel can vary from less than 75 ppm (part per

million) to more than 200 ppm. Vanadium can range from less than 200 ppm to more than 500 ppm.

[0088] The percentage of the hydrocarbon types found in bitumen can vary. For example, asphaltenes can vary from 10 wt % to 20 wt %.

[0089] "Catalyst" The definition of catalyst as used in the chemical industry is that it is a substance that initiates or accelerates a chemical reaction without itself being affected. The term "catalyst" encompasses pure, diluted, blended, mixed catalysts and catalysts provided on a support material, or by coating, impregnating and the like. Catalyst can be added directly to the oil sand feed stream and thus carried into the reactor where the reaction between hydrogen and bitumen and its components is promoted. Catalyst can also be utilized in the hydrotreater where the reactions between hydrogen and sulfur, nitrogen, complex aromatics, etc. is promoted and these contaminants can then be removed from the product stream. Catalysts for these purposes are usually formulated by the catalyst manufacturer to suit the requirements of the feed stock being processed. The invention can utilize a broad variety of catalysts including those having metals and can include elements such as nickel, cobalt, molybdenum, iron, and aluminum.

[0090] "Degrees API" The American Petroleum Institute, the United States Bureau of Mines and the National Institute of Standards and Technology selected the API scale in 1921 as the standard for petroleum products in the United States. It is an arbitrary scale but is accepted widely internationally in the petroleum industry as it provides a measure of the density of the liquid being described.

$$\text{Degrees API} = \frac{141.5}{\text{Specific gravity } 60^{\circ} \text{ F./} 60^{\circ} \text{ F.}} - 131.5$$

[0091] The above shows the relationship between degrees API and specific gravity. For example, if a substance had a specific gravity of 0.0 at 60° F. with reference to water at 60° F., solving the equation would give it an API gravity of 45.4°.

[0092] "High grade" or "excellent quality" oil sands contain 12% bitumen or higher.

[0093] "Low grade" is the opposite of high grade. The subject matter is of inferior or poor quality. For example, oil sands containing about 6-9 wt. % bitumen would be considered low grade; 9-11 wt. % bitumen oil sands would be average grade and 11 wt. % plus bitumen oil sands would be considered high grade. High grade product oil (SCO) would be a low sulfur (about 0.1 to 0.5 wt. % sulfur, or less), light crude oil possessing excellent qualities for further processing.

[0094] "Low Grade Oil Shale" can be shale that yields very little shale oil when processed. Since the kerogen in oil shale is a solid and cannot be extracted without a change in at least one of form, physical properties or phase, as for example, bitumen is extracted from oil sands, the quality or grade of oil shale can be measured by the amount of shale oil that is produced when conversion of the kerogen in the shale takes place. Oil shale producing less than about 15 gal/ton of shale oil is considered low grade.

[0095] “In-situ” production methods are used on bitumen deposits buried too deep for mining to be economical. The bitumen can be recovered by several means, including steam assisted gravity drainage (SAGD). In-situ production of bitumen means production takes place at its original site.

[0096] “In situ oil Sand” is oil sand which is in its original location, or oil sand which has not been chemically modified, reacted or mixed so as to lose its original chemical properties.

[0097] “In situ kerogen” is kerogen which is in its original location, or kerogen which has not been chemically modified, reacted or mixed so as to lose its original chemical properties. Additionally, In situ kerogen refers to kerogen contained in shale that is in its natural original site. The shale is in place, not having been mined or otherwise relocated. In situ production of kerogen requires it to be converted from a solid to a condensable vapor at its original site and this vapor when recovered and condensed becomes liquid shale oil.

[0098] “In situ Shale” shale which is in its original location, or shale which has not been chemically modified, reacted or mixed so as to lose its typical properties. In one embodiment oil shale contains water in a range of about 1 wt % to 7 wt %. The water concentration of shale can be higher (e.g., 10-15%) depending on the source.

[0099] “Kerogen” is a bituminous material in shale that yields oil (“kerogen oil”) upon heating. When kerogen is heated to temperatures over 700° F., it decomposes to produce a liquid oil similar to petroleum. i.e., naturally occurring erode oil. Unlike the bitumen contained in oil sands, the kerogen in oil shale is a solid that does not melt and is insoluble. To exploit it, kerogen must be converted from a solid to a non-solid state. This is accomplished by heating the kerogen containing oil shale to about 900° F. in the absence of oxygen to convert the kerogen to a condensable vapor which upon cooling becomes liquid shale oil or raw kerogen oil.

[0100] “Mesh”—In coarser sizes of screens, the term “mesh” means the distance between adjacent wires or rods. In the finer sizes of screen cloths the mesh means the number of openings per linear inch of screen.

[0101] “Oil shale”—Oil shales are fine grained sedimentary rocks containing relatively large amounts of organic matter from which significant amounts of shale oil can be extracted. Oil shale is a fine crained, calcareous rock containing kerogen, a solid hydrocarbon.

[0102] “Pressure” as used herein, the term pressure refers to the force acting per unit area within the various vessels and pipelines constituting the process plant. It is expressed as pounds per square inch (psig). All references herein to pressure are in units of pounds per square inch gage, psig, unless otherwise indicated.

[0103] “Sand” as used herein, unless otherwise specified, “sand” as a feedstock encompasses all types of sand either alone or in combination: e.g., sand, raw sand, tailings sand, reclaimed sand, spent sand, supplemental sand and ordinary sand.

[0104] “Raw Sand” is ordinary silica sand mined from sources other than tailings ponds. An embodiment provides the fluid bed medium in the reactor. The fluidized bed of

sand provides much surface area promoting contact and reaction between hydrogen and injected liquid bitumen.

[0105] “Reclaimed sand”—This term refers to sand recovered from current wet extraction operations directly from the process operations or from the settling ponds following the relatively rapid settling out of the coarse sand.

[0106] “Run of Mine” is the term applied to oil sand and shale mined before any processing or beneficiation takes place. The oil sands are “as mined”, dug from the pit and loaded onto trucks or other conveying equipment.

[0107] “SAGD” (steam assisted gravity drainage) refers to steam assisted gravity drainage method of bitumen extraction or production. This production method can apply to reserves too deep to be mined economically. In SAGD production, parallel horizontal wells are sunk into a reserve. One well injects steam to heat the surrounding oil sands and reduce the viscosity of the bitumen. A second, lower well extracts the bitumen as it flows by gravity to the production well.

[0108] “Wet extraction” refers to the current industry process of bitumen extraction using hot water as the main processing medium. Advances have been made since the early days when very hot water was used. Today, a combination of chemicals addition and lower water temperature of about 120° F. are being used but current mining operations using water extraction can be described as wet extraction methods.

[0109] “Tailings”—The term tailings as used herein refers to the refuse material comprising sand, clay traces of bitumen, etc. remaining after the bitumen has been extracted from oil sands. This material is discarded by the process as a water slurry into tailings ponds where the heavier sand settles to the bottom of the pond. The problem with tailings ponds is that the finer clay particles remain in suspension for a long period taking years to settle and consolidate.

[0110] “Oil sand” and “Tar Sand”—These terms are used synonymously interchangeably but modern industry usage more commonly recites “oil sand”. Oil sand (tar sand) contains the heavier asphaltic components of petroleum. It can be considered the end product of evaporation of petroleum’s volatile components. Tar sand or oil sand is a combination of clay, sand, water and bitumen. It is sand whose interstices remains filled with asphalt after lighter portions of crude oil have escaped. Tar or oil sands as found in nature consist of particles or grains of sand, each grain being surrounded with a film of water upon which a film of bitumen, a dark, asphalt like oil is deposited. The bitumen content of oil sands can vary broadly from less than 1 wt. % to as high as 20 wt. %. The average bitumen content of oil sands below about 7-8 wt. % bitumen are currently considered to be of low quality.

[0111] “Upgrading” is a term used to refer to the treatment of crude oil to improve its quality. As used in the process the main product of upgrading is synthesized crude oil that can be later refined like (conventional) natural erode oil into a broad range of products. It is the conversion of bitumen from a viscous, tar-like substance into a high quality, light sweet crude oil with no residual bottoms. Upgrading is the process of converting heavy oil or bitumen into synthetic crude oil. The process converts bitumen which can have an API gravity of 8-100 to an SCO of 30-34° API. In the process,

heavy long chain hydrocarbons are broken down or cracked into lighter short chain hydrocarbons sulfur and nitrogen compounds are converted to hydrogen sulfide and ammonia and removed. Also, metals such as nickel and vanadium can be removed in the process. One objective of upgrading is to reduce the viscosity of the oil and control the amount of solids and water. The full upgrading process reduces sulfur, nitrogen and metals content to produce a crude oil substitute of high quality that can be refined by current commercial refinery operations.

**[0112]** “Feedstock” and “feed” refer to a material which enters a unit operation or process. In the invention a variety of feedstocks are processed to produce products including SCO. For example, the reactor feed system in one embodiment can provide both solid and liquid feedstocks (feeds) to the reactor. Feedstocks feeds can be fed to a unit operation or process separately or in mixtures with other materials and feeds. In one embodiment feeds of both solids and liquids are fed to the reactor separately through different inlets. Solid feeds can include at least one of an oil sand, an oil shale, a sand, as well as other solids or semi-solids which can include coal, coke and very heavy oils, whether separately or in mixtures. The liquid feeds can include bitumen, kerogen, heavy oil, viscous hydrocarbon containing material, whether separately or in mixtures. Feedstocks to the invention can essentially be any hydrocarbon bearing material. Gases can also be feedstocks. In one embodiment a hydrogen containing gas is fed to the reactor. Hydrocarbon containing gases can also be fed to the reactor in an embodiment.

**[0113]** “Fluidizing medium” refers to the flow stream which serves to fluidize the bed. In one embodiment the fluidizing medium is a hydrogen feed stream.

**[0114]** “Fluidized bed” herein encompasses traditional fluidized bed technology utilized in reactors, as well as a fluidized bed reactor design employed in embodiments of the reactor disclosed herein. Embodiments of the “fluidized bed” of the present invention encompass the fluidized state of any material fluidized in the reactor. The reactor of the present invention contains a fluidized bed which can be comprised substantially of the feed materials fed to the reactor. The fluidized bed can be formed when the feed material present in the reactor is fluidized by the fluidizing medium. The fluidized bed can be continuously replenished as feed present in the reactor is reacted and the produced gases and solids exit the reactor. The fluidized bed can be comprised of any one, or more, of the reactor feed(s) disclosed herein. Fluidized feed alone can comprise the fluidized bed. Catalyst can be added to the reactor as a reactor feed and can flow-through the reactor to exit with the spent sand, or spent shale. This can be referred to as a flow-through fluidized bed reactor design. The fluidized bed of an embodiment of the reactor comprises any one, or more, of a hydrocarbon containing fluidizable feed, an optional catalyst and a fluidizing medium which fluidizes the fluidizable feed forming a fluidized bed. The fluidized feed reacts (optionally in the presence of catalyst) and the reaction products which result from the reaction of the feed comprising the fluidized bed itself exit the reactor. As reaction products exit the reactor, the fluidized bed is maintained by fluidizable feed entering the reactor and becoming fluidized forming newly fluidized material of the fluidized bed. The fluidized bed of the reactor can be a continuously reacting

and continuously replenished. Thus, material of the fluidized bed can be continuously reacted in reaction and replenished with fluidizable feed. The fluidized bed of the reactor can be designed free of (not including) any fixed catalyst beds. No fixed catalyst beds, or traditional fixed beds, are required by the reactor for normal operation.

**[0115]** “Reaction process” refers to the chemical and physical activities occurring in the reactor including, but not limited to, the reaction between hydrogen and various feeds, the physical agitation or fluidization of the reactor contents by the flow of hydrogen and other gases, the separation of gases and solid particles, etc.

**[0116]** “PSA tail gas” includes at least any gas evolved from regeneration of the adsorption medium of the PSA system, as well as any other one or more gases utilized during the regeneration process. “PSA tail gas” broadly includes any gas which exits the PSA system during a regeneration process, or during process transition from regeneration to online operation. “PSA tail gas” can include any one or more of the following hydrogen, methane, ethane, propane, butane, light hydrocarbons, C<sub>1</sub>-C<sub>4</sub> hydrocarbons. Further, the term “PSA tail gas” broadly encompasses any gases which exit from the PSA system during any PSA system operation which can be recovered for further processing in the process. Such operations can include system pressurization, purging, abnormal operation, pressure relief, let-down, reduced capacity operation or excess capacity operation, or any operation which produced exiting hydrocarbon or hydrogen gas which can serve as a feed source of any other portion of the process including the hydrogen plant. “PSA tail gas” can comprise methane and/or ethane, as well as optionally including hydrogen. Other light hydrocarbons can be optionally included in “PSA tail gas”.

**[0117]** The invention can now be described with reference to the figures. **FIG. 1** is a flow chart to one embodiment of the present invention where oil sand (tar sand) is converted to oil. In accordance with the present invention, oil sand (tar sand) from the run of mine conveyor belt **101** is continuously fed to any suitable sizing equipment **102** for classifying oil sand (tar sand), at a temperature of about 50° F. Oil sand (tar sand) is composed of bitumen and sand.

**[0118]** The bitumen in the oil sand (tar sand) that is processed in the present invention normally contains heavy metals which catalytically help promote the endothermic and exothermic reactions in reactor **104**. However, it can be advantageous to add additional catalyst for reasons including reduced reaction time, reduced residence time requirement in the reactor, promote a greater degree of reaction completion. These activities lead to improved performance and more efficient operations resulting in economic benefits.

**[0119]** Endpoints of ranges are recognized to incorporate within their tolerance other values within the knowledge of one of ordinary skill in the art, including, but not limited to, those which are insignificantly different from the respective endpoint as related to this invention (in other words, endpoints are to be construed to incorporate values “about” or “close” or “near” to each respective endpoint). The range and ratio limits, recited herein, are combinable. For example, if ranges of 10-200 and 50-150 are recited for a particular parameter, it is understood that ranges of 10-50, 10-150, 50-200, or 150-200 are also contemplated.

**[0120]** The oil sand (tar sand) processed in accordance with the present invention is exemplified by the following, non-limiting example:

OIL SAND (TAR SAND) FEED EXAMPLE	
sand	84.6 wt. %
bitumen	10.4 wt. %
water	5 wt. %
carbon	83.1 wt. %
hydrogen	10.6 wt. %
sulfur	4.8 wt. %
nitrogen	0.4 wt. %
oxygen	1.1 wt. %
nickel	75 PPM
vanadium	200 PPM
Total	100 wt. %

[0121] In the present invention dry oil sand (tar sand) having an average particle size of that of sand is conveyed through conduit 103 as the feed for fluidized bed reactor 104, discussed in greater detail in FIG. 2. Oil sand (tar sand) particles which are oversized are either recycled to the sizing equipment 102, or conveyed to any suitable equipment for reducing the size of the oversized feed. In the present invention, the phrase “dry oil sand (tar sand)” means, under atmospheric conditions, a friable, non-sticky, easily run of mine oil sands.

[0122] Oil sand (tar sand) is fed through pressure feeder rotary valves 104A which are circumferentially positioned adjacent and around the upper end of the fluidized bed reactor 104, which is described in detail greater in FIG. 2. The rotary feeders 104A are positioned at an angle of between 20 and 60 degrees relative to the vertical reactor axis in order to “fan feed” the fluidizable sized oil sand (tar sand) into the top of the reactor 104. More uniform dispersion of the oil sand (tar sand) in the fluidized bed reactor can be obtained when three or more rotary feed valves 104A are positioned equidistantly around the circumference of the reactor. Although three feeders 104A are preferred, the size of the reactor and the degree of fanning desired can control the number of valve feeders. Thus, there can be 4, 5, 6, 7 or more valve feeders used present invention.

[0123] High pressure hydrogen is conveyed through lines 138 to the feeders 104A, at a pressure of between 625 psig and 700 psig, preferably about 635 psig, to assist in injecting, feeding and dispersing the oil sand (tar sand) into reactor 104. Another portion of the hydrogen gas feed from line 137 for fluidizing medium (hydrogen feed) is diverted through lines 139 and injected into the separator section 104B, at the bottom end of the reactor.

[0124] The process performed in fluidized bed reaction 104 involves hydrocracking, which is an endothermic reaction, and hydrogenation, which is an exothermic reaction, which reactions are conducted to favor the production of liquid fuels and minimize the production of gas yields. The reactor operates at temperatures within the equipment design range of about 800° F. to about 1500° F., with one embodiment having a reaction temperature of 900° F.-12000F, and another embodiment having a reaction temperature of about 915° F. Embodiments operating lower than 900° F., and even at 800° F. are less likely for, or avoid, cracking the large fragments of hydrogenated bitumen in the oil sand (tar sand).

[0125] It is advantageous to conduct the endothermic hydrocracking and exothermic hydrogenating processing of oil sand (tar sand) in reactor 104 in a predominantly hydrogen gas environment. The hydrogen atmosphere in reactor 104 is maintained at about 600 psig by fresh make-up hydrogen conveyed through line 130 from a hydrogen plant and a hydrogen recycle stream 129 which contains cleaned-up hydrogen. The volume of recycle hydrogen to fresh make-up hydrogen is preferably at least about 26 to 1.

[0126] Advantageously all the high pressure hydrogen for the process of the present invention, for reaction in reactor 104 and the various heat exchange operations, is provided by the steam powered compressor 132. Compressor 132 receives fresh make-up hydrogen which is conveyed through line 130 and recycle hydrogen which is conveyed through lines 129, 140, 142, 144 and 131. Compressor 132 is powered by steam conveyed through line 162 from direct fired heater 135. 184 is the fired heater stack gases line.

[0127] Reactor 104 operates in a highly agitated fashion insuring almost instant and complete reaction between the bitumen components and hydrogen. The residence or retention time of the oil sand (tar sand) in reactor 104 is about 15 minutes, but can be between 10 and 20 minutes, depending on the throughput and efficiency of the reactor process. The pressure drop from the bottom to the top of the reactor 104 is about 35 psi.

[0128] Overhead products from reactor 104 are discharged from reactor 104 through cyclone separators 104C, while solids are discharged through separator section 104B located at the lower end of reactor 104. The cyclones separators 104C discharge an overhead stream, e.g., gas and vapor reaction components, off-gas and product, through their upper ends into line 110, while separated solids are discharged through the lower ends of the dip legs. The cyclone separators 104C extend about 20 feet down into the reactor 104 and establish the bed height in the reactor 104.

[0129] The hot spent oil sand (tar sand) is continuously discharged at a pressure of about 635 psig and a temperature of about 800° F. through lock hopper valving arrangement 104B in the lower end of reactor 104 into line 105 which conveys the discharged material to spent sand heat exchangers 106 and 108. Spent sand is discharged through line 109.

[0130] The reactor overhead stream from the cyclone separators 104C is discharged into line 110, at a temperature of about 800° F. and a pressure of about 600 psig. The overhead stream discharged from the reactor 104 still contains dust and dry waste particles, and is first conveyed through line 110 to cyclone separator 111 where solids are separated and removed through line 150. The gaseous effluent from separator 111 is conveyed through line 112 to an electrostatic precipitator 113 for the final cleanup. The cleaned overhead stream from precipitator 113 is removed and conveyed through line 114, and separated solids are discharged through line 151. Cyclone separator 111 and electrostatic precipitator 113 are of conventional design and one of ordinary skill in the art practicing the present invention can select suitable devices for performing the described operation.

[0131] The cleaned stream from the precipitator 113, product, vaporous components, and off gas, are conveyed to in-and-out heat exchanger 115 through line 114. In the

in-and-out exchanger **115** the cleaned stream from line **114** is brought into indirect heat exchange relationship with hydrogen being conveyed through line **133**, from compressor **132**, i.e., recycle and fresh make-up hydrogen, whereby heat is transferred from the cleaned stream to the hydrogen in line **133** prior to the hydrogen stream entering the fired heater **104**. The cooled and cleaned stream, products, vaporous components, off-gases, from heat exchanger **115** is discharged into line **116** while hydrogen is discharged into line **134** which conveys the hydrogen to the direct fired heater **135**.

[0132] The cooled stream being conveyed through line **116** is introduced into condenser **117** and is discharged at a temperature of about 100° F. into line **118**. The vapor and gas stream from the condenser is conveyed through line **118** at a temperature of 100° F. and is introduced into separator **119** where vapors and liquid are separated and discharged.

[0133] Since the gas stream has been cooled down to about 100° F. and is still at a pressure of 480 psig, all carbon compounds C<sub>3</sub> and above have been collected and removed from the separator **119** through flow line **155** to storage. Sour water from the separator is discharged through flow line **154**. The crude oil product stream in line **155** is a mixture of naphtha and gas oils having an A.P.I. of approximately 33.5 and is a light sweet crude. The gas stream in line **120** is conveyed to a scrubbing system, e.g., at least one amine absorption column **121** where sulfur components, e.g., hydrogen sulfide and/or sulfur dioxide gases, are absorbed and discharged through line **122** and conveyed to a suitable sulfur recovery plant. Carbon dioxide gas can also be absorbed by the scrubbing system. The amine absorption system having amine absorber **121** is described in greater detail in FIG. 5.

[0134] The only gases not absorbed and removed in the amine absorption system having amine absorber **121** are unreacted recycle hydrogen and C<sub>1</sub>+C<sub>2</sub> hydrocarbons which are conveyed through line **129** to heat exchangers **106** so that the spent oil sand (tar sand) is cooled and the recycle hydrogen and C<sub>1</sub>+C<sub>2</sub> hydrocarbons is heated and discharged into line **140**. The C<sub>1</sub> and C<sub>2</sub> hydrocarbons in line **129** can not be absorbed nor condensed but can be recycled with the unreacted hydrogen after processing in units **141**, **143** and **145** discussed hereinafter. The C<sub>1</sub> and C<sub>2</sub> hydrocarbons can reach equilibrium within the reactor **104** at about 2% and can then add to the production of crude oil per ton of oil sand (tar sand). A small offset can be the increase in the recycle stream.

[0135] As discussed above, the spent sand from the reactor **104** is discharged into a succession of heat exchangers **106** and **108**. The first heat exchanger **106** cools the sand from 792° F. to 400° F. using cool recycle hydrogen being conveyed through line **129**. The cooled spent sand is conveyed in line **107** from heat exchanger **106** and introduced into a second heat exchanger **108** so that the sand is cooled by cold air introduced through line **180** from blower **181** and through line **182**, before discharging. The air heated by the spent sand is discharged into line **183** which conveys the heated air to fired heater **135** for combustion therein. Although two heat exchangers are shown, the invention contemplates using more if necessary.

[0136] The heated and partial recycle hydrogen stream conveyed through line **140** is introduced into cyclone **141**,

discharged into line **142** which conveys the stream to precipitator **143**, and then through line **144** for introduction into reboiler **145**.

[0137] FIG. 2 schematically shows the pressurized, continuously operating fluid bed reactor **204** in accordance with the present invention. Sized and screened oil sand (tar sand) or shale are conveyed through lines **203** and fed through pressure feeder rotary valves **204A** into the top of the reactor **204**. A portion of the gases processed in compressor **132** (FIG. 1), and heated in fired heater **135** (FIG. 1) are conveyed by line **236** and introduced into fluidized bed reactor **204** in an upward direction to fluidize the bed of the reactor **204**. Another portion of the hydrogen gas from line **133** is conveyed through line **237** to oil sand (tar sand) feed valves **204A** through lines **238**. Another portion of the hydrogen gas feed from line **237** is diverted through lines **239** and injected into the separator section **204B**, at the bottom end of reactor **204**. Hydrogen conveyed in lines **239** is injected into the separator section **204B** of reactor **204** through injectors which are located at the ends of flow lines **239** (not shown) and aid in heat retention in the reactor system and spent sand discharge through line **205**.

[0138] Fluidizable feed enters reactor **204** and forms a fluidized bed when it is in contact with the fluidizing medium. The fluidized bed can be comprised substantially of fluidizable feed and its reaction products. The reaction process is continuous and fluidized feed is continuously reacted into reaction products which exit the reactor. The feed material comprising the fluidized bed is continuously reacted and the fluidized bed is continuously replenished by the fluidizable feed entering the reactor and coming into contact with the fluidizing medium to become fluidized and one with the fluidized bed. Catalyst may be optionally added as a feed material.

[0139] High temperature and high pressure hydrogen (make-up and recycle) after passing through the direct fired heater **135**, is introduced into reactor **204** from line **236**. Reaction products and unreacted hydrogen exit the reactor through internal cyclones **204C** ensuring even flow out of the reactor. Although two cyclone separators are shown, the invention contemplates using as many as necessary to provide even flow of product gases from reactor **204** and bed height maintenance. The hot reactor effluent stream in line **210** is then conveyed to physical and chemical units, described in FIG. 1 for cleanup heat recovery and product separation.

[0140] As discussed above with reference to FIG. 1, a portion of the fresh make-up and cleaned recycle hydrogen from the compressor is conveyed to a direct fired heater. FIG. 3 schematically shows a fired heater **335** (**135**) that is designed to balance out the total energy to operate the reactor system. Preheated air conveyed through feed lines **383** (**183**) is combusted with fuel in the radiant section of fired heater **335** (**135**) and elevates the temperature of the recycle and make-up hydrogen that is conveyed through line **334** (**134**). The fuel that is combusted is obtained from the C<sub>3</sub> fraction, e.g. propane, or natural gas produced or purchased from the described process or other sources. The hydrogen stream in lines **334** (**134**) has been preheated in the reactor in-out exchanger **115** to approximately 750° F. Since the hydrogen stream is circulated through the radiant section of the heater **335** the temperature of the hydrogen stream is

elevated to a temperature of about 1200° F. Circulation of the hydrogen stream through line 133, 134, exchanger 115 and fired heater 335 (135) is maintained by compressor 132 so that the 1200° F. hydrogen stream can be introduced into reactor 104 (FIG. 1) or 204 (FIG. 2).

[0141] Waste heat from the radiant section of direct fired heater 335 is recovered in convection section 335A (135A), 335B (135B) and 335C (135C). Steam separated in drum 360 (160) is discharged into line 361 (161) and introduced into convection section 335A (135A) where the steam temperature is raised from about 596° F. to about 800° F. After passing through convection section 335A (135A), the superheated, high pressure steam is conveyed through line 362 (162) to drive the steam turbine 163 (FIG. 4). Reduced temperature and pressure steam from turbine 163 is conveyed to steam condenser 165 and the condensate recirculated via line 166 and pump 166A (FIG. 1). The flow from pump 166A is conveyed through line 168 (368) and combined with make-up water from line 167. The water being conveyed in line 268 (168) is introduced into convection section 335C (135C), heated and discharged through line 369 (169) for further processing, e.g., deaeration.

[0142] Steam drum 360 (160) separates steam which is conveyed to radiant section 335A (135A) through line 161 to produce superheated steam for the turbine compressor 163.

[0143] The steam circulation loop includes steam drum 360 (160), line 370 (170), recirculation pump 371 (171) and lines 372-373 (172-173) which convey boiler water through radiant section 335B (135B) and back into drum 360 (160). Water for the boiler system is provided through feed line 467 (167) which flows into line 368 (168) which is in communication with line 169 through convection section 335a (135a) to deaeration.

[0144] As discussed above, convection section 335A (135A) superheats steam which is conveyed through line 362 (162) to drive compressor turbine 163, which drives compressor 132. Steam is generated in convection section 335B (135B) and make-up water and turbine condensate for boiler feed water are preheated in convection section 335C (135C).

[0145] FIG. 4, schematically shows a compressor 432 (132) driven by a high pressure steam turbine 463 (163) to maintain circulation of gases to operate the reactor system 104. Make-up hydrogen 430 (130) and recycle hydrogen 431 (131), at approximately 450 psig and 100° F. are pressurized by the compressor 432 (132) to approximately 670 psig and 122° F. and discharged into line 133 which conveys and introduces the high pressure hydrogen into the in-out exchanger 115 to be further heated by exchange with reactor product gases.

[0146] High pressure steam in line 162, 362, at 1500 psig and 800° F. drives the turbine 463 (163). Exhaust steam 464 (164) is condensed in condenser 465 (165), and along with make-up water 467 (167) is fed to the fired heater convection section 135C, 335C for preheating and reuse as boiler feed water make-up.

[0147] The product separation of FIG. 1, components can be described in greater detail with reference to FIG. 5, which schematically shows the product separation from the circulating gas stream and removal of acid gasses in an

amine system. Partially cooled reactor effluent gases 516 (116) from the in-out exchanger 115 are further cooled in product condenser 517 (117) and conveyed through line 518 (118) to separator 519 (119) where condensed liquids are removed as product raw crude 555 (155). Overhead gases are conveyed through line 520 (120) to an amine absorber 5A (121) where acid gasses H<sub>2</sub>S, CO<sub>2</sub> and SO<sub>2</sub> are absorbed by a counter current circulating amine solution. The recycle gases 5B flow from the top of the absorber 5A (121) to recycle hydrogen stream 129.

[0148] The rich amine solution 5C (121C) exits the bottom of the absorber, flows through an amine exchanger 5D (121D) where it is heated by exchange with hot stream amine solution 5L (121L) and enters the top of an amine stripper 5F (121F) through line 121E. The lean amine solution passes through exchanger 121D to exchanger 121N via line 121M and enters 121 via lean amine feed line 121P.

[0149] Absorbed acid gases are stripped from the amine solution by the application of heat to the solution in reboiler 545 (145) (having reboiler feed line 146 and reboiler return line 147) and are conveyed through flow line 522 (122) from the stripper to sulfur recovery off-site. Hot recycle gases are conveyed through line 544 (144) from the spent sand cooler 545 (145) (106 via cyclone 141 and precipitator 143) to provide heat for reboiler 545 (145) and the partially cooled recycled gases SG (121G) are further cooled by cooler SH (121H) and then flow through line 531 (131) to the suction side of compressor 132. Solids removed by cyclone 141 and precipitator 143 are discharged via lines 152 and 153 respectively.

[0150] Lean amine solution SJ (121J) is circulated by amine circulation pumps 5K (121K) through the amine exchanger SD (121D) and amine cooler SN (121N) to the top of the amine absorber SA (121A) to repeat the gas cleanup process.

#### EXAMPLE 1

[0151] The overall mass balance for the process according to the present invention is shown in FIG. 6, where 1000 tons/br of oil sand (tar sand) at 50° F. are reacted with hydrogen to produce 665 bbl/r of synthetic crude oil. The following Table provides the feed and product values for processing 1000 tons/hr. of oil sand (tar sand).

[0152] Raw Materials Products

[0153] 1000 TONS/HR. OIL SAND (TAR SAND) 665 BBL/HR SCO

[0154] 1.6 MMSCF/HR HYDROGEN 5.2 MMSCF/HR STACK GAS

[0155] 3.3 MMSCF/HR AIR 6600 LBS/HR SULFUR

[0156] 0.5 MMSCF/HR NATURAL GAS 850 TONS/HR SPENT SAND

[0157] Reactor Dimensions and Mass and Energy Balances

[0158] Reactor 104

[0159] Column Diameter 20.00 ft

[0160] Cross Section Area 34.16 ft.sup.2

[0161] Void Fraction 0.85 (At Fluidization)

- [0162] Cross Section of Sand 47.12 ft.sup.2
- [0163] Cross Section of Gas 267.04 ft.sup.2
- [0164] Reactor Volume 27394.26 ft.sup.3
- [0165] Bed Diameter 20.00 ft
- [0166] Bed Height 87.20 ft
- [0167] Time-Space Constant 0.25 hr
- [0168] Pressure Drop 35.00 psi
- [0169] Oil Sand (Tar Sand) Feed
- [0170] Sand Flow Rate 1000.00 tons/hr
- [0171] Density of sand 21.68 lbs./ft.sup.3
- [0172] Volumetric sand flow 16436.55 ft.sup.3/hr
- [0173] Sand Velocity 5.81 ft/minute
- [0174] Hold-up 15.00 minutes
- [0175] Hydrogen
  - [0176] Hydrogen Flow Rate 238661.44 lbs/hr
  - [0177] (45226343 SCF/hr)
  - [0178] Cp of H<sub>2</sub> 3.50 btu/lb-° F. (@ 900° F.)
  - [0179] Hydrogen Recycle Ratio 26.52
  - [0180] Hydrogen Flow Rate 45.28 SCF/hr
  - [0181] Hydrogen velocity 3.02 ft/s
- [0182] Off Gas
  - [0183] Gas Production 0.40 MMSCF/hr
  - [0184] MW 30.30 g/mole
  - [0185] Cp of flue gas 0.55 btu/lb-° F.
- [0186] Off Gas Composition
  - [0187] CO 0.30%
  - [0188] CO<sub>2</sub> 0.20%
  - [0189] H<sub>2</sub> S 31.00%
  - [0190] NH<sub>3</sub> 2.50%
  - [0191] C<sub>3</sub> 66.00%
- [0192] Energy Balance
- [0193] Over-All Considerations
  - [0194] Heat of Reaction 75.00 btu/lb. Bitumen
  - [0195] Cp Sand 0.19 btu/ton-° F.
  - [0196] Cp Bitumen 0.34 btu/lb-° F.
  - [0197] Cp Tarsand (sand+Bitumen) 426.70 btu/ton-° F.
  - [0198] Sand Feed Temperature 50.00° F.
  - [0199] Sand temperature 50.00° F. at reactor inlet
  - [0200] Reaction temperature 800.00° F.
  - [0201] Sand Feed 1,000.00 tons/hr
- [0202] Oil Sand (Tar Sand) Reactor
- [0203] Reactor Conditions
  - [0204] Heat required in reactor 356.03 MMbtu/hr
- [0205] Heat generated in Reactor 22.50 MMbtu/hr
- [0206] Additional Heat Required 335.24 MMbtu/hr
- [0207] Minimum H<sub>2</sub> for reaction 9000.00 lbs./hr
- [0208] (1.71 MMSCF/hr)
- [0209] Additional H<sub>2</sub> Supplied 229736.15 lbs./hr
- [0210] (43.53 MMSCF/hr)
- [0211] Total H<sub>2</sub> Supplied 238736.15 lbs./hr
- [0212] (45.24 MMSCF/hr)
- [0213] C<sub>1</sub>-C<sub>2</sub> Flow within H<sub>2</sub> Stream 4594.72 lbs/hr
- [0214] (at equilibrium-2%) (0.08 MMSCF/hr)
- [0215] Entering H<sub>2</sub> Temperature 1200.00° F.
- [0216] Cp H<sub>2</sub> 3.50 btu/lb-° F.
- [0217] Heat Supplied by C<sub>1</sub>-C<sub>2</sub> 1.01 MMbtu/hr
- [0218] Heat Supplied by H<sub>2</sub> 334.23 MMbtu/hr
- [0219] H<sub>2</sub> Recycle ratio 26.53
- [0220] Reactor Bottoms Cooler:
  - [0221] Assures Efficient Removal of Exiting Solids
  - [0222] Cold Hydrogen Cooler Stream 1,148.68 lbs./hr
  - [0223] (0.22 MMSCF/hr)
  - [0224] Heat Removed 2.73 MMbtu/hr
  - [0225] Entering Hydrogen Temperature 121.64° F.
  - [0226] Exiting Sand Temperature 791.60° F.
- [0227] Sand Cooler
- [0228] Sand
  - [0229] Sand Flow Rate 850.00 tons/hr
  - [0230] Temperature of Entering Sand 791.60° F.
  - [0231] Temperature of Spent Sand 180.00° F.
  - [0232] Cp Sand 0.19 btu/lb-° F.
  - [0233] Heat Removed 198.59 MMbtu/hr
- [0234] Hydrogen Coolant Flow
  - [0235] Hydrogen Flow 238736.15 lbs/hr
  - [0236] (45.24 MMSCF/hr)
  - [0237] Heat to Be Removed 182.96 MMb/hr
  - [0238] Entering Hydrogen Temperature 100.00° F.
  - [0239] Exiting Hydrogen Temperature 318.96° F.
- [0240] Air Coolant
  - [0241] Air Required for Combustion 250000.00 lbs/hr
  - [0242] (3.27 MMSCF/hr)
  - [0243] Cp Air 0.25 btu/lb-° F.
  - [0244] Entering Air Temperature 50.00° F.
  - [0245] Exiting Air Temperature 300.00° F.
  - [0246] Heat Removed 15.63 MMbtu/hr

- [0247] Amine Reboiler
- [0248] Hydrogen Supply
  - [0249] Entering Hydrogen Temperature 318.96° F.
  - [0250] Exiting Hydrogen Temperature 100.00° F.
- [0251] Amine Boil-Off
  - [0252] Heat Available to the system 182.96 MMbtu/hr
- [0253] In-Out Heat Exchanger
- [0254] Hydrogen to be Heated
  - [0255] Hydrogen Flow 238736.15 lbs/hr
  - [0256] (45.24 MMSCF/hr)
  - [0257] Inlet H<sub>2</sub> Temperature 121.64° F.
  - [0258] Exiting H<sub>2</sub> Temperature 750.00° F.
  - [0259] Total Heat Required 525.05 MMbtu/hr
- [0260] Off Gas Heat Supply
  - [0261] Off Gas flow rate 31978.89 lbs/hr
  - [0262] 0.40 MMSCF/hr
  - [0263] Condensables in vapor phase 214941.75 lbs/hr
  - [0264] MW 30.30 lb/lb-mole
  - [0265] Cp Vapor 0.55 btu/lb-° F.
  - [0266] Cp Liquid 0.45 btu/lb-° F.
  - [0267] Cp Non-Condenses 3.00 btu/lb-° F.
  - [0268] Heat of Vaporization 65.00 btu/lb
  - [0269] Hydrogen Recycle Flow 229736.15 hrs/hr in Stream (\*43.53 MMSCF/hr)
  - [0270] Inlet Temperature 800.00° F.
  - [0271] Exit Temperature 350.00° F.
- [0272] Product Condenser/Cooler Product Side
  - [0273] Entering Temperature 350.00° F.
  - [0274] Exiting Temperature 100.00° F.
  - [0275] Condensate 24941.75 lbs/hr
  - [0276] (665.29 bbl/hr)
  - [0277] Heat Removal H<sub>2</sub> 201.02 MMbtu/hr
  - [0278] Off Gas 4.40 MMbtu/hr
  - [0279] Condensate 38.15 MMbtu/hr
  - [0280] Total 243.57 MMbtu/hr
- [0281] Cooler Requirement 243.57 MMbtu/hr
- [0282] Compressor Hydrogen Side
  - [0283] Flow Rate 755412.69 SCF/min
  - [0284] (45.32 MMSCF/hr)
  - [0285] Pressure Out 670.00 psig
  - [0286] Pressure In 450.00 psig
  - [0287] DP 220.00 psi
  - [0288] gamma (Cp/Cv) 1.40
- [0289] # Stages 3
- [0290] Temperature Inlet 100.00° F.
- [0291] Mechanic Efficiency 0.80\*100%
- [0292] Pb/Pa 1.14
- [0293] Power Requirement per Stage 6366.67 hp
- [0294] Total Power Required 19100.00 hp
- [0295] Outlet Temperature 121.64° F.
- [0296] Steam Supply
  - [0297] Pressure 1500.00 psig
  - [0298] Temperature 800.00° F.
  - [0299] Degree Superheat 200.00° F.
  - [0300] Saturation Temperature 596.20° F.
  - [0301] Steam Heat Vue 1364.00 btu/lb
  - [0302] Flow Rate 10894.28 lbs/hr
- [0303] Fired Heater Products to be Heated
  - [0304] Hydrogen Flowrate 238736.15 lbs/hr
  - [0305] (45.24 MMSCF/hr)
  - [0306] Hydrogen Temperature 750.00° F.
  - [0307] Water Flow Rate 10894.28 lbs/hr
  - [0308] Water Temperature 75.00° F.
  - [0309] Heat Duty 517.83 MMbtu/hr
  - [0310] C<sub>3</sub>'s (Fuel Produced by the Process)
  - [0311] Flow Rate 4263.85 lbs/hr
  - [0312] (0.04 MMSCF/hr)
  - [0313] Heat of Combustion 20000.00 btu/lb
  - [0314] Cp 0.60 btu/lb-° F.
  - [0315] Temperature in 75.00° F.
  - [0316] Heat Supplied
  - [0317] (After temperature correction) 79.84 MMbtu/hr
- [0318] Make-Up Methane
  - [0319] Combustion Temperature 2200.00° F.
  - [0320] Heat Remaining to 437.99 MMbtu/hr be supplied by Methane
  - [0321] Flow Rate 21653.89 lbs/hr
  - [0322] (0.51 MMSCF/hr)
  - [0323] Heat of Combustion 20227.00 btu/lb
  - [0324] (After temperature correction)
  - [0325] Temperature in 75.00° F.
- [0326] Combustion Air
  - [0327] Air Required for Combustion 200000.00 lbs/hr
  - [0328] (2.61 MMSCF/hr)
  - [0329] Air Supplied 25% Excess 250000.00 lbs/hr
  - [0330] (3.27 MMSCF/hr)

- [0331] Compressor Suction
- [0332] Cooler (5H)
- [0333] Outflows
- [0334] Hydrogen
- [0335] Flowrate 200000.00 lbs/hr
- [0336] Temperature 100.00° F.
- [0337] Required Coolant Supply 22.42 MMbtu/hr
- [0338] Material Balance
- [0339] Oil Sand (Tar Sand) Reactor (104)
- [0340] In Flows
- [0341] Sand
- [0342] Flowrate 1000.00 tons/hr
- [0343] Temperature 50.00° F.
- [0344] Pressure 14.70 psia
- [0345] (Force Fed)
- [0346] Hydrogen
- [0347] Flowrate 45.23 MMSCF/hr
- [0348] Temperature 1200.00° F.
- [0349] Pressure 635.00 psig
- [0350] C<sub>1</sub>-C<sub>2</sub> s
- [0351] Flowrate 0.08 MMSCF/hr
- [0352] Temperature 1200.00° F.
- [0353] Pressure 635.00 psig
- [0354] Out Flows
- [0355] Sand
- [0356] Flowrate 850.00 tons/hr
- [0357] Temperature 190.00° F.
- [0358] Pressure 600.00 psig
- [0359] Off Gas
- [0360] Flowrate 43.92 MMSCF/hr
- [0361] Temperature 800.00° F.
- [0362] Pressure 600.00 psig
- [0363] Composition wt %
- [0364] H<sub>2</sub> 81.98
- [0365] CO 0.05
- [0366] CO<sub>2</sub> 0.04
- [0367] H<sub>2</sub> S 5.60
- [0368] NH<sub>3</sub> 0.45
- [0369] C<sub>3</sub> 11.92
- [0370] Product
- [0371] Flowrate 214937.52 lbs./hr
- [0372] (Vapor Phase)
- [0373] Temperature 800.00° F.
- [0374] Pressure 600.00 psig
- [0375] Sand Cooler (106, 108) In Flows
- [0376] Sand
- [0377] Flowrate 850.00 tons/hr
- [0378] Temperature 791.92° F.
- [0379] Pressure 600.00 psig
- [0380] Hydrogen
- [0381] Flowrate 45.23 MMSCF/hr
- [0382] Temperature 100.00° F.
- [0383] Pressure 500.00 psig
- [0384] Air
- [0385] Flowrate 3.27 MMSCF/hr
- [0386] Temperature 50.00° F.
- [0387] Pressure 30.00 psig
- [0388] Out Flows
- [0389] Sand
- [0390] Flowrate 850.00 tons/hr
- [0391] Temperature 200.00° F.
- [0392] Pressure 480.00 psig
- [0393] Hydrogen
- [0394] Flowrate 45.23 MMSCF/hr
- [0395] Temperature 313.94° F.
- [0396] Pressure 480.00 psig
- [0397] Air
- [0398] Flowrate 3.27 MMSCF/hr
- [0399] Temperature 300.00° F.
- [0400] Pressure 20.00 psig
- [0401] In-Out Heat Exchanger (115) In Flows
- [0402] Hydrogen
- [0403] Flowrate 45.23 MMSCF/hr
- [0404] Temperature 147.60° F.
- [0405] Pressure 670.00 psig
- [0406] Off Gas
- [0407] Flowrate 43.92 MMSCF/hr
- [0408] Temperature 800.00° F.
- [0409] Pressure 600.00 psig
- [0410] Composition wt %
- [0411] H<sub>2</sub> 81.98
- [0412] CO 0.05
- [0413] CO<sub>2</sub> 0.04
- [0414] H<sub>2</sub> S 5.60
- [0415] NH<sub>3</sub> 0.45

- [0416] C<sub>3</sub> 11.92
- [0417] Product
- [0418] Flowrate 214937.52 lbs./hr
- [0419] (Vapor Phase)
- [0420] Temperature 800.00° F.
- [0421] Pressure 600.00 psig
- [0422] Out Flows
  - [0423] Hydrogen Flowrate 45.23 MMSCF/hr
  - [0424] Temperature 750.00° F.
  - [0425] Pressure 650.00 psig
  - [0426] Off Gas Flowrate 43.92 MMSCF/hr
  - [0427] Temperature 368.63° F.
  - [0428] Pressure 580.00 psig
  - [0429] Off Gas Composition as Above
- [0430] Product
  - [0431] Flowrate 214937.52 lbs./hr
  - [0432] (Vapor Phase)
  - [0433] Temperature 368.63° F.
  - [0434] Pressure 580.00 psig
- [0435] Product Condenser/In Flows
  - [0436] Cooler (117)
    - [0437] Off Gas
      - [0438] Flowrate 43.92 MMSCF/hr
      - [0439] Temperature 368.63° F.
      - [0440] Pressure 580.00 psig
      - [0441] Off Gas Composition as Above
    - [0442] Product
      - [0443] Flowrate 214937.52 lbs./hr
      - [0444] (Vapor Phase)
      - [0445] Temperature 368.63° F.
      - [0446] Pressure 550.00 psig
- [0447] Out Flows
  - [0448] Off Gas
    - [0449] Flowrate 43.92 MMSCF/hr
    - [0450] Temperature 100.00° F.
    - [0451] Pressure 540.00 psig
    - [0452] Off Gas Composition as Above
  - [0453] Product
    - [0454] Flowrate 214937.52 lbs./hr
    - [0455] (as condensate)
    - [0456] Temperature 100.00° F.
    - [0457] Pressure 540.00 psig

[0458] Amine System (121, FIG. 5) In Flows

- [0459] Hydrogen
- [0460] Flowrate 45.23 MMSCF/hr
- [0461] Temperature 318.00° F.
- [0462] Pressure 470.00 psig

[0463] Out Flows

- [0464] Hydrogen
- [0465] Flowrate 45.23 MMSCF/hr
- [0466] Temperature 100.00° F.
- [0467] Pressure 450.00 psig

EXAMPLE 2

[0468] FIG. 7 shows another embodiment of the present invention. In this embodiment, a oil sand (tar sand) feed is converted into a synthetic crude oil. Run of mine oil sand (tar sand) from trucks is dumped into receiving, screening, and sizing equipment 702 via tar sand feed 701 for classifying oil sand (tar sand) at ambient temperature. The oil sand (tar sand) comprises bitumen and sand. The oil sand (tar sand) is crushed into relatively large fluidizable pieces that are capable of passing through a one inch mesh, or that are about one inch or less in size. In this embodiment, crushing the oil sand (tar sand) into fines or pieces less than sand size is preferably avoided to facilitate fines removal from the product stream. Limiting the amount of crushing can also reduce heat generation that can adversely affect oil sand (tar sand) processing. Oil sand (tar sand) pieces typically comprise an agglomeration of sand particles, each sand particle surrounded by a film of water and an outer layer of bitumen. On contacting a hot fluidizing flow of hydrogen during later reaction steps, the water film can rapidly evaporate assisting the oil sand (tar sand) pieces to disintegrate into a finely fluidized dispersion of sand particles and bitumen in hydrogen.

[0469] The crushed oil sand (tar sand) is conveyed through conduit 703 to feed lock hoppers 704 as the feed for fluidized bed reactor 705. The feed flow through conduit 703 and between feed lock hoppers 704 is controlled by ball valves ("ball valves") 703A. 704 designates a lock hopper and 703A designates a lock hopper valve. The bitumen in the oil sand (tar sand) can contain heavy metals, such as nickel, which can catalytically promote endothermic and exothermic reactions in reactor 705. However supplemental catalyst such as, for example, nickel, cobalt and molybdenum can be added through catalyst feed conduit 704A the catalyst injection line to one of the feed lock hoppers 704 to assist catalysis provided by the heavy metals in the mined oil sand (tar sand) or shale. 704B is the reactor feed line. The reactor 705 and related equipment are shown in more detail in FIG. 8.

[0470] Recycle hydrogen in conduit 725 and fresh make-up hydrogen in conduit 725A are conveyed to compressor 732. A first mixture of recycle hydrogen and makeup hydrogen exits compressor 732 in line 733, is cooled by heat exchanger 754 via feed line 733A, passes through line 757 to feed lock hoppers 704. This cooled first hydrogen mixture helps to prevent the oil sand (tar sand) from gumming by keeping the oil sand (tar sand) cool and forces the crushed

oil sand (tar sand) into the reactor **705** which operates at a pressure of about 600 psig. Preferably, the first hydrogen mixture reaches the lock hoppers **704** at a temperature of about 100° F. or less, and maintains the oil sand (tar sand) at a temperature of about 100° F. or less. The oil sand (tar sand) is fed from feed lock hoppers **704** through conduit **704B** and into reactor **705** through a feed inlet **705H** the feed inlet nozzle, assisted by the first hydrogen mixture at 670 psig pressure in line **757**. There are three feed lock hoppers in this embodiment, but the number can vary in other embodiments. The oil sand (tar sand) can be fed into the reactor approximately horizontally, near the bottom of the reactor, and just above ceramic support grid **705C**. Equipment for treating mined oil sand (tar sand) or shale feed material and for feeding the material into reactor, such as the equipment described above, can be referred to as a feed introducing system. Equipment for feeding oil sand (tar sand) or shale feed material into the reactor, such as the feed lock hoppers **704**, conduit **703** and ball valves **703A**, can be referred to as a feeder device.

[**0471**] On entering the reactor **705** (detail in **FIG. 8**), the oil sand (tar sand) is contacted and heated by a second hydrogen mixture. The second hydrogen mixture flows from fired heater **735** and into a gas inlet **705I** the fluidizing medium inlet nozzle at the bottom of the reactor **705B** through ceramic lined conduit line **736** at a temperature of about 1500° F. and about 635 psig pressure. The second hydrogen mixture passes through a slotted fire brick or ceramic grid **705C** before contacting the entering oil sand (tar sand). The flow rate and velocity of the second hydrogen mixture are sufficient to fluidize the oil sand (tar sand) and to heat the oil sand (tar sand) to a desired reaction temperature. The heated oil sand (tar sand) and the second hydrogen mixture react in the reactor **705** in a fluidized bed **705E** at the desired reaction temperature of about 900° F. to about 1000° F., and at a pressure of about 600 psig. The second hydrogen mixture flow rate typically exceeds the minimum needed for complete oil sand (tar sand) reaction with hydrogen by a factor of about 15 to about 26, and preferably by a factor of about 21. Adjustment of the second hydrogen mixture flow rate can require adjustment of other reaction parameters to maintain the fluidized bed **705E** at desired pressures and temperatures. The oil sand (tar sand) reacts with the hydrogen mixture in the fluidized bed **705E** by endothermic hydrocracking and exothermic hydrogenating reactions. Reaction products include substantially sulfur-free hydrocarbon that are condensable into hydrocarbon liquids at standard temperature and pressure.

[**0472**] Fluidizable feed enters reactor **705** and forms a fluidized bed when it is in contact with the fluidizing medium. The fluidized bed is comprised substantially of fluidizable feed and its reaction products. The reaction process is continuous and fluidized feed is continuously reacted into reaction products which exit the reactor. The feed material comprising the fluidized bed is continuously reacted and the fluidized bed is continuously replenished by the fluidizable feed entering the reactor and coming into contact with the fluidizing medium to become fluidized and form the fluidized bed (become one with the fluidized bed). Catalyst may be optionally added as a feed.

[**0473**] In one embodiment, the fluidized bed reactor can be free of (not including) a contained fixed catalyst bed and having a fluidized bed comprising substantially of fluidiz-

able feed, said reactor having a feed inlet for the fluidizable feed, the reactor having a fluidizing medium inlet for a gas comprising hydrogen. The fluidizing medium in contact with said fluidizable feed forms a fluidized feed. The fluidized feed forms the fluidized bed in which said fluidized bed comprises at least said fluidized feed. The fluidizable feed can be any fluidizable feed disclosed herein.

[**0474**] Embodiments of the reactor of this process can be of the flow-through type allowing for the flow of products out of the reactor and for the entrance of feeds forming the fluidized bed to be free of (not including) a fixed catalyst bed, or other fixed bed.

[**0475**] Reaction products including synthetic crude oil and unreacted hydrogen mixture exit the reactor **705** through a product stream outlet **705F** the reactor outlet nozzle as an overhead or product stream through cyclone separators **705A** and into exit conduit line **710**. Solids entrained in the overhead product stream, such as sand particles and fines, are trapped by the cyclone separators **705A** and are deposited near the ceramic screen **705C** over reactor bottom **705B**, where they are again entrained in the fluidized bed **705E**. Eventually, the spent sand and solids exit the reactor **705** through a conduit line **705D** the spent sand outlet line. The overhead stream flows through a hydrogen recycling system wherein hydrogen is removed from the remainder of the overhead stream, treated, and returned to the reactor.

[**0476**] It is advantageous to conduct the endothermic hydrocracking and exothermic hydrogenation reactions in a predominantly hydrogen gas environment. The first and second hydrogen mixtures are mixtures of fresh make-up hydrogen and recycle hydrogen which are fed to a compressor **732** via conduit lines **725A** and **725** respectively. The recycle hydrogen contains hydrogen and up to 5 mole percent of combined methane and ethane. The amount of combined methane and ethane in the recycle hydrogen is maintained by a purge in a hydrogen recycle system connected to the reactor **705**. The volume of recycle hydrogen to fresh make-up hydrogen is preferably about 21:1, but can vary from about 15:1 to about 26:1.

[**0477**] The reactor **705** is operated so as to highly agitate the reactants and ensure rapid and complete reaction between the bitumen components and hydrogen in the reactor **705**. The residence or reaction time of the oil sand (tar sand) in reactor **705** is about 10 minutes, but can be between 5 and 20 minutes, depending on the throughput and efficiency of the reactor process. The pressure drop from the bottom to the top of the fluidized bed **705E** is about 35 psi.

[**0478**] Spent sand, at a temperature of about 950° F., overflows from reactor **705** into conduit line **705D** through a spent solids outlet **705G** the spent sand outlet nozzle. The height of the conduit line **705D** can establish the maximum height of the fluidized bed **705E**. The sand then flows through spent sand lock hoppers **706**, through conduit line **707** and into rotary coolers **708** which cool the sand from a temperature of 950° F. to about 665° F. Spent sand is discharged through line **709**. The cooled sand can be discharged and used, for example, for land reclamation.

[**0479**] The rotary coolers **708** can use ambient air fed through air intake **778** to cool the spent sand. The air exits the rotary coolers **708** through line **779** at a temperature of about 625° F. and passes through a cyclone **780** to remove

entrained fines. The fines are discharged through conduit line **785**. The cooling air is heated by the spent sand, then passes to the fired heater **735** via blower **782** and conduit lines **781** and **783** where the air is used as preheated combustion air.

[**0480**] The number of feed lock hoppers **704** and spent sand lock hoppers **706** is controlled by the size of the reactor, thus more or less than the three feed lock hoppers **704** and more or less than three spent sand lock hoppers can be used in the present invention.

[**0481**] The reactor overhead stream from the cyclone separator **705A** is discharged into line **710** the reactor product gas outlet line, and then to hot gas clean-up **711**. The overhead stream in line **710** exits the reactor **705** at about 950° F. and enters the hot gas clean-up **711**. Ceramic bag collectors or filters in the hot gas clean-up **711** remove and collect fines remaining in the overhead stream. The filters are periodically pulsed by a back flow of a 670 psig, 875° F. hydrogen mixture taken from in-out heat exchanger **715** via conduit line **734A**. Collected fines and solids are removed from the bottom of the hot gas clean-up **711** and are collected in hot gas clean-up lock hoppers **712**. The fines can be combined with spent sand and used for land reclamation. The disposal of the dry sand and fines resulting from this invention is environmentally preferable to existing wet disposal systems.

[**0482**] The substantially solids-free overhead stream flows from the hot gas clean-up through hot gas clean-up overhead line **713** to the in-out heat exchanger **715**. The in-out heat exchanger **715** is an indirect heat exchanger wherein heat is transferred from the overhead stream to a portion of the hydrogen mixture exiting compressor **732** via conduit **733**. The heated hydrogen mixture is conveyed via a conduit line **734** to the fired heater **735**. The cooled overhead stream exits the in-out heat exchanger **715** through line **716**. **736** is the fluidizing medium inlet line to the reactor **705**.

[**0483**] The overhead stream in line **716** enters condenser **717** where condensable vapors and gases are condensed. The overhead stream exits the condenser **717** in line **718** at a temperature of about 100° F. and passes to a first separator **719** where the sour water is separated from the liquid product and removed via line **786**.

[**0484**] The partially cooled overhead product stream in line **716** enters condenser **717** where condensable vapors and gasses are condensed. The cooling medium can be circulated cooling water from an on-site cooling tower. Condensed liquids and the non-condensable gases at about 100° F. flow via line **718** to a separator **719** where water is separated from the condensed product liquids and gas stream. The amount of sour water produced varies depending upon the water content of the feed materials being processed in the reactor **705**. Any entrained oxygen in the reactor feed is converted to water in the predominately hydrogen atmosphere in the reactor. The waste water exits the separator **719** via line **786** and is treated to conform to environmental requirements. Stream **723** is fed to absorber **724**.

[**0485**] The condensed liquid product from separator **719** flows via conduit line **720** to the letdown vessel **721** where pressure reduction takes place and dissolved light gases, methane and ethane, are flashed off. The flashed off gases are

conveyed via conduit line **722** to the fuel system of the fired heater **735** for combustion. The liquid hydrocarbon products are removed from let-down tank **721** via flow line **790** as a light low sulfur (about 0.1 to 0.5 wt. % sulfur, or less) synthetic crude oil product stream which in one embodiment includes a mixture of naphtha and gas oils having an API gravity of approximately 30° to 34°. As discussed earlier, the quality of the product SCO including sulfur content can be modified by altering process variables and the inclusion or non-inclusion of various catalyst formulations to suit the requirements of the refiner.

[**0486**] The crude oil product in line **790** can be sent directly to storage without further treatment or it can be sent to a hydrotreater for further upgrading to reduce sulfur content to 0.1 wt. % and improve API gravity to about 34°.

[**0487**] The absorber **724** can comprise, for example, a counter current circulating ethanol amine solution in intimate contact with the remaining overhead stream. The remaining overhead stream can comprise gases such as, for example, H<sub>2</sub> S, CO<sub>2</sub>, recycle hydrogen, and C<sub>1</sub> and C<sub>2</sub> hydrocarbons. H<sub>2</sub> S and CO<sub>2</sub> are removed from the remaining overhead stream by the absorber **724**. Remaining hydrogen, C<sub>1</sub> and C<sub>2</sub> hydrocarbons form the recycle hydrogen mixture and flow through line **725** to compressor **732**.

[**0488**] The rich amine solution having absorbed H<sub>2</sub> S, CO<sub>2</sub>, SO<sub>2</sub> and NH<sub>3</sub> is discharged from the absorber **724** through line **740** and flows through an amine heat exchanger **741**. In the amine heat exchanger **741** the rich amine solution is heated by exchange with hot amine solution in line **750** which is returning from amine regenerator **743** to the absorber **724**. The heated rich amine solution flows through line **742** and enters the top of the amine regenerator **743**. Absorbed acid gases are stripped from the rich amine solution by further heating the rich solution using steam from a steam reboiler **745**. Heat for the reboiler **745** is supplied by steam from the fired heater **735** steam recovery system. Reboiler **745** is fed amine solution by reboiler feed line **746** and discharges it back to regenerator **743** through reboiler return line **747**.

[**0489**] Lean amine solution is discharged from the regenerator **743** in line **748** and is circulated by an amine circulation pump **749** through amine exchanger **741** and amine cooler **752** to the top of the amine absorber **724**. The effluent of **741** is fed by line **751** to amine cooler **752** which in turn feeds Absorber **724** via line **753**.

[**0490**] Recycle hydrogen, and C<sub>1</sub> and C<sub>2</sub> hydrocarbons flow through line **725** to compressor **732** and are mixed with make up fresh hydrogen in line **725A** at a pressure of 450 psig and a temperature of about 100° F. The recycle gas stream is also at a pressure of 450 psig, which is the lowest pressure in the system. The compressor **732** is driven by a high pressure steam turbine **763**. High pressure steam supply in line **762** comes from the fired heater steam system at 900 to 1500 psig and a temperature of 800° F., which is superheated by 200° F. in the fired heater **735**. Exhaust steam in line **764** is condensed in condenser **765** and along with make up water is fed to the fired heater **735** for preheating and reuse as boiler feed water make up.

[**0491**] The compressor **732** pressurizes the recycle hydrogen mixture and make-up hydrogen from 450 psig to approximately 670 psig and 187° F. and discharges the

hydrogen mixture into line 733. A portion of the hydrogen mixture in line 733 is the first hydrogen mixture and is delivered to heat exchanger 754 via line 733A. Another portion of the hydrogen mixture in line 733 is the second hydrogen mixture and is delivered to the in-out heat exchanger 715.

[0492] The heat exchanger 754 cools the first hydrogen mixture from about 187° F. to 100° F. A portion of the first hydrogen mixture in line 757 flows into line 756 and to a C<sub>1</sub> and C<sub>2</sub> hydrocarbon pressure swing adsorption ("PSA") system 755. The PSA system helps to maintain the C<sub>1</sub> and C<sub>2</sub> hydrocarbon level in the first and the second hydrocarbon mixture at about 2 vol %-3 vol %, or 2 vol %-6 vol %, also 6 vol %-8 vol % can be experienced. C<sub>1</sub> and C<sub>2</sub> hydrocarbons purged from the first hydrocarbon mixture are discharged through line 758 and combined with the gas in line 722 which is delivered to the fired heater 735. Purified hydrogen produced by the PSA 755 flows through line 756A and back to the suction of compressor 732 via line 725. 757 is the lock hopper motivation gas line.

[0493] The second hydrogen mixture is preheated to 875° F. in the in-out heat exchanger 715 by the overhead stream at 935° F. Preheated air conveyed through feed line 783 is combusted with fuel in the fired heater 735 and elevates the temperature of the second hydrogen mixture that is conveyed through line 734 from in-out heat exchanger 715. The fuel that is combusted is obtained from the natural gas line 759 and purge gas line 722. The hydrogen mixture circulates through the fired heater 735 and exits through line 736. The second hydrogen mixture provides the heat to maintain reaction in the reactor 705.

[0494] Waste heat from the radiant section of direct fired heater 735 is recovered in convection sections 735A, 735B and 735C. Steam and water are discharged from a steam drum 760 through the fired heater 735 convection section. 772 is the circulating boiler water line. Steam and water is returned to the drum via line 773. Steam separated from water in drum 760 is discharged into line 761 and introduced into convection section 735A where the steam temperature is raised from about 596° F. to about 800° F. After passing through convection section 735A, the superheated, high pressure steam is conveyed through line 762 to drive the steam turbine 763. Reduced temperature and pressure steam from turbine 763 is conveyed to steam condenser 765 and the condensate recirculated via line 767. The flow from pump 766A is conveyed through line 767 and combined with make-up water. The water being conveyed in line 767 is introduced into convection section 735C, heated and discharged through line 736 for further processing, such as deaeration. 769 is the preheated boiler feed water line to deaeration. 784 is the fired heater stack gases line.

[0495] The following table shows material flows and operating conditions an operating reactor system.

[0496] Fluidized Bed Reactor:

- [0497] Reactor (fluidized bed) Temperature 950° F.
- [0498] Reactor (fluidized bed) Pressure 600 psig
- [0499] H<sub>2</sub> Recycle Ratio 21.09
- [0500] Catalyst Flow Rate into Reactor 1255.07 lbs/hr
- [0501] Oil sand (tar sand) Flow into Reactor 2520 tons/hr

- [0502] Oil sand (tar sand) Feed Inlet Temperature 50° F.
- [0503] Hydrogen Mixture Flow Rate into Reactor 60.4 MMSCF/hr
- [0504] Hydrogen Mixture Gas Inlet Temperature 1500° F.
- [0505] Rotary Coolers:
  - [0506] Air Temperature at Intake 50° F.
  - [0507] Air Temperature (exiting) 623° F.
  - [0508] Sand Entering Temperature 950° F.
  - [0509] Sand Exiting Temperature 665° F.
- [0510] In-Out Heat Exchanger:
  - [0511] Overhead Stream Entering Volumetric 58.31 MMSCF/hr
  - [0512] Flow Rate
  - [0513] Overhead Stream Entering Temperature 950° F.
  - [0514] Overhead Stream Exiting Temperature 516° F.
  - [0515] Hydrogen Mixture Entering Flow Rate 60.4 MMSCF/hr
  - [0516] Hydrogen Mixture Entering Temperature 185° F.
  - [0517] Hydrogen Mixture Exiting Temperature 875° F.
  - [0518] FIRED HEATER:
    - [0519] Fuel Consumption (natural gas equivalent) 1.2 MMSCF/hr
    - [0520] Vapor Let-Down & PSA Off Gas Fuel Supply 0.56 MMSCF/hr
    - [0521] (natural gas equivalent)
    - [0522] Make-up Fuel Supply 0.64 MMSCF/hr
    - [0523] Combustion Air Entering Flow Rate 26.59 MMSCF/hr
    - [0524] (@65% excess)
    - [0525] Steam Production Rate (@ 1500 psig) 228,996 lbs/hr
    - [0526] Hydrogen Mixture Entering Flow Rate 60.4 MMSCF/hr
    - [0527] Hydrogen Mixture Entering Temperature 875° F.
    - [0528] Hydrogen Mixture Exiting Temperature 1500° F.
  - [0529] COMPRESSOR:
    - [0530] Power Required From Turbine 40,148 h.p.
    - [0531] Steam Flow Rate to Turbine 228,996.1 lbs/hr
    - [0532] Steam Pressure Entering Turbine 1500 psig
    - [0533] Steam Temperature Entering Turbine 800° F.
    - [0534] (200° F. superheat)
    - [0535] Hydrogen Mixture Entering Flow Rate 60.7 MMSCF/hr
    - [0536] Hydrogen Mixture Compressor Entering 100° F.
    - [0537] Temperature

- [0538] Hydrogen Mixture Compressor Entering 450 psig
- [0539] Pressure
- [0540] Hydrogen Mixture Compressor Exiting 185° F.
- [0541] Temperature
- [0542] Hydrogen Mixture Compressor Exiting 670 psig
- [0543] Pressure
- [0544] Product Condenser/Separator:
- [0545] Product Fluid Stream Entering Flow Rate 58.3 MMSCF/hr
- [0546] Product Fluid Stream Entering Temperature 516° F.
- [0547] Recycle Hydrogen Mixture Exiting 100° F.
- [0548] Temperature
- [0549] Synthetic Crude Oil Flow Rate 1255 bbl/hr
- [0550] Amine System:
- [0551] Amine Recirculation Flow Rate 50,400 lbs/hr
- [0552] Ammonia Production 1478 lbs/hr
- [0553] Elemental Sulfur Production 17260 lbs/hr

[0554] One aspect of the invention is the ability to process various feedstocks (feeds) having hydrocarbon content. Examples of feedstock alternatives and example combinations in which they are present are provided in Table 1: In Table 1, the alternative feedstocks are shown to possess at least the components indicated as "X" in each respective alternative. The alternative feed components included in Table 1 include bitumen, oil sand, sand, raw kerogen oil and oil shale. The feedstocks can contain other compounds and are not limited to the components identified in table 1.

TABLE 1

Alternative Feedstock	Feedstocks And Feedstock Combinations				
	Bitumen	Oil Sand	Sand	Raw kerogen oil	Oil Shale
1	X				
2		X			
3				X	
4					X
5	X	X			
6	X	X	X		
7				X	X
8			X	X	X
9	X			X	
10		X			X
11	X	X	X		X
12	X	X	X	X	
13	X	X	X	X	X

[0555] The bitumen concentration of raw oil sand can vary. Whether an oil sand is considered an economically attractive feed can vary depending upon the market value of the SCO product which can be derived from a given feedstock(s). The present invention can process oil sands without limitation regarding bitumen content. Oil sands containing less than 8 wt % bitumen are considered to be of lower quality. However, the invention can process low grade and

very poor quality oil sands. In one embodiment, the bitumen feed to the reactor can be processed concurrently with very low grade oil sands to produce an economically valuable product. The processing range of oil sand feedstocks can have bitumen in a range from little or no bitumen to the richest deposits available (e.g., 0 wt % to more than 18 wt %, or higher).

[0556] The production rate of the plant can be maintained constant by adjusting the rate of hot liquid bitumen injection via conduit 7103. Low grade oil sands would require a higher rate of SAGD or wet extraction bitumen injection to maintain a constant production level of SCO for example, 60,000 bbl/day or 30,000 bbl/day depending on plant design capacity. Bench scale (5 bbl/day-5,000 bbl/day) and pilot plant scale (5,000-30,000 bbl/day) designs producing very small or medium size capacities can be operated.

[0557] The invention can be built new or can be retrofitted into an existing plant. Further, in some embodiments, the operation of existing plants can be modified to achieve the present invention.

[0558] In one embodiment, run of mine oil sand (from trucks, or other source(s)) is dumped into receiving equipment which can include screening and sizing equipment 7002 for classifying oil sand. The run of mine sand (oil sand) enters the process at ambient temperature.

[0559] The oil sand is screened to remove rocks and debris which can include boulders, stones tree trunks and other debris such as animal bones or manmade items. The oil sand is crushed into fluidizable pieces that are in the range of ¼ to 1 inch mesh. Such pieces are oil sand pieces that greater than ¼ inch and smaller than 1 inch in size. The pieces in one embodiment are of particle sizes that would not pass through a ¼ inch mesh screen but would fall through a 1 inch mesh screen. Smaller or larger pieces can be used in other embodiments.

[0560] Limiting the amount of crushing (e.g., not crushing the pieces too finely) reduces heat generation which can adversely affect oil sand feeding to the reactor. Such limitation can avoid the build-up of heat in the material being crushed. Crushing generates heat because work is performed on the material. Less crushing generates less heat. Heat generated in crushing oil sand can cause the bitumen contained in the oil sand to become more tacky and sticky. Sticky oil sand does not flow well. It bridges in hoppers and plugs flow lines. The flow of oil sand into the reactor can be impeded if the oil sand is too sticky.

[0561] The run of mine (optionally strip mined) oil sand provides both raw bitumen and sand.

[0562] The crushed oil sand is conveyed through conduit 7003 to feed hoppers 7004 which provide the mechanical means for feeding solids at ambient atmospheric conditions into the pressurized fluid bed reactor 7005. The oil sand is fed through a lock hopper system 7004 used for feeding the reactor 7005. Solid feed motivation is provided by injecting cool and higher pressure hydrogen via conduit 7057 into the lock hoppers 7004. The feed flow through conduit 7003 and between feed lock hoppers 7004 is controlled by ball valves 7003A. The valves' interior surfaces are hard lined for handling very abrasive materials.

[0563] Upon being fed to the reactor, the oil sand is contacted with the gas phase present in the reactor. Upon

contacting the gas phase present in the reactor which includes a hot fluidizing flow of hydrogen the bitumen of the oil sand can experience a change of phase of its hydrocarbon component (oil sand bitumen) and/or chemical reaction with hydrogen or other hydrocarbons present in the reaction mixture. During this process collectively termed "reaction process", the water film of the oil sand bitumen vaporizes (in some instances vaporization is explosive) causing the oil sand pieces to disintegrate, exposing more unreacted material and surface area for reaction.

[0564] At the desired reactor conditions, the bitumen in the oil sand and injected liquid feed reacts with the hydrogen by endothermic hydrocracking and exothermic hydrogenation. These reactions are approximately in balance and once the reactor feed is brought up to about 900° F. by the hot hydrogen stream 7036 from the fired heater 7035, the reactions can be self-sustaining. This is considered to result from the endothermic hydro cracking reactions absorb heat, while the exothermic hydrogenation reactions produce heat. The production of heat approximately offsets the absorption of heat and therefore the reactions are approximately in balance with respect to heat requirements. The endothermic and exothermic reactions, i.e., The exothermic reactions support the endothermic reactions resulting in self-sustaining reactions.

[0565] Fluidizable feed enters reactor 7005 and forms a fluidized bed when it is in contact with the fluidizing medium. The fluidized bed is comprised substantially of fluidizable feed and its reaction products. The reaction process is continuous and fluidized feed is continuously reacted to produce reaction products which exit the reactor. The feed material comprising the fluidized bed is continuously reacted and the fluidized bed is continuously replenished by the fluidizable feed entering the reactor and coming into contact with the fluidizing medium to becoming fluidized and form the fluidized bed (become one with the fluidized bed). Catalyst may be optionally added as a feed.

[0566] In one embodiment, the fluidized bed reactor is free of a contained fixed catalyst bed and has a fluidized bed comprising substantially feed. The reactor has a feed inlet for a fluidizable feed and a fluidizing medium inlet for a gas comprising hydrogen. The fluidizing medium in contact with said fluidizable feed forms a fluidized feed, and the fluidized feed forms the fluidized bed in which the fluidized bed comprises at least the fluidized feed. The fluidizable feed can be any fluidizable feed disclosed herein.

[0567] Embodiments of the reactor of this process can be of the flow-through type allowing for the flow of products out of the reactor and the entrance of feeds forming the fluidized bed free of (not including) a fixed bed of catalyst, fixed bed of inerts or other fixed bed of other material.

[0568] The vaporized and/or reacted hydrocarbons of the oil sand bitumen exit the reactor 7005 through the reactor overheads line 7016.

[0569] The oil sand which has been processed produces spent sand which can be considered to be oil sands from which an amount of bitumen has been extracted.

[0570] The spent sand exiting the reactor via 7005D and 7006 is the medium by which coke and heavy metals are ejected from the process and thereby avoid entering into the

product SCO. The heavy metals combine with the coke and the coke forms a light coating on the spent sand particles.

[0571] Hydrogen is heated to about 1500° F. in fired heater 7035 and is introduced to the reactor by conduit 7036 providing fluidization of the fluidizing medium. Hot liquid bitumen at about 300° F. (optionally from SAGD, wet extraction operations, or other processes) is separately introduced via conduit 7103 by pump 7102 to the reactor where intimate contact with gaseous hydrogen feed to the fluidized bed occurs. Intimate contact refers to the high likelihood of reactants being exposed to hydrogen in the reactor environment. The hot bitumen feed temperature is not limited and can be of any temperature where the bitumen can be pumped. One embodiment employs a temperature range of about 200° F. to about 350° F. The result of this contact is the vaporization of the bitumen feed in to gaseous bitumen which undergoes further reaction or exits the reactor through vapor overheads line 7010.

[0572] Reactor product vapors and reaction gasses exit the reactor via conduit 7010 and high grade 34° API, low sulfur (about 0.1 to 0.5 wt. % sulfur, or less; one embodiment experiences about 0.2 wt. %) oil is condensed and processed downstream.

[0573] The invention can process any type of bitumen which can be fed to the reactor The invention will process a broad range of material ranging from raw sand which contains no bitumen to oil sand containing in excess of 18 wt % bitumen.

[0574] Another embodiment of the invention processes a bitumen feedstock including bitumen which has been sourced or produced from oil sand bitumen which has been extracted from under ground via in situ processes. Another embodiment of the invention processes a bitumen feed stock including bitumen which has been sourced or produced from oil sand in situ. This production technique applies to oil sand too deep in the ground to be mined by open pit mining.

[0575] Another embodiment of the invention processes a bitumen feedstock including bitumen which is derived from oil sand, whether in situ or previously mined, which has been extracted by a hot water extraction process before upgrading.

[0576] Another embodiment of the invention processes a bitumen feedstock including bitumen which is derived from oil sand, previously mined wherein the bitumen is separated from the oil sand by a hot water extraction process.

[0577] Bitumen is also produced in situ by other thermal methods such as cyclic steam stimulation or "huff and puff" technology. This involves injecting steam into the deeply buried oil sands for several weeks to melt the bitumen. The steam injection is then stopped and hot bitumen is pumped to the surface through the same well bore. The production phase between cycles of steam injection can last for several months.

[0578] Another in situ production process for recovering heavy oil and bitumen is using hot water and saturated hydrocarbon solvent vapor (e.g., having ethane, propane or butane) in conjunction with horizontal wells to mobilize and recover viscous oil and bitumen from hydrocarbon deposits. The process involved injecting a hydrocarbon solvent into the reservoir along with a displacement gas to mobilize the

hydrocarbons in the reservoir and move them toward the production well. This can be referred to as a vapor production process.

[0579] The invention can be used to remediate any materials contaminated with hydrocarbons. For example sand, earth, soil, rocks, and other solids can all respectively be contaminated with liquid hydrocarbons. Feeding such remediation feedstocks into the reaction will remediate the materials. Further, remediation feed could be fed alone or mixed with other feedstocks such as oil sand or shale.

[0580] Remediation feedstocks are not limited to solids or mixtures of liquids and solids, liquid remediation is also possible. For example, motor oil or other liquid hydrocarbons can be blended with bitumen, or fed separately to the reactor of the invention.

[0581] Kerogen oil can be processed by the invention the same way as bitumen.

[0582] Kerogen oil (raw or processed) can also be blended with bitumen feed and fed to the reactor.

[0583] In various embodiments of the invention, any solid feeds can be mixed and any liquid feeds can be mixed. This provides for varied blends of solid and liquid feed to be fed to the reactor. In some embodiments the blend compositions can be changed over time and additionally can be changed over the course of a run in some embodiments without requiring shutdown.

[0584] Referring to FIG. 7-3, the bitumen feedstock or bitumen feedstock blend is in bitumen storage tank 7100 from which it is pumped by pump 7101 to reactor 7005. As discussed above, the bitumen fed to reactor 7005 experiences vaporization and/or reaction.

[0585] In one embodiment, referencing FIG. 7-5, run of mine oil shale feed 7600 and kerogen oil feed 7103 are processed in reactor 7005 in the presence of a reactor gas phase including hydrogen.

[0586] Run of mine oil shale (from trucks, or other source(s)) is dumped into receiving equipment which can include screening and sizing equipment 7002 for classifying oil shale. The run of mine shale (oil shale) enters the process at ambient temperature.

[0587] The oil shale is screened and can be crushed into fluidizable pieces that are in the range of ¼ to 1 inch mesh. The run of mine (optionally strip mined) oil shale provides both raw kerogen oil and shale

[0588] The crushed oil shale is conveyed through conduit 7003 to feed hoppers 7004 which provide the mechanical means for feeding solids at ambient atmospheric conditions into the pressurized fluid bed reactor 7005. The oil shale is fed through a lock hopper system 7004 be used for feeding the reactor 7005. Solid feed motivation is provided by injecting cool and higher pressure hydrogen via conduit 7057 into the lock hoppers 7004. The feed flow through conduit 7003 and between feed lock hoppers 7004 is controlled by special ball valves 7003A. The valves' interior surfaces are hard lined (abrasion resistant) for handling very abrasive materials.

[0589] Upon being fed to the reactor, the oil shale is contacted with the gas phase present in the reactor. Upon contacting the gas phase present in the reactor which

includes a hot fluidizing flow of hydrogen the kerogen of the oil shale can experience a change of phase of its constituents (oil shale kerogen) and/or chemical reaction. This process is collectively termed as the "reaction process"

[0590] At the desired reactor conditions, the kerogen in the oil shale and injected liquid feed reacts with the hydrogen by endothermic hydrocracking and exothermic hydrogenation. These reactions are approximately in balance and once the reactor feed is brought up to about 900° F. by the hot hydrogen stream 7036 from the fired heater 7035, the reactions are self-sustaining. In one embodiment, desired reactor conditions can be a temperature of from 840° F. to 915° F., or up to 1000° F., or higher, and having pressures of about 550 psig to 635 psig, or up to 700 psig.

[0591] The vaporized and/or reacted hydrocarbons of the oil shale kerogen exit the reactor 7005 through the reactor overheads line 7001.

[0592] The oil shale which has been processed produces spent shale. The spent shale exiting the reactor via 7005D and 7006 is the medium by which coke and heavy metals are ejected from the process and thereby avoid entering into the product SCO. The heavy metals combine with the coke and the coke forms a light coating on the spent shale particles.

[0593] Hydrogen is heated to about 1500° F. in fired heater 7035 and is introduced to the reactor by conduit 7036 providing fluidization of the fluidizing medium. Hot kerogen at about 300° F. is separately introduced via conduit 7103 to the reactor where intimate contact with gaseous hydrogen present in the fluidized bed occurs. The result of this contact is the vaporization of the kerogen feed in to gaseous hydrocarbons which can undergo further reaction or exits the reactor through vapor overheads line 7010.

[0594] Product vapors and gasses exit the reactor via conduit 7010 and high grade 34° API, low sulfur (about 0.1 to 0.5 wt. % sulfur, or less; one embodiment experiences about 0.2 wt. %) oil is condensed from the gas stream.

[0595] In one embodiment of the invention, kerogen oil which has been extracted from oil shale is processed. Alternatively, kerogen oil can be derived from oil shale, whether in situ or previously mined.

[0596] The above-identified kerogen containing feedstocks can be processed separately, or in a mixture.

[0597] Kerogen is found in oil shale. Oil shale occurs in many parts of the world. The US contains the world's largest deposits of oil shale as well as some of the richest in the western states of Colorado, Utah and Wyoming. Large shale deposits are also found in the eastern US.

[0598] Kerogen content in shale varies but is not identified as such since it is a solid embedded in the shale. The richness of oil shale is normally expressed by the amount of raw oil obtained from the processing of the shale. Western shales generally yield 30 to 40 gal of raw kerogen oil per ton of shale. Eastern shales yield about 10-20 gal per ton of shale.

[0599] Kerogen containing feedstocks can be combined with bitumen and processed by the invention.

[0600] Feed lock hoppers 7004 and spend sand or shale lock hoppers 7006 feed solid materials at ambient conditions into reactor 7005 at about 600 psig pressure and to remove spent solids from the pressurized reactor 7005 to ambient

conditions on a continuous operating basis. Hydraulically operated special ball valves **7003A** open and shut on a pre-determined sequence for pressuring and depressuring the hoppers to permit solids to flow into and out of the reactor **7005** without the loss of pressure in the reactor. The number of feed lock hopper systems and spent material lock hoppers is dependant on the size and throughput of reactor **7005**.

[**0601**] The reactor **7005** is operated so as to highly agitate the reactants to promote rapid and intimate contact between the bitumen or kerogen components and hydrogen to ensure complete reaction takes place. The residence or reaction time of the oil sand or shale is between 3 and 15 minutes depending on the throughput and efficiency of the reactor process which is dictated by the analysis of the various feed components. The pressure drop across the fluidized bed in reactor **7005** is about 25 to 40 psi.

[**0602**] In one embodiment, reactor **7005** contains a bed which is fluidized and which can include either the oil sand feed, or include other fluidizable materials. In some embodiments having oil sand or sand as the fluidization material, the fluidized sand particles provide an inert, non-reactive carrier or base promoting contact between the hydrogen and the liquid bitumen for reaction to take place.

[**0603**] Spent sand or shale from the fluid bed in reactor **7005** is displaced by incoming feed to the bottom of the reactor and overflows at the top of the reactor into conduit line **7005D**. The height of the conduit line above the reactor grid **7005C** establishes the maximum depth of the fluidized bed in reactor **7005**. The hot spent sand or shale then flows through lock hoppers **7006**, through conduit line **7007** and into rotary coolers **7008** where the solids are cooled from reactor temperature to about 300° F. The cooler discharge **7009** can then be back hauled by conventional equipment for land reclamation or be slurried with water and/or pumped to a settling location.

[**0604**] On entering the reactor **7005**, oil sand or sand is contacted and heated by the combined recycle and make-up hydrogen introduced via line **7036** from the fired heater **7035**. The fired heater elevates the temperature of the hydrogen stream to about 1500° F. at about 635 psig and after entering the bottom of the reactor **7005B**, the hydrogen passes through a slotted fire brick or ceramic grid **7005C** which supports the fluid bed of sand and aids in evenly distributing the hydrogen stream into the fluidized bed. The hydrogen flow rate and velocity of between 2 and 10 ft./sec. is sufficient to fluidize the oil sand or sand bed. The high temperature hydrogen heats the oil sand or sand from ambient to a reaction temperature of about 850° F. to 925° F. at a pressure of about 500 to 650 psig.

[**0605**] The volume of recycle hydrogen compared to fresh make-up hydrogen i.e. the recycle ratio, is a function of a number of operating variables including:

[**0606**] Hydrogen velocity to fluidize the contents of the reactor **7005** to create a fluid bed.

[**0607**] The amount of heat to bring the oil sand, bitumen or shale feed up to reaction temperature of about 900° F. Hydrogen is the ideal medium by which heat is conveyed into the reactor **7005** having the highest heat capacity of any gas.

[**0608**] Temperature of the hydrogen from the fired heater **7035** in line **7036** normally heated to about 1500° F.

[**0609**] Residence time of the reactants in the reactor **7005** (reaction time) for optimum conversion of the bitumen or kerogen to product vapors.

[**0610**] The recycle ratio is preferably about 21:1 but can vary from about 5:1 to about 30:1 depending upon which of the process embodiments is being operated. The combined recycle hydrogen and make-up hydrogen are compressed by the high speed centrifugal compressor **7032** and discharged into line **7033** at a pressure of about 670 psig and 187° F. The compressor **7032** discharge stream is divided into two streams: a smaller stream flows via conduit line **7033A** to gas cooler **7054**; a much larger stream representing about 85-90 wt. % of the total discharge is delivered to the in-out heat exchanger **7015**.

[**0611**] The cooled hydrogen gas in line **7057** exiting from gas cooler **7054** is again divided into two streams. The larger portion of this stream is delivered to the reactor feed lock hopper system **7004** at about 100° F. and is used to motivate the oil sand feed through the lock hoppers **7004** into the reactor **7005**. When sand only or oil shale is fed, cooling of this motivation stream is not necessary and this gas is then by-passed around gas cooler **7054** and used at compressor discharge conditions of 670 psig and 187° F.

[**0612**] The smaller portion of the cooled 100° F. hydrogen stream exiting gas cooler **7054** representing about 6 wt. % of the recycle gas entering compressor **7032** flows via conduit line **7056** to the pressure swing adsorption (PSA) system **7055**.

[**0613**] While the bitumen feed or the bitumen in the oil sand can contain heavy metals, such as nickel, which can catalytically promote endothermic and exothermic reactions in reactor **7005**, supplemental catalyst, such as nickel, cobalt or molybdenum, can be added through feed conduit **7004A** to feed lock hopper **7004** to promote additional catalytic reaction. The reactor and related equipment are shown in more detail in **FIG. 8-0**.

[**0614**] The invention can include the following unit operations:

Hot gas clean-up system	7011
In-out heat exchanger	7015
Product separation	7019,
	7021
Hydrotreater	7200
Amine adsorption system	7400
Steam turbine	7063
Hydrogen circulation compressor	7032
Pressure swing adsorption (PSA)	7055
Fired heater	7035
Spent sand coolers	7008

[**0615**] In an embodiment having these unit operations, high grade synthetic crude oil can be produced at about 850-950° F. temperature and pressure of about 600 psig.

[**0616**] The recycle hydrogen **7025** and make up hydrogen **7025A** from the hydrogen plant **7300** are conveyed to compressor **7032**. A side stream of combined recycle and

make-up hydrogen exits compressor **7032** via lines **7033** and **7033A** and is cooled by heat exchanger **7054**. This cooled stream can be discussed later since it is used in various requirements.

[**0617**] On entering the reactor **7005**, oil sand or sand is contacted and heated by the combined recycle and make-up hydrogen introduced via line **7036** from the fired heater **7035**. The fired heater elevates the temperature of the hydrogen stream to about 1500° F. at about 635 psig and after entering the bottom of the reactor **7005B**, the hydrogen passes through a slotted fire brick or ceramic grid **7005C** which supports the fluid bed of sand and aids in evenly distributing the hydrogen stream into the fluidized bed. The hydrogen flow rate and velocity of between 2 and 10 ft./sec. is sufficient to fluidize the oil sand or sand bed. The high temperature hydrogen heats the oil sand or sand from ambient to a reaction temperature of about 850° F. to 925° F. at a pressure of about 500 to 650 psig.

[**0618**] The hydrogen flow rate from the fired heater **7035** typically exceeds the stoichiometric requirements for bitumen conversion with hydrogen by a factor of about 5 to 30. The recycle rate is governed by many variables such as temperature of the solids fed to the reactor and the ratio of oil sands or sand to hot bitumen feed. Since oil sand typically contains about 10 wt. % bitumen and 85 wt. % sand, increasing bitumen feed injection would require less sensible heat for a constant production rate and hence a lower ratio of recycle hydrogen to make-up hydrogen would be necessary.

[**0619**] Fresh make-up hydrogen produced on site in a conventional steam-methane reform hydrogen plant **7300** is admitted into compressor **7032** via conduit line **7025A** and joins the recycle hydrogen from the amine scrubbing system **7400** delivered to the compressor via line **7025**. Both streams enter the compressor **7032** at about 100° F. and 500 psig.

[**0620**] The ambient air used in rotary coolers **7008** to cool the spent sand or shale exits the coolers at a temperature of about 625° F. via conduit line **7079**. It is highly advantageous to the overall process efficiency to use this hot air for combustion air in the fired heater **7035** and in the hydrogen plant **7300**. For this purpose, the hot air is cleansed of entrained fines by passing through a cyclone **7080** followed by a bag collector. The fines are discharged through conduit line **7085** to be returned to land reclamation along with the spent sand or shale **7009**. Hot combustion air blower **7082** supplies the fired heater **7035** with its requirements via conduit line **7083** and the hydrogen plant **7300** via conduit line **7301**.

[**0621**] Reaction products include hydrocarbon vapors that are condensable into hydrocarbon liquids of excellent quality in terms of API gravity, sulfur content, and distillation. The product is stable and in one embodiment contains no material boiling above 975° F. It is pipe-linable and constitutes an excellent refinery feed.

[**0622**] Reactor effluent gases including product vapors, excess recyclable hydrogen, methane, ethane, propane, hydrogen sulfide and water vapor exit the reactor **7005** as an overhead or product stream through cyclone separators **7005A** and into exit conduit line **7010**. Solids entrained in the overhead product stream, such as sand particles, shale

and fines, are removed from the product stream by the cyclone separators **7005A** and recycled to the bottom of the reactor near the ceramic support grid **7005C** where they are again entrained in the fluidized contents of the reactor.

[**0623**] The reactor overhead stream from the cyclone separators **7005A** flows via line **7010** to the hot gas clean-up system **7011**. Here the hot gas stream at about 900° F. is filtered to remove and collect any fines that have escaped the reactor cyclone separators **7005A**. The ceramic or sponge iron filters are periodically and section by section back flushed or pulsated by a higher pressure hydrogen flow via line **7034A** from the in-out heat exchanger **7015**. Collected fines and solids are removed from the bottom of hot gas filter **7011** by lock hoppers **7012** which operate in similar manner as the lock hoppers on the reactor **7005**. The fines are combined with spent sand or shale and used for land reclamation. The disposal of the dry sand or shale and fines resulting from the invention is environmentally preferable to existing wet disposal systems.

[**0624**] The filtered solids free product stream flows from the hot gas clean-up system through line **7013** to the in-out heat exchanger **7015**. This exchanger is an indirect, shell and tube, heat exchanger wherein heat is recovered from the reactor overhead product stream and transferred to the purified recycle and fresh make-up replenished hydrogen stream being returned to the reactor. This heat exchange is a major part of total process waste heat recovery and is vital to the overall process efficiency. The heat duty of in-out exchanger **7015** is nearly equal to that of the fired heater **7035**. The heated product free hydrogen stream is conveyed via conduit line **7034** to the fired heater **7035** for additional heating prior to admission to the reactor. The cooled overhead product stream exits the in-out heat exchanger **7015** through line **7016** at about 450° F.

[**0625**] The PSA system **7055** utilizing resins removes the light hydrocarbons, methane and ethane, and also ammonia from the entering gas stream by adsorption of these gases onto the resin. The purified hydrogen stream leaves the PSA system **7055** through line **7056A** and flows back to the suction of the compressor **7032** via line **7025**.

[**0626**] When the PSA resin has reached its capacity to adsorb the undesirable gases, the stream being purified is diverted to another vessel in the system. The pressure is reduced in the vessel removed from service so that it can be regenerated and returned to service later. The adsorbed C<sub>1</sub>, C<sub>2</sub> and ammonia gases are removed from the resin by a hydrogen gas purge stream which is then passed through a chiller to condense and knock out the ammonia. The ammonia is removed from the purge gas stream to avoid burning it with the fuel in the fired heater **7035** and creating air pollutants. The PSA purge stream cleansed of ammonia is discharged from the PSA system **7055** through line **7058** and combines with the flash vapors in line **7022** from the product let-down vessel **7021**. These purged and flashed off gases flow together to the fuel system for combustion in heater **7035**. Purging the recycle hydrogen stream **7025** of C<sub>1</sub>, C<sub>2</sub>, and ammonia gases is necessary to prevent their accumulation in the system where they act as inerts and suppress the partial vapor pressure of hydrogen. Purging of the system is regulated to maintain a maximum concentration of about 6 wt. % inerts in the recycle stream.

[**0627**] The main compressor discharge stream of purified recycle and make-up hydrogen delivered to the in-out

exchanger **7015** is heated from about 187° F. to about 875° F. by exchange with the reactor overhead stream. The gas exits the in-out exchanger and flows via line **7034** to the radiant section of fired heater **7035**. The temperature of the reactor feed gas is elevated to about 1500° F. as it circulates through the fired tubes of heater **7035** and then flows via conduit line **7036** to the reactor **7005** to repeat the cycle.

[**0628**] Producing the hydrogen recycle involves filtration of the hot exiting vapors (e.g., 800° F. to 1000° F.) to remove particulates, acid gas removal in an amine absorption system, removal of reaction inert gases in a pressure swing adsorption (PSA) system and recompression of the purified recycle stream from a high pressure level (400 psig-700 psig) to minimize energy requirements.

[**0629**] The hot reactor effluent vapors at about 915° F. (range 850° F. to 1000° F.) can contain entrained particles of sand or shale that have escaped the internal cyclones. These entrained particulates have to be removed from the reactor effluent stream to avoid contamination of the liquid product that is eventually condensed down stream in the process.

[**0630**] The hydrogen recycle stream from the amine system and the hydrogen stream from the PSA system returning to the compressor suction for recompression are said to have been purified by the removal of the acid gases and purging of methane and ethane.

[**0631**] The inlet to the compressor can have the lowest pressure in the entire process circuit. It is desirable to have the compressor suction pressure as high as possible relative to the compressor discharge pressure in order to conserve energy. Less work is necessary to be performed by the compressor if the pressure differential or pressure drop across the gas circulating system is kept as low as possible. The compressor discharge pressure is normally operated at about 670 psig and the suction pressure at about 500 psig but can go as low as 450 psig.

[**0632**] Heat in the combustion gases from the radiant section of the direct fired heater **7035** is recovered in the convection sections **7035C**, **7035B** and **7035A** whose function is to preheat boiler feed water, generate high pressure steam and superheat steam, respectively. Boiler feed water from the water treating section and condensate returned from various steam condensing services in the plant is preheated in section **7035C** and flows to a de-aerator via line **7069** and is then pumped to steam drum **7060**. Boiler water is circulated from the steam drum **7060** via line **7072** to convection section **7035B** where high pressure steam is generated in the boiler tubes by exchange with the flue gases exiting the fired heater **7035**. A mixture of water and steam is returned via line **7073** to the boiler drum **7060** to separate the steam from the water. Steam system pressure can be designed to suit the operator and can vary from 800 psig to 1500 psig. In this example, operating pressure was chosen to be 1500 psig.

[**0633**] High pressure steam at 1500 psig flows from steam drum **7060** via line **7061** to convection section **7035A** where it is superheated from saturation temperature 596° F. to about 800° F. providing about 200° F. of superheat. The superheated steam flows via line **7062** to the high speed turbine **7063** which drives the centrifugal hydrogen compressor **7032** that circulates hydrogen around the process circuit. Exhaust steam from turbine **7063** is condensed in

condenser **7065** and the resulting condensate is returned to the steam system via condensate pump **7066A** and line **7067** for reuse as boiler feed water.

[**0634**] Optionally, additional heat recovery, not shown in the flow diagram, can be achieved down stream of in-out exchanger **7015** and before the partially cooled overhead product stream is admitted via conduit line **7016** to condenser **7017**. Low pressure steam can be generated for electric power generation, driving small turbines, pumps, and heating requirements within the plant complex. Any waste heat recovered at this point is advantageous to the overall process efficiency and applies to all three embodiments of this invention.

[**0635**] The partially cooled overhead product stream in line **7016** enters condenser **7017** where condensable vapors and gasses are condensed. The cooling medium is circulated cooling water from an on-site cooling tower. Condensed liquids and the non-condensable gases at about 100° F. flow via line **7018** to a separator **7019** where water is separated from the condensed product liquids and gas stream. The amount of sour water produced varies depending upon the water content of the feed materials being processed in the reactor **7005**. Any entrained oxygen in the reactor feed is converted to water in the predominately hydrogen atmosphere in the reactor. The waste water exits the separator **7019** via line **7086** and is treated to conform to environmental requirements.

[**0636**] The condensed liquid product from separator **7019** flows via conduit line **7020** to the letdown vessel **7021** where pressure reduction takes place and dissolved light gases, methane and ethane, are flashed off. The flashed off gases are conveyed via conduit line **7022** to the fuel system of the fired heater **7035** for combustion. The liquid hydrocarbon products are removed from let-down tank **7021** via flow line **7090** as a light low sulfur (about 0.1 to 0.5 wt. % sulfur, or less) synthetic crude oil product stream comprising a mixture of naphtha and gas oils having an API gravity of approximately 30° to 34°. As discussed earlier, the quality of the product SCO including sulfur content can be modified by altering process variables and the inclusion or non-inclusion of various catalyst formulations to suit the requirements of the refiner.

[**0637**] The crude oil product in line **7090** can be sent directly to storage without further treatment or it can be sent to a conventional hydrotreater **7200** for further upgrading to reduce sulfur content to 0.1 wt. % and improve API gravity to about 34°. Low sulfur (about 0.1 to 0.5 wt. % sulfur, or less) product from hydrotreater **7200** is stored separately and can be used as the primary fuel in the fired heater **7035** and the hydrogen plant **7300** reformer to reduce requirements for natural gas to hydrogen production only.

[**0638**] In an embodiment, this process invention provides the option of improving product quality by either adding catalyst with the feed to the reactor **7005** to promote catalytic hydrogenation or to further upgrade the product in the plant's integrated hydrotreater **7200**. Operational flexibility is provided by these options to produce a product with broader market appeal.

[**0639**] The hydrotreater **7200** receives hot hydrogen from the high pressure discharge side of compressor **7032** via the fired heater **7035** in conduit line **7201**. Relatively cool

hydrogen directly from the compressor discharge is provided via conduit line **7202** for temperature control since the hydrotreater process is exothermic. Excess hydrogen and the acid gas, hydrogen sulfide, produced in the hydrotreater **7200** flow, after being cooled to 100° F., via conduit line **7203** to the amine scrubbing system **7400** to remove the acid gas and recycle the excess hydrogen. Economic benefits can be realized by integrating the hydrotreater into this process invention as such integration makes possible the options of:

- [**0640**] Sharing of a common hydrogen circulation compressor **7032**,
- [**0641**] Sharing of a common amine scrubbing system **7400**,
- [**0642**] Sharing a common source of hydrogen **7300**, and/or
- [**0643**] Operational flexibility previously mentioned *vis a vis* catalyst addition to the reactor **7005**.

[**0644**] The reactor overhead stream now free of the condensed product and water vapors and comprising of excess hydrogen, light hydrocarbon gases methane and ethane and acid gases exit separator **7019** via conduit line **7023** at 100° F. and about 525 psig to flow to the amine scrubbing system **7400**. Here the stream is joined by the hydrotreater acid gas stream **7203** and the combined streams enter in to the acid gas absorber column **7024**. As the combined recycle gases rise up through the packed absorber column **7024** a counter current flow of cold di-ethanol amine solution (DEA) absorbs the acid gases, mainly H<sub>2</sub>S with trace amounts of CO<sub>2</sub> and the scrubbed recycle hydrogen along with the light hydrocarbon gases exit the absorber to flow through line **7025** to compressor **7032**.

[**0645**] The rich amine solution having absorbed the acid gases is discharged from the bottom of the absorber **7024** via line **7040** and flows through the lean-rich exchanger **7041**. Here the cold rich amine solution is heated by exchange with hot lean amine solution returning from amine regenerator **7043** to the absorber **7024**. The preheated rich amine solution flows through line **7042** and enters near the top of amine regenerator **7043**. In the regenerator **7043** a combination of pressure reduction and heat strips the acid gases from the amine solution as it descends the packed or trayed tower. Heat is applied to the amine solution in steam reboiler **7045** and with the reduction of pressure the absorbed acid gases are released to exit the regenerator via line **7044**. The hydrogen sulfide is then processed in a sulfur recovery plant where it is converted to elemental sulfur. The process is highly efficient recovering more than 99.98 wt. % of the acid gas sulfur.

[**0646**] Lean hot amine solution flows from regenerator **7043** in line **7048** to amine circulation pump **7049** which pressurizes the amine solution sufficiently to flow via line **7050** through lean-rich exchanger **7041** where it is cooled by exchange with the cold rich amine solution from the absorber **7024**. The operating principles in lean-rich exchanger **7041** and in-out exchanger **7015** are similar in that energy conservation is effected by recovering heat from streams that have to be alternately heated and cooled. The lean amine solution is further cooled in amine cooler **7052** and returned to the top of amine absorber **7024** via conduit line **7053** to repeat the cycle.

[**0647**] Optionally, the following design elements can be included in the invention:

- [**0648**] Integrating hydrogen plant **7300**,
- [**0649**] A common boiler feed water treatment system for the fired heater **7035** and the hydrogen plant reformer,
- [**0650**] Combustion air is preheated in rotary sand cooler **7003** for the fired heater **7035** and hydrogen plant reformer,
- [**0651**] Fuel requirements for the fired heater **7035** and hydrogen plant reformer can be met by the high grade low sulfur (about 0.1 to 0.5 wt. % sulfur, or less) product oil produced by this process,
- [**0652**] A common cooling water system can be shared, An integrated steam system is possible maximizing waste heat recovery for power generation,
- [**0653**] A common control room integrates all plant operations minimizing manpower requirements and improves operating efficiency, and/or
- [**0654**] Integration of support facilities such as administration, operations, maintenance, etc. is possible.

[**0655**] While particular embodiments of the present invention have been illustrated and described herein, the present invention is not limited to such illustrations and descriptions. It is apparent that changes and modifications can be incorporated and embodied as part of the present invention within the scope of the following claims.

[**0656**] Product quality can be varied by altering process variables such as temperature, pressure, reactor residence time, hydrogen recycle ratio, amount and variations in formulation of catalyst addition to the reactor feed, etc. The product quality can be modified to more adequately meet the requirements of a particular refiner's feed stock.

[**0657**] It is also possible to achieve the design of the present invention by retrofitting an existing process or facility to achieve the present invention.

### EXAMPLE 3

[**0658**] The process illustrated in **FIG. 7-6** and **FIG. 7-7** encompasses an embodiment of the present invention in which PSA tail gas is recovered ("PSA tail gas recovery embodiment"). Reference characters are provided in **FIG. 7-6** and **FIG. 7-7** in which like reference characters match like reference characters in **FIG. 7-4** and **7-5** and in the written description for like equipment. PSA tail gas can be burned for its heat value and/or can be recovered as feed for an integrated hydrogen plant. Reference characters are provided in **FIG. 7-6** and **FIG. 7-7** illustrating the PSA tail gas recycle process.

[**0659**] This PSA tail gas recovery embodiment of the process includes the PSA tail gas being fed to an integrated hydrogen plant. Feeding the PSA tail gas to an integrated hydrogen plant allows for the production of hydrogen from one or more of the light hydrocarbons which respectively can be present in the PSA tail gas including C<sub>1</sub>, C<sub>2</sub>, C<sub>3</sub>, C<sub>4</sub> light hydrocarbons (such as methane and/or ethane). The composition of PSA tail gas can comprise at least methane

and/or ethane, as well as comprising optionally other light hydrocarbons and optionally comprising hydrogen.

[0660] Recovering the PSA tail gas allows for the amount of tail gas to be burned as fuel in the fired heater to be reduced or eliminated entirely. Further, feeding the PSA tail gas to the integrated hydrogen plant allows for the production of hydrogen gas and the realization of its economic and chemical value, rather than burning the steam for its heat of combustion value. The percentage of PSA tail gas to be recovered by feeding to the integrated hydrogen plant from the total PSA tail gas recovered is in a range of from 0 vol % to 100 vol percent of PSA tail gas produced.

[0661] In one embodiment, PSA tail gas is both fed to the integrated hydrogen plant and burned in the fired heater. Where PSA tail gas is both fed to the integrated hydrogen plant, as well as burned in the fired heater, the PSA tail gas stream can be proportioned between the integrated hydrogen plant and the fired heater in any ratio.

[0662] The PSA tail gas recovery embodiment of the process can be a continuous process for recovering hydrocarbons evolved in the regeneration of a PSA system rather than burning it as a source of heat. One embodiment is a continuous process for recovering methane and ethane which is produced during regeneration. Regeneration can include a step of flowing a gas through the regeneration system which enhances the evolution of light hydrocarbons from the adsorption medium. In one embodiment, evolution of light hydrocarbons from the adsorption medium is achieved by utilizing a gas having hydrogen in a range of 0 vol % to 100 vol % to evolve, or purge, the light hydrocarbons from the adsorption medium during regeneration.

[0663] The light hydrocarbons, e.g., methane and/or ethane, or C<sub>1</sub>-C<sub>4</sub> hydrocarbons, in the PSA tail gas system along with any purge gasses, such as hydrogen, in the PSA process are routed to an injected hydrogen plant to be steam reformed to produce hydrogen.

[0664] Since the purge gas, e.g., hydrogen, is mixed with the recovered light hydrocarbons, e.g., methane and/or ethane; C<sub>1</sub>-C<sub>2</sub> hydrocarbons, the purge hydrogen can flow with the hydrogen produced in the integrated hydrogen plant reformer. The integrated hydrogen plant product gas stream 7025A is referred to as the hydrogen make-up stream for the process.

[0665] The PSA tail gas recovery embodiment can produce H<sub>2</sub> make-up gas (hydrogen make-up stream) in quantities great enough to reduce and/or entirely eliminate the need for the process to be supplied with natural gas from an external source. In other words, the plant can be self-sufficient in some embodiments in its ability to provide all of the hydrogen required for its normal operation from the light hydrocarbons recovered from the PSA tail gas stream, or the PSA system 7055. In such operation, the PSA system evolves sufficient light hydrocarbons, e.g., methane and/or ethane; C<sub>1</sub>-C<sub>4</sub> hydrocarbons that no external sources of hydrogen are necessary for normal operation of the plant. Thus, during normal operation this embodiment eliminates the need for supplying the process with external sources of natural gas (e.g., gas comprising methane and/or ethane), or other light hydrocarbon gases to the hydrogen plant.

[0666] Dependence on purchased natural gas can be reduced or eliminated in the PSA tail gas recovery embodiment of the process.

[0667] Sufficient PSA tail gas is recovered to eliminate the need for external sources of natural gas, e.g., gas including methane and/or ethane, or other light hydrocarbon gases as feed for the hydrogen plant. SCO can be produced in the process from any feedstock disclosed herein, including e.g., oil sand, shale, bitumen, or heavy oil, without the need for external sources of natural gas, e.g., gas including methane and/or ethane, or other light hydrocarbon gases as feed for the hydrogen plant.

[0668] FIG. 7-6 illustrates a slip stream of approximately 6 vol % of the total recycle gas stream 7033 is taken from the discharge of the main compressor 7032 via line 7033A and cooled from compressor discharge conditions to about 100° F. in cooler 7054. Compressor discharge temperature is a function of the compression ratio and will vary but in a typical application of this process can average about 185° F. The gas stream at this point comprises approximately 91 vol % hydrogen, 6 vol % methane and 3 vol % ethane on average with the components ranging 85-95 vol %, 3-10 vol %, and 2-5 vol % respectively. The cooled gas flows via line 7056 to the pressure swing adsorption (PSA) system 7055 where the methane and ethane components of the gas are adsorbed and the purified gas stream comprising approximately 99.9 vol % hydrogen (range 96-100 vol %) flows via line 7056A back to line 7025 which is the suction line to the compressor 7032.

[0669] During the regeneration cycle of the PSA system, the adsorbed methane and ethane gases along with a purge stream of hydrogen exit the PSA system 7055 via line 7058. The gas stream at this point comprises approximately 62 vol % hydrogen, 25 vol % methane and 13 vol % ethane on average, ranging 52-72 vol %, 15-35 vol %, and 7-18 vol % respectively. In some embodiments of the process, the PSA tail gas in line 7058 would combine with other purge gases in line 7022 and flow to the fired heater 7035 fuel system represented by line 7022. In a different embodiment, the PSA system 7055 tail gas flows via line 7058 to the integrated hydrogen plant 7300. Here the PSA tail gas is recompressed to the requirements of the hydrogen plant, typically in the range of 300 psig-500 psig, and becomes process feedstock for conversion to hydrogen thus replacing purchased natural gas.

[0670] The amount of PSA tail gas produced is a function of the operating conditions in the Chattanooga reactor 7005. The amount of PSA tail gas produced can be controlled by varying the operating conditions of reactor 7005. Higher temperatures produce higher amounts of light hydrocarbons. Sufficient PSA tail gas can be produced to meet all of the process feed requirements of the hydrogen plant 7300. Any excess tail gas produced can be let down to the plant's fuel system to displace product oil used for fuel. The methane and ethane in the PSA tail gas provide the hydrocarbon component for producing hydrogen via steam reforming. The hydrogen in the tail gas is processed unchanged in the hydrogen plant 7300 and can eliminate the need to separate the hydrogen from the light hydrocarbon gases prior to steam reforming.

[0671] PSA tail gas stream 7058A is fed to tail gas compressor 7500 at a pressure of about 5 psig. The PSA tail gas pressure can range from about 3.0 psig to about 15 psig. Tail gas compressor 7500 discharges the compressed tail gas through as compressed tail gas stream 7058B having a

pressure of 450 psig and which can range in pressure from 300 psig to 600 psig. Compressed tail gas stream **7058** is fed to hydrogen plant **7300**.

[**0672**] The process utilizing PSA tail gas from PSA system **7055** fed to hydrogen plant **7300** can encompass processing any feed to the reactor system disclosed herein. Material balances are provided below for three non-limiting processing cases utilizing different non-limiting examples of process feeds. These non-limiting feed cases are:

- [**0673**] Oil Sand,
- [**0674**] Bitumen Heavy Oil and Dry Sand, and
- [**0675**] Oil shale.

[**0676**] The process disclosed herein in its variations and embodiments can utilize and amount of the SCO which is produced by the process as fuel. The process generally can utilize an amount of the SCO produced in an amount in a range of from 0 wt % to 100 wt %, or alternatively having a maximum value of SCO used as fuel being the fuel feed limit of all of the combustion apparatus (e.g. fired heater) of the process. The amount of product SCO produced by the process which can be burned as fuel in the process will vary with the operational demands of the plant. For example, the amount of SCO which can be burned as fuel in the fired heater will depend upon the feed parameters of the fired heater design. The process can burn SCO as production operations demand. Using an amount of SCO produced by the process as fuel in the process reduces or eliminates fuel oil needed to be purchased from external sources. In one embodiment, the process is self-sufficient in fuel oil requirements, i.e., the SCO product which is utilized for fuel in the process eliminates (makes 0 wt %) the need to supply fuel oil from an external source to the process.

[**0677**] The material balance for a non-limiting oil sand feed case is as follows in Table 2 below.

**TABLE 2**

Oil Sand Feed Case Material Balance Oil Sand Processing With PSA Tail Gas Recycle			
	Lower Production	Normal Production	Increased Production
<u>Raw Material Data</u>			
Oil Sand Feed (tons/hr)	5,460	7,800	10,140
wt % Bitumen in Oil Sand	8.00	9.34	15.00
Make-up Hydrogen (mmscf/hr)	6.86	9.80	12.74
Hydrogen Recycle Ratio	16.00	25.39	30.00
Fuel (33.5° API Crude Product) (bbl/hr)	448.7	641	833.3
<u>Product Data</u>			
Product 33.5° API Crude Product (bbl/hr)	2,199	3,141	4,083
Sulfur (lbs/hr)	31,754	45,363	58,972
NH3 (lbs/hr)	2,720	3,885	5,051
Net 33.5° API Crude Product (bbl/day)	42,000	60,000	78,000
<u>PSA Tail Gas To Hydrogen Plant</u>			
Hydrogen (mmscf/hr)	3.38	4.83	6.28
Methane (mmscf/hr)	1.38	1.97	2.56
Ethane (mmscf/hr)	0.70	1.00	1.30

[**0678**] The material balance for a non-limiting bitumen heavy-oil and dry sand processing case is as follows in Table

3 below. The process is not limited to bitumen heavy-oil and any heavy oil can be processed by this process. The “dry sand” of this case is sand which is substantially free of bitumen.

**TABLE 3**

Bitumen Heavy-Oil and Dry Sand Feed Case Material Balance Bitumen Heavy Oil and Dry Sand Processing			
	Lower Production	Normal Production	Increased Production
<u>Raw Material Data</u>			
Dry Sand Feed (tons/hr)	468	668	868
Bitumen Feed	468	668	868
Make-up Hydrogen (mmscf/hr)	3.68	5.25	6.83
Hydrogen Recycle Ratio	16	6.4	30
Fuel (33.5° API Crude Product) (bbl/hr)	53	75	98
<u>Product Data</u>			
Product 33.5° API Crude Product (bbl/hr)	1,824	2,605	3,387
Sulfur (lbs/hr)	31,835	45,479	59,123
NH3 (lbs/hr)	2,727	3,895	5,064
Net 33.5° API Crude Product (bbl/day)	42,504	60,720	78,936
<u>PSA Tail Gas To Hydrogen Plant</u>			
Hydrogen (mmscf/hr)	3.38	4.83	6.28
Methane (mmscf/hr)	1.38	1.97	2.56
Ethane (mmscf/hr)	0.70	1.00	1.30

[**0679**] The material balance for a non-limiting oil shale feed case is as follows in Table 4 below.

**TABLE 4**

Non-Limiting Oil Shale Feed Case Material Balance Oil Shale Processing			
	Lower Production	Normal Production	Increased Production
<u>Raw Material Data</u>			
Oil Sand Feed (tons/hr)	3,400	4,857	6,314
wt % Kerogen in Oil Sand	10	15	20
Make-up Hydrogen (mmscf/hr)	6.86	9.80	12.74
Hydrogen Recycle Ratio	16	26	30
Fuel (33.5° API Crude Product) (bbl/hr)	449	641	833
<u>Product Data</u>			
Product 33.5° API Crude Product (bbl/hr)	2,199	3,141	4,083
Sulfur (lbs/hr)	15,917	22,739	29,561
Net 33.5° API Crude Product (bbl/day)	42,000	60,000	78,000
<u>PSA Tail Gas To Hydrogen Plant</u>			
Hydrogen (mmscf/hr)	3.38	4.83	6.28
Methane (mmscf/hr)	1.38	1.97	2.56
Ethane (mmscf/hr)	0.70	1.00	1.30

[**0680**] In one embodiment of the process, the PSA tail gas recovery satisfies all hydrocarbon feed requirements of the integrated hydrogen plant and the burning of SCO product oil satisfies all fuel oil combustion feed requirements for the fired heater resulting in a process which is entirely self-sufficient and self-sustaining in regard to requiring no external sources of natural gas (or light hydrocarbons; methane

and/or ethane; C<sub>1</sub>-C<sub>4</sub> hydrocarbons) as feed to the integrated hydrogen plant during normal operations and requiring no external sources of fuel oil for combustion purposes in the fired heater during normal operation.

[0681] Although the invention has been described in conjunction with specific embodiments, many alternatives and variations will be apparent to those skilled in the art in light of this description and the annexed drawings. Accordingly, the invention is intended to embrace all of the alternatives and variations that fall within the spirit and the scope of the appended claims.

We claim:

1. A process for producing synthetic crude oil, comprising the steps of:

providing a reactor having a first reactor feed stream comprising a bitumen, and a gas reactor feed comprising hydrogen,

contacting said bitumen with said hydrogen in said reactor thereby producing a reactor gas stream,

providing a pressure swing adsorption system producing a PSA tail gas,

feeding said PSA tail gas to an integrated hydrogen plant.

2. The process for producing synthetic crude oil according to claim 1, further comprising the steps of:

producing a synthetic crude oil product,

burning an amount of said synthetic crude oil product.

3. The process of claim 1, wherein said first feed stream comprises an oil sand.

4. The process of claim 3, wherein said oil sand comprises a Canadian Athabasca oil sand

5. The process of claim 1, wherein said feed comprises heavy oil.

6. The process of claim 1, wherein said process is self-sufficient in production of gas feed to the integrated hydrogen plant.

7. The process of claim 1, wherein said process burns a produced synthetic crude oil in the fired heater.

8. The process of claim 1, in which said reactor has a fluidized bed comprising an oil sand.

9. A process for producing synthetic crude oil, comprising the steps of:

providing a reactor, a first reactor feed stream comprising a kerogen, and a gas reactor feed comprising hydrogen,

contacting said kerogen with said hydrogen in said reactor thereby producing a reactor gas stream,

providing a pressure swing adsorption system producing a PSA tail gas,

feeding said PSA tail gas to an integrated hydrogen plant.

10. The process for producing synthetic crude oil according to claim 9, further comprising the steps of:

producing a synthetic crude oil product,

burning an amount of said synthetic crude oil product.

11. The process of claim 9, wherein said first feed stream comprises an oil shale.

12. The process of claim 9, wherein said process is self-sufficient in production of gas feed to the integrated hydrogen plant.

13. The process of claim 10, wherein said process burns a produced synthetic crude oil in the fired heater.

14. The process of claim 9, in which said reactor has a fluidized bed comprises an oil shale.

15. An apparatus for producing a synthetic crude oil according to the process of claim 1.

16. A system for producing a synthetic crude oil according to the process of claim 1.

17. A control system for producing a synthetic crude oil according to the process of claim 1.

18. An apparatus for producing a synthetic crude oil according to the process of claim 9.

19. A system for producing a synthetic crude oil according to the process of claim 9.

20. A control system for producing a synthetic crude oil according to the process of claim 9.

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