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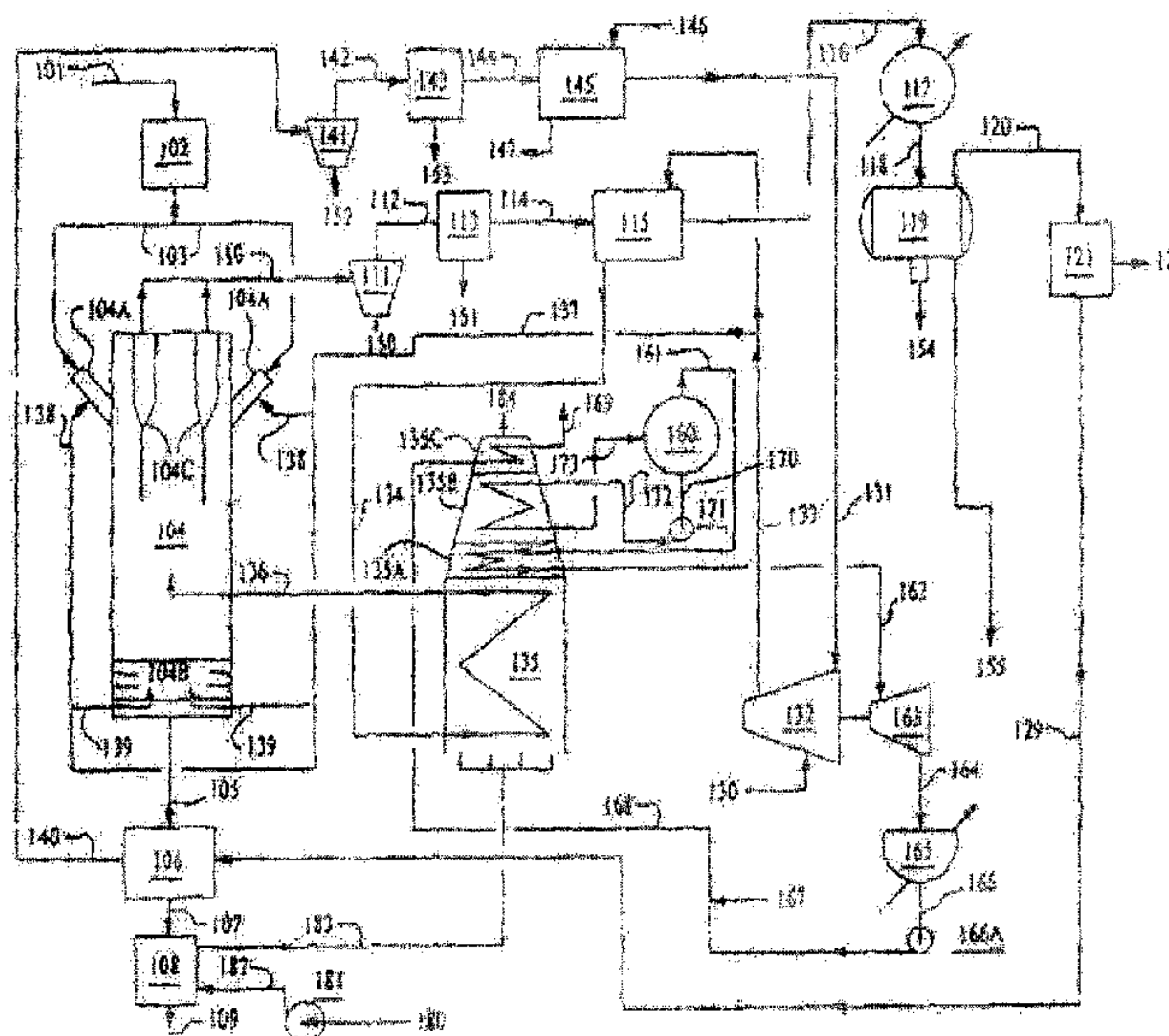
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(54) Titre : PROCEDE ET DISPOSITIF DE TRANSFORMATION DU SCHISTE BITUMINEUX OU DES SABLES
BITUMINEUX EN PETROLE

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



(57) Abrégé/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 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 % H₂).



ABSTRACT OF THE DISCLOSURE

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 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 % H₂).

Title Of The Invention

Process and apparatus for converting oil shale or oil sand (tar sand) to oil

Field Of The Invention

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.

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

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 flotation. In these systems, tailing oil recovery is necessary to clear the sand for disposal, however the sand is not completely cleared of bitumen.

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.

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

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. Moreover, with the present invention, less gas and substantially no coke are produced.

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.

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.

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

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.

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 % H₂).

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.

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.

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

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 % H₂) in a fluidized bed reactor, wherein the fluidizing medium is a feed stream including hydrogen gas.

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.

In each of the following three embodiments:

- Extraction produced bitumen and raw sand or tailings pond sand; or
- *in situ* produced bitumen and raw sand or tailings pond sand; or
- *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.

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 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.

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.

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.

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.

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 % H₂).

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.

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.

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.

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

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.

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.

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.

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.

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.

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.

Brief Description Of The Drawings

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

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

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

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

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

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

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

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

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

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

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

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

FIG. 9 shows a process flow diagram for Example 3

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

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 % H₂) 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.

The concentration of a hydrogen feed can vary in this invention. One embodiment utilizes pure hydrogen feed (100% H₂). 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.

While the reactor can react with any concentration of hydrogen in the feed stream (1 vol %-100 vol % H₂). Some embodiments use hydrogen feed streams with H₂ concentrations

of 75 vol %-94 vol % and methane and ethane concentrations of 6 vol % to 25 vol %. One embodiment utilizes a hydrogen feed stream having 94 vol % H₂ and 6 vol % methane and ethane mixture.

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.

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 failings 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.

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. Our present design provides four lock-hopper feed systems equally spaced around the circumference of each reactor.

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.

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.

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, $C_1 + C_2$ 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.

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.

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.

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. Our present design provides four lock-hopper feed systems equally spaced around the circumference of each reactor.

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).

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.

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 915° F, the normal temperature before the in-out heat exchanger. This second temperature can however vary (300° F to 900° F).

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 450°.

The overhead gas stream can be cooled to 100° F from about 450° 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.

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).

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.

The following definitions are employed in this disclosure.

“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:

19 wt. % aliphatics	(which can range from 5 wt. %-30 wt. %, or higher)
30 wt. % aromatics	(which can range from 15 wt. %-50 wt. %, or higher)
32 wt. % resins	(which can range from 15 wt. %-50 wt. %, or higher)
19 wt. % asphaltenes	(which can range from 5 wt. %-30 wt. %, or higher)

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

In addition bitumen can contain some water and nitrogen compounds ranging from less than 0.4 wt. % to in excess of 0.7 wt. %. The metals content, while small, must be

removed to avoid contamination of the product SCO. 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.

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

“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.

“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.

$$\text{Degrees API} = \frac{141.5}{\text{Density at 60°F}} - 131.5$$

Specific gravity 60°F/60°F

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°.

“High grade” or “excellent quality” oil sands contain 12% bitumen or higher.

“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.

“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.

“*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.

“*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.

“*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.

“*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.

“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.

“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.

“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 grained, calcareous rock containing kerogen, a solid hydrocarbon.

“Pressure” as used herein, the term pressure refers to the force acting per unit area within the various vessels and pipelines constituting our 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.

“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.

“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.

“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.

“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.

“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.

“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.

“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.

“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.

“Upgrading” is a term used to refer to the treatment of crude oil to improve its quality. As used in our 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 8-10° 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.

“Feedstock” and “feed” refer to a material which enters a unit operation or process. In the invention a variety of deedstocks 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.

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

“Reaction process” refers to the chemical and physical activities occurring in the reactor including, but not limited to, the reaction between hydrogen and various foods, 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.

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.

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.

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.

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
 100 wt. % 100 wt. %

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.

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.

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.

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-1200°F, 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).

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.

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.

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.

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.

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.

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.

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 134.

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.

Since the gas stream has been cooled down to about 100°F and is still at a pressure of 480 psig, all carbon compounds C₃ and above have been condensed are 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 sulfur dioxide gases, are absorbed and discharged through line 122 and conveyed to a suitable sulfur recovery plant. The amine absorption system 121 is described in greater detail in FIG. 5.

The only gases not absorbed and removed in absorption system 121 are unreacted recycle hydrogen and $C_1 + C_2$ 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 $C_1 + C_2$ hydrocarbons is heated and discharged into line 140. The C_1 and C_2 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_1 and C_2 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.

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.

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 exchanger 145.

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.

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.

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 C₃ 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 is maintained by compressor 132 so that the 1200°F hydrogen stream can be introduced into reactor 104 (FIG. 1) or 204 (FIG. 2).

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. Reduced temperature and pressure steam from turbine 163 is conveyed to steam condenser 165 and the condensate recirculated via line 166 and pump 166A. 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 is introduced into convection

section 335C (135C), heated and discharged through line 369 (169) for further processing, e.g., deaeration.

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.

The steam circulation loop include steam drum 360 (160), line 370 (170), recirculation pump 371 (171) and lines 372-373 (172-173) which conveys 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 468. Line 468 is similar to flow line 168, 368 which communication with line 169 through connection section 335a (135a) to discharge.

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).

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.

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.

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_2S , CO_2 and SO_2 are absorbed by a counter current circulating amine solution. The recycle gases 5B flow from the top of the absorber 5A to recycle hydrogen stream 129.

The rich amine solution 5C exits the bottom of the absorber, flows through an amine exchanger 5D where it is heated by exchange with hot stream amine solution 5L and enters the top of an amine stripper 5F. Absorbed acid gases are stripped from the amine solution by the application of heat to the solution in reboiler 545 (145) 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 145 to provide heat for reboiler 545 (145) and the partially cooled recycled gases 5G are further cooled by cooler 5H and then flow through line 531 (131) to the suction side of compressor 132.

Lean amine solution 5J is circulated by amine circulation pumps 5K through the amine exchanger 5D and amine cooler 5N to the top of the amine absorber 5A to repeat the gas cleanup process.

EXAMPLE 1

The overall mass balance for the process according to the present invention is shown in FIG. 6, where 1000 tons/hr 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).

RAW MATERIALS PRODUCTS

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

1.6 MMSCF/HR HYDROGEN 5.2 MMSCF/HR STACK GAS

3.3 MMSCF/HR AIR 6600 LBS/HR SULFUR

0.5 MMSCF/HR NATURAL GAS 850 TONS/HR SPENT SAND

REACTOR DIMENSIONS AND MASS AND ENERGY BALANCES

REACTOR 104

Column Diameter 20.00 ft

Cross Section Area 34.16 ft.²

Void Fraction 0.85 (At Fluidization)

Cross Section of Sand 47.12 ft.²

Cross Section of Gas 267.04 ft.²

Reactor Volume 27394.26 ft.³

Bed Diameter 20.00 ft

Bed Height 87.20 ft

Time-Space Constant 0.25 hr

Pressure Drop 35.00 psi

OIL SAND (TAR SAND) FEED

Sand Flow Rate 1000.00 tons/hr

Density of sand 21.68 lbs./ft.³

Volumetric sand flow 16436.55 ft.³/hr

Sand Velocity 5.81 ft/minute

Hold-up 15.00 minutes

HYDROGEN

Hydrogen Flow Rate 238661.44 lbs/hr

(45226343 SCF/hr)

Cp of H₂ 3.50 btu/lb-°F. (@ 900°F.)

Hydrogen Recycle Ratio 26.52

Hydrogen Flow Rate 45.28 SCF/hr

Hydrogen velocity 3.02 ft/s

OFF GAS

Gas Production 0.40 MMSCF/hr

MW 30.30 g/mole

Cp of flue gas 0.55 btu/lb-°F.

OFF GAS COMPOSITION

CO 0.30%

CO₂ 0.20%

H₂ S 31.00%

NH₃ 2.50%

C₃ 66.00%

ENERGY BALANCE

OVER-ALL CONSIDERATIONS

Heat of Reaction 75.00 btu/lb. Bitumen

Cp Sand 0.19 btu/ton-°F.

Cp Bitumen 0.34 btu/lb-°F.

Cp Tarsand (sand + Bitumen) 426.70 btu/ton-°F.

Sand Feed Temperature 50.00°F.

Sand temperature 50.00°F.

at reactor inlet

Reaction temperature 800.00°F.

Sand Feed 1,000.00 tons/hr

OIL SAND (TAR SAND) REACTOR

REACTOR CONDITIONS

Heat required in reactor 356.03 MMbtu/hr

Heat generated in Reactor 22.50 MMbtu/hr

Additional Heat Required 335.24 MMbtu/hr

Minimum H₂ for reaction 9000.00 lbs./hr

(1.71 MMSCF/hr)

Additional H₂ Supplied 229736.15 lbs./hr

(43.53 MMSCF/hr)

Total H₂ Supplied 238736.15 lbs./hr

(45.24 MMSCF/hr)

C₁ -C₂ Flow within H₂ Stream 4594.72 lbs/hr

(at equilibrium-2%) (0.08 MMSCF/hr)

Entering H₂ Temperature 1200.00°F.

Cp H₂ 3.50 btu/lb-°F.

Heat Supplied by C₁ -C₂ 1.01 MMbtu/hr

Heat Supplied by H₂ 334.23 MMbtu/hr

H₂ Recycle ratio 26.53

REACTOR BOTTOMS COOLER:

Assures Efficient Removal of

Exiting Solids

Cold Hydrogen Cooler Stream 1,148.68 lbs./hr

(0.22 MMSCF/hr)

Heat Removed 2.73 MMbtu/hr

Entering Hydrogen Temperature 121.64°F.

Exiting Sand Temperature 791.60°F.

SAND COOLER

SAND

Sand Flow Rate 850.00 tons/hr

Temperature of Entering Sand 791.60°F.

Temperature of Spent Sand 180.00°F.

Cp Sand 0.19 btu/lb-°F.

Heat Removed 198.59 MMbtu/hr

HYDROGEN COOLANT FLOW

Hydrogen Flow 238736.15 lbs/hr
(45.24 MMSCF/hr)

Heat to Be Removed 182.96 MMb/hr

Entering Hydrogen Temperature 100.00°F.

Exiting Hydrogen Temperature 318.96°F.

AIR COOLANT

Air Required for Combustion 250000.00 lbs/hr
(3.27 MMSCF/hr)

Cp Air 0.25 btu/lb-°F.

Entering Air Temperature 50.00°F.

Exiting Air Temperature 300.00°F.

Heat Removed 15.63 MMbtu/hr

AMINE REBOILER**HYDROGEN SUPPLY**

Entering Hydrogen Temperature 318.96°F.

Exiting Hydrogen Temperature 100.00°F.

AMINE BOIL-OFF

Heat Available to the system 182.96 MMbtu/hr

IN-OUT HEAT EXCHANGER**HYDROGEN TO BE HEATED**

Hydrogen Flow 238736.15 lbs/hr
(45.24 MMSCF/hr)

Inlet H₂ Temperature 121.64°F.

Exiting H₂ Temperature 750.00°F.

Total Heat Required 525.05 MMbtu/hr

OFF GAS HEAT SUPPLY

Off Gas flow rate 31978.89 lbs/hr

0.40 MMSCF/hr

Condensables in vapor phase 214941.75 lbs/hr

MW 30.30 lb/lb-mole

Cp Vapor 0.55 btu/lb-°F.

Cp Liquid 0.45 btu/lb-°F.

Cp Non-Condenses 3.00 btu/lb-°F.

Heat of Vaporization 65.00 btu/lb

Hydrogen Recycle Flow 229736.15 hrs/hr
in Stream (*43.53 MMSCF/hr)

Inlet Temperature 800.00°F.

Exit Temperature 350.00°F.

PRODUCT CONDENSER/COOLER PRODUCT SIDE

Entering Temperature 350.00°F.

Exiting Temperature 100.00°F.

Condensate 24941.75 lbs/hr

(665.29 bbl/hr)

Heat Removal H₂ 201.02 MMbtu/hr

Off Gas 4.40 MMbtu/hr

Condensate 38.15 MMbtu/hr
 Total 243.57 MMbtu/hr
 COOLER REQUIREMENT 243.57 MMbtu/hr
 COMPRESSOR HYDROGEN SIDE
 Flow Rate 755412.69 SCF/min
 (45.32 MMSCF/hr)
 Pressure Out 670.00 psig
 Pressure In 450.00 psig
 DP 220.00 psi
 gamma (Cp/Cv) 1.40
 # Stages 3
 Temperature Inlet 100.00°F.
 Mechanic Efficiency 0.80 *100%
 Pb/Pa 1.14
 Power Requirement per Stage 6366.67 hp
 Total Power Required 19100.00 hp
 Outlet Temperature 121.64°F.
 STEAM SUPPLY
 Pressure 1500.00 psig
 Temperature 800.00°F.
 Degree Superheat 200.00°F.
 Saturation Temperature 596.20°F.
 Steam Heat Vue 1364.00 btu/lb
 Flow Rate 10894.28 lbs/hr
 FIRED HEATER PRODUCTS TO BE HEATED
 Hydrogen Flowrate 238736.15 lbs/hr
 (45.24 MMSCF/hr)
 Hydrogen Temperature 750.00°F.
 Water Flow Rate 10894.28 lbs/hr
 Water Temperature 75.00°F.
 Heat Duty 517.83 MMbtu/hr
 C₃ 'S (FUEL PRODUCED
 BY THE PROCESS)
 Flow Rate 4263.85 lbs/hr
 (0.04 MMSCF/hr)
 Heat of Combustion 20000.00 btu/lb
 Cp 0.60 btu/lb-°F.
 Temperature in 75.00°F.
 Heat Supplied
 (After temperature correction) 79.84 MMbtu/hr
 MAKE-UP METHANE
 Combustion Temperature 2200.00°F.
 Heat Remaining to 437.99 MMbtu/hr
 be supplied by Methane
 Flow Rate 21653.89 lbs/hr
 (0.51 MMSCF/hr)

Heat of Combustion 20227.00 btu/lb
 (After temperature correction)
 Temperature in 75.00°F.
 COMBUSTION AIR
 Air Required for Combustion 200000.00 lbs/hr
 (2.61 MMSCF/hr)
 Air Supplied 25% Excess 250000.00 lbs/hr
 (3.27 MMSCF/hr)
 COMPRESSOR SUCTION
 COOLER (5H)
 OUTFLOWS
 Hydrogen
 Flowrate 200000.00 lbs/hr
 Temperature 100.00°F.
 Required Coolant Supply 22.42 MMbtu/hr
 MATERIAL BALANCE
 OIL SAND (TAR SAND) REACTOR (104)
 IN FLOWS
 Sand
 Flowrate 1000.00 tons/hr
 Temperature 50.00°F.
 Pressure 14.70 psia
 (Force Fed)

Hydrogen
 Flowrate 45.23 MMSCF/hr
 Temperature 1200.00°F.
 Pressure 635.00 psig
 C₁-C₂ s
 Flowrate 0.08 MMSCF/hr
 Temperature 1200.00°F.
 Pressure 635.00 psig
 OUT FLOWS
 Sand
 Flowrate 850.00 tons/hr
 Temperature 190.00°F.
 Pressure 600.00 psig
 Off Gas
 Flowrate 43.92 MMSCF/hr
 Temperature 800.00°F.
 Pressure 600.00 psig
 Composition wt %
 H₂ 81.98
 CO 0.05
 CO₂ 0.04

H₂ S 5.60
NH₃ 0.45
C₃ 11.92
Product
Flowrate 214937.52 lbs./hr
(Vapor Phase)
Temperature 800.00°F.
Pressure 600.00 psig
SAND COOLER (106, 108) IN FLOWS
Sand
Flowrate 850.00 tons/hr
Temperature 791.92°F.
Pressure 600.00 psig
Hydrogen
Flowrate 45.23 MMSCF/hr
Temperature 100.00°F.
Pressure 500.00 psig
Air
Flowrate 3.27 MMSCF/hr
Temperature 50.00°F.
Pressure 30.00 psig
OUT FLOWS
Sand
Flowrate 850.00 tons/hr
Temperature 200.00°F.
Pressure 480.00 psig
Hydrogen
Flowrate 45.23 MMSCF/hr
Temperature 313.94°F.
Pressure 480.00 psig
Air
Flowrate 3.27 MMSCF/hr
Temperature 300.00°F.
Pressure 20.00 psig
IN-OUT HEAT EXCHANGER (115) IN FLOWS
Hydrogen
Flowrate 45.23 MMSCF/hr
Temperature 147.60°F.
Pressure 670.00 psig
Off Gas
Flowrate 43.92 MMSCF/hr
Temperature 800.00°F.
Pressure 600.00 psig
Composition wt %
H₂ 81.98
CO 0.05

CO₂ 0.04
 H₂ S 5.60
 NH₃ 0.45
 C₃ 11.92
 Product
 Flowrate 214937.52 lbs./hr
 (Vapor Phase)
 Temperature 800.00°F.
 Pressure 600.00 psig
 OUT FLOWS
 Hydrogen Flowrate 45.23 MMSCF/hr
 Temperature 750.00°F.
 Pressure 650.00 psig
 Off Gas Flowrate 43.92 MMSCF/hr
 Temperature 368.63°F.
 Pressure 580.00 psig
 Off Gas Composition as Above
 Product
 Flowrate 214937.52 lbs./hr
 (Vapor Phase)
 Temperature 368.63°F.
 Pressure 580.00 psig
 PRODUCT CONDENSER/ IN FLOWS
 COOLER (117)
 Off Gas
 Flowrate 43.92 MMSCF/hr
 Temperature 368.63°F.
 Pressure 580.00 psig
 Off Gas Composition as Above
 Product
 Flowrate 214937.52 lbs./hr
 (Vapor Phase)
 Temperature 368.63°F.
 Pressure 550.00 psig
 OUT FLOWS
 Off Gas
 Flowrate 43.92 MMSCF/hr
 Temperature 100.00°F.
 Pressure 540.00 psig
 Off Gas Composition as Above
 Product
 Flowrate 214937.52 lbs./hr
 (as condensate)
 Temperature 100.00°F.
 Pressure 540.00 psig
 AMINE SYSTEM (121, FIG. 5) IN FLOWS

Hydrogen
 Flowrate 45.23 MMSCF/hr
 Temperature 318.00°F.
 Pressure 470.00 psig
 OUT FLOWS
 Hydrogen
 Flowrate 45.23 MMSCF/hr
 Temperature 100.00°F.
 Pressure 450.00 psig

EXAMPLE 2

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 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.

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. 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 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. The reactor 705 and related equipment are shown in more detail in FIG. 8.

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, 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, 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 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.

On entering the reactor 705, 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 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.

Reaction products including synthetic crude oil and unreacted hydrogen mixture exit the reactor 705 through a product stream outlet 705F 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 at the bottom of the reactor 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 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.

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.

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.

Spent sand, at a temperature of about 950°F, overflows from reactor 705 into conduit line 705D through a spent solids outlet 705G. 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. The cooled sand can be discharged and used, for example, for land reclamation.

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.

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.

The reactor overhead stream from the cyclone separator 705A is discharged into line 710, 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 pulsated by a back flow of a 670 psig, 875°F hydrogen mixture taken from in-out heat exchanger 715 via conduit line 734A. Collected fine 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.

The substantially solids-free overhead stream flows from the hot gas clean-up through 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.

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 sour water is purged from the overhead stream via line 786.

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.

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.

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°.

The absorber 724 can comprise, for example, a counter current circulating ethanol amine solution in intimate contact with the remaining overhead stream. The remaining fluid stream can comprise gases such as, for example, H₂ S, CO₂, recycle hydrogen, and C₁ and C₂ hydrocarbons. H₂ S and CO₂ are removed from the remaining fluid stream by the absorber 724. Remaining hydrogen, C₁ and C₂ hydrocarbons form the recycle hydrogen mixture and flow through line 725 to compressor 732.

The rich amine solution having absorbed H₂ S, CO₂, SO₂ and NH₃ 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.

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.

Recycle hydrogen, and C₁ and C₂ 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.

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.

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 C₁ and C₂ hydrocarbon pressure swing adsorption ("PSA") system 755. The PSA system helps to maintain the C₁ and C₂ 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. C₁ and C₂ hydrocarbon purged from the first hydrocarbon mixture is discharged through line 758 and combined with the gas in line 22 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.

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.

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 into the fired heater 735. 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.

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

FLUIDIZED BED REACTOR:

Reactor (fluidized bed) Temperature 950°F.

Reactor (fluidized bed) Pressure 600 psig

H₂ Recycle Ratio 21.09

Catalyst Flow Rate into Reactor 1255.07 lbs/hr

Oil sand (tar sand) Flow into Reactor 2520 tons/hr

Oil sand (tar sand) Feed Inlet Temperature 50°F.

Hydrogen Mixture Flow Rate into Reactor 60.4 MMSCF/hr

Hydrogen Mixture Gas Inlet Temperature 1500°F.

ROTARY COOLERS:

Air Temperature at Intake 50°F.

Air Temperature (exiting) 623°F.

Sand Entering Temperature 950°F.

Sand Exiting Temperature 665°F.

IN-OUT HEAT EXCHANGER:

Overhead Stream Entering Volumetric 58.31 MMSCF/hr
Flow Rate

Overhead Stream Entering Temperature 950°F.

Overhead Stream Exiting Temperature 516°F.

Hydrogen Mixture Entering Flow Rate 60.4 MMSCF/hr

Hydrogen Mixture Entering Temperature 185°F.

Hydrogen Mixture Exiting Temperature 875°F.

FIRED HEATER:

Fuel Consumption (natural gas equivalent) 1.2 MMSCF/hr

Vapor Let-Down & PSA Off Gas Fuel Supply 0.56 MMSCF/hr
(natural gas equivalent)

Make-up Fuel Supply 0.64 MMSCF/hr

Combustion Air Entering Flow Rate 26.59 MMSCF/hr
(@ 65% excess)

Steam Production Rate (@ 1500 psig) 228,996 lbs/hr

Hydrogen Mixture Entering Flow Rate 60.4 MMSCF/hr

Hydrogen Mixture Entering Temperature 875°F.

Hydrogen Mixture Exiting Temperature 1500°F.

COMPRESSOR:

Power Required From Turbine 40,148 h.p.

Steam Flow Rate to Turbine 228,996.1 lbs/hr

Steam Pressure Entering Turbine 1500 psig

Steam Temperature Entering Turbine 800°F.
(200°F. superheat)

Hydrogen Mixture Entering Flow Rate 60.7 MMSCF/hr

Hydrogen Mixture Compressor Entering 100°F.

Temperature

Hydrogen Mixture Compressor Entering 450 psig
Pressure

Hydrogen Mixture Compressor Exiting 185°F.
Temperature

Hydrogen Mixture Compressor Exiting 670 psig
Pressure

PRODUCT CONDENSER/SEPARATOR:

Product Fluid Stream Entering Flow Rate 58.3 MMSCF/hr

Product Fluid Stream Entering Temperature 516°F.

Recycle Hydrogen Mixture Exiting 100°F.
Temperature

Synthetic Crude Oil Flow Rate 1255 bbl/hr

AMINE SYSTEM:

Amine Recirculation Flow Rate 50,400 lbs/hr

Ammonia Production 1478 lbs/hr
 Elemental Sulfur Production 17260 lbs/hr

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: Feedstocks And Feedstock Combinations

Alternative Feedstock	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

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).

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.

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.

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.

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 $\frac{1}{4}$ to 1 inch mesh. Such pieces are oil sand pieces that greater than $\frac{1}{4}$ inch and smaller than 1 inch in size. The pieces in one embodiment are of particle sizes that would not pass through a $\frac{1}{4}$ inch mesh screen but would fall through a 1 inch mesh screen. Smaller or larger pieces can be used in other embodiments.

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.

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

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.

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.

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..

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

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.

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.

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.

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.

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.

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.

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.

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.

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.

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.

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.

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.

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

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

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.

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.

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.

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.

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

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.

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"

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.

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

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.

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.

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.

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.

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

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.

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.

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

Feed lock hoppers 7004 and spent 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.

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.

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.

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.

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.

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:

- Hydrogen velocity to fluidize the contents of the reactor 7005 to create a fluid bed.
- 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.
- Temperature of the hydrogen from the fired heater 7035 in line 7036 normally heated to about 1500°F.
- Residence time of the reactants in the reactor 7005 (reaction time) for optimum conversion of the bitumen or kerogen to product vapors.

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.

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.

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.

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.

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

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.

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.

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.

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.

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.

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.

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.

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.

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.

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.

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.

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 C₁, C₂ 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 C₁, C₂, 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.

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.

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.

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.

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.

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.

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.

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.

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.

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.

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.

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.

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.

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:

- Sharing of a common hydrogen circulation compressor 7032,
- Sharing of a common amine scrubbing system 7400,
- Sharing a common source of hydrogen 7300, and/or
- Operational flexibility previously mentioned visa vie catalyst addition to the reactor 7005.

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_2S with trace amounts of CO_2 and the scrubbed recycle hydrogen along with the light hydrocarbon gases exit the absorber to flow through line 7025 to compressor 7032.

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.

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.

Optionally, the following design elements can be included in the invention:

- Integrating hydrogen plant 7300,
- A common boiler feed water treatment system for the fired heater 7035 and the hydrogen plant reformer,
- Combustion air is preheated in rotary sand cooler 7003 for the fired heater 7035 and hydrogen plant reformer,
- 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,
- A common cooling water system can be shared,
- An integrated steam system is possible maximizing waste heat recovery for power generation,
- A common control room integrates all plant operations minimizing manpower requirements and improves operating efficiency, and/or
- Integration of support facilities such as administration, operations, maintenance, etc. is possible.

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.

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.

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

Please See Next Page

Example 3

Please See Next Page

Tar Sand Production Design Index

Process Overview	2
Key Cost Data	3
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Material and Energy Balance	7
Process Flow Diagram	14

Process Overview

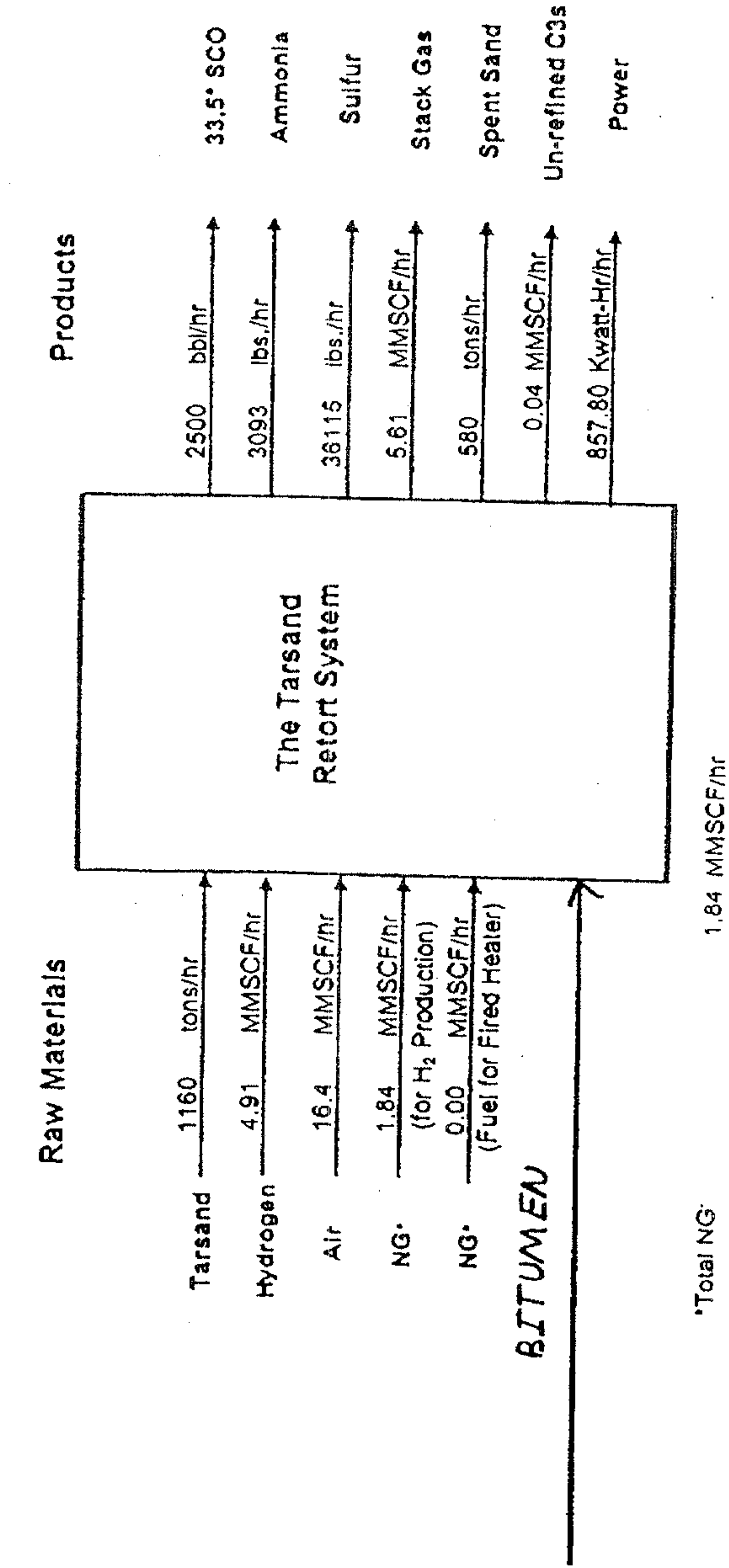
60,000 barrel Daily Production

Added insitu-Bitumen, Design Considerations changed to reflect single reactor processing.
Major changes included: C3 excess due to lower heat duty in Fired Heater and Heat Recovery in the Sand Cooler for use in power generation.

Reduced C/H ratio from 1.81 to 1.68.

Effects: Reduced makeup Hydrogen from 9.26 MMSCF/hr to 6.01 MMSCF/hr

Corrected Hydrogen losses to include Hydrogen Sulfide production.
Added 20% dilutant sand at 50°F to aid in handling.
Corrected PSA mass balance.
Optimized Sand cooler to increase combustion air temperature.
Effects: Slight increase in Hydrogen Make-up, produced a more realistic model.



Key Data**Raw Materials**

Bitumen	580	Tons/hr
Processed Tarsands (0% Bitumen)	580	Tons/hr
Recycle Hydrogen	28.54	MMSCF/Hr.
Make-up Hydrogen	4.91	MMSCF/Hr.
Combustion Air	16.35	MMSCF/Hr.
Hydrogen Recycle Ratio	5.81	
Natural Gas Equivalent (H2 Production)	1.84	MMSCF/Hr.
Natural Gas Equivalent (Fired Heater)	0.00	MMSCF/Hr.

Products

33.5° Crude	2500	bbl/hr
Light Gas (C1-C3) Produced	1.55	MMSCF/Hr.
Sulfur	36,115	lbs/hr
H ₂ S	38,357	lbs/hr
NH ₃	3,093	lbs/hr
Unrefined C3 Stream	0.04	MMSCF/Hr.

Substituting Crude for Natural Gas results in the Streams below.

Raw Materials

Bitumen	580	Tons/hr
Processed Tarsands (0% Bitumen)	580	Tons/hr
Recycle Hydrogen	28.54	MMSCF/Hr.
Make-up Hydrogen	4.91	MMSCF/Hr.
Combustion Air	16.35	MMSCF/Hr.
Hydrogen Recycle Ratio	5.81	
Natural Gas for H2 Production	1.29	MMSCF/Hr.

Products

33.5° Crude	2412	bbl/hr
Light Gas (C1-C3) Produced	1.55	MMSCF/Hr.
Sulfur	36,115	lbs/hr
H ₂ S	38,357	lbs/hr
NH ₃	3,093	lbs/hr

Design Considerations

Reactor Design

Column Diameter	15.00	ft	
Cross Sectional Area	176.71	ft ²	
Void Fraction	0.85		(At Fluidization)
Cross Section of Sand	26.51	ft ²	
Cross Section of Gas	150.21	ft ²	
Reactor Volume	26212.91	ft ³	
Bed Diameter	15.00	ft	
Bed Height	148.33	ft	
Time-Space Constant (Tau)	0.17	hr	$\tau = \frac{V(\text{reactor}) * (1-\text{Void})}{F(\text{sand})}$
Pressure Drop	35.00	psi	

Heavy Oil Considerations

% Insitu-Bitumen in Feed	50%	
Bitumen Temperature	300	°F
Bitumen Content of Raw Tarsand Feed	0.00%	
Tarsand Temperature	50	°F
Insitu-Bitumen Flow	580	tons/hr
Tarsand Flowrate	580	tons/hr
Heavy Oil Density	75	lbs/ft ³

TarSand Considerations

Bitumen Content	50.00%	
Sand Flow Rate	1160.00	tons/hr-per reactor
Sand Flow Rate	2,320,000	lbs./hr
Inlet Temperature	175.00	°F
Sand Density	121.68	lbs./ft ³
Estimated Density of sand at 50.00% Bitumen	98.34	lbs./ft ³
Volumetric sand flow	23591.62	ft ³ /hr
Sand Velocity	14.83	ft/minutes
Hold-up	10.00	minutes
Cp Sand	0.19	btu/lb-°F
Cp Bitumen as 10° Crude	0.34	btu/lb-°F
Cp Tarsand	0.27	btu/lb-°F

Organic Considerations

Bitumen Content	50%	
Organic Flow Rate (including sand)	1160.00	tons/hr
Sand Flow Rate	2,320,000	lbs./hr
Inlet Temperature	175.00	°F
Volumetric sand flow	23591.62	ft ³ /hr
Sand Velocity	14.83	ft/minute
Hold-up	10.00	minutes
Cp Tarsand (sand+Bitumen)	307.98	btu/ton-°F

Hydrogen Considerations

Hydrogen Flow Rate to two reactors	150626.56	lbs/hr	28.54 MMSCF/hr
Cp of H ₂	3.50	btu/lb-°F @900°F	
Hydrogen Recycle Ratio	5.81		
Hydrogen Flow Rate	28.54	MMSCF/hr	
Hydrogen Velocity	3.39	ft/s	

Off Gas Considerations

Gas Production	1.55	MMSCF/hr
MW	30.30	g/mole
Cp of off gas	0.55	btu/lb-°F

Off Gas Composition	wt%
CO	0.30%
CO ₂	0.20%
H ₂ S	31.00%
NH ₃	2.50%
C3	66.00%

By-product Production Rate	
Sulfur	36,115 lbs/hr
H ₂ S	38,357 lbs/hr
NH ₃	3,093 lbs/hr

Stream Composition & Properties

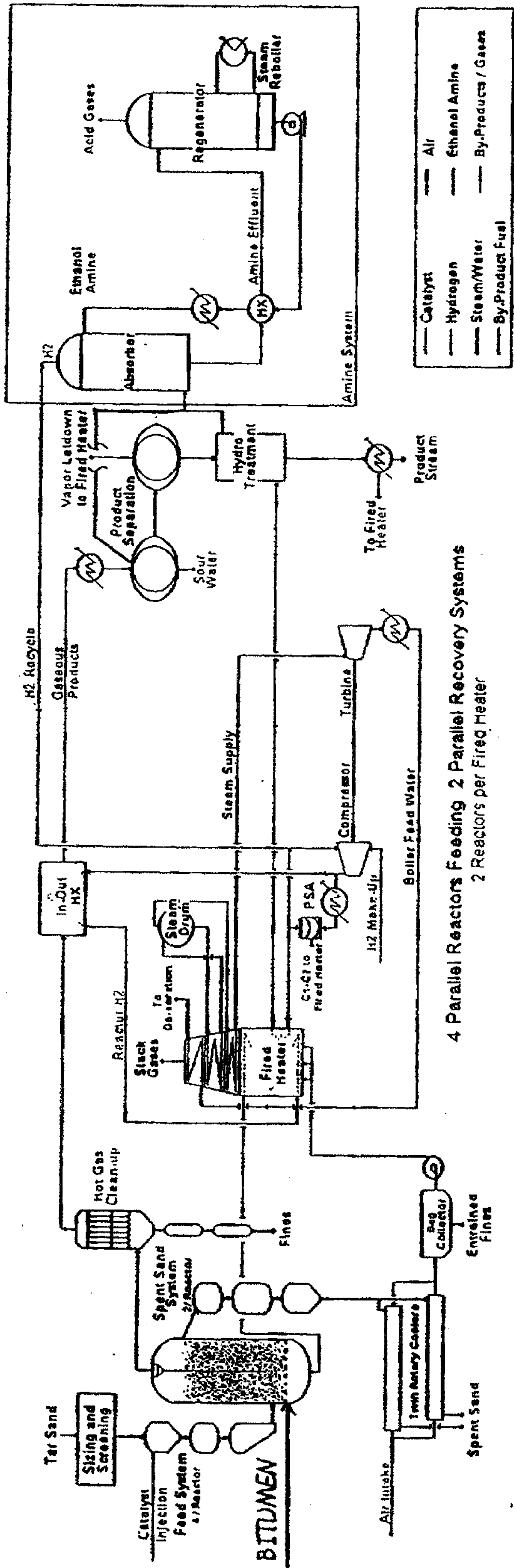
Tarsand Composition		Bitumen Composition		Product Composition		Reactor Effluent (wt%)	
Bitumen	50.0%	C	82.52%	C	86.20%	Water	10.61%
Water	4.0%	H	10.53%	H	12.80%	Off Gas	7.07%
Sand	46.0%	S	4.77%	S	0.10%	Light Gas Oil	44.55%
Off Gas Composition		N	0.40%	N	0.05%	Heavy Gas Oil	29.70%
CO	0.30%	O	1.09%	O	.	Hydrogen	8.06%
CO ₂	0.20%	Fines	0.70%	Fines	.		
H ₂ S	31.00%	H/C ratio	1.45	H/C ratio	1.68		
NH ₃	2.50%	MW	570-620	MW	210-235		
C ₃	66.00%	<u>Hydrocarbon Type</u>		<u>Hydrocarbon Type</u>			
		Aliphatics	19%	Saturated	78%		
		Aromatics	30%	Aromatic	22%		
		Resin	32%	Resin	0%		
		Asphaltenes	19%	Asphaltenes	0%		
		<u>Metals (ppm)</u>		<u>Metals (ppm)</u>			
		Vanadium	200	Vanadium	trace		
		Nickel	75	Nickel	trace		

Carbon Balance

In			
Tarsand Carbon		949,402	lbs/hour
Outs			
Product		839,491	lbs/hour
Residue	7%	67,008	lbs/hour
C1-C2	6%	50,256	lbs/hour
Carbon Difference		(7,352)	lbs/hour

Stream Definitions

# of PSA Units	2.00 units		
Hydrogen Flow In			
Process Fraction	6% of Compressor inlet		
Flowrate	11.37 MMSCF/hr		
Composition (mol%)			
H2	91%	10.34	
CH4	6%	0.68	
C2H6	3%	0.34	
Hydrogen Flow Out (to Compressor inlet)			
Flowrate	9.55 MMSCF/hr		
Composition (mol%)			
H2	99.90%	9.54	
CH4	0.10%	0.01	
C2H6	0.00%	0.00	
Tail Gas (to Fired Heater)			
Flowrate	1.82 MMSCF/hr		
Mol. Wt.	12.43		
Composition	%mole	%lb.	
H2	44.3%	7.1%	4249.32 lbs.
CH4	55.7%	37.0%	28389.59 lbs.
C2H6	18.8%	45.3%	26993.14 lbs.
Additional Hydrogen Makeup Required		0.8052 MMSCF/hr	



Reactor Totals for 4 Reactors		
Reactor Temperature	915.00 °F	
Reactor Pressure	800.00 psi	
H ₂ Recycle Ratio	5.81	
Tar Sand Supply		
Mass	4540.00 Ton/hr	
Temperature	175.00 °F	
Nitrogen Supply		
Volume	114.2 MMSCF/hr	
Temperature	1500.00 °F	
Air-Sand Cooler (2 per Reactor)		
Air Temperature In	50.00 °F	
Air Temperature Out	373.20 °F	
Sand Temperature In	915.00 °F	
Sand Temperature Out	550.00 °F	

In-Out Heat Exchanger (2 units)		
Gaseous Products		
Volume	15.30 MMSCF/hr	
Temperature In	915 °F	
Temperature Out	387 °F	
Hydrogen Supply		
Volume	14.3 MMSCF/hr	
Temperature In	185.00 °F	
Temperature Out	875.00 °F	
Fired Heater (2 units)		
Make-up Required (NG Equivalent)	0.00 MMSCF/hr	
Combustion Air @ 65% excess	6.18 MMSCF/hr	
Steam Generated (1500 psi)	56.775 lb/hr	
Hydrogen Supply		
Volume	14.3 MMSCF/hr	
Temperature In	875.00 °F	
Temperature Out	1500.00 °F	

Compressor Split between 2 units		
Power Required	19,908 n.p.	
Required Steam		
Mass	113546.9 lb/hr	
Pressure	1500 psi	
Temperature (200° Superheat)	800 °F	
Hydrogen Supply		
Volume	30.1 MMSCF/hr	
Temperature In	100 °F	
Pressure In	450 psi	
Temperature Out	185 °F	
Pressure Out	670 psi	

Product Condenser / Separator		
Gaseous Products		
Volume	30.8 MMSCF/hr	
Temperature In	387 °F	
Temperature Out	100 °F	
Condensate	33.5° Crude	
Volume	2500 bbl/hr	
Amine System		
Amine Required	92,800 lb/hr	
Ammonia Production	3093 bbl/hr	
Elemental Sulfur Production	38115 bbl/hr	
Hydrogen Supply		
Volume	28.5 MMSCF/hr	
Temperature In	100 °F	
Temperature Out	100 °F	

Material and Energy Balance

Over-all Considerations

Heat of Reaction	75.00 btu/lb. Bitumen
Organic Feed Temperature	175.00 °F
Sand temperature at reactor inlet	175.00 °F
Reaction temperature	915.00 °F
Sand Feed	1160.00 tons/hr
Number of reactors	1.00

Reactor

Reactor Conditions				
Heat required in reactor		403.57 MMbtu/hr	(latent and sensible heats)	
Heat Generated in Reactor		87.00 MMbtu/hr		
Additional Heat Required		312.49 MMbtu/hr		
Minimum H2 Supplied for Fluidization		20877.91 lbs./hr	3.96	MMSCF/hr
Hydrogen required by reaction		5920.03 lbs./hr	1.12	MMSCF/hr
H2 Required by By-Products		1401.10 lbs./hr	0.27	MMSCF/hr
Hydrogen Supplied for PSA Purge		560.26 lbs./hr	0.11	MMSCF/hr
Additional H2 Supplied		140933.26 lbs./hr	26.71	MMSCF/hr
Total H2		150626.56 lbs./hr	28.54	MMSCF/hr
C1-C2 Flow	at 9%	12683.99 lbs./hr	0.30	MMSCF/hr
Entering H2 Temperature		1500.00 °F		
Cp H2		3.50 btu/lb-°F		
Heat Supplied by C1-C2		4.08 MMbtu/hr		
Heat Supplied by H2		308.41 MMbtu/hr		
Heat Required		0.00 MMbtu/hr	Set equal to 0 by hitting Control G.	
H2 Reactor Recycle ratio		5.81		
Reactor Bottoms Cooler: Assures Efficient Removal of Exiting Solids (removed from design)				
Cold Hydrogen Cooler Stream Fraction		0	0.00 lbs./hr	0.00 MMSCF/hr
Heat Removed		0.00 MMbtu/hr		
Entering Hydrogen temperature		185.00 °F		
Exiting Sand Temperature		915.00 °F		

Total Reactor Requirements

Total Hydrogen Required	150626.56 lbs./hr	28.54 MMSCF/hr
Total Heat Required	403.57 MMbtu/hr	
Total Heat Generated	87.00 MMbtu/hr	

Sand Cooler

# Sand Coolers	1.00 units
----------------	------------

Sand

Sand Flow Rate	580.00 tons/hr
Temperature of Entering Sand	915.00 °F
Temperature of Spent Sand	550.00 °F
Cp Sand	0.19 btu/lb-°F
Heat Removed	80.87 MMbtu/hr

Air Coolant

Air Required for Combustion	1000855.86 lbs./hr	13.08 MMSCF/hr
Cp Air	0.25 btu/lb-°F	
Entering Air Temperature	50.00 °F	
Exiting Air Temperature	373.20 °F	

Steam Generation for Power Production

Sand Temperature IN	550.00 °F	
Sand temperature OUT	300.00 °F	
Heat Available	55.39 MMbtu/hr	
50 lb Steam Generated (@282°F and $H_{v,20}$ of 923 btu/lb)	77144.85 lbs/hr	
Turbine Base Steam rate	67.00 lbs/HP-hr	
Estimated Power Generated	1151.42 HP-hr/hr	857.80 Kwatt-hr/hr

In-Out Heat Exchanger

Hydrogen to be heated		
Hydrogen Flow	150626.56 lbs./hr	28.54 MMSCF/hr
Inlet H2 Temperature	185.00 °F	
Exiting H2 Temperature	875.00 °F	
Total Heat Required	363.76 MMbtu/hr	
Off Gas Heat Supply		
Off Gas flow rate	123651.72 lbs./hr	1.55 MMSCF/hr
Condensables in vapor phase	1298469.06 lbs./hr	2.34 MMSCF/hr
MW off-gas	30.30 lb/lb-mole	
Cp Vapor	0.55 btu/lb-°F	
Cp Liquid	0.45 btu/lb-°F @70°F	
Cp NonCondensables	3.00 btu/lb-°F	
Heat of Vaporization	65.00 btu/lb	
Hydrogen Recycle Flow in Stream	140933.26 lbs./hr	26.71 MMSCF/hr
Inlet Temperature	915.00 °F	
Exit Temperature	366.76 °F	

Product Condenser / Cooler

# Product Condensers	1.00 units		
Product Side			
Entering Temperature	366.76 °F		
Exiting Temperature	100.00 °F		
Product	864000.00 lbs./hr	2500.00 bbl/hr	60,000.00
Coke Residue	67008.14 lbs./hr		
Heat Removal	H2	131.58 MMbtu/hr	
	Off Gas	18.14 MMbtu/hr	
	Condensate	240.27 MMbtu/hr	
	Total	389.99 MMbtu/hr	
Cooler requirement			
Entering Temperature	50.00 °F		
Exiting Water Temperature	75.00 °F		
Water Flow Rate	15,599,774 lbs./hr	258,189 ft3/hr	

Compressor

Hydrogen Side		
Flow Rate	501940.86 SCF/min	30.12 MMSCF/hr
Pressure Out	670.00 psi	
Pressure In	450.00 psi	
DP	220.00 psi	
γ (Cp/Cv)	1.40	
# Stages	3	
Temperature Inlet	100.00 °F	
Mechanical Efficiency	51%	
Pb/Pa	1.14	
Power Requirement per Stage	6635.91 hp	
Total Power Required	19907.72 hp	
Outlet Temperature	185.00 °F	
Steam Supply		
Pressure	1500.00 psi	
Temperature	800.00 °F	
Degree Superheat	200.00 °F	
Saturation Temperature	596.20 °F	
Steam Heat Value	1364.00 btu/lb	
Flow Rate	113549.90 lbs./hr	

PSA: Light Gas Scrubber

# of PSA Units		2.00 units
Hydrogen Flow In		
Process Fraction	6%	of Compressor inlet
Flowrate	3.00	MMSCF/hr
Composition (mol%)		
H ₂	91%	2.73
CH ₄	6%	0.18
C ₂ H ₆	3%	0.09
Hydrogen Flow Out (to Compressor inlet)		
Flowrate	2.52	MMSCF/hr
Composition (mol%)		
H ₂	99.90%	2.52
CH ₄	0.10%	0.00
C ₂ H ₆	0.00%	0.00
Tail Gas (to Fired Heater)		
Flowrate	0.48	MMSCF/hr
Mol. Wt.	12.43	
Composition		
	%mole	%lb.
H ₂	44.3%	7.1%
CH ₄	37.0%	47.6%
C ₂ H ₆	18.8%	45.3%
Additional Hydrogen Makeup Required		0.2123 MMSCF/hr

Fired Heater

# Fired Heaters		1.00 units
Products to be Heated		
Hydrogen Flowrate	150626.56 lbs./hr	28.54 MMSCF/hr
Hydrogen Temperature	875.00 °F	
Water Flow Rate	113549.90 lbs./hr	
Water Temperature	200.00 °F	
Heat Duty	596.04	MMBtu/hr
Fuel to Fired Heater (Heat of combustion temperature corrected.)		
Off-gas lights (C-3's, etc.)	Flowrate	0.14 MMSCF/hr
	Heat Supplied	308.72 MMBtu/hr
PSA Tail Gas	Flowrate	0.51 MMSCF/hr
	Heat Supplied	367.35 MMBtu/hr
	Make-up H ₂	0.53 MMSCF/hr
Make-up Natural Gas Equiv.	Flowrate	0.00 MMSCF/hr (as Natural Gas)
	Heat Supplied	0.00 MMBtu/hr
C3's (Fuel produced by the process)		
Flow Rate	16486.89 lbs./hr	0.14 MMSCF/hr
Heat of Combustion	20000.00 btu/lb	
Cp	0.60 btu/lb-°F	
Temperature in	75.00 °F	
Heat Supplied (After temperature correction)	308.72	MMBtu/hr
Tail Gas from PSA		
Flow Rate	15724.56 lbs./hr	0.51 MMSCF/hr
Heat of Combustion	25130.79 btu/lb	
Cp	0.88 btu/lb-°F	
Temperature in	185.00 °F	
Heat Supplied (After temperature correction)	367.35	MMBtu/hr
Make-up Methane (If Needed)		
Combustion Temperature	2200.00 °F	
Heat Remaining to be supplied by Methane	0.00	MMBtu/hr
Flow Rate	0.00 lbs./hr	0.00 MMSCF/hr
Heat of Combustion (After temperature correction)	20227.00 btu/lb	
Temperature in	75.00 °F	
Combustion Air		
Air Required for Combustion	500427.93 lbs./hr	8.18 MMSCF/hr
Air Supplied 65% Excess	1000855.86 lbs./hr	16.35 MMSCF/hr
Excess Heat (If Produced)		
Excess Heat from process (combustion of C3 Stream)	80.03 MMBtu/hr	0.04 MMSCF/hr (as C3 stream)

Determination of CO₂ Production

Fired Heater Combustion (Includes purge stream from PSA)

Assuming:	CH ₄ + 2O ₂	→	CO ₂ + 2H ₂ O	
then	0.67	MMSCF/hr CH ₄	→	672,478 SCF/hr from fired heater
Assuming:	2C ₂ H ₆ + 7O ₂	→	4CO ₂ + 6H ₂ O	
then	0.34	MMSCF/hr C ₂ H ₆	→	682,027 SCF/hr from fired heater

Hydrogen Plant Combustion and Production

Assuming:	CH ₄ + 2O ₂	→	CO ₂ + 2H ₂ O	
then	1.84	MMSCF/hr CH ₄	→	1,835,819 SCF/hr from hydrogen plant
CO ₂ Produced during processing			2,560 lbs. →	22,051 SCF/hr from Production
				3,212,375 SCF/hr CO ₂ Produced

1284.95 SCF CO ₂ per bbl.
• or •
149.18 Lbs. CO ₂ per bbl.

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.

Claims

We claim:

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

providing a reactor,

providing a first feed stream comprising an oil sand,

providing a second feed stream comprising a hydrocarbon containing liquid,

providing a gas feed comprising hydrogen,

contacting said oil sand and said hydrocarbon containing liquid with said hydrogen in said reactor,

producing a reactor gas stream.

Claim 2. The process of claim 1, wherein said first feed stream comprises a Canadian Athabasca oil sand

Claim 3. The process of claim 1, wherein said first feed stream comprises an oil sand comprising a bitumen in a concentration of 10 wt. % or greater.

Claim 4. The process of claim 1, further comprising the step of heating tailings or raw sand to a temperature in a range of about 800 °F to about 1000 °F.

Claim 5. The process of claim 1, wherein said first feed stream comprises *in situ* bitumen, and further comprising the step of locating said reactor less than about 10 miles of the wellhead producing said bitumen.

Claim 6. The process of claim 1, wherein said first feed stream is fed in a continuous stream to the reactor

Claim 7. The process of claim 1, wherein said first feed stream comprises reclaimed sand.

Claim 8. The process of claim 1, wherein said second feed stream is at a temperature in a range of about 150 °F to about 400 °F when fed to said reactor.

Claim 9. The process of claim 1, wherein said second feed stream is at a temperature in a range of about 200 °F to about 300 °F when fed to said reactor.

Claim 10. The process of claim 1, wherein said second feed stream is at a temperature of about 300°F when fed to said reactor.

Claim 11. The process according to claim 1, wherein said second feed stream comprises bitumen produced by wet extraction methods

Claim 12. The process according to claim 1, wherein said second feed stream comprises bitumen produced by steam assisted gravity drainage

Claim 13. The process according to claim 1, wherein said second feed stream comprises liquid bitumen.

Claim 14. The process of claim 1, wherein said gas feed comprises hydrogen in a range of 50 wt. % to 100 wt. %

Claim 15. The process of claim 1, wherein said gas feed comprises hydrogen in a range of 90 wt. % to 100 wt. %

Claim 16. The process of claim 1, wherein said gas feed comprises about 94 wt. % hydrogen.

Claim 17. The process of claim 1, wherein said gas feed comprises about 10 wt. % or less of methane.

Claim 18. The process of claim 1, wherein said gas feed comprises about 10 wt. % or less of ethane.

Claim 19. The process of claim 1, wherein said gas feed comprises about 6 wt. % or less of methane and ethane in combination.

Claim 20. Process of claim 1, wherein said gas feed comprises recycle hydrogen.

Claim 21. The process of claim 20, wherein said recycle hydrogen is heated in a range of about 800°F to about 1500°F.

Claim 22. The process of claim 20, wherein said recycle hydrogen is heated in a range of about 900°F to about 1200°F.

Claim 23. The process of claim 20, wherein said recycle hydrogen is heated to about 1500°F.

Claim 24. The process of claim 1, further comprising the step of:
condensing said reactor gas stream,
producing a synthetic crude oil having an API gravity in the range of about 25°API to about 40°API .

Claim 25. The process according to 24, wherein said API gravity is in the range of about 28°API to about 32°API.

Claim 26. The process according to 24, further comprising:

providing a reactor having a catalyst, and

wherein said API gravity is in the range of about 32°API to about 36°API.

Claim 27. The process of claim 24, wherein said synthetic crude oil has an API gravity of about 34.

Claim 28. The process of claim 24, wherein said synthetic crude oil has a sulfur concentration in a range of about 0.1 wt. % to about 0.5 wt. %.

Claim 29. The process of claim 24, wherein said synthetic crude oil has a sulfur concentration of about 0.2 wt. % sulfur or less.

Claim 30. The process of claim 1, wherein said first feed stream comprises oil sand pieces which are in a range of about ¼ to about 1 inch mesh.

Claim 31. The process of claim 1, wherein said process is a continuous process

Claim 32. The process of claim 1, wherein said reactor is a fluidized bed reactor.

Claim 33. The process of claim 1, wherein said contacting occurs at a temperature in a range of 850°F to 1000°F

Claim 34. The process of claim 1, further comprising the step of producing a processed sand effluent stream.

Claim 35. The process of claim 1, further comprising the step of:
providing a synthetic crude oil feed to a refinery process unit operation.

Claim 36. The process of claim 1, further comprising the step of:
providing a synthetic crude oil feed to a chemical process unit operation.

Claim 37. The process of claim 1, wherein at least one of said first feed stream or said second feed stream comprises bitumen processed to be different from its *in situ* form.

Claim 38. The process of claim 1, wherein at least one of said first feed stream or said second feed stream comprises an *in situ* bitumen.

Claim 39. The process of claim 1, further comprising the step of, providing the heat necessary to bring the first feed stream to a temperature in a range of 800°F to 1200°F.

Claim 40. The process of claim 1, further comprising the step of, providing the heat necessary to bring the second feed stream to a temperature in a range of 800°F to 1200°F

Claim 41. The process of claim 1, further comprising the step of:
providing a fluidized bed reactor.

Claim 42. The process of claim 41, further comprising the step of:

providing a fluidizing medium to said reactor which comprises hydrogen.

Claim 43. The process of claim 1, wherein at least said first feed stream or said second feed stream is fed cocurrently with the gas feed stream.

Claim 44. The process of claim 1, further comprising the step of:

operating said reactor at a temperature of about 900°F and about 600 psig.

Claim 45. The process of claim 1, further comprising the step of preheating said reactor to a temperature of about 900°F and about 600 psig prior to feeding said second feed stream.

Claim 46. The process of claim 1, further comprising the step of preheating said reactor to a temperature of about 900°F and about 600 psig prior to feeding said second feed stream, wherein said second feed stream is preheated to about 300°F.

Claim 47. The process of claim 1, wherein at least one of said first feed or said second feed is transported at least in part through a steam-traced line from another bitumen extraction process.

Claim 48. The process according to claim 1, further comprising the step of: employing a catalyst in said reactor.

Claim 49. The process of claim 1, further comprising the step of: feeding said first feed stream and said second feed stream to said reactors through separate reactor feed inlets.

Claim 50. The process of claim 41, further comprising the step of:
providing a fluid bed medium separately from the second feed stream to said fluid
bed reactor.

Claim 51. The process of claim 41, further comprising the step of:
providing a fluid bed medium separately from the first feed stream and second
feed stream to said fluid bed reactor.

Claim 52. The process of claim 41, further comprising the step of:
providing a fluid bed medium separately from the first feed stream to said fluid
bed reactor.

Claim 53. The process of 41, wherein said fluid bed reactor comprises a fluid
bed having a shale.

Claim 54. The process of 41, wherein said fluid bed reactor comprises a fluid
bed having a sand.

Claim 55. The process of claim 1, wherein a spent sand from the reactor is
slurried with the water and transported back to its original or a different location for disposal.

Claim 56. The process of claim 1, wherein a spent sand comprises a coke
from the reaction process.

Claim 57. The process of claim 1, further comprising the step of:
processing reclaimed sand and extracted bitumen in said reactor.

Claim 58 A process for producing synthetic crude oil, comprising the steps of:

providing a reactor,
providing a first feed stream comprising kerogen,
providing a second feed stream comprising a hydrocarbon containing liquid,
providing a gas feed comprising hydrogen,
contacting said kerogen with said hydrogen in said reactor,
producing a reactor gas stream.

Claim 59. The process of claim 58, further comprising the step of:
providing said first feed stream comprising oil shale having kerogen.

Claim 60. The process of claim 58, wherein said gas feed comprises
hydrogen in a range of 50 wt. % to 100 wt. %

Claim 61. The process of claim 58, wherein said gas feed comprises
hydrogen in a range of 90 wt. % to 100 wt. %

Claim 62. The process of claim 58, wherein said gas feed comprises about 94
wt. % hydrogen.

Claim 63. The process of claim 58, wherein said gas feed comprises about 6
wt. % or less of methane.

Claim 64. The process of claim 58, wherein said gas feed comprises about 6
wt. % or less of ethane.

Claim 65. The process of claim 58, wherein said gas feed comprises about 6 wt. % or less of methane and ethane in combination..

Claim 66. The process of claim 58, wherein said kerogen has been processed to be different from *in situ* kerogen.

Claim 67. The process of claim 58, wherein said kerogen is *in situ* produced kerogen.

Claim 68. The process of claim 58, further comprising the step of:
heating said first feed to a temperature in a range of about 800 °F to about 1200 °F.

Claim 69. The process of claim 58 , further comprising the step of:
heating said second feed to a temperature in a range of about 800 °F to about 1200 °F.

Claim 70. The process of claim 58, further comprising the step of:
condensing said reactor gas stream,
producing a synthetic crude oil having an API gravity in the range of about 25°API to about 40°API.

Claim 71. The process of claim 70, wherein said synthetic crude oil has an API gravity of about 34°API.

Claim 72. The process of claim 70, wherein said synthetic crude oil has a sulfur concentration in a range of about 0.1 wt % to about 0.5 wt %.

Claim 73. The process of claim 70, wherein said synthetic crude oil has a sulfur concentration of about 0.2 wt % sulfur or less.

Claim 74. The process of claim 58, wherein said first feed stream comprises oil shale pieces which are in a range of about 1/4 to about 1 inch mesh.

Claim 75. The process of claim 58, wherein said process is a continuous process.

Claim 76. The process of claim 58, wherein said first feed stream is continuously fed to said reactor.

Claim 77. The process of claim 58, wherein said reactor is a fluidized bed reactor.

Claim 78. Process of claim 58, wherein said gas feed comprises recycle hydrogen.

Claim 79. The process of claim 78, wherein said recycle hydrogen is heated in a range of about 800 °F to about 1500 °F.

Claim 80. The process of claim 78, wherein said recycle hydrogen is heated to about 1500°F.

Claim 81. The process of claim 58, further comprising the step of producing an effluent stream of processed oil shale which is not in a powdered form.

Claim 82. The process of claim 58, further comprising the step of producing an effluent stream of processed oil shale comprising oil shale pieces which are greater than 1/4 inch mesh.

Claim 83. The process of claim 58, wherein said first feed stream comprises surface mined oil shale comprising kerogen.

Claim 84. The process of claim 58, wherein said first feed stream comprises an oil shale comprising kerogen in a concentration of 5 wt % or greater.

Claim 85. The process of claim 58, further comprising the step of: providing a synthetic crude oil feed to a refinery process unit operation.

Claim 86. The process of claim 58, further comprising the step of: providing a synthetic crude oil feed to a chemical process unit operation.

Claim 87. The process of claim 58, wherein at least one of said first feed stream comprises oil shale processed to be different from its *in situ* form.

Claim 88. The process of claim 58, wherein said second feed stream comprises a bitumen.

Claim 89. The process of claim 58, wherein said second feed stream comprises an *in situ* bitumen.

Claim 90. The process of claim 58, wherein said second feed stream comprises a bitumen different from its *in situ* form.

Claim 91. The process of claim 77, comprising a fluidizing medium which comprises hydrogen.

Claim 92. The process of claim 58, wherein said first feed stream is fed cocurrently with the gas feed stream.

Claim 93. The process of claim 58, further comprising the step of operating said reactor at a temperature of about 900°F and about 600 psig.

Claim 94. The process of claim 58, further comprising the step of preheating said reactor to a temperature of about 900°F and about 600 psig prior to feeding said second feed stream.

Claim 95. The process of claim 58, further comprising the step of preheating said reactor to a temperature of about 900°F and about 600 psig prior to feeding said first feed stream, wherein said second first feed stream is preheated to about 300°F.

Claim 96. The process of claim 58, wherein at least said second feed stream is transported at least in part through a steam-traced line from an extraction facility.

Claim 97. The process of claim 58, further comprising the step of: processing reclamation material in said reactor.

Claim 98. The process of claim 58, further comprising the step of: processing reclaimed sand and oil shale in said reactor.

Claim 99. The process of claim 58, wherein the first feed stream comprises both kerogen and bitumen.

Claim 100. The process of claim 59, wherein the first feed stream comprises an oil sand and an oil shale.

Claim 101. The process according to claim 58, further comprising the step of: employing a catalyst in said reactor.

Claim 102. The process of claim 77, further comprising the step of: providing a fluid bed medium separately from the first feed stream.

Claim 103. The process of claim 77, wherein said fluid bed medium is shale

Claim 104. The process of claim 77, wherein said fluid bed medium is sand.

Claim 105. An apparatus for producing synthetic crude oil according to the process of any one of claims 1 – 57.

Claim 106. An system for producing synthetic crude oil according to the process of any one of claims 1 – 57.

Claim 107. A control system for producing synthetic crude oil according to the process of any one of claims 1 – 57.

Claim 108. An apparatus for producing synthetic crude oil according to the process of any one of claims 58 – 104.

Claim 109. An system for producing synthetic crude oil according to the process of any one of claims 58 – 104.

Claim 110. A control system for producing synthetic crude oil according to the process of any one of claims 58 – 104.

Application number / numéro de demande: 2467372

Figures: 9,

Pages: _____

Drawing

Unscannable items
received with this application
(Request original documents in File Prep. Section on the 10th floor)

Documents reçu avec cette demande ne pouvant être balayés
(Commander les documents originaux dans la section de préparation des dossiers au
10^{ème} étage)

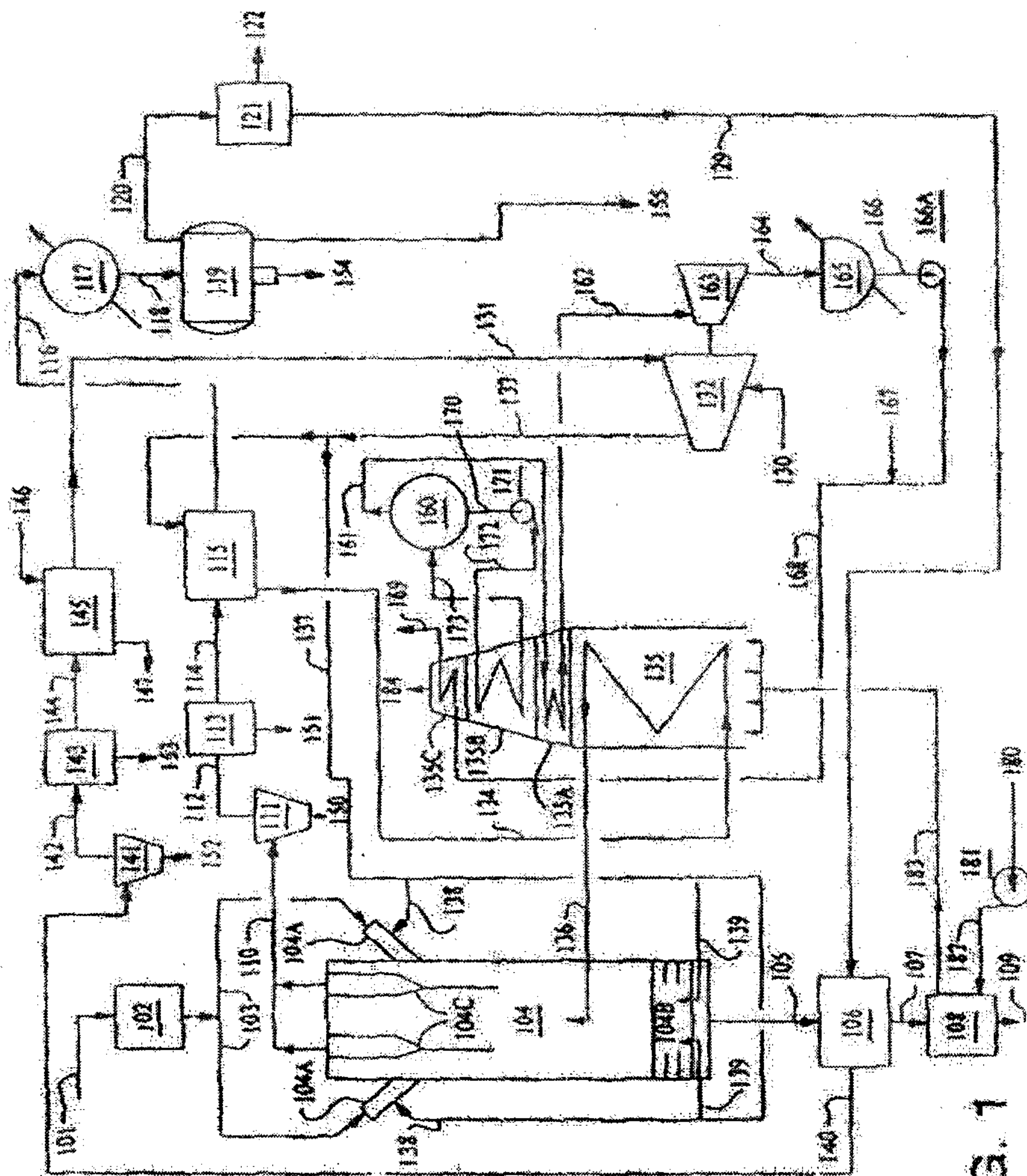


FIG. 1

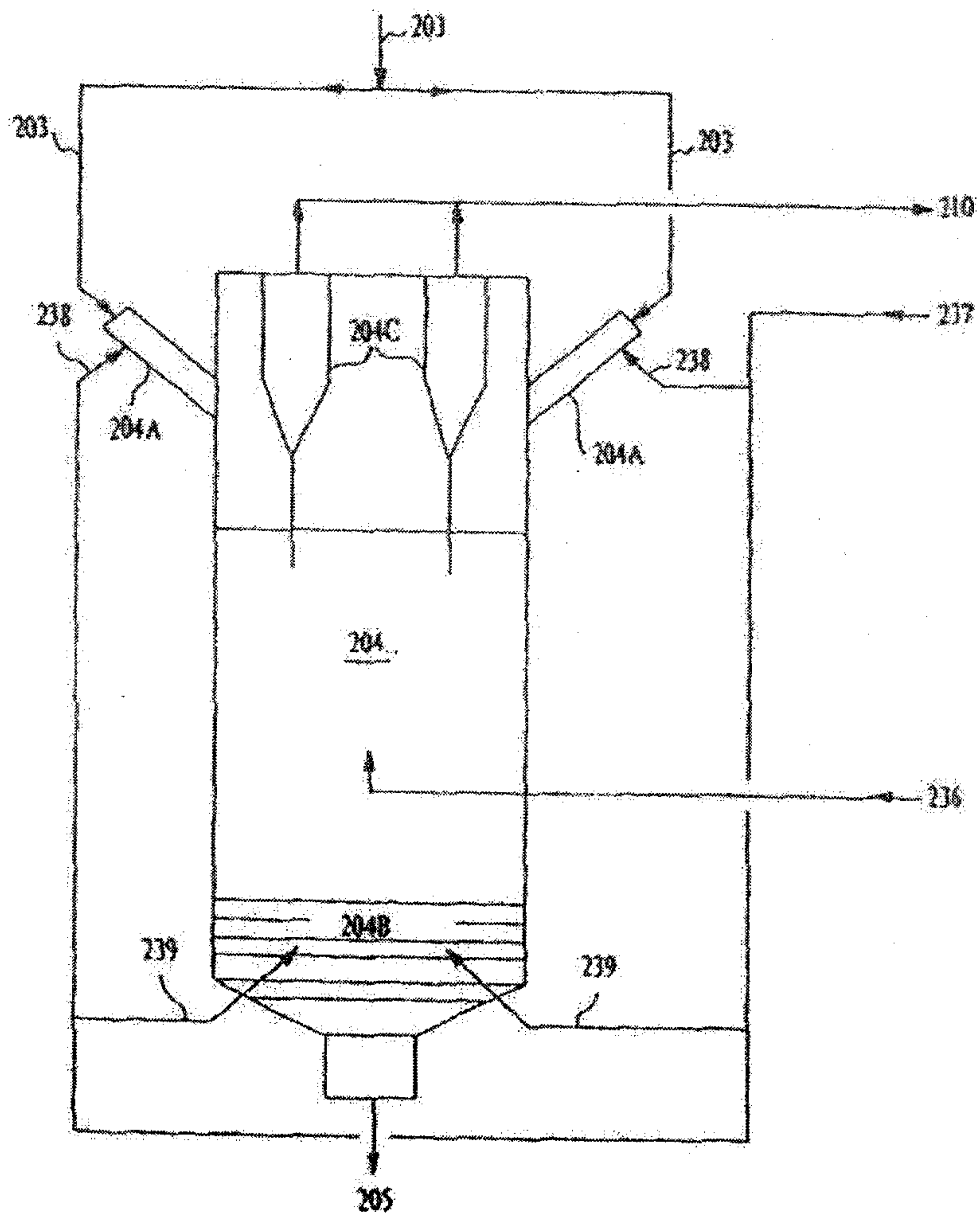


FIG. 2

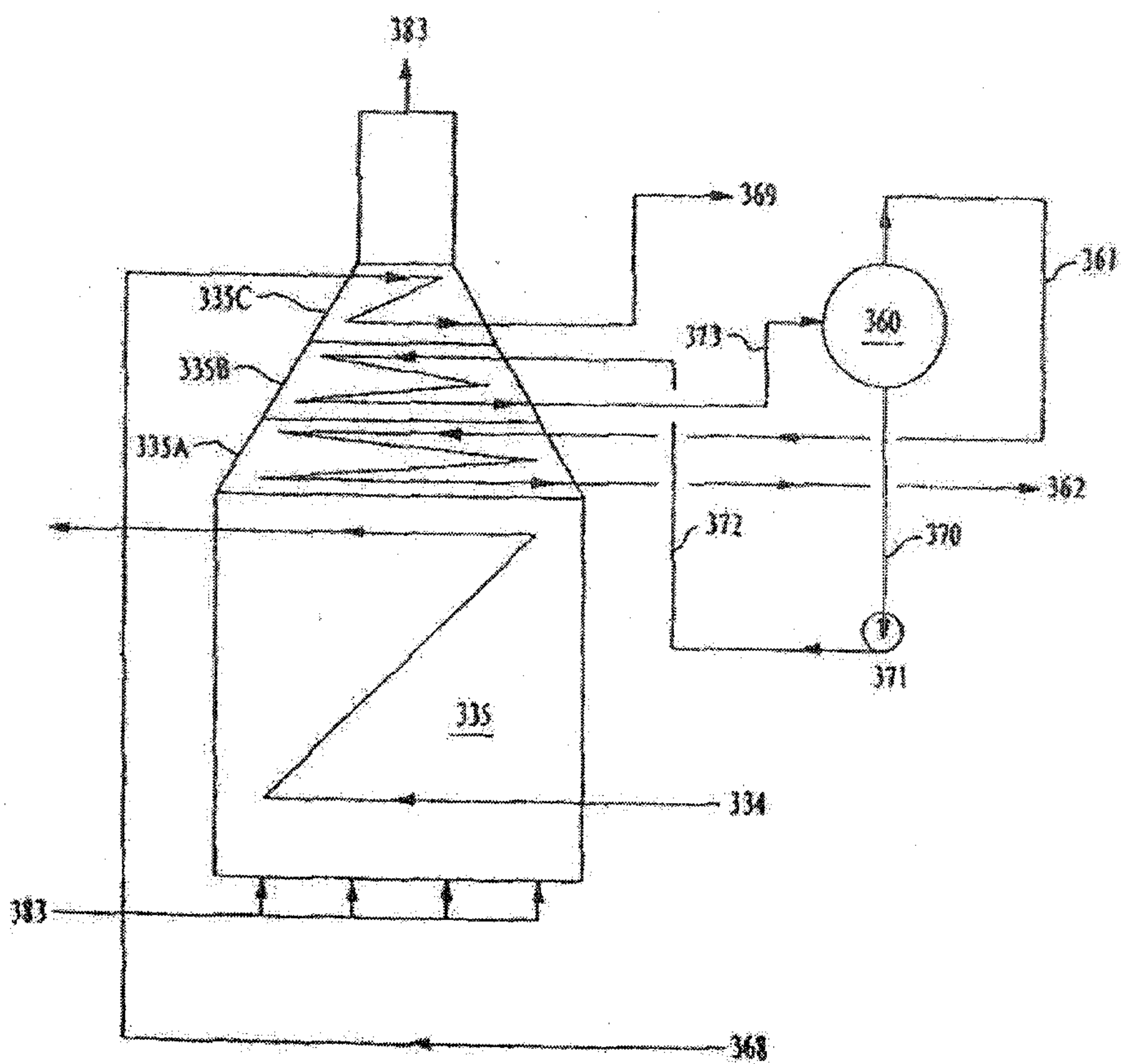
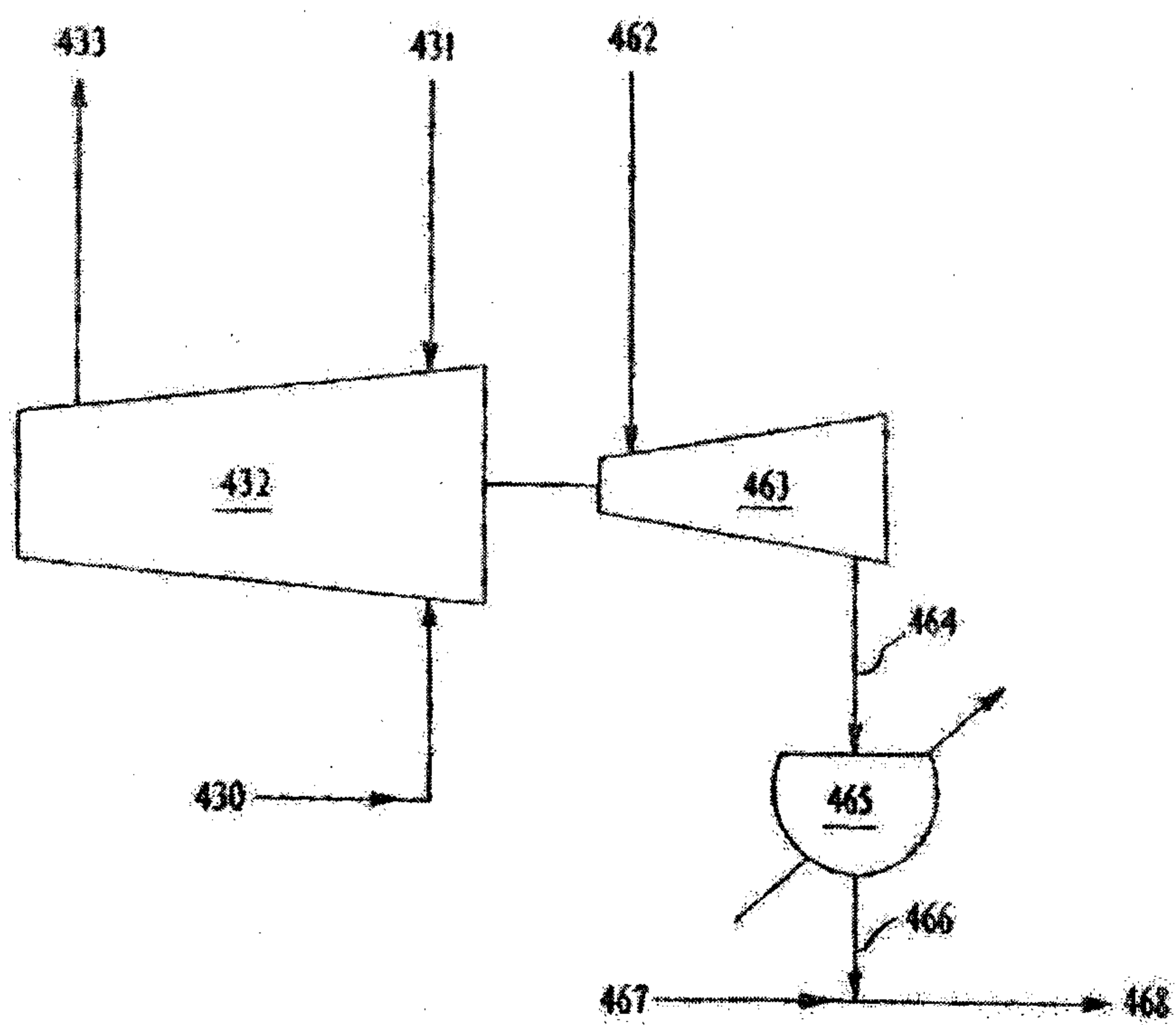


FIG. 3

**FIG. 4**

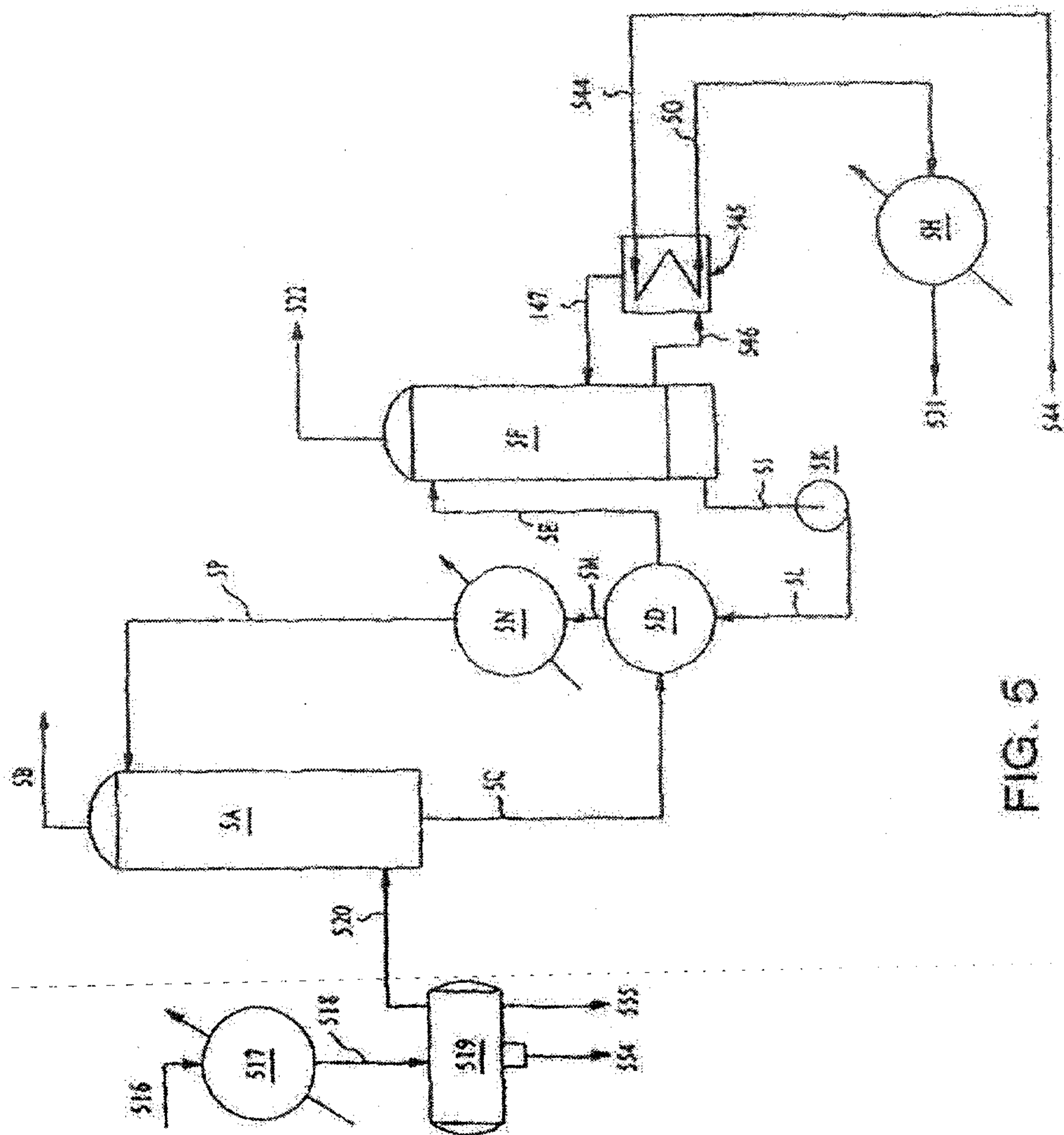


FIG. 5

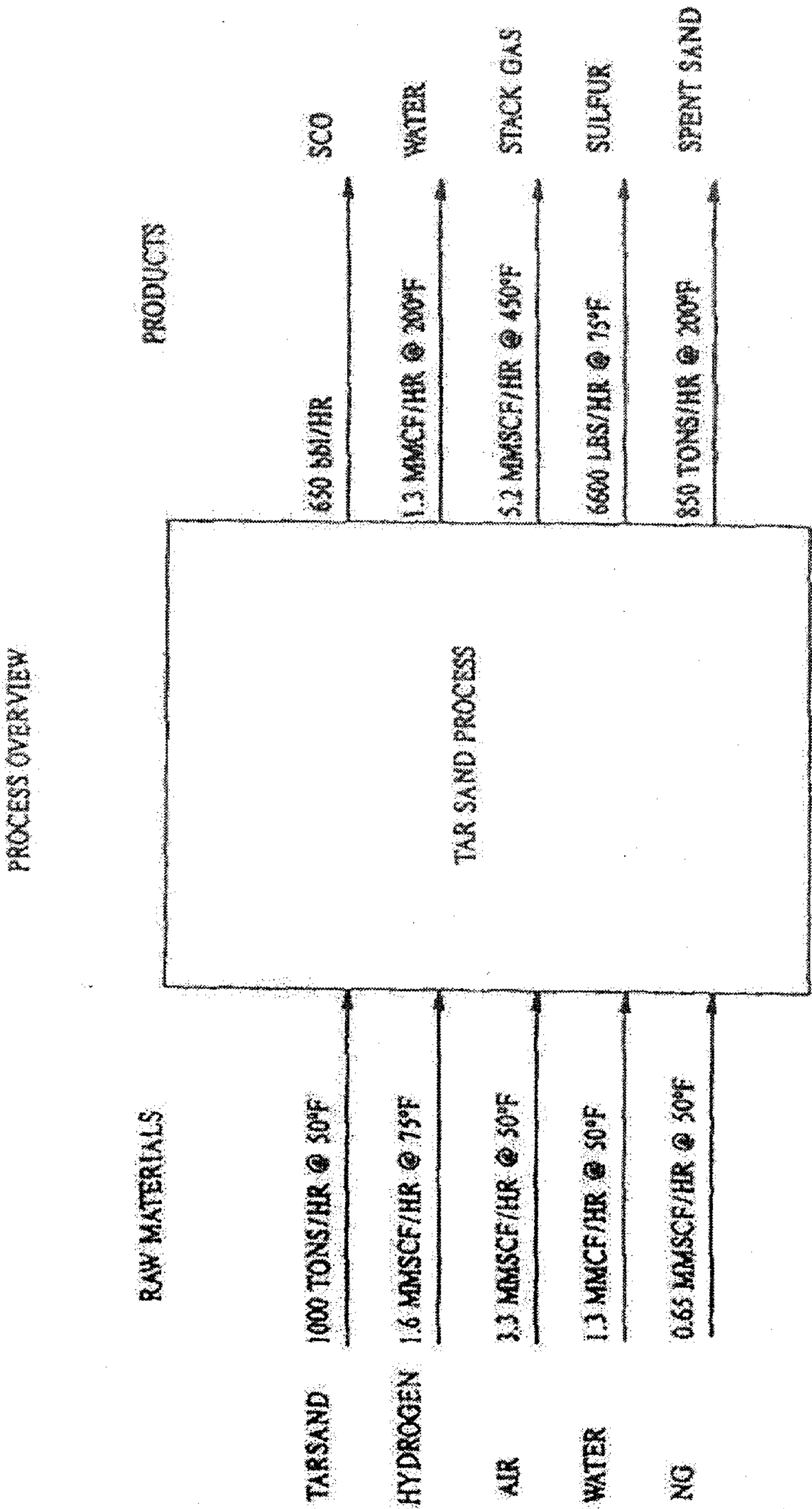


FIG. 6

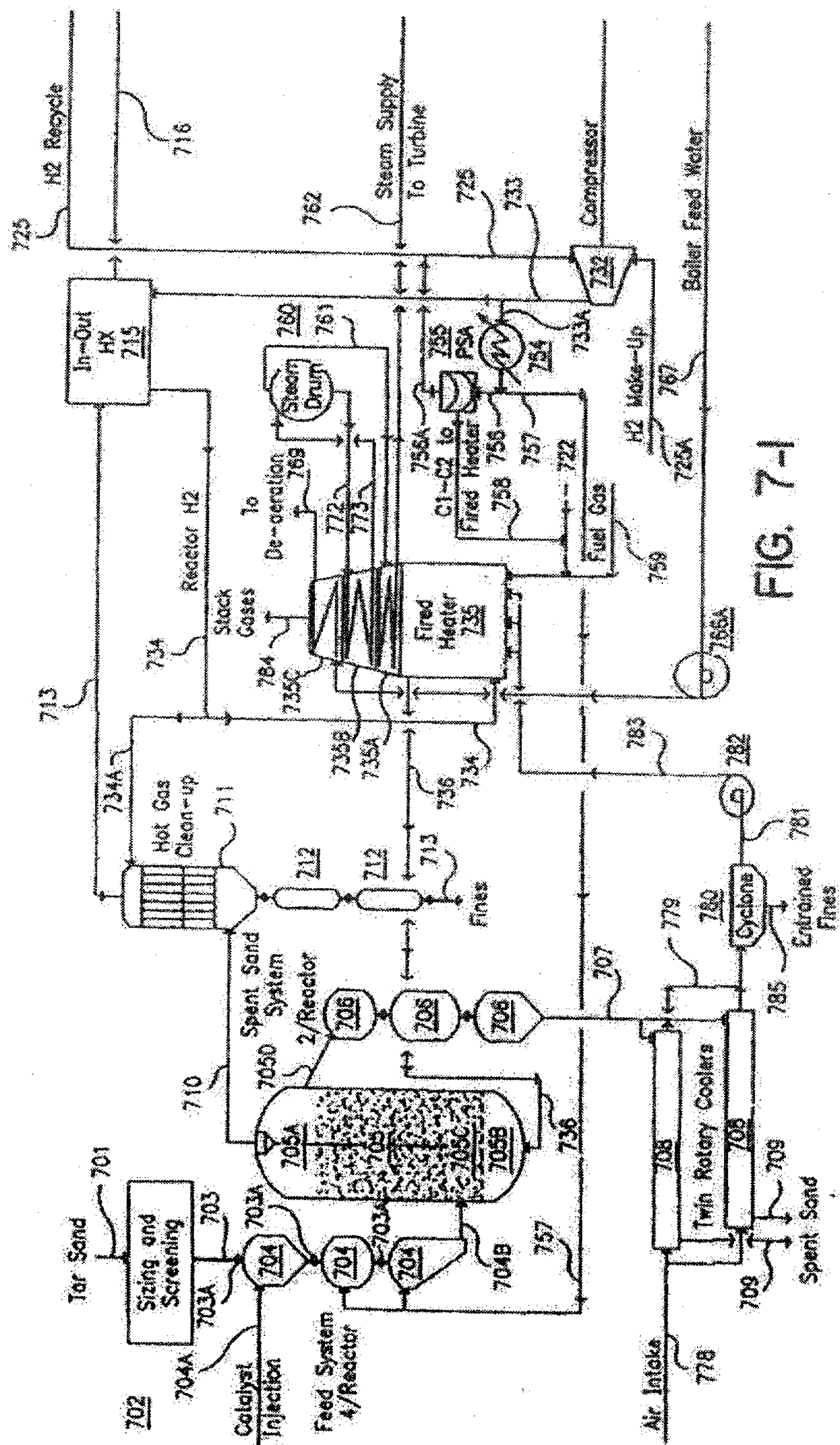


FIG. 7-I

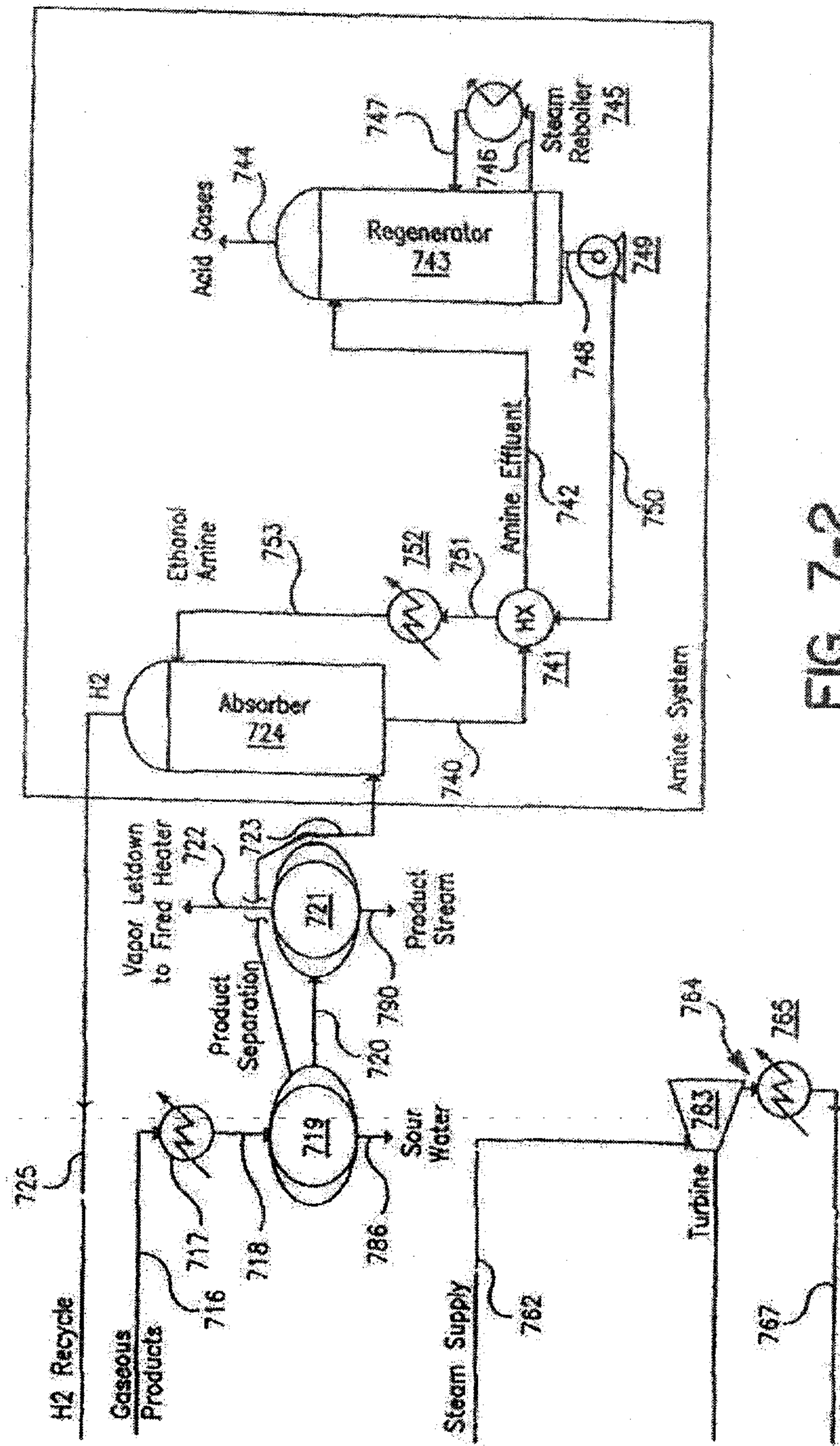


FIG. 7-2

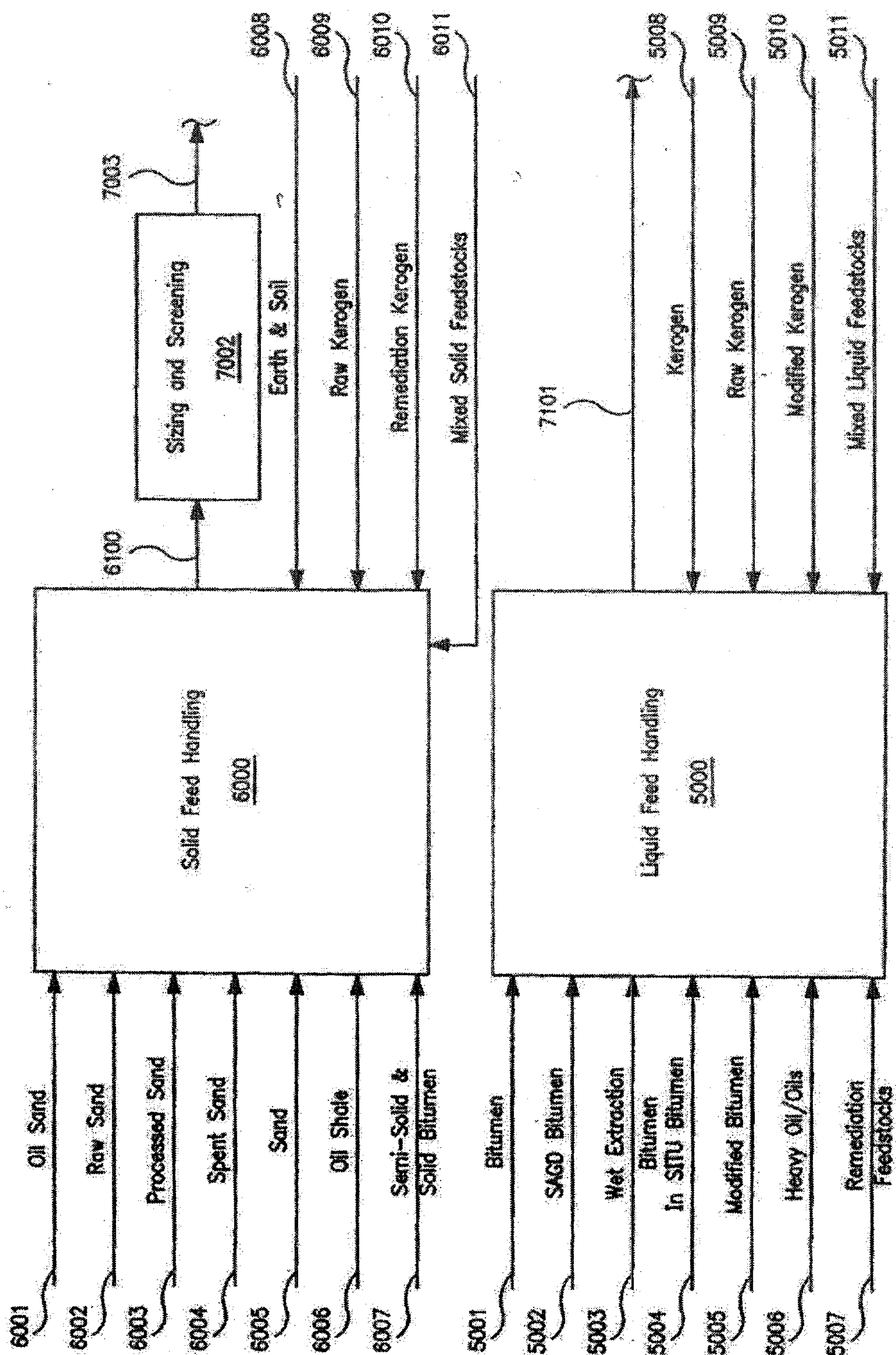


FIG. 7-3

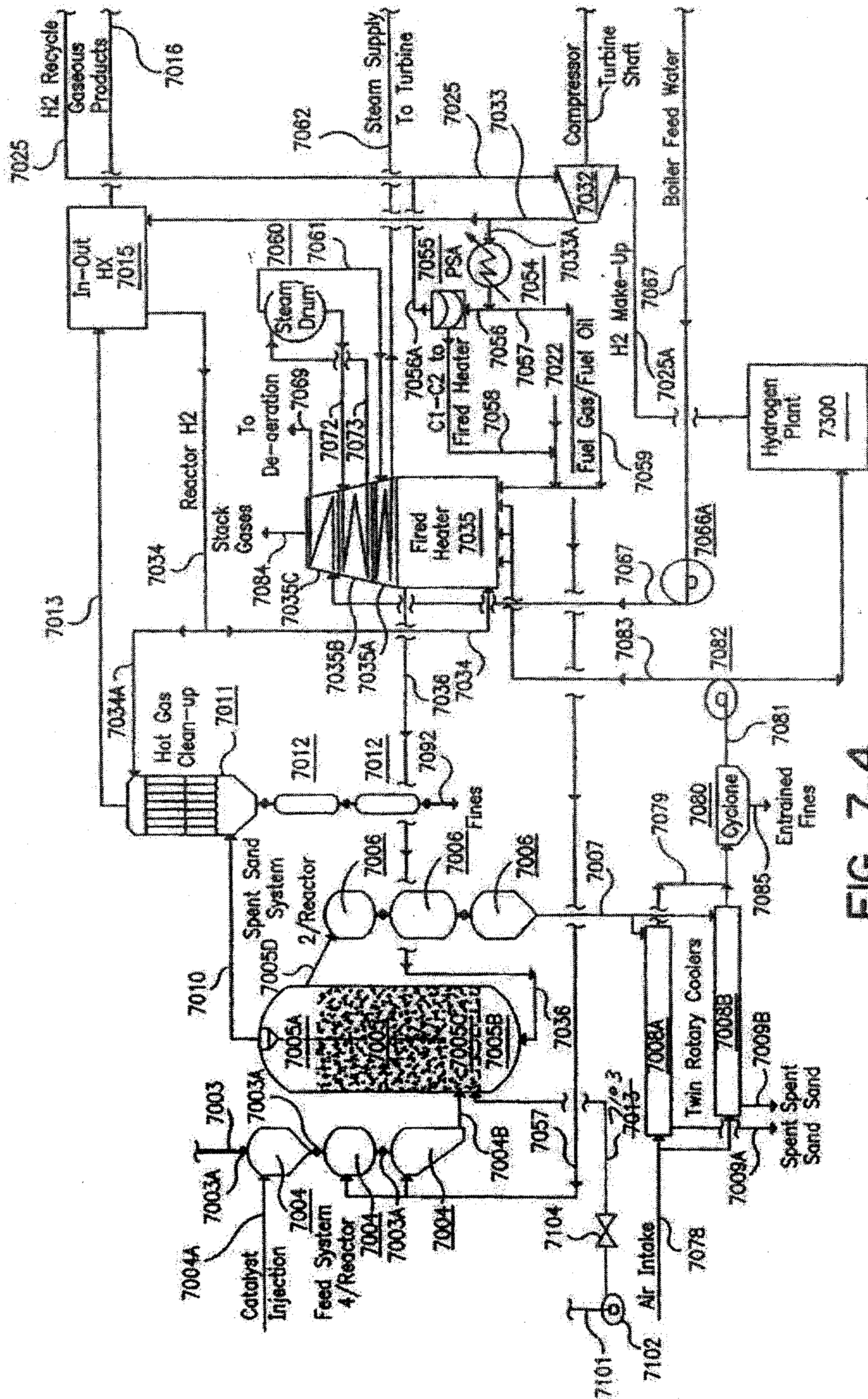


FIG. 7-4

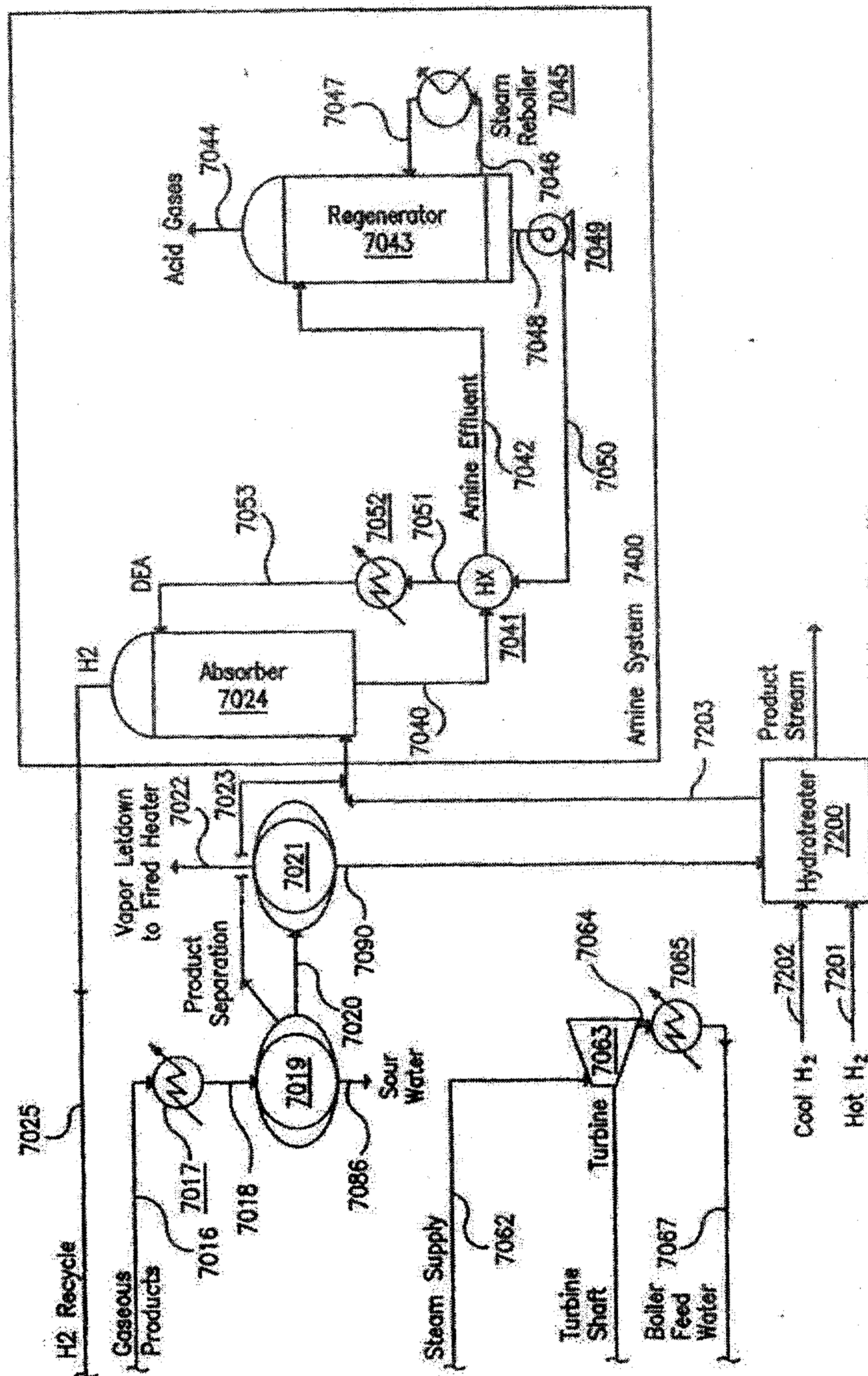


FIG. 7-5

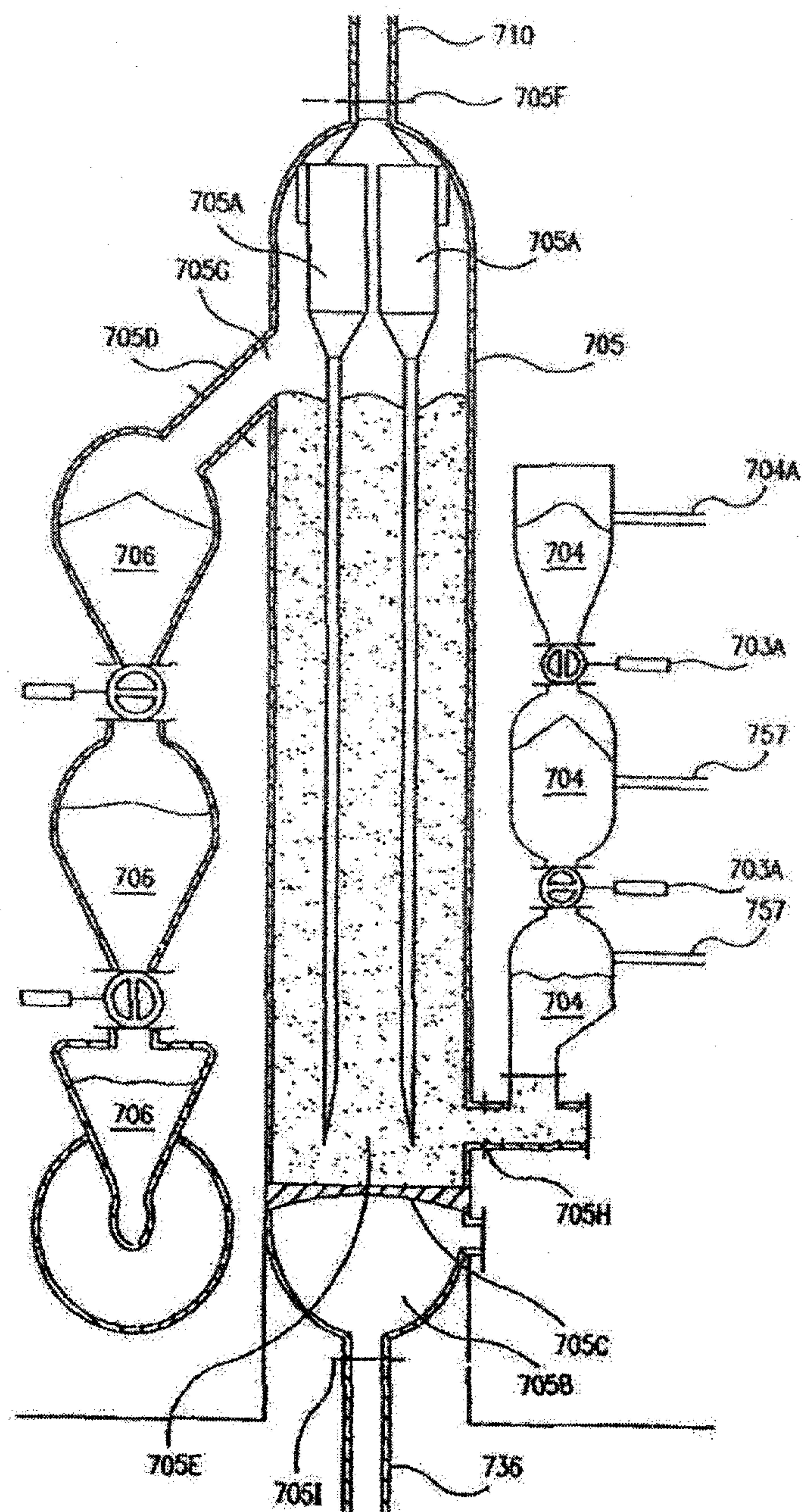
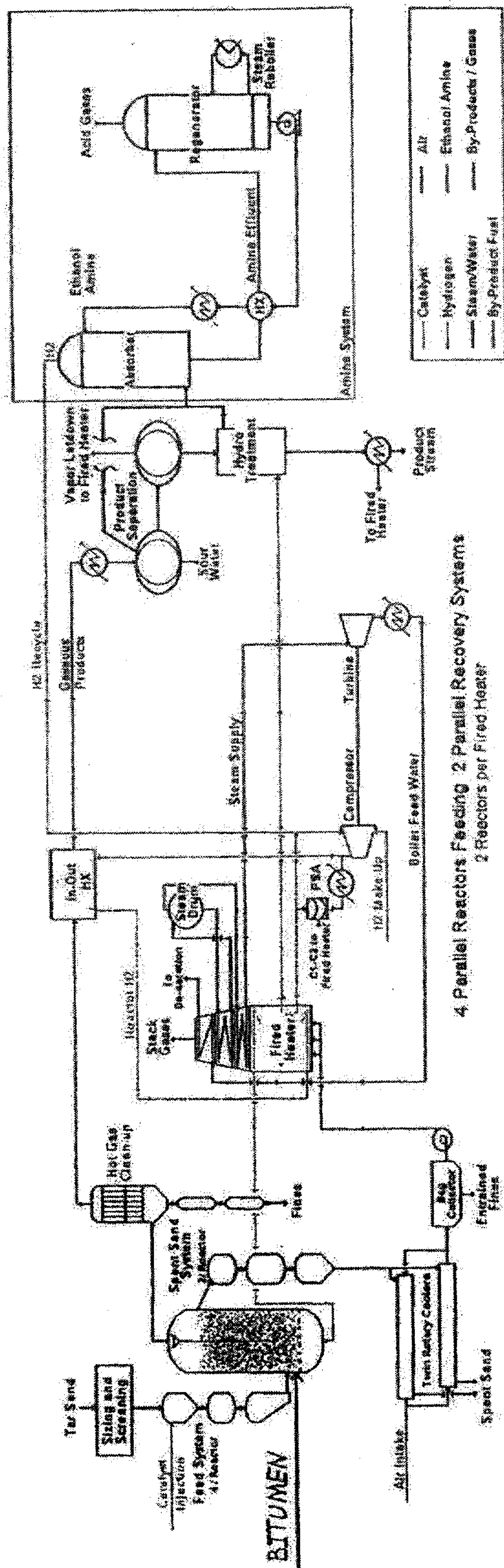


FIG. 8

Figure 9



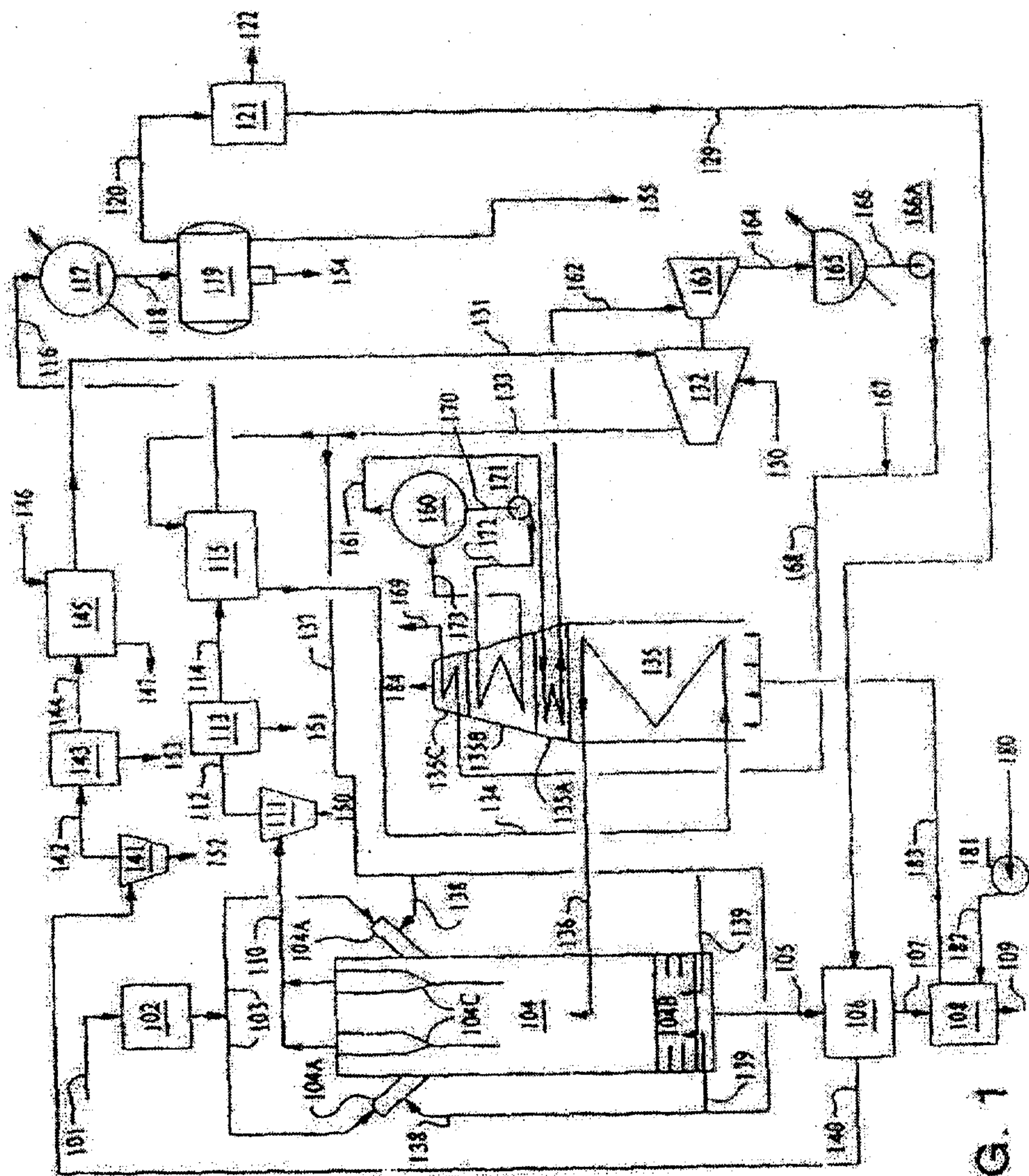


FIG. 1

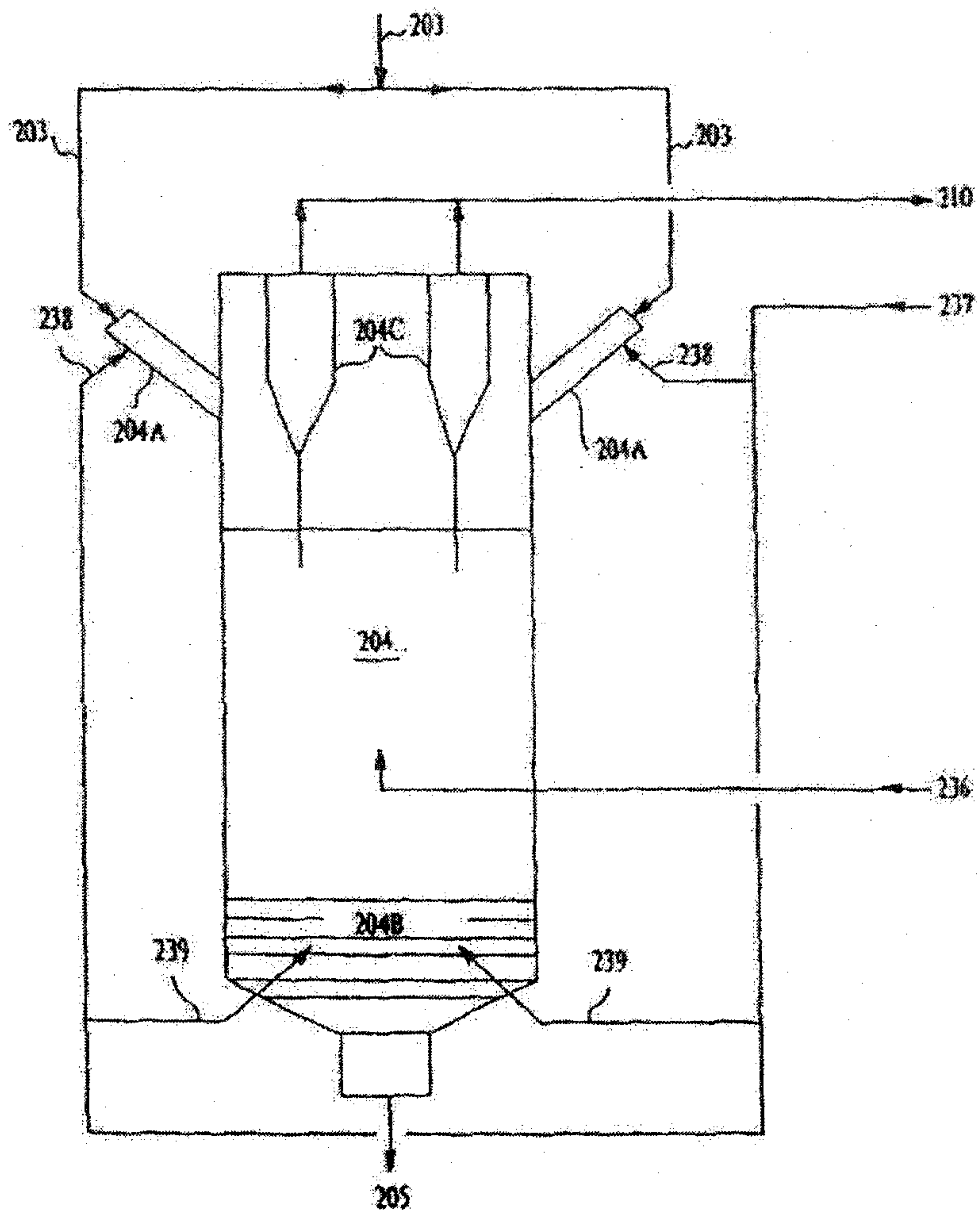


FIG. 2

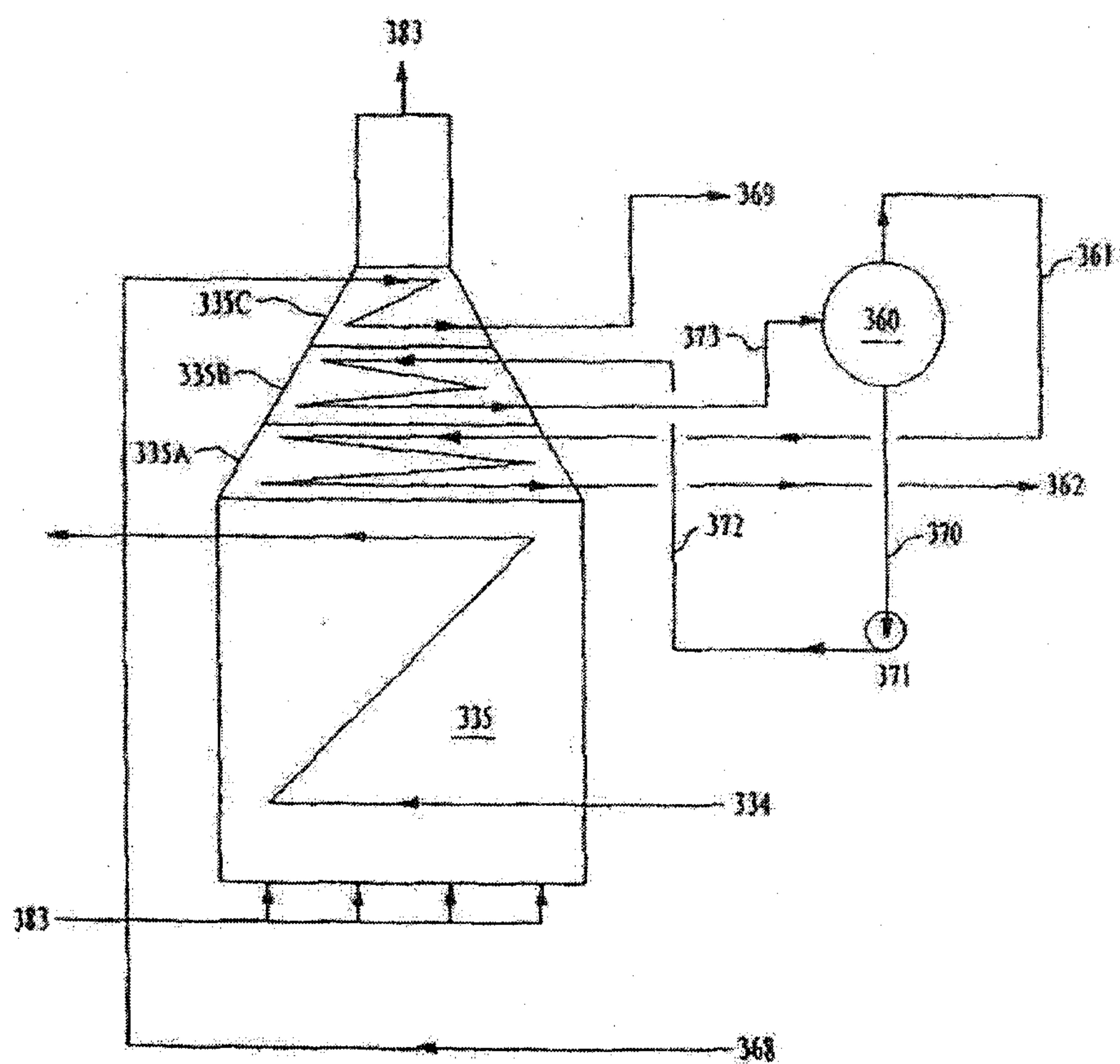


FIG. 3

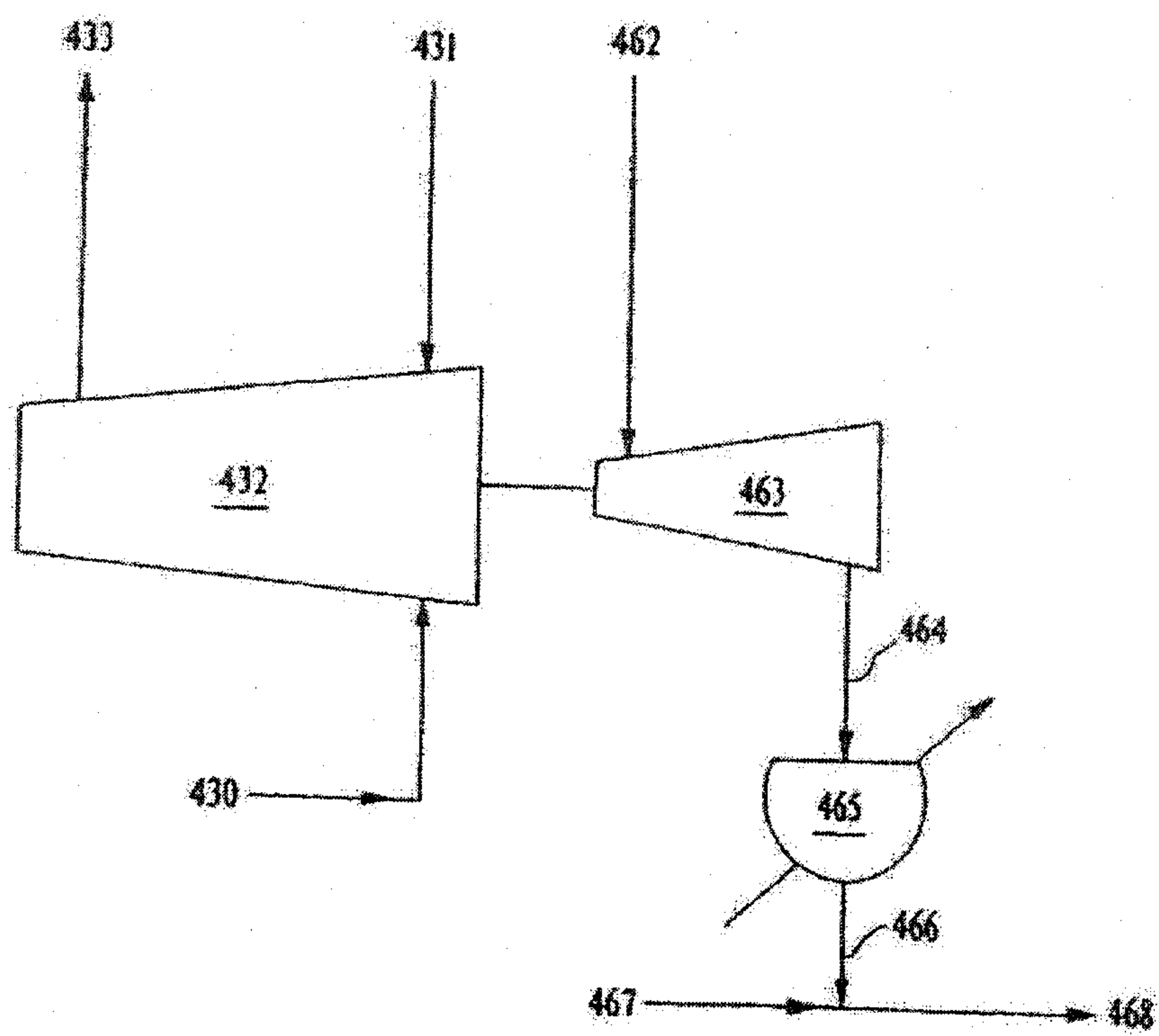


FIG. 4

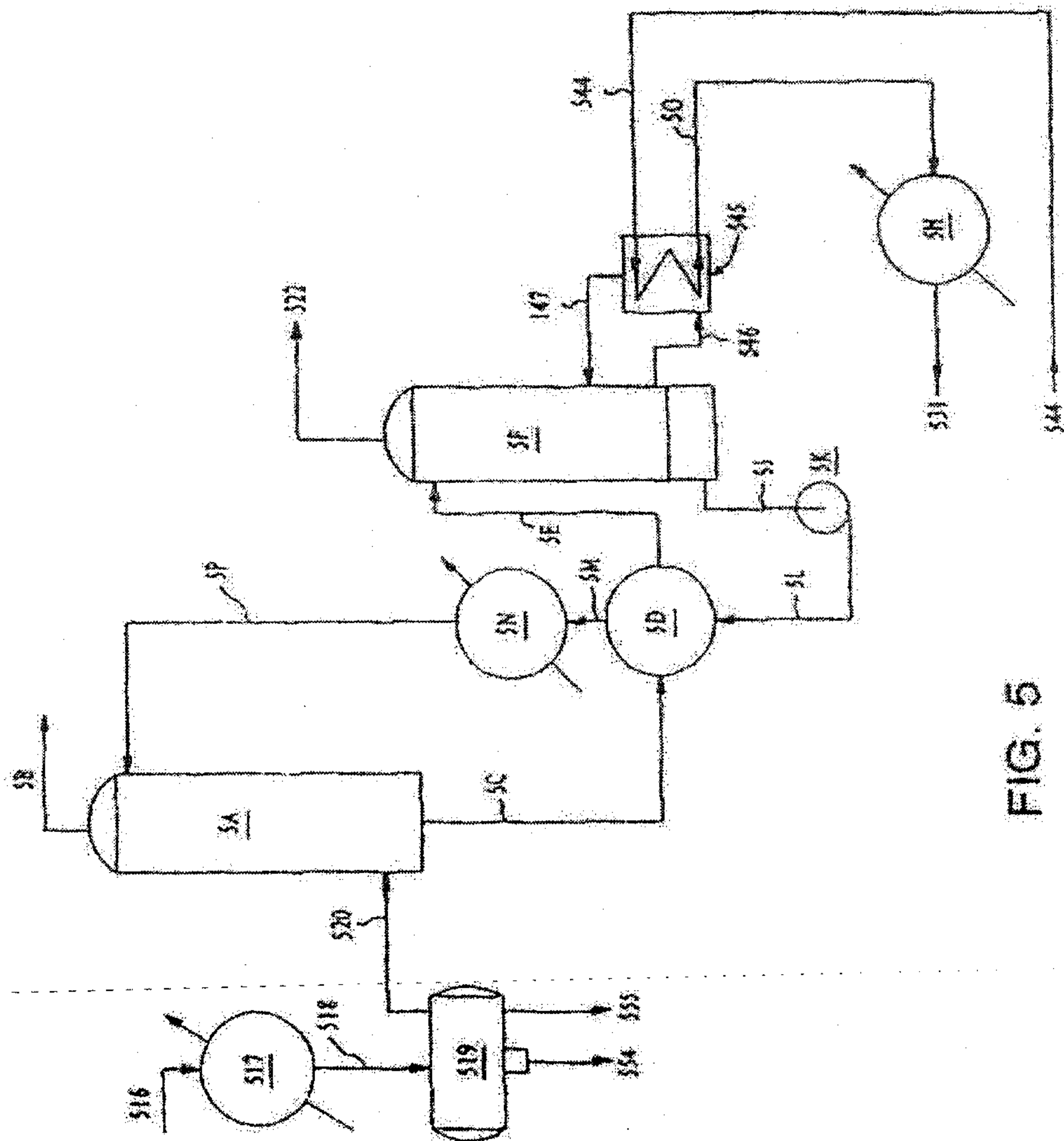


FIG. 5

PROCESS OVERVIEW

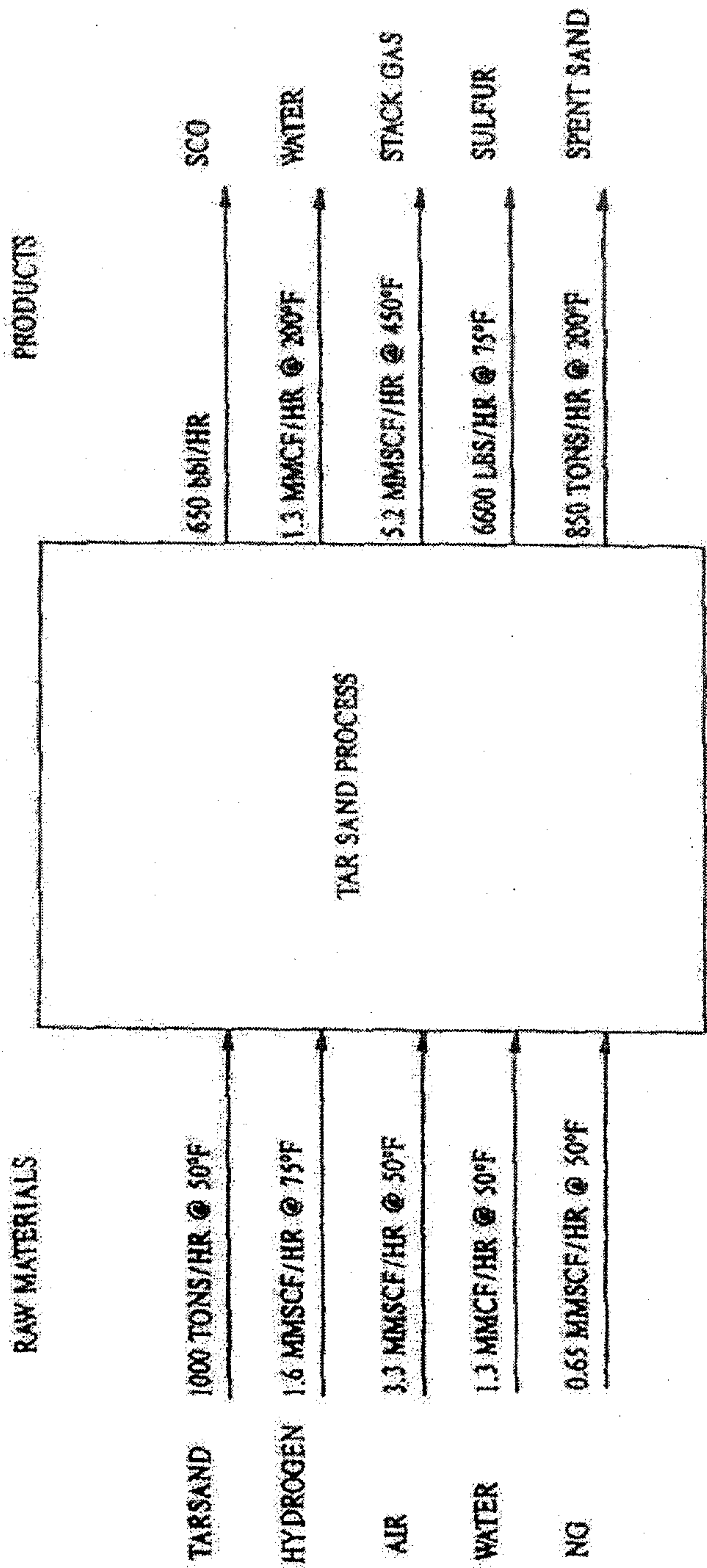


FIG. 6

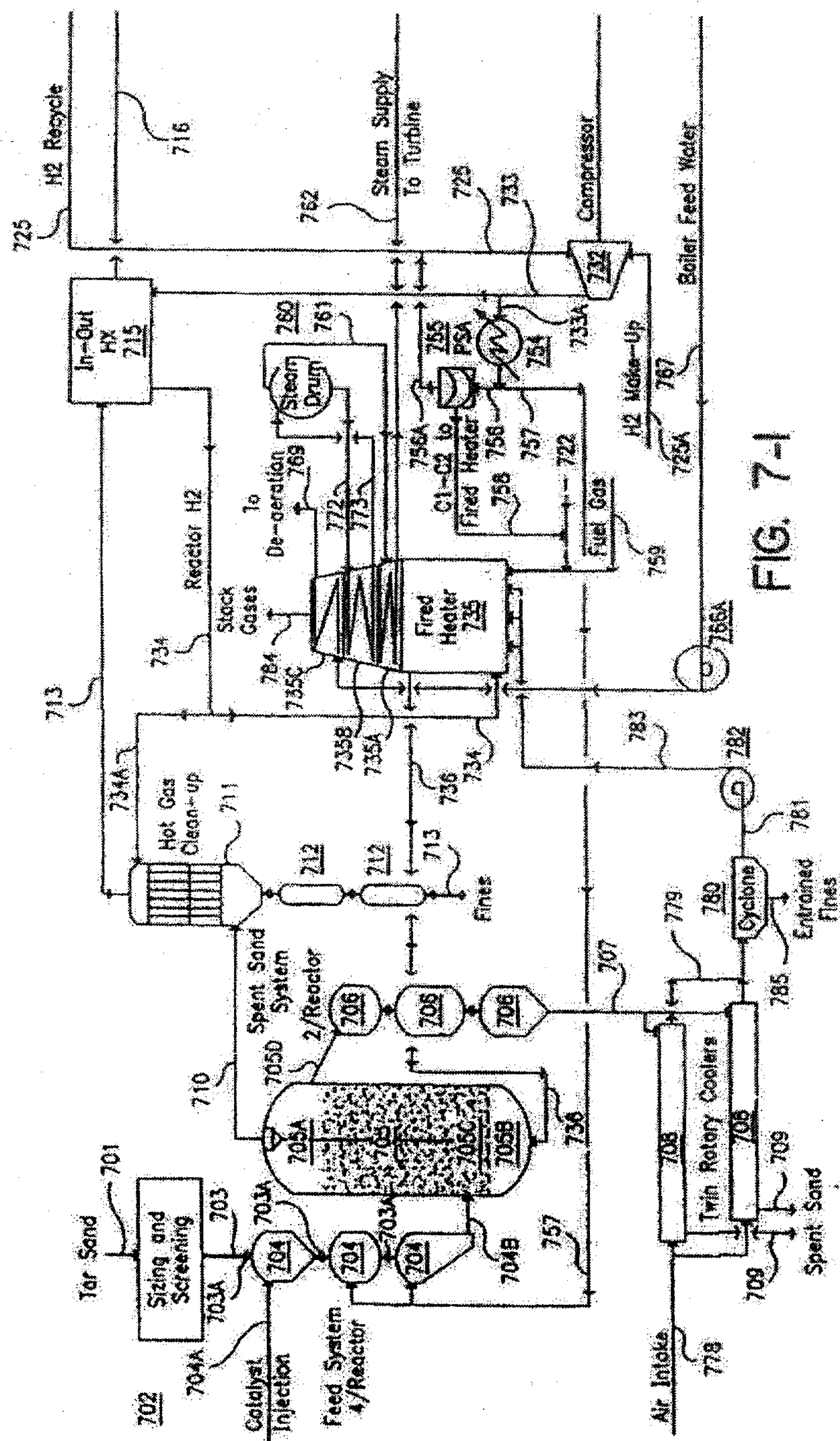


FIG. 7-1

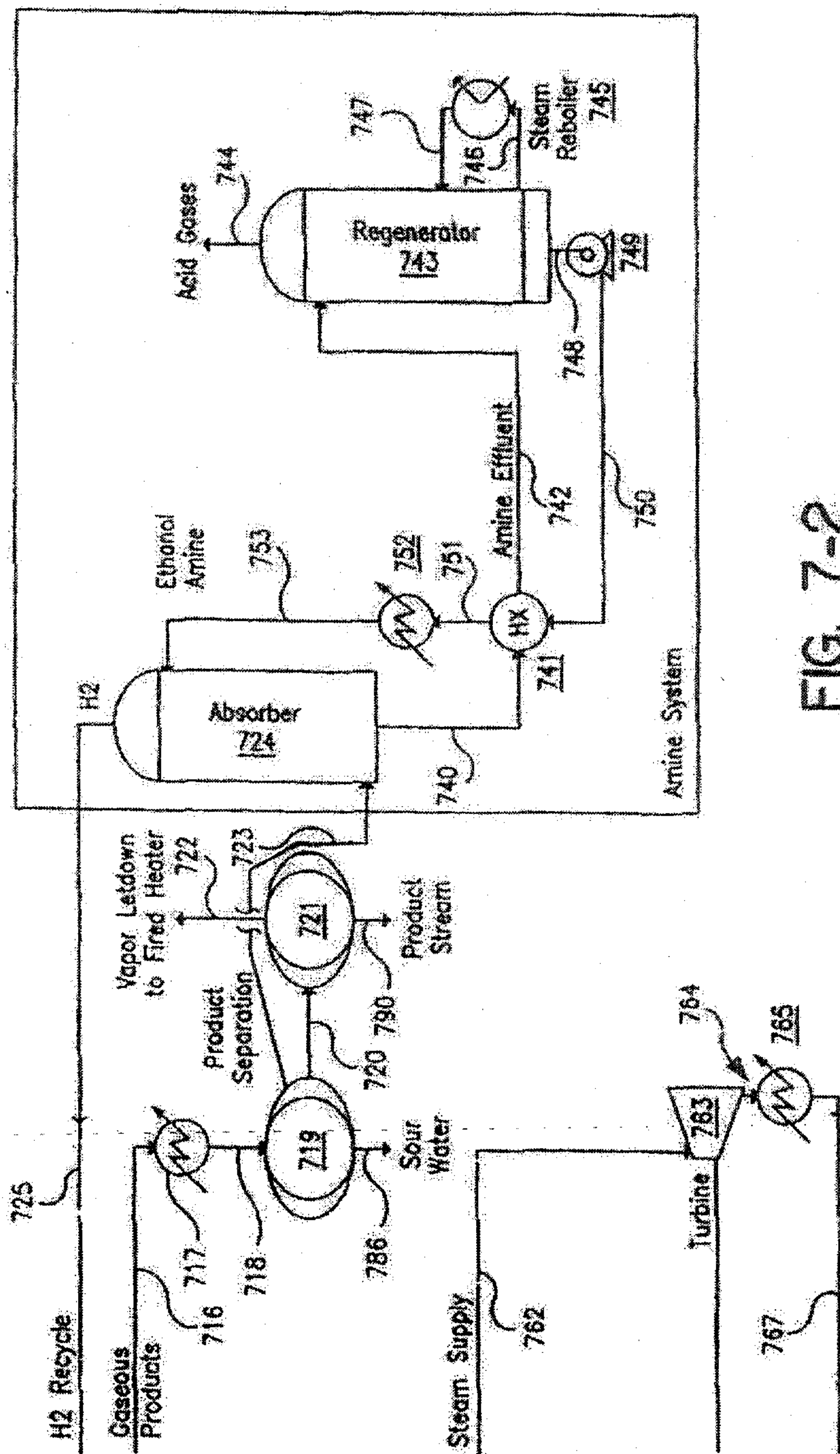


FIG. 7-2

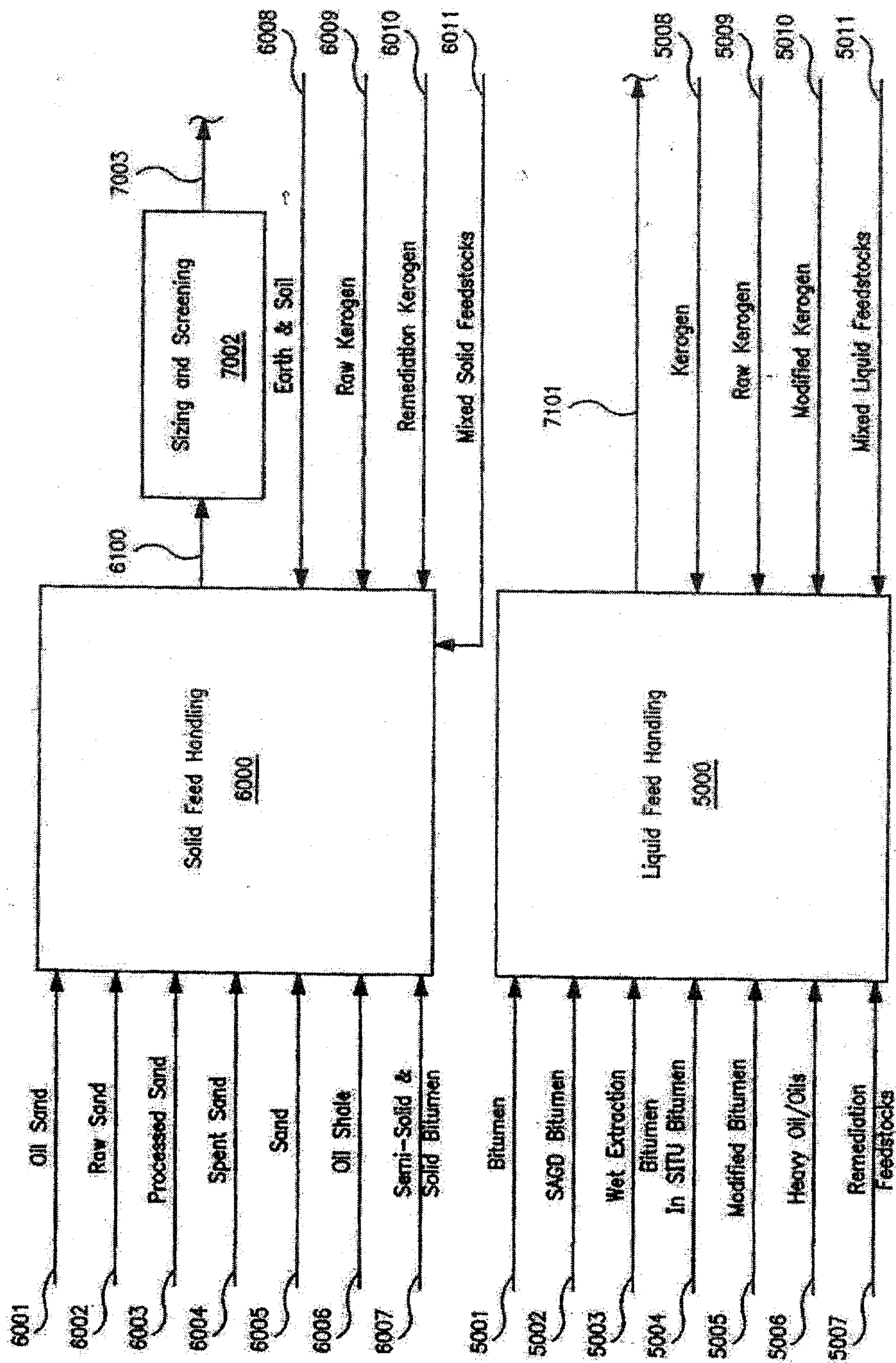
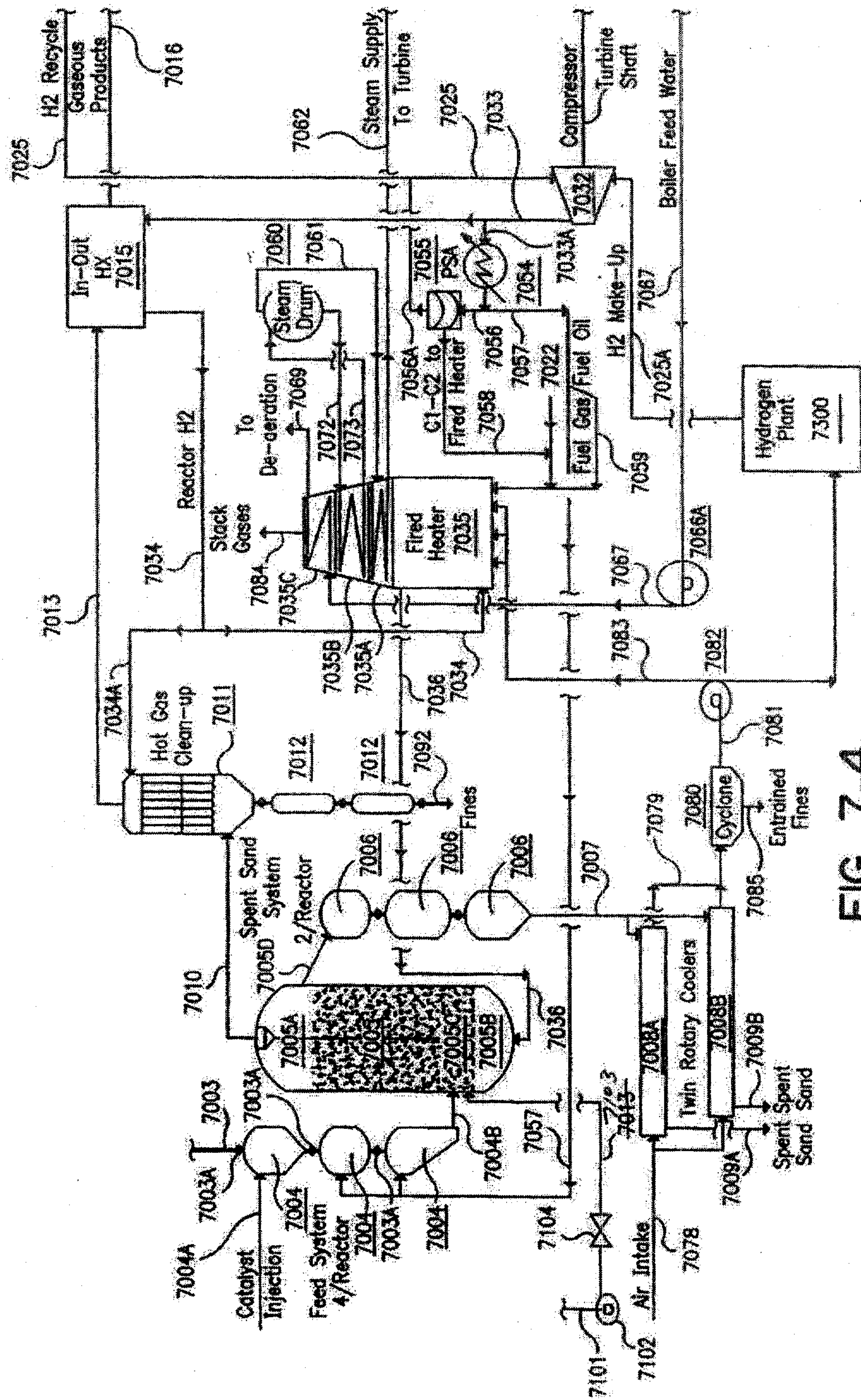


FIG. 7-3



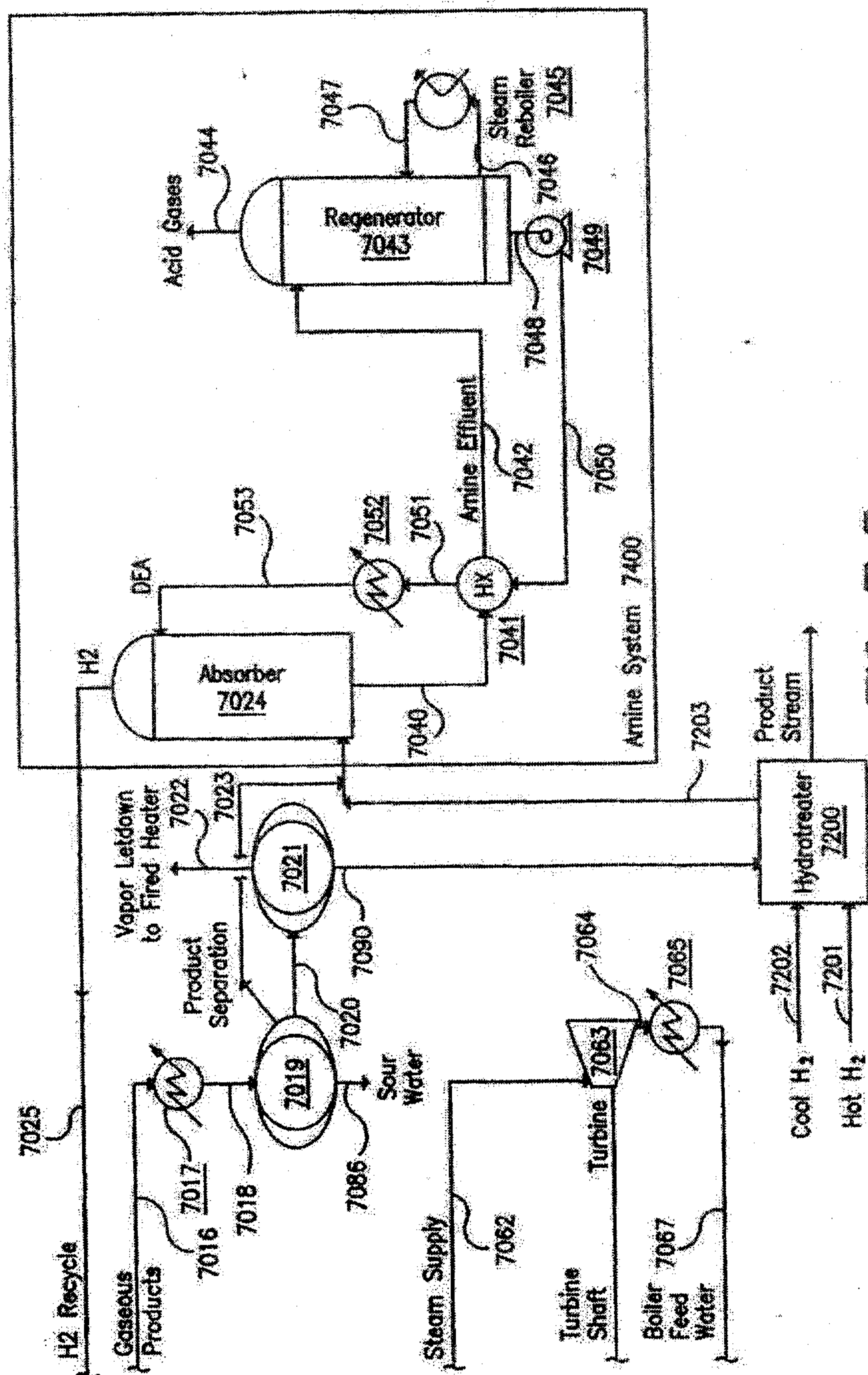


FIG. 7-5

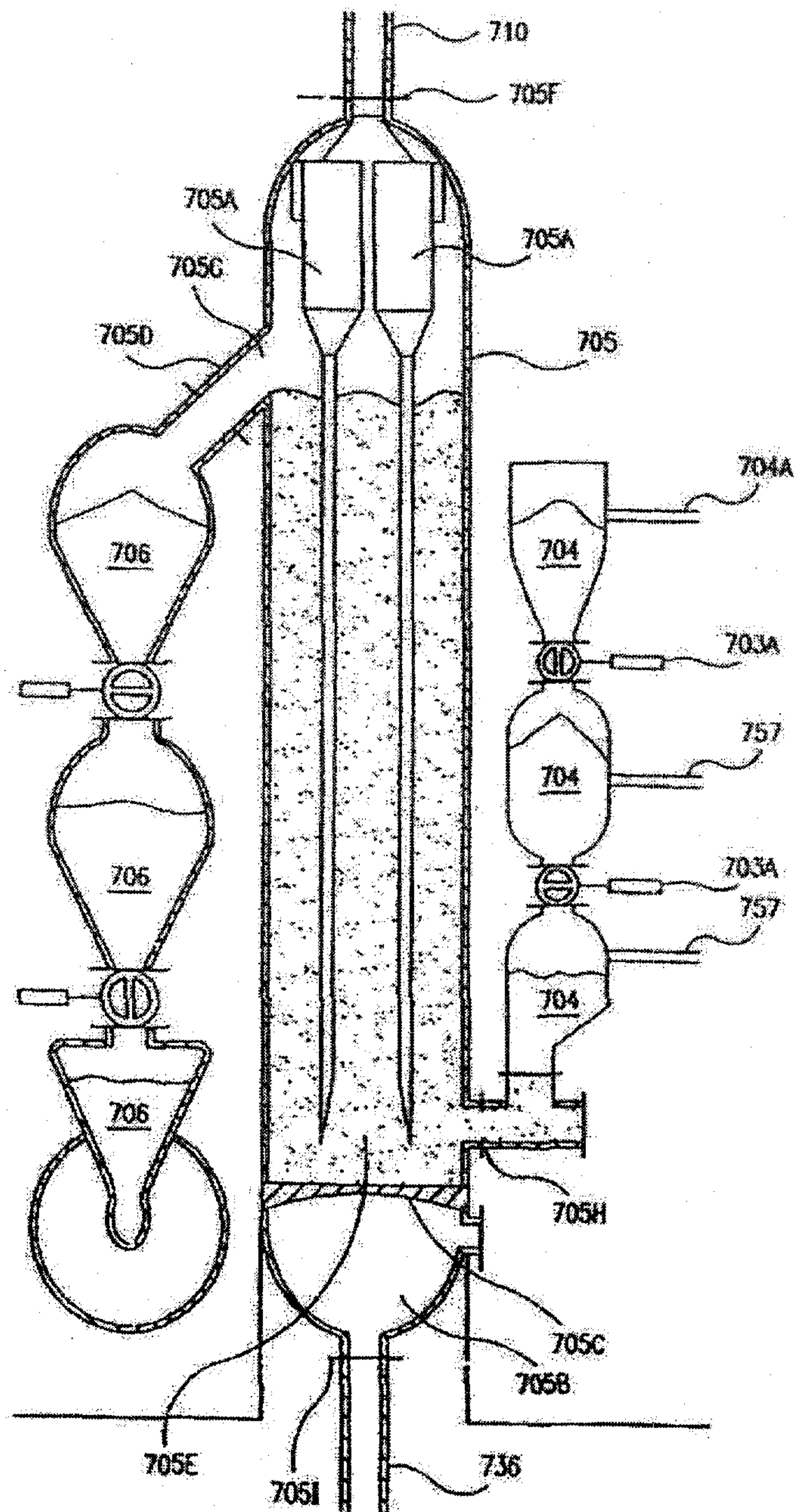


FIG. 8

