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(72) Inventeurs/Inventors:
KIRKBRIDE, CHALMER G., US;
DOYLE, JAMES A., US;
HILDEBRANDT, FRED, CA
(73) Propriétaire/Owner:
CHATTANOOGA CORPORATION, US
(74) Agent: BLAKE, CASSELS & GRAYDON LLP

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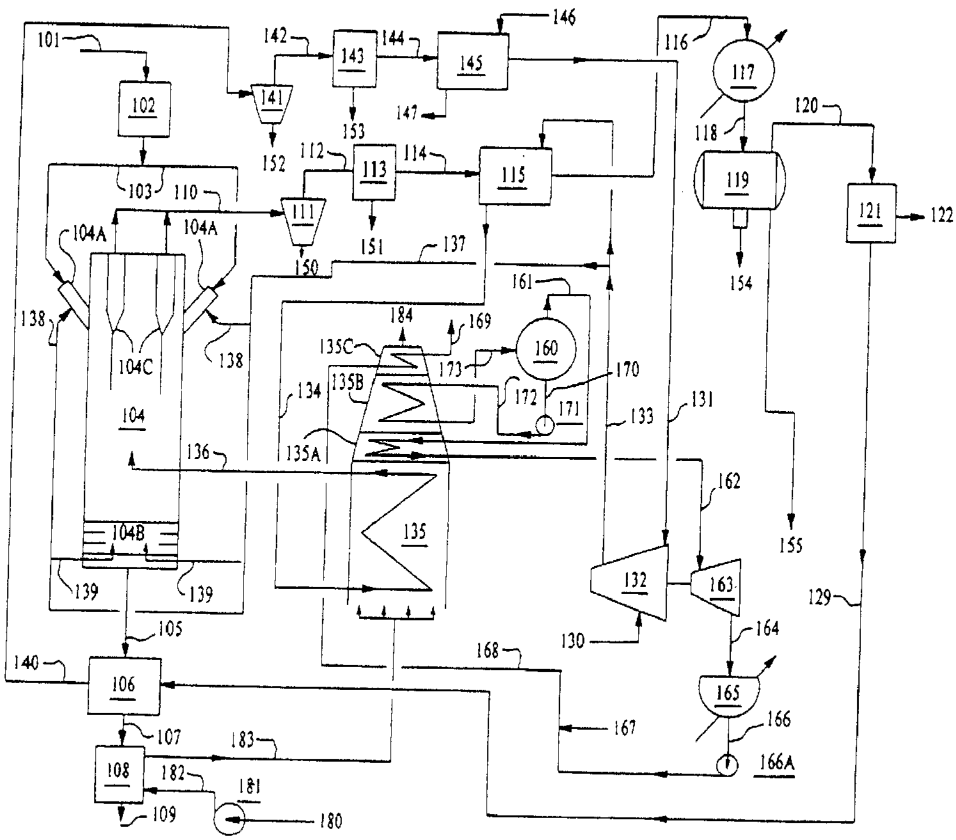
(54) Title: PROCESS AND APPARATUS FOR CONVERTING OIL SHALE OR TAR SANDS TO OIL



(57) Abrégé/Abstract:
The invention relates to a continuous process for producing synthetic crude oil from oil bearing material, e.g., oil shale or tar sand, through continuous feeding and calcining, hydrocracking and hydrogenating kerogen or bitumen.

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ABSTRACT

The invention relates to a continuous process for producing synthetic crude oil from oil bearing material, e.g., oil shale or tar sand, through continuous feeding and calcining, hydrocracking and hydrogenating kerogen or bitumen.

PROCESS AND APPARATUS FOR CONVERTING OIL SHALE OR TAR SANDS TO OIL

FIELD OF THE INVENTION

The present invention relates to a continuous process for producing synthetic crude oil (SCO) from oil shale or tar sand and an apparatus for its practice. More specifically, the present invention provides a process for treating dry tar sand or shale without prior beneficiation, in a reactor operating at elevated pressure and temperature conditions, in the presence of substantially only hydrogen gas. The spent shale or tar sand can then be used to prepare soil and construction compositions.

BACKGROUND OF THE INVENTION

There are some tar sand systems that are successful in making SCO, such as those in the Canadian Athabasca tar sand area that surface mine and process the tar sands, where they first separate sand (85 wt.%) from bitumen (15 wt.%) to avoid processing the sand in the reaction systems. The separated bitumen is converted to sweet, light crude oil by conventional refinery type 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 tar sand for flotation. 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 tar sands, e.g., fluid cokers, LC finer, tumblers (being phased out by hydro-pumping), beneficiation including: primary separation vessels with floatation cells and secondary separation systems necessary to recover the bitumen from the 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 required for floatation since the water must be reused to maximize clean-up to reduce environmental problems. These systems require large facilities along with the maintenance and reclamation required.

For example, U.S. Patent Nos. 5,340,467 and 5,316,467 to Gregoli, et al. relate to the recovery of hydrocarbons (bitumen) from tar sands. In the Gregoli, et al. patent process, tar sand is slurried with water and a chemical additive and then sent to a separation system. The bitumen recovery from tar sand processes described in U.S. Patent Nos. 5,143,598 to Graham et al. and 4,474,616 to Smith, et al. also involve the formation of aqueous slurries. Other processes involving slurries, digestion, or extraction processes are taught in U.S. Patent Nos. 4,098,674 to Rammler, et al., 4,036,732 to Irani, et al., 4,409,090 to Hanson, et al., 4,456,536 to Lorenz, et al. and Miller, et al.

In situ processing of tar sand is also known as seen from the teachings of U.S. Patent Nos. 4,140,179, 4,301,865 and 4,457,365 to Kasevich, et al. and 3,680,634 to Peacock, et al.

U.S. Patent No. 4,094,767 to Gifford relates to fluidized bed retorting of tar sands. In the process disclosed by the Gifford patent, raw tar sand is treated in a fluidized bed reactor in the presence of a reducing environment, steam, recycle gases and combustion gases. The conversion of the bitumen, according to the Gifford patent, is through vaporization and cracking, thereby leaving a coked sand product. The steam and oxygen, according to Gifford are "injected into the fluidized bed in the decoking area above the spent sand cooling zone, and below the input area in the cracking zone for fresh tar sand."

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 tar sand by using a single continuous reactor system to hydrocrack and hydrogenate the dry tar sand. Moreover, because the present invention directly hydrogenates dry tar sand, larger quantities of valuable sweet, light crude oil is obtained. Moreover, with the present invention, less gas and substantially no coke is produced.

BRIEF SUMMARY OF THE INVENTION

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

Accordingly, one aspect of the present invention is to provide a continuous process and an apparatus for its practice where oil bearing material such as the kerogen in oil shale or the bitumen in tar sand is continuously treated.

Another aspect of the present invention relates to the treatment of dry tar sand.

An object of the present invention is providing a process for converting tar sand to oil through the use of substantially only hydrogen.

Another object of the present invention is providing a heat recovery process whereby hydrogen provides the heat necessary to bring the raw tar sand up to reactor temperature.

A still further object of the present invention is providing a process where hydrogen is used for hydrocracking and hydrogenating the bitumen in the tar sand or oil shale.

A further objective of the present invention is providing a process for using recycle and make-up hydrogen as a heat transfer vehicle.

A still further object of the present invention is to produce dry, relatively clean sand as waste that will not pollute and can be used as excellent landfill for permanently improved and desirable land.

Objects and advantages of the invention are set forth in part herein and in part will be apparent herefrom, or may be learned by practice with the invention, the same being realized and attained by means of the flow charts, process steps, structures, instrumentalities and combinations pointed out in the appended claims. Accordingly, the invention resides in the novel steps, parts, structures, arrangements, combinations and improvements herein shown and described.

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 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.

DETAILED DESCRIPTION OF THE INVENTION

In the present invention the hydrocarbon content of the hydrocarbon bearing solids, e.g., dry tar sand or oil shale is reacted in a fluidized bed reactor with hydrogen and the process is operated to avoid decompression of the hydrogen. In the present invention, the hydrocarbon bearing solid does not include bituminous or anthracite coals or similar type material. A first portion of a substantially only hydrogen stream is used to feed the oil shale or 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 hydrogen stream that is used in the present invention is formed from fresh make-up hydrogen and recycle hydrogen generated during the process, or obtained from other hydrogen producing processes. A mixed fresh-make-up and recycle hydrogen stream is discharged from a compressor at a first temperature and pressure, and a portion is diverted for admixture with the fluidizable particles of 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 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 hydrogen stream at said second temperature is conveyed to a direct fired heater where the hydrogen stream is heated to a third temperature higher than said second temperature and then used as the fluidizing medium in the reactor to fluidize the 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 tar sand or the kerogen in the oil shale and hydrogen are reacted via endothermic and exothermic reactions to produce spent 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 first separated overhead product is conveyed to a series of additional separators to provide a particle free clean product stream. The cleaned product stream at a first temperature 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. 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. 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 cooled, absorbed overhead hydrogen stream is conveyed to a heat exchanger where it contacts spent tar sand or spent shale and its temperature is elevated due to heat transferred from the spent discharge. The hydrogen stream at the elevated temperature is conveyed to a cyclone separator, or other suitable separating devices to remove particles. It then flows to the amine system to regenerate the amine solution. It is eventually 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 portions of the hydrogen stream.

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

The bitumen in the 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 may be advantageous to add additional catalyst. The tar sand processed in accordance with the present invention is exemplified by the following, non-limiting example:

TAR SAND FEED (WATER-FREE BASIS)

sand		84.6 wt. %
bitumen		15.4 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 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. 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.

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 tar sand into the top of the reactor 104. More uniform dispersion of the 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 will control the number of valve feeders. Thus, there could be 4, 5, 6, 7 or more valve feeders used in the present invention.

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

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 of between 800°F and 900°F, preferably closer to 800°F to avoid cracking the large fragments of hydrogenated bitumen in the tar sand.

It is advantageous to conduct the endothermic hydrocracking and exothermic hydrogenating processing of tar sand in reactor 104 in a predominantly hydrogen gas environment. The hydrogen atmosphere in reactor 104 is maintained at about 600 psi 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. 184 is the fired heater stack gases line.

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 tar sand in reactor 104 is about 15 minutes, but could 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 the separator section located at the lower end of reactor 104. The cyclone 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 tar sand is continuously discharged at a pressure of about 635 psi and a temperature of about 800°F through lock hopper valving arrangement 104B in the lower end of reactor 104 into line 105 which conveys the discharged material to spent sand heat exchangers 106 and 108. Spent sand is discharged through line 109.

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 psi. 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 135.

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 psi, all carbon compounds C_3 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 having amine absorber 121 is described in greater detail in **FIG.** 5.

The only gases not absorbed and removed in the absorption system having amine absorber 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 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 will not be absorbed nor condensed but will be recycled with the unreacted hydrogen after processing in units 141, 143 and 145 discussed hereinafter. The C_1 and C_2 hydrocarbons will reach equilibrium within the reactor 104 at about 2 vol.% and will then add to the production of crude oil per ton of tar sand. A small offset will 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 reboiler 145.

FLUIDIZED BED REACTOR

FIG. 2 schematically shows the pressurized, continuously operating fluid bed reactor 204 in accordance with the present invention. Sized and screened 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 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.

DIRECT FIRED HEATER

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 135 that is designed to balance out the total energy required to operate the reactor system. Preheated air conveyed through feed lines 183 is combusted with fuel in the radiant section of fired heater 135 and elevates the temperature of the recycle and make-up hydrogen that is conveyed through line 134. The fuel that is combusted is obtained from the C_3 fraction, e.g. propane, or natural gas produced or purchased from the described process or other sources. The hydrogen stream in lines 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 135 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 135 is maintained by compressor 132 so that the 1200°F hydrogen stream can be introduced via line 136 into reactor 104 **(FIG.** 1) or 204 **(FIG.** 2).

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

Steam drum 160 separates steam which is conveyed to radiant section 135A through line 161 to produce superheated steam for the turbine compressor 163.

The steam circulation loop includes steam drum 160, line 170, recirculation pump 171 and lines 172-173 which conveys boiler water through radiant section 135B and back into drum

160. Water for the boiler system is provided through feed line 167 which flows into line 168 which is in communication with line 169 through convection section 135A to deaeration.

As discussed above, convection section 135A super heats steam which is conveyed through line 162 to drive compressor turbine 163, which drives compressor 132. Steam is generated in convection section 135B and make-up water and turbine condensate for boiler feed water are preheated in convection section 135C.

COMPRESSOR SYSTEM

FIG. 4, schematically shows a compressor 132 driven by a high pressure steam turbine 163 required to maintain circulation of gases to operate the reactor system 104. Make-up hydrogen 130 and recycle hydrogen 131, at approximately 450 psig and 100°F are pressurized by the compressor 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, at 1500 psig and 800°F drives the turbine 163. Exhaust steam 164 is condensed in condenser 165, and along with make-up water 167 is fed to the fired heater convection section 135C for preheating and reuse as boiler feed water make-up.

PRODUCT SEPARATION

The product separation of **FIG.** 1, components will 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 116 from the in-out exchanger 115 are further cooled in product condenser 117 and conveyed through line 118 to separator 119 where condensed liquids are removed as product raw crude 155. Overhead gases are conveyed through line 120 to an amine absorber 121 where acid gasses H_2S, CO_2 and SO_2 are absorbed by a counter current circulating amine solution. The recycle gases 121B flow from the top of the absorber 121 to recycle hydrogen stream 129.

The rich amine solution 121C exits the bottom of the absorber, flows through an amine exchanger 121D where it is heated by exchange with hot lean amine solution 121L and enters the top of an amine stripper 121F through line 121E. The lean amine solution passes through exchanger 121D to exchanger 121N via line 121M and enters 121 via lean amine feed line 121P. Absorbed acid gases are stripped from the amine solution by the application of heat to the solution in reboiler 145 (having reboiler feed line 146 and stripper return line 147) and are conveyed through flow line 122 from the stripper to sulfur recovery off-site. Hot recycle gases are conveyed through line 144 from the spent sand cooler 106 via cyclone 141 and precipitator 143 to provide heat for reboiler 145 and the partially cooled recycled gases 121G are further cooled by cooler 121H and then flow through line 131 to the suction side of compressor 132. Solids removed by cyclone 141 and precipitator 143 are discharged via lines 152 and 153 respectively.

Lean amine solution 121J is circulated by amine circulation pumps 121K through the amine exchanger 121D and amine cooler 121N to the top of the amine absorber 121 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 tar sand at 50°F are reacted with hydrogen to produce 665 bbl/hr of synthetic crude oil. The following Table provides the feed and product values for processing 1000 tons/hr. of tar sand.

RAW MATERIALS		PRODUCTS
1000 TONS/HR. 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 Sectional Area	314.16 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

TAR SAND FEED

Sand Flow Rate	1000.00 tons/hr
Density of sand	121.68 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 H_2	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	wt. %
CO	0.30%
CO_2	0.20%
H_2S	31.00%
NH_3	2.50%
C_3	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 at reactor inlet	50.00	°F
Reaction temperature	800.00	°F
Sand Feed	1,000.00	tons/hr

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 H_2 for reaction	9000.00	lbs./hr
	(1.71	MMSCF/hr)
Additional H_2 Supplied	229736.15	lbs./hr
	43.53	MMSCF/hr)
Total H_2 Supplied	238736.15	lbs./hr
	(45.24	MMSCF/hr)
C_1-C_2 Flow within H_2 Stream (at equilibrium –2 vol.%)	4594.72	lbs/hr
	(0.08	MMSCF/hr)
Entering H_2 Temperature	1200.00	°F
Cp H_2	3.50	btu/lb-°F
Heat Supplied by C_1-C_2	1.01	MMbtu/hr
Heat Supplied by H_2	334.23	MMbtu/hr
H_2 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)

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Heat to Be Removed	182.96	MMbtu/hr
Entering Hydrogen Temperature	100.00	°F
Exiting Hydrogen Temperature	318.96	°F

<u>AIR COOLANT</u>

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

<u>HYDROGEN SUPPLY</u>

Entering Hydrogen Temperature	318.96	°F
Exiting Hydrogen Temperature	100.00	°F

<u>AMINE BOIL-OFF</u>

Heat Available to the system	182.96	MMbtu/hr

IN-OUT HEAT EXCHANGER

<u>HYDROGEN TO BE HEATED</u>

Hydrogen Flow	238736.15	lbs/hr
	(45.24	MMSCF/hr)
Inlet H_2 Temperature	121.64	°F
Exiting H_2 Temperature	750.00	°F
Total Heat Required	525.05	MMbtu/hr

<u>OFF GAS HEAT SUPPLY</u>

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 @70°F
Cp Non-Condensables	3.00	btu/lb-°F
Heat of Vaporization	65.00	btu/lb
Hydrogen Recycle Flow in Stream	229736.15	lbs/hr
	(*43.53	MMSCF/hr)
Inlet Temperature	800.00	°F
Exit Temperature	350.00	°F

PRODUCT CONDENSER/COOLER

<u>PRODUCT SIDE</u>

Entering Temperature		350.00	°F
Exiting Temperature		100.00	°F
Condensate		214941.75	lbs/hr
		665.29	bbl/hr
Heat Removal	H_2	201.02	MMbtu/hr
	Off Gas	4.40	MMbtu/hr
	Condensate	38.15	MMbtu/hr
	Total	243.57	MMbtu/hr

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COOLER REQUIREMENT	243.57	MMbtu/hr

COMPRESSOR

HYDROGEN SIDE

Flow Rate	755412.69	SCF/min
	45.32	MMSCF/hr
Pressure Out	670.00	psi
Pressure In	450.00	psi
DP	220.00	psi
gamma (Cp/Cv)	1.40	
# Stages	3	
Temperature Inlet	100.00	°F
Mechanical 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	psi
Temperature	800.00	°F
Degree Superheat	200.00	°F
Saturation Temperature	596.20	°F
Steam Heat Value	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

C_3'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 be supplied by Methane	437.99	MMbtu/hr
Flow Rate	21653.89	lbs/hr
	(0.51	MMSCF/hr)
Heat of Combustion (After temperature correction)	20227.00	btu/lb
Temperature in	75.00	°F

COMBUSTION AIR		
Air Required for Combustion		200000.00 lbs/hr (2.61 MMSCF/hr)
Air Supplied	25 wt.% 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

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 psi
C_1-C_2's		
	Flowrate	0.08 MMSCF/hr
	Temperature	1200.00 °F
	Pressure	635.00 psi
OUT FLOWS		
Sand		
	Flowrate	850.00 tons/hr
	Temperature	190.00 °F
	Pressure	600.00 psi
Off Gas		
	Flowrate	43.92 MMSCF/hr
	Temperature	800.00 °F
	Pressure	600.00 psi
	Composition	wt%
	H_2	81.98
	CO	0.05
	CO_2	0.04
	H_2S	5.60
	NH_3	0.45
	C_3	11.92

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Product			
	Flowrate (Vapor Phase)	214937.52	lbs./hr
	Temperature	800.00	°F
	Pressure	600.00	psi

SAND COOLER (106, 108)

IN FLOWS

Sand			
	Flowrate	850.00	tons/hr
	Temperature	791.92	°F
	Pressure	600.00	psi
Hydrogen			
	Flowrate	45.23	MMSCF/hr
	Temperature	100.00	°F
	Pressure	500.00	psi
Air			
	Flowrate	3.27	MMSCF/hr
	Temperature	50.00	°F
	Pressure	30.00	psi

OUT FLOWS

Sand			
	Flowrate	850.00	tons/hr
	Temperature	200.00	°F
	Pressure	480.00	psi
Hydrogen			
	Flowrate	45.23	MMSCF/hr
	Temperature	313.94	°F
	Pressure	480.00	psi
Air			
	Flowrate	3.27	MMSCF/hr
	Temperature	300.00	°F
	Pressure	20.00	psi

IN-OUT HEAT EXCHANGER (115)

IN FLOWS

Hydrogen			
	Flowrate	45.23	MMSCF/hr
	Temperature	147.60	°F
	Pressure	670.00	psi
Off Gas			
	Flowrate	43.92	MMSCF/hr
	Temperature	800.00	°F
	Pressure	600.00	psi
	Composition	wt%	
	H_2	81.94	
	CO	0.05	
	CO_2	0.04	

H_2S	5.60	
NH_3	0.45	
C_3	11.92	

Product	Flowrate (Vapor Phase)	214937.52	lbs./hr
	Temperature	800.00	°F
	Pressure	600.00	psi

<u>OUT FLOWS</u>

Hydrogen	Flowrate	45.23	MMSCF/hr
	Temperature	750.00	°F
	Pressure	650.00	psi
Off Gas	Flowrate	43.92	MMSCF/hr
	Temperature	368.63	°F
	Pressure	580.00	psi

Off Gas Composition as Above

Product	Flowrate (Vapor Phase)	214937.52	lbs./hr
	Temperature	368.63	°F
	Pressure	580.00	psi

<u>PRODUCT CONDENSER/COOLER (117)</u>

<u>IN FLOWS</u>

Off Gas	Flowrate	43.92	MMSCF/hr
	Temperature	368.63	°F
	Pressure	580.00	psi

Off Gas Composition as Above

Product	Flowrate (Vapor Phase)	214937.52	lbs./hr
	Temperature	368.63	°F
	Pressure	550.00	psi

<u>OUT FLOWS</u>

Off Gas	Flowrate	43.92	MMSCF/hr
	Temperature	100.00	°F
	Pressure	540.00	psi

Off Gas Composition as Above

Product			
	Flowrate (as condensate)	214937.52	lbs./hr
	Temperature	100.00	°F
	Pressure	540.00	psi

AMINE SYSTEM (121, FIG. 5)

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

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 may be incorporated and embodied as part of the present invention within the scope of the following claims.

CLAIMS:

1. A process for producing an oil from a hydrocarbon bearing feed, comprising the steps of:

providing a feed comprising tar sand;

introducing said feed in a fluidizable form into a fluidized bed reactor;

introducing a fluidizing medium having a gas comprising hydrogen into the fluidized bed reactor;

fluidizing said feed with said fluidizing medium in the reactor to comprise a fluidized bed;

continuously reacting said feed with said hydrogen;

continuously producing a reactor product stream comprising a hydrocarbon; and producing an oil.

2. The process of Claim 1, further comprising the step of continuously discharging a spent solid from said fluidized bed reactor.

3. The process of any one of Claim 1 or 2, in which said fluidizing medium is 95 vol. % hydrogen or greater

4. The process of any one of Claim 1 or 2, in which said fluidizing medium is 98 wt. % hydrogen or greater.

5. The process of any one of Claim 1 or 2, in which said fluidizing medium is 99.8 vol. % hydrogen or greater.

6. The process of any one of Claims 1 to 5, in which said reactor product stream comprises an off gas having a hydrogen composition of 81 wt. % or greater.

7. The process of Claim 1 or 2, wherein said tar sand comprises bitumen.

8. The process of Claim 7, further comprising the step of continuously reacting said bitumen with said hydrogen to continuously produce said reactor product stream.

9. The process of Claim 2 in which said spent solid is spent tar sand.

10. A process for producing oil from a hydrocarbon bearing feed, comprising the steps of:

providing a feed comprising oil shale;

introducing said feed in a fluidizable form into a fluidized bed reactor;

introducing a fluidizing medium having a gas comprising hydrogen into the fluidized bed reactor;

fluidizing said feed with said fluidizing medium in the reactor to comprise a fluidized bed;

continuously reacting said feed with said hydrogen;

continuously producing a reactor product stream comprising a hydrocarbon; and producing an oil.

11. The process of Claim 10, further comprising the step of continuously discharging a spent solid from said fluidized bed reactor.

12. The process of Claim 10 or 11, wherein said oil shale comprises kerogen.

13. The process of Claim 12, further comprising the step of continuously reacting said kerogen with said hydrogen to continuously produce said reactor product stream.

14. The process of Claim 13, in which said spent solid is spent oil shale.

15. The process of any one of Claims 10 to 14, in which said fluidizing medium is 95 vol. % hydrogen or greater.

16. The process of any one of Claims 10 to 14, in which said fluidizing medium is 98 wt. % hydrogen or greater.

17. The process of any one of Claims 10 to 14, in which said fluidizing medium is 99.8 vol. % hydrogen or greater.

18. The process of any one of Claims 10 to 17, in which said reactor product stream comprises an off gas having a hydrogen composition of 81 wt. % or greater.

19. The process of any one of Claims 10 to 18, further comprising the step of supplying a hydrogen stream comprising a recycle gas stream comprising hydrogen and fresh make-up hydrogen to said fluidized bed reactor.

20. The process of any one of Claims 1 to 19, wherein a gas stream comprising hydrogen entrains the feed.

21. The process of any one of Claims 1 to 20, wherein said fluidizing medium is at a first temperature and is conveyed through a first heat exchanger to raise the temperature of said fluidizing medium to a second temperature prior to entering a fired heater.

22. The process of any one of Claims 1 to 21, wherein said fluidizing medium is conveyed to a fired heater which heats said fluidizing medium to a higher temperature than that of the reactor.

23. The process of Claim 22, wherein said heater provides superheated steam for gas compression.

24. The process of any one of Claims 1 to 23, wherein the fluidizing medium is introduced into the reactor at a flow rate which exceeds the minimum required for complete reaction of said feed with hydrogen by a factor of between 15 and about 26.

25. The process of any one of Claims 1 to 24, wherein the flow rate of hydrogen exceeds the minimum required for complete reaction of the feed with hydrogen by a factor of about 21.

26. The process of any one of Claims 1 to 24, wherein the fluidized bed has a fluidized bed temperature and the fluidizing medium entering the reactor has a fluidizing medium temperature, wherein the fluidizing medium temperature is greater than said fluidized bed temperature.

27. The process of any one of Claims 1 to 26, wherein the fluidizing medium comprises make-up hydrogen and a recycled hydrogen, and wherein the reactor product stream includes a recyclable unreacted hydrogen.

28. The process of any one of Claims 1 to 27, further comprising:

separating a gas mixture from the reactor product stream, the gas mixture containing unreacted hydrogen;

purifying the gas mixture to form a recycle gas stream comprising hydrogen in a concentration of 98 wt. % hydrogen or greater; and

returning at least a portion of the recycle gas stream to the reactor.

29. The process of Claim 28, further comprising the step of maintaining a combined level of methane and ethane in the recycle gas stream by control of the equilibrium point of the reaction for formation of said methane and ethane within the reactor.

30. The process of Claim 29, in which said combined level of methane and ethane in the recycle gas stream is 2 vol. % or less of the total volume of the recycle stream.

31. The process of any one of Claims 28 to 30, wherein the recycle gas stream pressure is not lower than about 450 psig.

32. The process of any one of Claims 28 to 31, further comprising the step of admixing make-up hydrogen with a the recycle gas stream prior to returning the recycle gas stream to the reactor.

33. The process of any one of Claims 1 to 3 and 10 to 15, wherein the fluidizing medium comprises at least about 95 vol. % hydrogen and wherein said hydrogen has a flow rate into the reactor of at least 26 times the flow rate required for complete reaction of said feed with hydrogen.

34. The process of any one of Claims 1 to 33, wherein the hydrogen flow rate into the reactor is 26 times the flow rate required for complete reaction of said feed with hydrogen, and wherein the reaction temperature is in a range of about 800°F to about 1200°F.

35. The process of any one of Claims 1 to 33, in which the feed is fed to the reactor at a plurality of locations.

36. The process of any one of Claims 1 to 35, wherein said step of introducing said feed includes admixing said feed with a portion of said fluidizing medium forming a dispersible feed.

37. The process of Claim 36, wherein said step of introducing said feed comprises injecting said dispersible feed at a plurality of locations at an upper portion of said reactor.

38. The process of any one of Claims 1 to 36, wherein the feed is introduced into the fluidized bed reactor near the bottom portion thereof.

39. The process of Claim 35 or 37, wherein said plurality of locations is two locations.

40. The process of Claim 35 or 37, wherein said plurality of locations comprises at least three locations.

41. The process of Claim 35 or 37, wherein said plurality of locations is three locations.

42. The process of Claim 35 or 37, wherein said locations are circumferentially positioned about the reactor.

43. The process of any one of Claims 1 to 42, wherein said step of introducing said feed utilizes fan feeding.

44. The process of any one of Claims 1 to 43, further comprising a step of reducing the size of said feed to produce a fluidizable feed prior to the feeding step.

45. The process of any one of Claims 1 to 44, further comprising a step of feeding said feed in admixture with a first gas stream comprising hydrogen to said reactor.

46. The process of any one of Claims 1 to 45, in which said step of reacting is conducted at a temperature between 800°F and 1200°F.

47. The process of any one of Claims 1 to 45, in which and said step of reacting is conducted at a temperature between about between 800°F and 900°F.

48. The process of any one of Claims 1 to 45, wherein said step of reacting is conducted at a temperature of about 800°F.

49. The process of any one of Claims 19 to 48, wherein the volume ratio of the recycle gas stream to fresh make-up hydrogen is at least 26:1.

50. The process of any one of Claims 1 to 49, wherein the fluidizing medium temperature on entering the reactor is about 1200°F.

51. The process of any one of Claims 1 to 49, wherein the fluidizing medium feed temperature on entering the reactor does not exceed 1200°F.

52. The process of any one of Claims 1 to 51, wherein said step of reacting is at 600 psig.

53. The process of any one of Claims 1 to 52, wherein said step of reacting comprises hydrocracking.

54. The process of any one of Claims 1 to 52, wherein said step of reacting comprises hydrogenation.

55. The process of any one of Claims 1 to 52, wherein said step of reacting comprises a hydrocracking reaction and a hydrogenation reaction.

56. The process of any one of Claims 1 to 55, wherein said reacting occurs in the presence of a catalyst.

57. The process of Claim 56, wherein said catalyst catalyzes hydrocracking.

58. The process of Claim 56, wherein said catalyst catalyzes hydrogenation.

59. The process of any one of Claims 56 to 58, wherein the catalyst is a heavy metal.

60. The process of any one of Claims 1 to 59, wherein said feed comprises a catalyst.

61. The process of any one of Claims 1 to 60, in which the feed has a residence time in the reactor between about 10 to about 20 minutes.

62. The process of any one of Claims 2, 9, 11, and 14, in which said solid is conveyed to at least one separation zone and a cleaned product stream is produced in said separation zone.

63. The process of Claim 62, in which said separation location includes first and second separation zones, and said first separation zone is located in the upper portion of said reactor and produces a first cleaned product stream which is conveyed to a second separation zone to produce a second cleaned product stream.

64. The process of Claim 63, wherein said second cleaned product stream is conveyed to a condenser to form a gas-liquid product stream.

65. The process of Claim 63, wherein said second cleaned product stream at a first temperature is conveyed to a first heat exchanger for cooling said second cleaned product stream to a second lower temperature.

66. The process of Claim 63, wherein said second cleaned product stream at said second lower temperature is conveyed to a condenser to form a gas-liquid product stream.

67. The process of Claim 66, wherein said gas-liquid product stream is conveyed to a separator wherein gas is separated therefrom.

68. The process of Claim 67, in which the gas separated from said reactor product stream is conveyed to an amine scrubber and is scrubbed to produce a first recycle gas stream containing hydrogen.

69. The process according to any one of Claims 2, 9, 11, 14, and 62, in which said spent solid is discharged adjacent to the top portion of said fluidized bed.

70. The process of any one of Claims 2, 9, 11, 14, and 62, wherein air for a fired heater is pre-heated with said spent solid.

71. The process of Claim 70, wherein said heater provides superheated steam for gas compression.

72. The process of any one of Claims 2, 9, 11, 14, and 62, further comprising the step of using said spent solid as land fill.

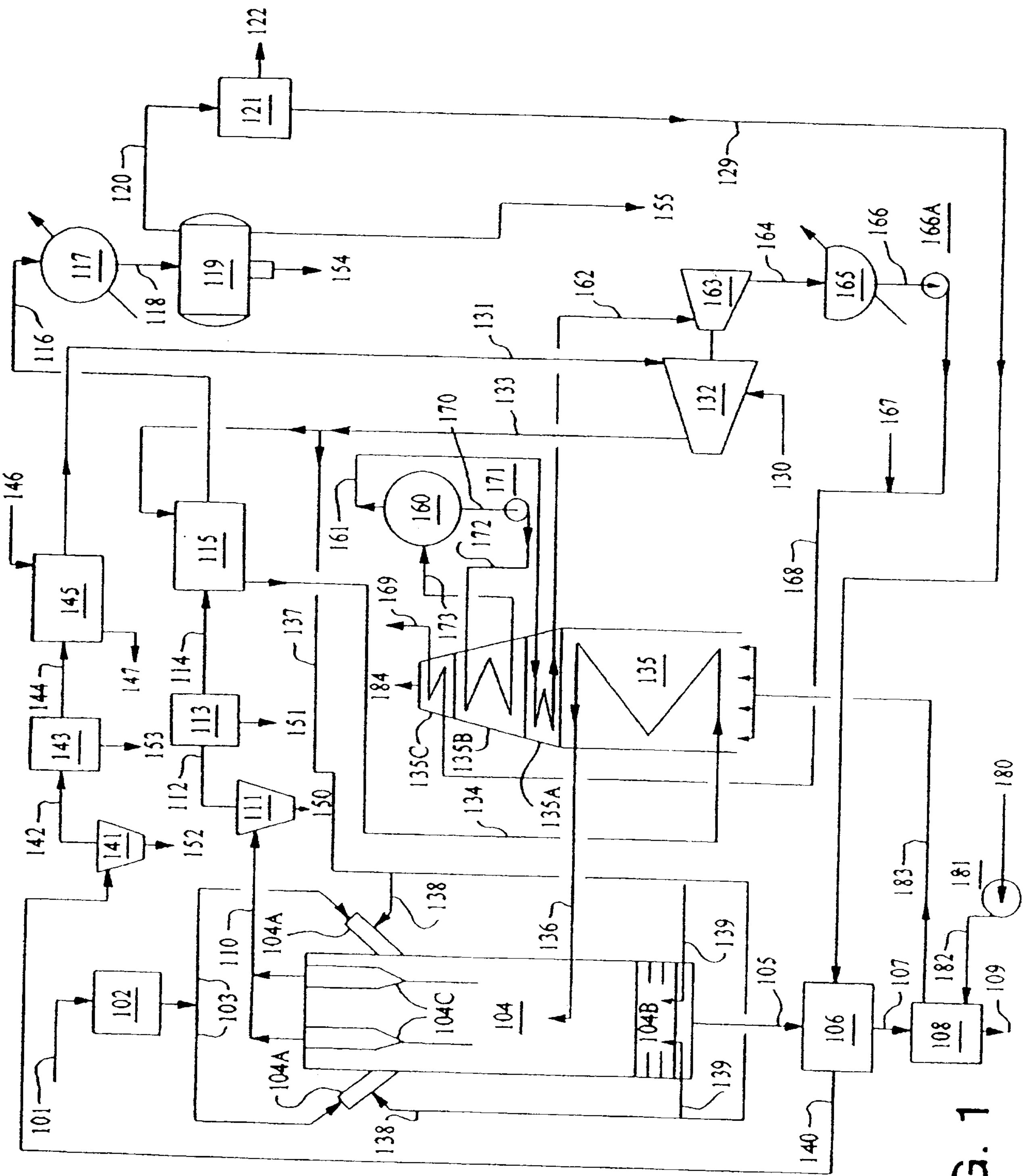

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FIG. 1

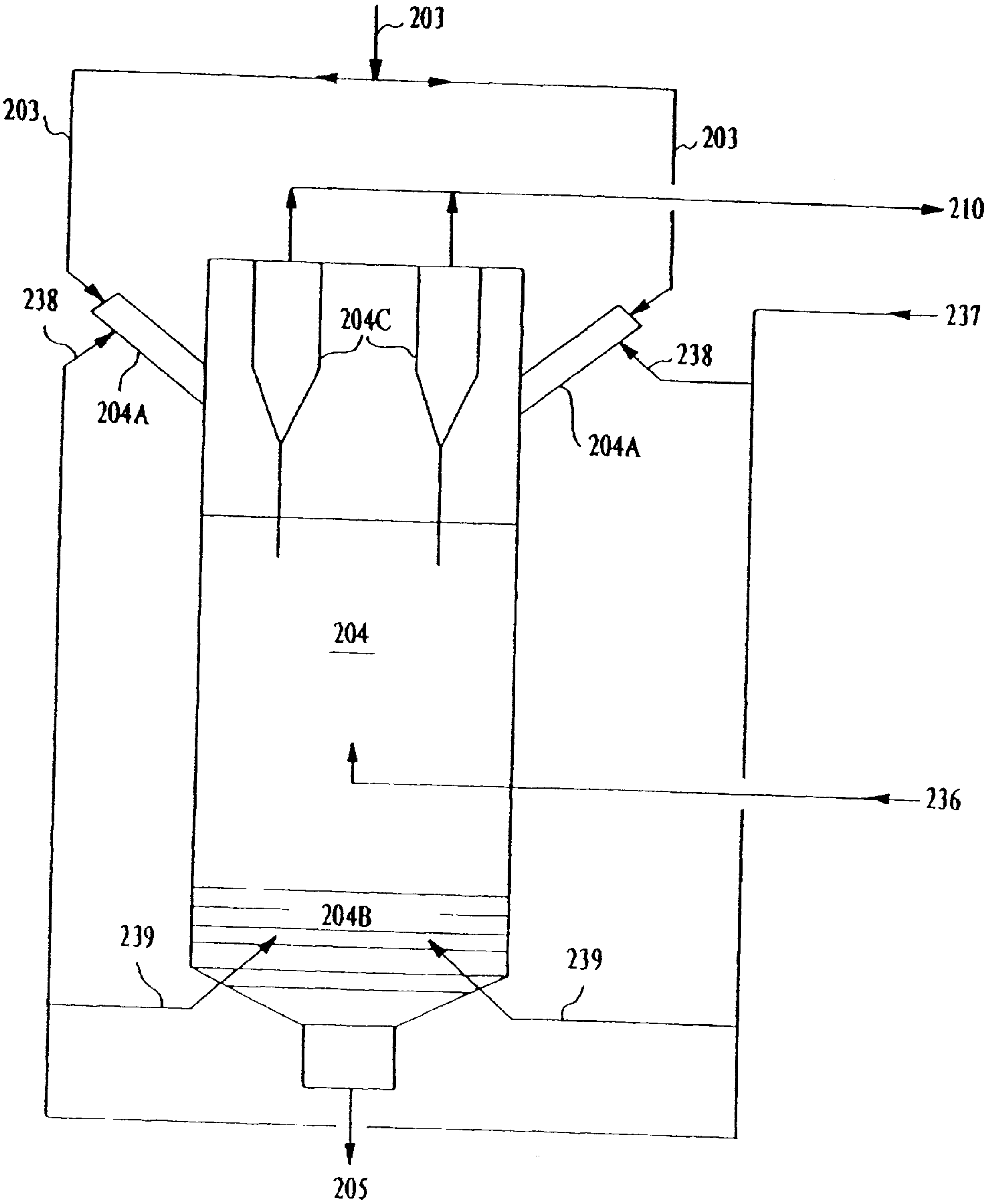
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FIG. 2

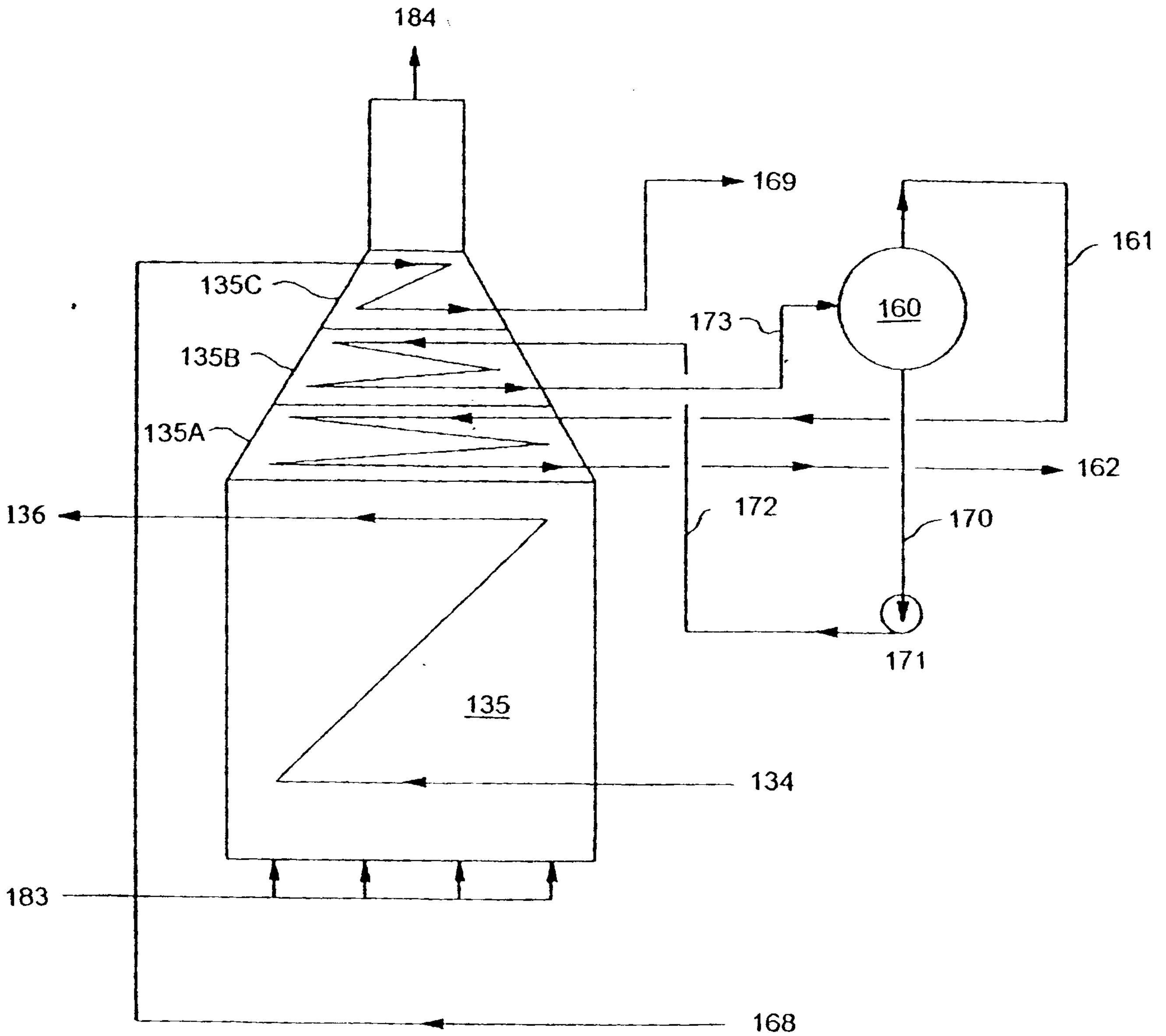
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FIG. 3

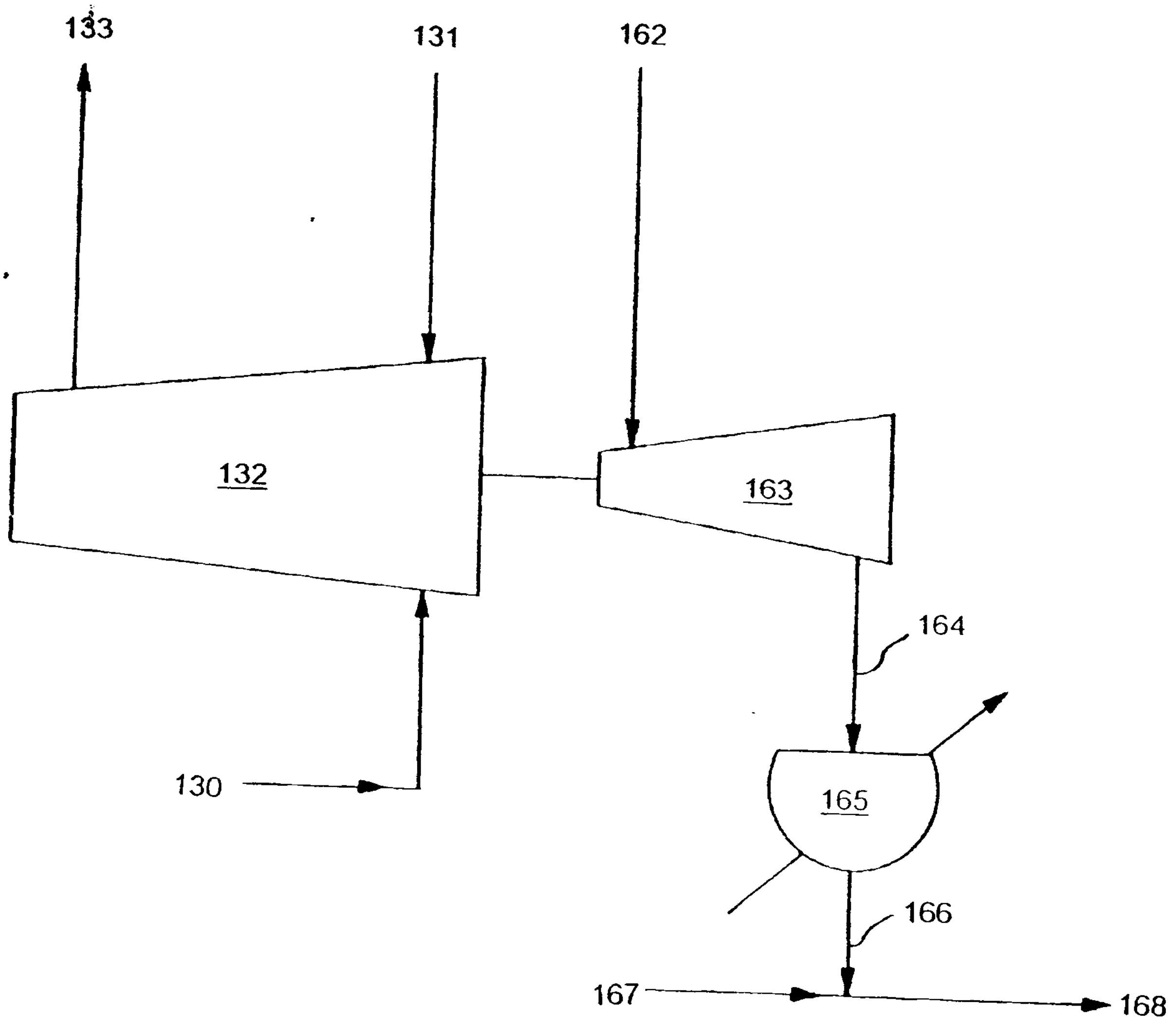
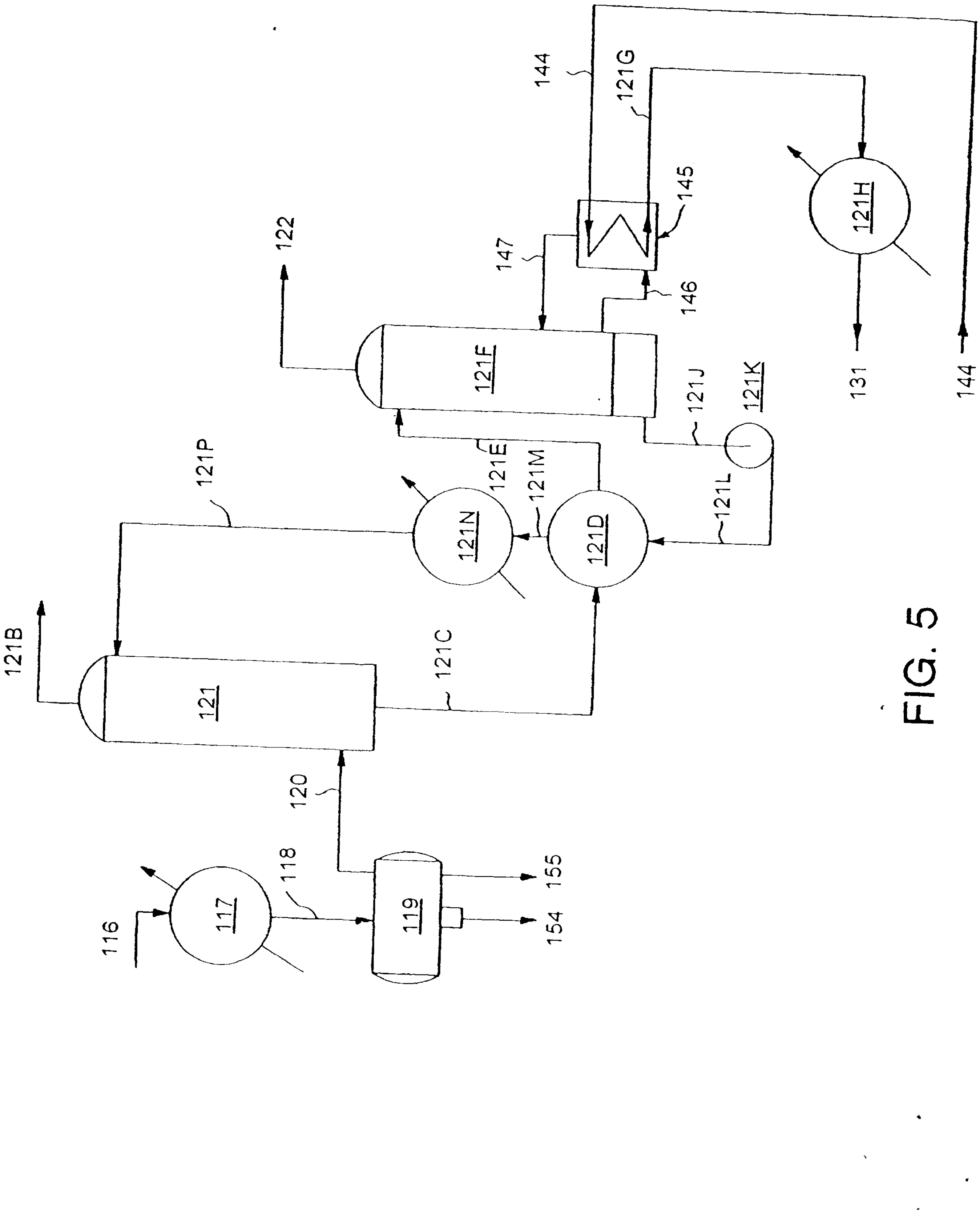


FIG. 4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121G
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FIG. 5

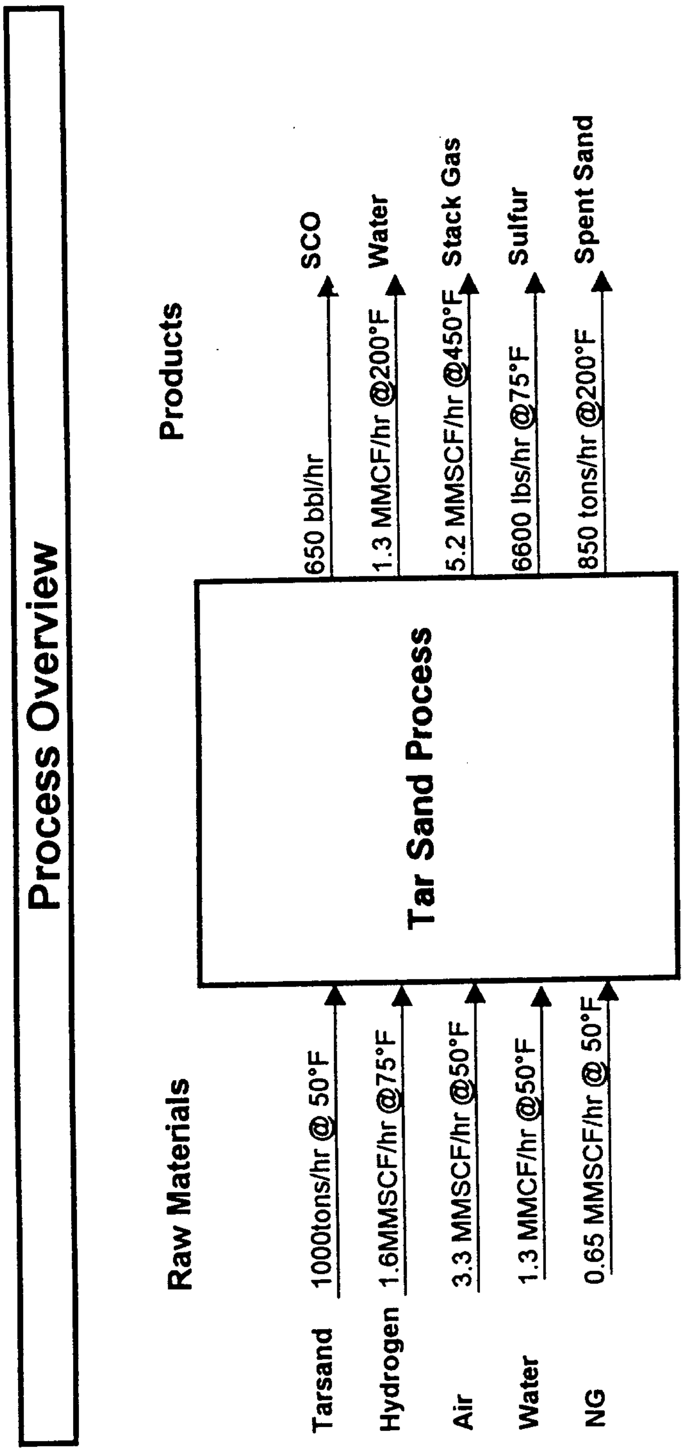
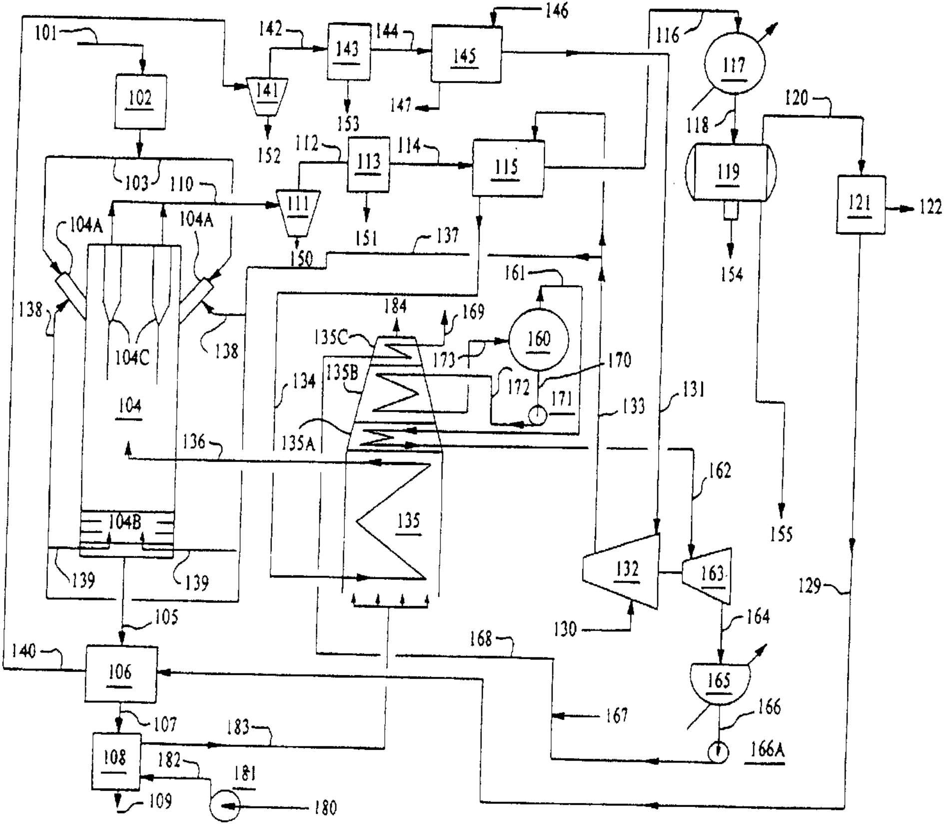

Process Overview
Raw Materials
Products
Tarsand
1000tons/hr @ 50°F
Hydrogen
1.6MMSCF/hr @75°F
Air
3.3 MMSCF/hr @50°F
Water
1.3 MMCF/hr @50°F
NG
0.65 MMSCF/hr @ 50°F
Tar Sand Process
650 bbl/hr
SCO
1.3 MMCF/hr @200°F
Water
5.2 MMSCF/hr @450°F
Stack Gas
6600 lbs/hr @75°F
Sulfur
850 tons/hr @200°F
Spent Sand

FIG. 6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