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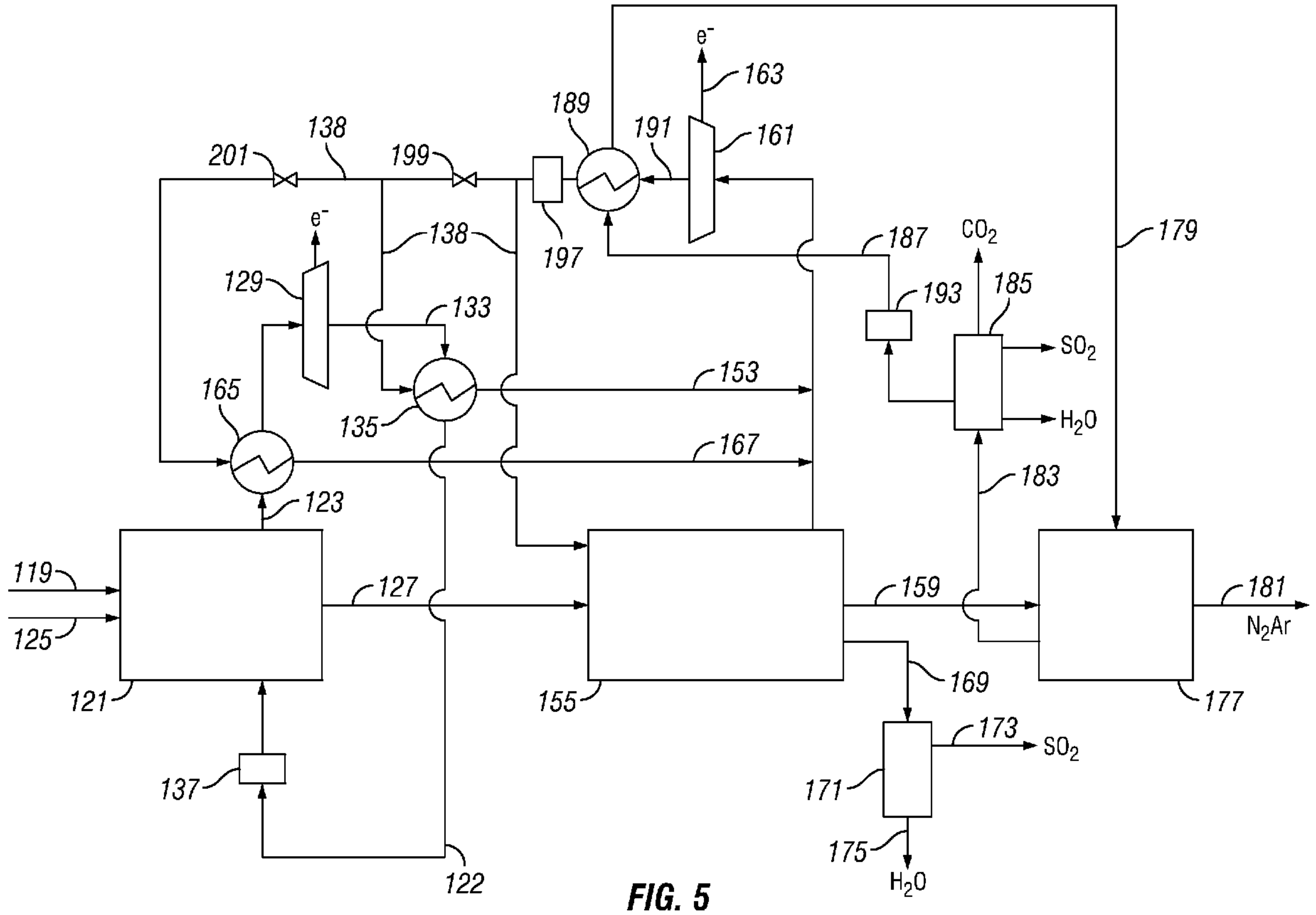

FIG. 5

(57) **Abrégé/Abstract:**
A process for producing power from a fuel stream containing at least 30 mol% hydrogen sulfide is provided. The fuel stream is combusted with an oxidant stream containing molecular oxygen to generate a combusted gas stream containing thermal power, where the molar ratio of molecular oxygen to hydrogen sulfide is at least 1:1. Electrical power is generated from the thermal power of the combusted gas stream.

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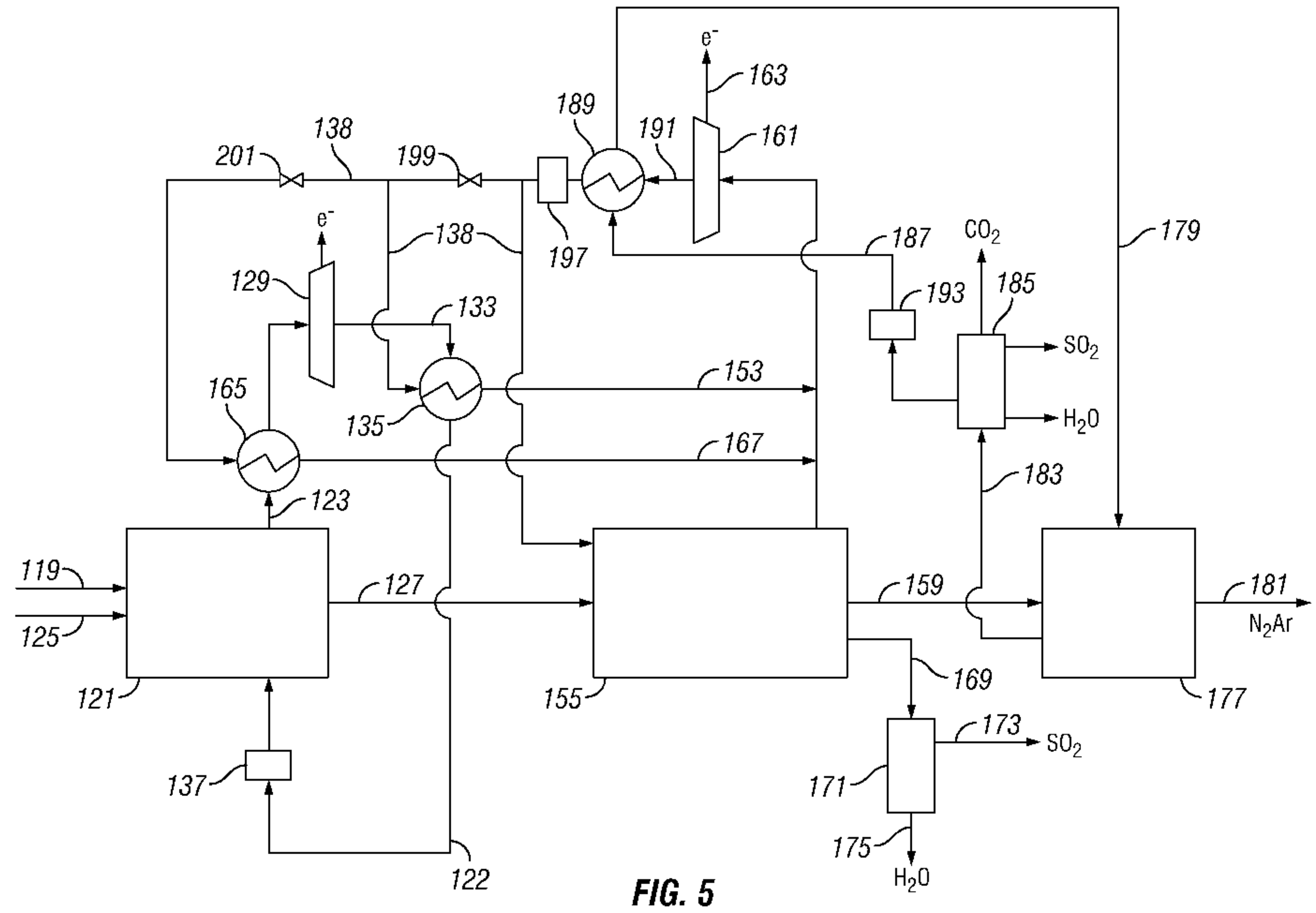

FIG. 5

(57) **Abstract**: A process for producing power from a fuel stream containing at least 30 mol% hydrogen sulfide is provided. The fuel stream is combusted with an oxidant stream containing molecular oxygen to generate a combusted gas stream containing thermal power, where the molar ratio of molecular oxygen to hydrogen sulfide is at least 1:1. Electrical power is generated from the thermal power of the combusted gas stream.

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METHOD OF PRODUCING POWER

Field of the Invention

The present invention relates to methods of generating power. In particular, the present invention is directed to a method of producing power from a hydrogen sulfide-rich fuel stream.

Background of the Invention

Natural gas is recovered from natural gas reservoirs and is combusted to produce electrical energy, typically after being transported via pipeline to an energy producing facility. The composition of the gas stream recovered from the reservoir impacts the economic viability of recovering and producing electrical power from the gas stream. Gas streams recovered from different reservoirs may have significantly different compositions. For example, sweet, non-acid gas contains large quantities methane, typically greater than 90 mole %, and may contain small quantities of hydrogen sulfide and carbon dioxide. Sweet, non-acid gas streams are the most preferred gas streams for production of electrical power, requiring little processing for conditioning and pipelining.

Acid gas contains significant quantities of carbon dioxide, for example, greater than 5 mole % CO_2 and is less desirable than sweet non-acid gas since energy must be expended to remove CO_2 from the gas stream prior to pipelining to avoid corroding the pipeline. The energy is frequently provided by burning a portion of the methane from which the CO_2 is separated, reducing the amount of natural gas that is supplied to the pipeline. Significant quantities of CO_2 in the gas stream also reduce the overall energy value of the gas stream relative to a sweet non-acid gas since CO_2 has no thermal energy value.

Sour gas contains significant quantities of hydrogen sulfide, at least 5 mole %, or at least 10 mole %, often at least 30 mole %, and sometimes greater than 50, 75, or even 90 mole % H_2S. Moderate quantities of hydrogen sulfide in a gas stream render the gas stream less desirable for recovery and production of natural gas since the hydrogen sulfide must be removed from the reservoir gas stream. This requires a portion of the energy of the gas stream to effect the separation, reducing the overall energy that may be captured from the gas stream. Large quantities of hydrogen sulfide in a gas stream, e.g. at least 30 mole % H_2S, preclude the development and recovery of the gas stream reservoir since the energy required to separate the H_2S from the natural gas becomes too great for the reservoir to be developed commercially.

Prior to about 1970, hydrogen sulfide from gas streams containing small to moderate amounts of hydrogen sulfide was flared to allow recovery of natural gas and associated hydrocarbon liquids. Flaring the hydrogen sulfide resulted in the emission of sulfur dioxide to the atmosphere, which is now limited by restrictions on sulfur emissions. The Claus process was introduced as a means to separate and capture H_2S as elemental sulfur from hydrogen sulfide containing natural gas streams, while capturing some of the heat energy value of the H_2S for use in the separation process. The Claus process burns 1/3 of the H_2S substoichiometrically with air and reacts the resulting sulfur dioxide with the remaining H_2S over a catalyst to recover elemental sulfur. The thermal energy produced by burning a portion of the hydrogen sulfide is generally used to separate the hydrogen sulfide from the gas stream.

The Claus process, however, is effective commercially only for gas streams containing moderate amounts of hydrogen sulfide, for example, up to 30 mole % H_2S. As shown in the chart in Fig. 1 for a natural gas field containing 35 mole % H_2S and minor amounts of carbon dioxide and nitrogen along with natural gas liquids and methane, less than 50% of the energy of the natural gas in the gas stream is available after energy losses associated with separating H_2S, CO_2, and natural gas liquids even including the energy provided by the Claus process. This leaves only half of the stream available as marketable gas to pay for field development, a complex processing operation, and delivery infrastructure—a cost that is frequently too high to justify developing and producing the sour natural gas field. Producing natural gas from sour gas reservoirs containing even higher levels of hydrogen sulfide is even less commercially desirable, and, at concentrations of about 65-70 mole % hydrogen sulfide and above, cannot be done without utilizing more energy to produce the natural gas than is present in the natural gas produced.

What is needed is a process to capture the energy of a gas stream containing a large quantity of hydrogen sulfide, e.g., gas streams containing at least 30 mole % hydrogen sulfide.

Summary of the Invention

In one aspect, the present invention is directed to a method for producing electrical power in which at least 90 vol.% of a gas stream containing at least 30 mole % hydrogen sulfide is combusted to generate thermal power contained in a combustion gas stream. The combustion gas stream comprises sulfur dioxide. Electrical power is generated from the thermal power of the combustion gas stream.

In another aspect, the present invention is directed to method for producing electrical power in which a gas stream containing at least 30 mole % hydrogen sulfide is separated from a subterranean geological formation and is combusted to generate thermal power contained in a combustion gas stream comprised of sulfur dioxide. Electrical power is generated from the thermal power of the combustion gas stream.

In an embodiment, sulfur dioxide is captured from the combustion gas stream and is injected into a geological subterranean formation.

Brief Description of the Drawings

Fig. 1 is a chart showing the energy capture efficiency of a process utilizing a prior art Claus process to capture sulfur from a gas stream containing 35 vol. % hydrogen sulfide.

Fig. 2 is a schematic of a system for practicing the process of the present invention for treating a feed gas stream high in hydrogen sulfide to produce electrical power.

Fig. 3 is a schematic of a system for practicing the process of the present invention for treating a feed gas stream high in hydrogen sulfide to produce electrical power, where the system includes a gas turbine and a heat recovery steam generator.

Fig. 4 is a schematic of a system for practicing the process of the present invention for treating a feed gas stream high in hydrogen sulfide to produce electrical power, where the system includes a gas turbine, a supplemental firing unit, and a heat recovery steam generator.

Fig. 5 is a schematic of a system for practicing the process of the present invention for treating a feed gas stream high in hydrogen sulfide to produce electrical power, where the system includes a binary power cycle.

Detailed Description of the Invention

The present invention describes methods for utilization of one or more feed streams containing high levels of hydrogen sulfide to produce electrical power. Electrical power is produced from such feed streams in the process of the present invention by combusting a fuel stream comprised of one or more of the feed streams—the fuel stream being comprised of at least 30 mol.% hydrogen sulfide—to produce thermal power, and subsequently generating electrical power from the thermal power produced by combusting the fuel stream.

Feed streams containing high levels of hydrogen sulfide are available from subsurface geological formations such as sour gas-rich reservoirs, and a fuel stream for use in the process of the present invention may be formed from one or more feed streams separated from a subsurface geological formation. The invention described herein allows for the processing of fuel streams previously deemed not suitable for commercial development. Such fuel streams contain at least about 30 mole %, or at least 35 mole %, or at least about 40 mole %, or at least about 50 mole %, or at least about 90 mole % hydrogen sulfide.

The fuel stream utilized in the process of the present invention may contain non-H_2S components. The fuel stream may contain substantial amounts of methane, other hydrocarbonaceous gases, and/or hydrocarbon liquids, and may contain up to 70 mole %, or up to 50 mole %, or up to 30 mole % hydrocarbonaceous gases, including methane, and hydrocarbon liquids. The fuel stream may also contain up to 30 mole % carbon dioxide, however, less carbon dioxide is preferable since carbon dioxide provides no thermal energy upon combustion of the fuel stream. The fuel stream may also include minor amounts of entrained hydrocarbon liquids, minor amounts of inert gases such as nitrogen and helium, and entrained particulates.

The fuel stream utilized in the process of the present invention may be formed from one feed stream from a subsurface geological formation or more than one feed stream from one or more subsurface geological formations. If the fuel stream is formed from more than one feed stream, an individual feed stream may contain less than 30 mole % hydrogen sulfide, however, the combined feed streams that form the fuel stream will contain at least 30 mole % hydrogen sulfide. Similarly, if the fuel stream is formed from more than one feed stream an individual feed stream may contain more than 70 mole % hydrocarbonaceous gases and liquids, however, the combined feed streams the form the fuel stream will contain at most 70 mole % hydrocarbonaceous gases and liquids. In like manner, if the fuel stream is formed from more than one feed stream an individual feed stream may contain more than 30 mole % carbon dioxide, however, the combined feed streams that form the fuel stream will contain at most 30 mole % carbon dioxide.

The fuel stream utilized in the process of the present invention will be formed primarily of gaseous components, but may contain small amounts of hydrocarbon liquids, non-hydrocarbon liquids, and particulate solids. The gaseous components of the fuel stream, at the temperature and pressure at which the fuel stream is provided to the process of the present invention, may comprise at least 90 mass %, or at least 95 mass %, or at least 99 mass % of the total components in the fuel stream. In an embodiment, the fuel stream may contain entrained hydrocarbon liquids (C_2+), where the entrained hydrocarbon liquids may comprise up to at most 10 mass%, or up to 5 mass%, or up to 1 mass% of the fuel stream. Alternatively, entrained hydrocarbon liquids (C_2+) may be separated from a feed stream prior to utilizing the feed stream as at least a portion of the fuel stream that is combusted to generate power. Preferably the fuel stream contains no solid particulates, and may be filtered to remove substantially all particulates prior to combustion.

The feed streams used to form the fuel stream utilized in the process of the present invention are preferably obtained from a subsurface geological formation. One or more feed streams used to form the fuel stream, however, may be provided from a source other than a subsurface geological formation, for example, from a processing plant, a reformer, or a refinery.

If a feed stream used in the process of the present invention is provided from a subsurface geological formation, the feed stream may be separated from the subsurface geological formation in accordance with conventional methods and means for separating gas streams from subsurface geological formations. For example, the feed stream may be separated from a subsurface geological formation in the same manner that natural gas is conventionally recovered from subsurface geological formations. Subsurface geological formations from which a feed stream may be separated include gas reservoirs, condensate reservoirs, and oil reservoirs.

Due to the difficulty and danger of pipelining a gas stream containing at least 30 mole % of hydrogen sulfide over substantial distances, the process of the present invention is preferably effected near the subsurface geological formation from which the feed stream is recovered. In particular, the process of the present invention is preferably effected without having to pipeline the feed stream to a processing facility located more than 25 miles, or more than 50 miles, or more than 100 miles from the location of the subsurface geological formation from which the feed stream is recovered.

In an embodiment of the present invention, the process may be effected at a location near one or more subsurface geological formations located near a power grid so that electrical power produced by the process of the present invention may be easily exported to a power grid or utilized to provide power to a load. If located near a power grid, preferably the power grid is located near a consumer of electrical power to reduce the amount of shrinkage (energy loss) due to transmission of the power over distance. If the process is effected at a location remote from an electrical power grid, electrical power produced by the process may be transmitted to a power grid utilizing efficient electrical transmission lines having low line power loss to reduce shrinkage. For example, electrical power produced by the process of the present invention may be transmitted from remote location to a power grid and to power consumers via a 765 kV 6 conductor transmission line having a line loss of from 0.3MW to 0.7 MW, or from 0.4MW to 0.6MW, per 100 miles at a 1000 MW load—which is comparable to losses of transporting a natural gas from a remote location via pipeline to an electrical power production facility. Alternatively, if the subsurface geological formation is in a remote location, electrical power consuming units may be located near the subsurface geological formation to utilize the electrical power generated from a feed stream separated from the subsurface geological formation. For example, computer server farms may be located near the subsurface geological formation to utilize the electrical power generated from a feed stream separated from the formation.

Referring now to Fig. 2, a feed stream 101 may be recovered from a subsurface geological formation 103. The feed stream 101 may be utilized directly as the fuel stream, or alternatively, the feed stream 101 may be processed to separate at least a portion, and preferably substantially all, of liquids and/or solids that may be entrained in the feed stream 101. To separate liquids and/or solids that may be entrained in the feed stream 101, the feed stream 101 may be passed to a separator 105. The separator 105 may be a simple closed tank where the force of gravity serves to separate liquids from the gas. The separator 105 may also include an expander, where the feed stream is expanded to cool and condense the liquids from the gas, where the expansion is preferably a limited expansion so the resulting feed stream retains significant pressure. The separator may include a filter or a membrane effective to separate particulate solids, preferably substantially all particulate solids, from the feed stream 101, preferably while maintaining significant pressure in the feed stream 101. If the separator 105 includes an expander, the filter or membrane for separating particulate solids from the feed stream 101 is preferably positioned in the separator 105 to remove particulate solids from the feed stream 101 prior to expanding the feed stream 101 through the expander. Entrained hydrocarbonaceous liquids 107, other condensable liquids such as water 108, and solids 110 may be separated from the feed stream 101 in the separator 105 to produce a feed stream 111 that may be utilized as a portion of the fuel stream for combustion or as the entire fuel stream for combustion.

In an embodiment, further natural gas liquids (C_2-C_5) may be separated from a feed stream 111 by absorbing the natural gas liquids from the feed stream 111 in a lean oil unit 109 after separation of heavier hydrocarbon liquids, water, and particulate solids from the feed stream 101. The feed stream 111 may be contacted with an oil in the lean oil unit 109 to absorb non-methane hydrocarbons which may be recovered from the lean oil unit 109 as natural gas liquids 115 and to produce a feed stream 117 from which most or all of the natural gas liquids have been removed.

The feed stream 117, or the feed stream 111, or the feed stream 101 forms at least a portion of, or all of, the fuel stream 119 that is combusted in the process of the present invention. At least a portion of, or all of, the fuel stream 119 is combusted in combustor 121 to generate a combusted gas stream containing thermal power. An oxidant stream 125 is provided to the combustor 121 for combustive reaction with the fuel stream 119 to produce the combusted gas stream.

The oxidant stream 125 is provided in an amount effective to provide a stoichiometric amount, or a slight stoichiometric excess, or a substantial stoichiometric excess of molecular oxygen relative to combustible fuel components in the fuel stream. The oxidant stream 125 is provided in an amount, or at a rate, sufficient to provide at least 1.0 mole, or at least 1.5 moles, or at least 2 moles of molecular oxygen per mole of hydrogen sulfide in the fuel stream. If the fuel stream also comprises methane, the oxidant stream may be provided in an amount, or at a rate, sufficient to provide at least 2 moles of molecular oxygen per mole of methane in addition to the amount of molecular oxygen provided relative to hydrogen sulfide.

As used herein, “oxidant” refers to a composition that contains molecular oxygen that may be combustively reacted with hydrogen sulfide as a fuel source. Examples of oxidants include oxygen, oxygen admixed with steam, oxygen admixed with carbon dioxide, air, and/or enriched air. “Enriched air” refers to air having an oxygen content greater than about 21 percent by volume. Enriched air may be used to increase, relative to air, the combustion temperature of the fuel stream 119 at a constant fuel input rate and/or to facilitate post combustion processing of the combustion effluent gases.

The combustor 121 may comprise a combustion zone and a steam generator, where the steam generator may be boiler, a superheated steam boiler, or a supercritical steam generator.

The fuel stream 119 and the oxidant stream 125 may be combustively reacted in the combustion zone to produce the combusted gas stream. Steam may be produced by exchanging heat between the combusted gas stream and a liquid aqueous stream 122, preferably water, and/or steam at the steam generator in the combustor 121.

The fuel stream 119, or the portion of the fuel stream 119 that is to be combusted, and the oxidant stream 125 are provided to the combustor 121, and are mixed and combustively reacted in the combustion zone of the combustor 121. Combustion of the fuel stream 119 generates a significant amount of thermal energy that is contained in the combusted gas stream produced by the combustion of the fuel stream. The thermal energy contained in the combusted gas stream may be transferred to the liquid aqueous stream/steam in the steam generator portion of the combustor 121. Preferably the thermal energy is transferred from the combusted gas stream to the liquid aqueous stream/steam by indirectly contacting the combusted gas stream with the liquid aqueous stream/steam, for example by passing the combusted gas stream through a coil surrounded by the liquid aqueous stream and/or steam or by passing the combusted gas stream over a coil containing the liquid aqueous stream and/or steam. Sufficient thermal energy may be transferred from the combusted gas stream to the liquid aqueous stream and/or steam to generate superheated or supercritical steam and a cooled combusted gas stream 127. The cooled combusted gas stream 127 is a flue gas stream that comprises sulfur dioxide and water.

In an embodiment, the combustor 121 is a pressurized steam generator, and the fuel stream 119 and oxidant stream 125 are provided to the pressurized steam generator at relatively high pressure, e.g., at least 1.7 MPa, or at least 3.4 MPa, or at least 6.9 MPa. The fuel stream and the oxidant stream should be provided to the pressurized steam generator at substantially similar pressures to prevent an overpressurized stream blowing an underpressurized stream back out of the pressurized boiler or steam generator. The one or more feed streams that comprise the fuel stream 119 may be at a pressure of at least 1.7 MPa when separated from the geological subterranean formation and often may be at a high pressure of at least 6.9 MPa or at least 10 MPa, or at least 13.8 MPa, or at least 15 MPa, or at least 20 MPa when separated from the geological subterranean formation. One or more of the feed streams may be passed through an expander in the separator 105 to reduce the pressure of the feed stream 101 comprising the fuel stream 119. If the fuel stream 119 has a very high pressure relative to the oxidant stream 125, the fuel stream may be expanded in one or more expanders 124 to reduce the pressure of the fuel stream to be substantially the same as the pressure of the oxidant stream. In an embodiment of the process of the invention the fuel stream 119 may be at a pressure of from 1.7 MPa to 13.8 MPa when provided to the combustor 121 after any expansion of one or more feed streams in the separator 105 and/or expansion of the fuel stream in one or more expanders 124. One or more of the expanders 124 and/or an expander in separator 105 may be utilized to generate mechanical rotational power upon expansion of the fuel stream 119 through the expander(s) which may be used to generate electrical power.

The oxidant stream 125 may be compressed using a conventional compressor 126 to compress the oxidant stream to a pressure of from 1.7 MPa to 13.8 MPa prior to being provided to the combustor 121 comprising a pressurized steam generator, and preferably is compressed to about the pressure of the fuel stream 119 entering the pressurized boiler or steam generator. Power to operate the oxidant compressor 126 may be provided to the compressor 126 by a shaft 128 coupled to an expander 124 through which the fuel stream 119 is expanded. It is preferred to substantially compress the oxidant stream to maximize the power produced relative to the size of the combustor 121.

Combustion in a combustor 121 comprising a pressurized steam generator may be further preferred when the fuel stream contains substantial quantities of methane and/or carbon dioxide so that carbon dioxide present in the cooled combusted gas stream 127 may be more easily separated from the cooled combusted gas stream. Carbon dioxide contained in the cooled high pressure combusted gas stream 127 may be easily separated from the cooled combusted gas stream, for example, by pressure swing absorption, due to the high pressure of the cooled combusted gas stream.

In the steam generator, whether pressurized or not, a substantial portion, preferably substantially all, of the thermal power generated by combustion of the fuel stream 119 and produced in the combusted gas stream may be captured in superheated steam, e.g., at least 80%, or at least 85%, or at least 90%, up to 95%, or up to 97% of the thermal power generated by combustion of the fuel stream may be captured in steam. A substantial portion, preferably substantially all, of the thermal power in the combusted gas stream may be captured as steam having a selected temperature and/or pressure profile. A substantial portion, preferably substantially all, of the thermal power in the combusted gas stream may be used to generate steam at pressures ranging from 0.34 MPa to 34.5 MPa, or from 13.8 MPa to 34.5 MPa, or from 22.2 MPa to 34.5 MPa, or from 30 MPa to 34.5 MPa; and temperatures ranging from 240° C to 650°C, or from 335°C to 650°C, or from 375°C to 650°C. Preferably the steam generated by capturing the thermal power from the combusted gas stream is supercritical steam (temperature of at least 374°C and a pressure of at least 22.15 MPa) or ultrasupercritical steam (temperature of at least 374°C and a pressure of at least 30 MPa).

Electrical power 131 may be generated from the thermal power captured in the steam generated in the combustor 121. The steam 123 generated in the combustor 121 may be passed through one or more steam expanders 129 to expand the steam 123 and form an expanded steam stream 133 and to generate mechanical rotational power by turning a shaft. The mechanical rotational power may be converted to electrical power utilizing conventional electrical power generation equipment and processes for converting mechanical power to electrical power. Expansion of the steam 123 through the one or more expanders may generate substantial electrical power, where the electrical power generated may be at least 100 MW_e, or at least 200 MW_e, or at least 300 MW_e, or at least 400 MW_e. The one or more steam expanders 129 are preferably steam turbines.

The steam 123 is cooled by Joule-Thompson cooling as it passes through the one or more steam expanders 129, converting the thermal power contained in the steam 123 into mechanical power, which is then converted to electrical power. The cooled expanded steam stream 133 may have a temperature of from 100°C to 150°C, or from 110°C to 130°C, and may have a pressure corresponding to the vapor pressure of water at the temperature of the steam, e.g. from 0.101 MPa to 0.5 MPa.

The expanded steam stream 133 may be cooled further to liquefy the steam into the liquid aqueous stream 122 that may be recycled into the combustor 121. The expanded steam stream 133 may be cooled by passing the expanded steam stream into a cooling tower 135, where the steam is condensed to the liquid aqueous stream having a temperature that is low enough so that the liquid aqueous stream 122 may be pumped back into the combustor 121 through pump 137 without causing cavitation in the pump, preferably at most 95°C, or at most 90°C, or at most 85°C, and typically from 75-85°C at a pressure of 0.101 MPa, or at a temperature of up to 150°C at a pressure of up to 0.5 MPa. Alternatively, and more preferably, the expanded steam stream 133 is cooled using a heat exchanger 135 where heat is exchanged between the expanded steam stream 133 and a fluid 138 having a boiling point at least 50°C cooler than the boiling point of the liquid of the liquid aqueous stream 122 at 0.101 MPa and having a latent heat of evaporation of at least 350 kJ/kg to condense water from the expanded steam stream 133 and capture the latent heat of condensation of water of the liquid aqueous stream as thermal power, as described in further detail below. The liquid aqueous stream 122 produced by cooling the expanded steam stream 133 may then be recycled back to the combustor 121 by pumping the liquid aqueous stream 122 into the combustor 121 with a boiler water pump 137.

In another embodiment, shown in Fig. 3, the combustor 121 may be comprised of a gas turbine 139 and a steam generator 141. The fuel stream 119 and the oxidant stream 125 may be combustively reacted in the gas turbine 139 to produce a combusted gas stream containing thermal power. In an embodiment, the oxidant stream 125 and the fuel stream 119 may be provided for combustion in the gas turbine in relative amounts, or at relative rates, such that the molar ratio of molecular oxygen and hydrogen sulfide provided for combustion is at least 2:1. Providing the oxidant stream and the fuel stream for combustion at a molar ratio of at least 2:1 molecular oxygen: hydrogen sulfide may limit the temperature of the resulting combusted gas stream relative to combustion of the streams at substantially stoichiometric amounts of molecular oxygen and hydrogen sulfide, where limiting the temperature of the combusted gas stream may be desirable to avoid damaging the gas turbine or to avoid the necessity of utilizing a gas turbine constructed with specialized high heat resistant metallurgy. The fuel stream 119 and the oxidant stream 125 may be provided to the gas turbine 139, respectively, at a pressure of from 1 MPa to 5 MPa, where preferably the pressures at which the fuel stream 119 and the oxidant stream 122 are provided to the gas turbine are within 0.5 MPa of each other.

At least a portion of the thermal power contained in the combusted gas stream may be utilized to generate electrical power by expanding the combusted gas stream in the gas turbine 139 to generate rotational mechanical power that may be converted to electrical power 144. Expansion of the combusted gas stream in the gas turbine 139 may generate mechanical rotational power by turning a shaft. The mechanical rotational power may be converted to electrical power 144 utilizing conventional electrical power generation equipment and processes for converting rotational mechanical shaft power to electrical power.

The expanded combusted gas stream 145 is cooler than the combusted gas stream due to the Joule-Thompson effect, however, the expanded combusted gas stream may still contain a significant amount of thermal power. In an embodiment of the process of the present invention, the expanded combusted gas stream has a temperature above the autoignition temperature of hydrogen sulfide—at least 260°C—and more preferably has a temperature of at least 300°C, or at least 400°C, or at least 500°C, or at least 600°C, and typically has a temperature of from 500°C to 650°C.

The expanded combusted gas stream 145 may be passed to a steam generator 141 to generate a steam stream 123 from the thermal power contained in the expanded combusted gas stream 145 and to generate a flue gas stream containing sulfur dioxide and water formed of the cooled expanded combusted gas stream . The steam stream 123 may be produced by indirectly exchanging heat between the expanded combusted gas stream 145 and a liquid aqueous stream 122, preferably water, and/or steam in the steam generator 141. A substantial portion, preferably substantially all, of the thermal power in the expanded combusted gas stream 145 may be captured in superheated steam, e.g., at least 80%, or at least 85%, or at least 90%, up to 95%, or up to 97% of the thermal power in the expanded combusted gas stream 145 may be captured in superheated steam. A substantial portion, preferably substantially all, of the thermal power in the expanded combusted gas stream 145 may be captured as steam having a selected temperature and/or pressure profile. A substantial portion, preferably substantially all, of the thermal power in the expanded combusted gas stream 145 may be used to generate steam at pressures ranging from 0.34 MPa to 34.5 MPa, or from 13.8 MPa to 34.5 MPa, or from 22.2 MPa to 34.5 MPa, or from 30 MPa to 34.5 MPa; and temperatures ranging from 240° C to 650°C, or from 335°C to 650°C, or from 375°C to 650°C. Preferably the steam generated by capturing the thermal power from the expanded combustion gas stream 145 is supercritical steam (temperature of at least 374°C and a pressure of at least 22.15 MPa) or ultrasupercritical steam (temperature of at least 374°C and a pressure of at least 30 MPa).

Electrical power 131 may be generated from the steam produced in the steam generator 141 as described above with respect to generating power from steam produced in combustor 121—the steam 123 may be passed through one or more steam expanders 129 to generate rotational mechanical power that may be converted to electrical power 131. The resulting expanded steam stream 133 may be cooled in heat exchanger 135 by heat transfer with fluid 138 to condense water and produce the liquid aqueous stream 122, as described above, where the fluid 138 has a boiling point at least 50°C lower than the boiling point of the liquid aqueous stream 122 at 0.101 MPa and has a latent heat of evaporation of at least 350 kJ/kg, as described in further detail below. The resulting liquid aqueous stream 122 may be recycled back into the stream generator 141 via pump 137. Expansion of the steam 123 through the one or more steam expanders 129 may generate substantial electrical power, where the electrical power generated may be at least 100 MW_e, or at least 200 MW_e, or at least 300 MW_e, or at least 400 MW_e. The one or more steam expanders 129 are preferably steam turbines.

In another embodiment, as shown in Fig. 4, the combustor 121 may be comprised of a gas turbine 139, a supplemental firing unit 147, and a steam generator 141. The fuel stream 119, or a portion thereof 143, and the oxidant stream 125 may be combustively reacted in the gas turbine 139 to produce a combusted gas stream containing thermal power as described above. Further, the combusted gas stream may be expanded in the gas turbine 139 to generate rotational mechanical power that may be converted to electrical power 144 and an expanded combusted gas stream 145, as described above.

The oxidant stream 125 and the fuel stream 119, or portion thereof 143, may be provided for combustion in the gas turbine 139 in relative amounts, or at relative rates, such that the molar ratio of molecular oxygen to hydrogen sulfide that are provided for combustion is at least 2:1. The resulting expanded combusted gas stream 145 exiting the gas turbine 139 will contain at least 5 vol.% or at least 10 vol% molecular oxygen, and may contain from 5 vol% to 20 vol% molecular oxygen, and typically may contain from 8 vol% to 12 vol% molecular oxygen. The relatively large volume of oxidant stream needed to provide molecular oxygen in a molar ratio of at least 2:1 relative to hydrogen sulfide in the fuel stream inhibits excessive heating within the gas turbine 139 and ensures that sufficient molecular oxygen is contained in the expanded combusted gas stream 145 for further combustive reaction in the supplemental firing unit 147 without requiring the addition of further oxidant.

The expanded combusted gas stream 145 and a supplemental fuel stream 149, preferably a sulfur-containing fuel and most preferably a portion of the fuel stream 119, may be combustively reacted in the supplemental firing unit 147 to produce a second combusted gas stream 151 containing thermal power. The second combusted gas stream 151 may have a temperature of from 600°C to 1000°C.

The supplemental fuel stream 149 may be any fuel that may be combusted with the expanded combusted gas stream 145, where the expanded combusted gas stream 145 contains at least 5 vol.% molecular oxygen. The supplemental fuel stream 149 may be natural gas or coal. Preferably, the supplemental fuel stream 149 is a sulfur-containing fuel. Sulfur-containing fuels include hydrogen sulfide and high-sulfur content coal. Most preferably, the supplemental fuel stream 149 comprises or consists of a portion of the fuel stream 119.

The supplemental fuel stream 149 is preferably provided to the supplemental firing unit 147 in an amount sufficient to consume all, or substantially all, the molecular oxygen in the expanded combusted gas stream 145 to prevent the formation of oleum. The supplemental fuel stream 149 may be provided in an amount effective to consume at least 99% of the molecular oxygen in the expanded combusted gas stream 145.

The second combusted gas stream 151 may be passed to a steam generator 141 to generate a steam stream 123 from the thermal power contained in the second combusted gas stream 151 and to generate a flue gas stream containing sulfur dioxide and water formed of the cooled second combusted gas stream. Steam may be produced by exchanging heat between the second combusted gas stream 151 and a liquid aqueous stream 122, preferably water, and/or steam in the steam generator 141. A substantial portion, preferably substantially all, of the thermal power in the second combusted gas stream 151 may be captured in superheated steam, e.g., at least 80%, or at least 85%, or at least 90%, up to 95%, or up to 97% of the thermal power in the second combusted gas stream 151 may be captured in superheated steam. A substantial portion, preferably substantially all, of the thermal power in the second combusted gas stream 151 may be captured as steam having a selected temperature and/or pressure profile. A substantial portion, preferably substantially all, of the thermal power in the second combusted gas stream 151 may be used to generate steam at pressures ranging from 0.34 MPa to 34.5 MPa, or from 13.8 MPa to 34.5 MPa, or from 22.2 MPa to 34.5 MPa, or from 30 MPa to 34.5 MPa; and temperatures ranging from 240° C to 650°C, or from 335°C to 650°C, or from 375°C to 650°C. Preferably the steam generated by capturing the thermal power from the second combusted gas stream 151 is supercritical steam (temperature of at least 374°C and a pressure of at least 22.15 MPa) or ultrasupercritical steam (temperature of at least 374°C and a pressure of at least 30 MPa).

In an embodiment of the invention, the supplemental firing unit 147 and the steam generator 141 may form a single unit. In the combined supplemental firing unit and steam generator, heat may be exchanged between the second combusted gas stream and the liquid aqueous stream 122 and/or steam as the second combusted gas stream is formed immediately upon combustion of the expanded combusted gas stream 145 and the supplemental fuel stream 149. The temperature of the second combusted gas stream may be maintained at a selected temperature, preferably below 800°C, by the immediate exchange of heat between the second combusted gas stream and the liquid aqueous stream and/or steam.

Electrical power 131 may be generated from the steam produced in the steam generator 141 as described above with respect to generating power from steam produced in combustor 121—the steam 123 may be passed through one or more steam expanders 129 to generate rotational mechanical power that may be converted to electrical power 131. The resulting expanded steam stream 133 may be cooled by heat transfer with fluid 138 to condense water and produce the liquid aqueous stream 122, as described above, where the fluid 138 has a boiling point at least 50°C lower than the boiling point of the liquid aqueous stream 122 at 0.101 MPa and has a latent heat of evaporation of at least 350 kJ/kg, as described in further detail below. The resulting liquid aqueous stream 122 may be recycled back into the steam generator 141 via pump 137. Expansion of the steam 123 through the one or more steam expanders 129 may generate substantial electrical power, where the electrical power generated may be at least 100 MW_e, or at least 200 MW_e, or at least 300 MW_e, or at least 400 MW_e. The one or more steam expanders 129 are preferably steam turbines.

Referring now to Fig. 5, additional electrical power may be generated by capturing the latent heat of condensation of water by condensing water from the expanded steam stream 133. Fuel stream 119 and oxidant stream 125 are combusted in combustor 121, and steam stream 123 is generated by exchanging heat between a liquid aqueous stream 122 and a combusted gas produced by the combustion, as described in the various embodiments of combustion and heat exchange described above. Electrical power 131 is generated by expanding the steam stream 123 through the steam expander 129 to produce the expanded steam stream 133, where the steam stream 123 may have a temperature of from 600°C to 1000°C, and the expanded steam stream 133 may have a temperature of from greater than 100°C to 150°C, typically from 120°C to 135°C.

The expanded steam stream 133 may be cooled to condense water therefrom and produce the liquid aqueous stream 122 by exchanging heat with a portion of a fluid 138 in heat exchanger 135, producing a heat transfer gas 153 from the fluid 138. The fluid 138 may have a boiling point at 0.101 MPa that is at least 40°C, or at least 50°C, lower than the boiling point of the liquid aqueous stream 122, which may be water, and has a latent heat of evaporation of at least 350 kJ/kg, preferably at least 500 kJ/kg, and most preferably at least

1000 kJ/kg. Preferably the fluid 138 has a boiling point at 0.101 MPa that is from -50°C to 65°C, or from 0°C to 50°C. The fluid 138 may be selected from the group consisting of anhydrous ammonia, anhydrous carbon dioxide, carbon disulfide, sulfur dioxide, diethyl ether, dimethyl ether, methylene chloride, methanol, and acetone. In an embodiment, the liquid is inorganic.

The heat exchanger 135 may provide means for the expanded steam stream 133 and the fluid 138 to exchange heat indirectly, for example by conduction. The heat exchanger 135 may comprise a coil through which the fluid 138 may pass and over which the expanded steam stream 133 may flow to transfer thermal power from the steam stream 133 to the fluid 138 to convert the liquid to a heat transfer gas 153. Water from the expanded stream stream 133 may condense on the coil in the heat exchanger 135, permitting the transfer of thermal power of the latent heat of condensation of water from the expanded steam stream 133 to the liquid/heat transfer gas in the coil.

Preferably, the fluid 138 is provided to the heat exchanger 135 at a rate and a temperature relative to the rate and temperature of the expanded steam stream 133 effective to induce condensation of water from the expanded steam stream 133 and to convert the expanded steam stream 133 to the liquid aqueous stream 122 having a temperature of at most 95°, or at most 90°C, or at most 85°C, and preferably from 70°C to 85°C so that the liquid aqueous stream 122 may be pumped without significant cavitation through the pump 137. The fluid 138 may be provided to the heat exchanger at a temperature of at least 0°C, and from 0°C to 50°C, or from 0°C to 25°C, and preferably from 0°C to 15°C.

The fluid 138 should also be provided to the heat exchanger 135 at a rate and a temperature relative to the rate and temperature of the expanded steam stream 133 effective to permit the heat exchange between the fluid 138 and the expanded steam stream 133 to vaporize the fluid 138 and produce a heat transfer gas 153. The resulting heat transfer gas 153 may have a temperature of from 20°C up to 120°C, typically from 50°C to 70°C.

A significant portion of the thermal power transferred from the expanded steam stream 133 to the fluid 138 in the process of converting the expanded steam stream 133 to the liquid aqueous stream 122 is derived from the latent heat of condensation of water in the expanded steam stream 133 as water is condensed from the expanded steam stream. In an embodiment of a process of the present invention, at least 20%, or at least 30%, or at least 40% of the total electrical power produced by the process is derived from thermal power produced by the latent heat of condensation of water from the expanded steam stream 133 and transferred to the fluid 138 or the heat transfer gas 153.

The fluid 138 may also be utilized to capture thermal power from the flue gas stream 127 exiting the combustor 121. The flue gas stream 127, typically comprising sulfur dioxide, water, nitrogen and smaller quantities of carbon dioxide and argon, may exit the combustor 121 at a temperature of from greater than 100°C to 200°C, typically from 115°C to 150°C, where the flue gas stream is formed of the combusted gas stream, the expanded combusted gas stream, or the second combusted gas stream cooled by heat exchange with the liquid aqueous stream 122 in the steam generator. The fluid 138, having a boiling point at least 50°C lower than the boiling point of the liquid aqueous stream 122 and having a latent heat of evaporation of at least 350 kJ/kg, may be indirectly contacted with the flue gas stream 127 in an evaporator 155 to transfer thermal power from the flue gas stream 127 to the fluid 138—cooling the flue gas stream 127 and converting the fluid 138 to a heat transfer gas 157. The fluid 138 may be provided to the evaporator 155 for heat transfer duty at a temperature of from greater than 0°C to 50°C, typically from 3°C to 10°C. The fluid 138 may be indirectly contacted with the flue gas stream 127 to transfer the thermal energy from the flue gas stream 127 to the fluid 138/heat transfer gas 157 by passing the flue gas stream 127 through a coil surrounded by the fluid 138 and/or heat transfer gas 157 or by passing the flue gas stream 127 over a coil containing the fluid 138 and/or the heat transfer gas 157.

Sufficient thermal energy may be transferred from the flue gas stream 127 to the fluid 138 and/or heat transfer gas 157 to generate a heat transfer gas 157 having a temperature of from 100°C to 150°C, typically from 110°C to 130°C, at a pressure of 1MPa to 15 MPa, or from 2 MPa to 10 MPa. The flue gas stream 127 may be cooled by the heat transfer to produce a cooled flue gas stream 159 having a temperature of from greater than 0°C to 25°C, typically from 3°C to 10°C.

The heat transfer gas 157 containing thermal power derived by heat transfer with the flue gas stream 127 in the evaporator 155 and the heat transfer gas 153 containing thermal power derived by heat transfer with the expanded steam stream 133 may be combined and then expanded through an expander 161 to transfer at least a portion of the thermal power into rotational mechanical power which may be converted to electrical power 163. The electrical power produced by expanding the heat transfer gas 157 through the expander may be at least 50 MW, or at least 100 MW, or at least 150 MW, or at least 200 MW. The amount of electrical power 163 produced by expanding the combined heat transfer gases 157 and 153 through the expander 155 may be at least 25%, or 30%, or 35%, or 40%, or 50% of the total electrical power produced by the process of the present invention. The amount of electrical power 163 produced by expanding the combined heat transfer gases 157 and 153 through the expander 161 may be up to 100%, or up to 90%, or up to 80% of the electrical power produced by expansion of the combusted gas and expansion of the steam stream. A large portion of the electrical power 161 produced by expansion of the heat transfer gases 153 and 157 is derived from capture of the latent heat of condensation of water from the expanded steam stream since the latent heat of condensation of water provides from 1 MW_t to 1.3MW_t per 1 MW_e produced by expansion of the combusted gas stream and expansion of the steam stream.

In an embodiment of the process of the invention, a portion of the fluid 138 may be indirectly contacted with the steam stream 123 in a heat exchanger 165 to form a supercritical heat transfer gas 167 from the fluid 138. The steam stream 123 may have a temperature of from 600°C to 1000°C, and may easily heat the fluid 138 to a supercritical state, e.g. if the liquid is anhydrous ammonia the ammonia heat transfer gas 167 may be heated to a temperature greater than 130°C up to 150°C at a pressure of at least 11.4 MPa. The fluid 138 may be provided for heating with the steam stream 123 in the heat exchanger 165 in sufficient quantity or at a sufficient rate so that, upon combination with the heat transfer gas 153 and with the heat transfer gas 157 the combined heat transfer gases 153, 157, and 165 are in a supercritical state. Expansion of a supercritical heat transfer gas in the expander 161 may be desirable to avoid damage to the expander that might be caused by condensation of the heat transfer gas as it is expanded through the expander 161.

In a preferred embodiment, the flue gas stream is scrubbed to capture sulfur dioxide, water, and any carbon dioxide prior to venting the effluent gases from the system. As discussed above, the flue gas stream 127 is indirectly contacted with the fluid 138 in the evaporator 155 to transfer thermal power from the flue gas stream 127 to the liquid—forming the heat transfer gas 157 and a cooled flue gas stream 159. The cooled flue gas stream 159 may have a temperature of from 0°C to 50°C, more preferably the cooled flue gas stream 159 may have a temperature of from 3°C to 10°C. An aqueous liquid stream 169 comprising sulfur dioxide and water from the flue gas stream 127 may be separated from the cooled flue gas stream 159 in the evaporator 155 as a result of the heat transfer to the fluid 138. The aqueous liquid stream 169 may be drained from the evaporator 155 and fed to a separator 171. The aqueous liquid stream 169 may be heated in the separator to a temperature of at least 12ºC, or at least 15ºC up to 50ºC, preferably up to 25ºC to separate sulfur dioxide gas 173 from liquid water 175.

The cooled flue gas stream 159 may be introduced into a scrubber 177 and may be scrubbed with a solvent 179 to separate sulfur dioxide and carbon dioxide (if any) from the cooled flue gas stream 159 and into the solvent. The cooled flue gas stream 159 may be scrubbed with the solvent 179 utilizing conventional liquid/gas scrubbing equipment. The scrubbed flue gas 181 may contain nitrogen and argon, and may be vented to the atmosphere.

The solvent 179 may be any solvent which is effective to physically or chemically separate sulfur dioxide from a gas stream, preferably is also effective to physically or chemically separate carbon dioxide from a gas stream, and preferably has a freezing point below -80ºC. For example, the solvent 177 may be methanol, a methanol:water mixture, N-methyl-2-pyrrolidine, propylene carbonate, dimethyl ethers of polyethylene glycol, and ethanolamines. Preferably the solvent is anhydrous methanol.

The solvent 179 may be fed to the scrubber 177 at a temperature and pressure at which the selected solvent is effective to absorb sulfur dioxide, and preferably carbon dioxide as well, from the cooled flue gas stream 159. The solvent may be fed to the scrubber at a temperature of from -100ºC to 0ºC, and may be mixed with the cooled flue gas stream in the scrubber at a pressure of from 1.7 MPa to 10.3 MPa. The solvent 179 may be provided at a temperature within 5ºC of the freezing point of fluid 138 by exchanging heat with cold fluid 191 produced by expansion of the heat transfer gas in the expander 161, as described in further detail below. In preferred embodiment, the solvent is methanol, and the cooled flue gas stream 159 is scrubbed with the methanol solvent at a temperature of from -50ºC to -85ºC and a pressure of from 3.4 MPa to 6.9 MPa.

The solvent 183 containing sulfur dioxide and carbon dioxide (if any) scrubbed from the cooled flue gas stream 159 may be fed from the scrubber 177 to a separator 185 in which sulfur dioxide and carbon dioxide (if any) may be separated from the solvent and captured. Sulfur dioxide and carbon dioxide may be separated from the solvent 183 by heating the solvent to a temperature at which sulfur dioxide gas and carbon dioxide gas separate from the liquid solvent and/or by reducing the pressure on the solvent to a pressure at which sulfur dioxide gas and carbon dioxide gas dissociate from the liquid solvent. The temperature of the solvent exiting the scrubber may be from 5°C to 40°C, or from 5°C to 20°C, and the temperature of the solvent may be raised to a temperature of greater than 40°C up to 60°C to separate sulfur dioxide and carbon dioxide from the solvent. In a preferred embodiment, the temperature of the solvent may be raised to a temperature at which carbon dioxide is separated from the solvent, and then raised to a second temperature at which sulfur dioxide is separated from the solvent, wherein the carbon dioxide and the sulfur dioxide separated from the solvent are captured separately.

The heated solvent 187 may be cooled for recirculation to the scrubber 177 by exchanging heat in a heat exchanger 189 with the cold fluid 191 produced by expansion of the heat transfer gas 153, 157, and 167 through the expander 161. The outlet pressure of the expander 161 may be selected and controlled to produce the cold fluid 191 at a temperature near the freezing point of the fluid 138. For example, if the fluid 138 is anhydrous ammonia, the outlet pressure of the expander 161 may be selected to produce the cold fluid 191 at a temperature of from -75°C to -80°C. The heated solvent 187 may be cooled to the temperature selected to absorb sulfur dioxide and carbon dioxide from the cooled flue gas stream 159 by selecting an appropriate feed rate of the solvent from the separator 185 to the heat exchanger 189, where pump 193 may be utilized to pump the heated solvent 187 to the heat exchanger 189 at the selected rate.

The cold fluid 191 is heated by exchanging heat with the heated solvent 187 in heat exchanger 189 to produce the fluid 138 for use in cooling the flue gas stream 127 in the evaporator 155 and for cooling the expanded steam stream 133 in heat exchanger 135. The rate the supercooled fluid is fed to the heat exchanger 189 may be selected to produce the fluid 138 at a temperature of from 0°C to 50°C, or from 0°C to 25°C, and preferably from 0°C to 15°C. The resulting fluid 138 may be pumped to the evaporator 155, heat exchanger 135, and heat exchanger 165 via pump 197, and the relative amounts of fluid 138 pumped to each element may be controlled by metering devices 199 and 201.

The process of the present invention is effective to produce substantial electrical power from a fuel stream containing significant quantities of hydrogen sulfide.

To facilitate a better understanding of the present invention, the following examples of certain aspects of some embodiments are given. In no way should the following examples be read to limit, or define, the scope of the invention.

EXAMPLE 1

Calculations were performed demonstrating that combusting a fuel stream containing 100 mol% H_2S and an oxidant stream containing about 3 moles of molecular oxygen per mole of hydrogen sulfide in the fuel stream in a gas turbine to produce a combusted gas stream containing thermal energy, and expanding the combusted gas stream in the gas turbine to produce mechanical power that is converted to electrical power, is effective to produce electrical power on a scale comparable to producing electrical power from a natural gas stream.

The mass flow rates of a fuel stream formed of 100 vol.% (100 mol%) hydrogen sulfide and an oxidant stream formed of air, respectively, required to be provided for combustion and subsequent expansion of the combusted gas stream in a simple cycle gas turbine to produce an amount of electrical power were calculated and compared with the mass flow rates of a fuel stream formed of 100 vol.% methane and an oxidant stream formed of air, respectively, required to be provided for combustion and subsequent expansion of the combusted gas stream to produce an identical amount of electrical power in the same simple cycle gas turbine. Relevant characteristics of each of the fuel streams are provided in Table 1 and the composition of the oxidant stream at 0°C and 0.101 MPa is provided in Table 2.

TABLE 1

Fuel Composition	**Lower Heating Value (kJ/kg)**	**Auto-ignition temperature (°C)**	**Flammability limits (vol% in air)**
100 vol.% CH_4	49862	538	5.3-14.0
100 vol.% H_2S	15226	260	4.3-45.5

TABLE 2

Component	**Vol. %**	**L/m^3**	**g/m^3**	**Weight fraction**
Molecular Nitrogen	78.09	780.9	976.1	0.7551
Molecular Oxygen	20.95	209.5	299.3	0.2315
Argon	0.93	9.3	16.6	0.0128
Carbon Dioxide	0.038	0.38	0.8	0.0006

The mass rate flow calculations were based in part on specifications provided for the production of electrical power using a General Electric 7FA Simple Cycle gas turbine. In particular, the heat rate in units of kJ/kW-h, the gas turbine mass flow rate in units of kg/s, and the gas turbine electrical power generation in megawatts (MW_e) for the GE 7FA Simple

Cycle gas turbine that were necessary to perform the calculations are provided in General Electric publication GE12985H, and are shown in Table 3.

TABLE 3

System	Heat rate (kJ/kW-h)	Mass flow rate (kg/s)	Electrical power generation (MW_e)
GE 7FA Simple Cycle gas turbine	9873	436	183

The mass flow rate of each respective fuel stream provided to the simple cycle gas turbine to produce 183 MW_e of electrical power was determined from the power to be produced, the efficiency of the gas turbine's conversion of thermal power to electrical power, and the lower heating value of the fuel composition. In particular, each fuel stream mass flow rate was calculated according to formula (1):

(1)

$$\text{Fuel Stream Mass Flow Rate} = \frac{(\text{Electrical Power Generation } (\frac{kJ}{h}) * \frac{1}{\text{Turbine Efficiency for converting Themal Power to Electrical Power}})}{\text{Lower Heating Value } (\frac{kJ}{kg})},$$

where the Electrical Power Generation is provided in Table 3 as 183 MW_e (where 183 MW_e = 658,800,000 kJ/h) and the Lower Heating Value for each fuel is provided in Table 1. The Turbine Efficiency for conversion of Thermal Power to Electrical Power for the GE 7FA Simple Cycle gas turbine was determined according to the formula (2):

(2)

$$\text{Turbine Efficiency} = \frac{3600 \frac{kJ}{kWh}}{\text{Heat Rate} \frac{kJ}{kWh}}$$

which is $\frac{3600 \text{ kJ/kWh}}{9873 \text{ kj/kWh}} = 0.365$, or 36.5%.

Using the formulas above, the mass flow rate of the fuel stream containing 100 vol.% H_2S required to be provided for combustion in the GE 7FA Simple Cycle gas turbine to produce 183 MW_e was calculated to be 118660 kg/h, and the mass flow rate of the fuel stream containing 100 vol.% CH_4 required to be provided for combustion in the gas turbine to produce the same amount of electrical power was calculated to be 36235 kg/h.

The mass flow rates of the oxidant stream to be provided for combustion in the gas turbine with each of the fuel streams to produce 183 MW_e were calculated from the total mass flow rate in the gas turbine as provided in Table 3 (436 kg/s) and the calculated mass flow rate of each respective fuel stream according to formula (3)

(3)

Oxidant stream mass flow rate (kg/h) =

total mass flow rate in gas turbine (kg/h) – mass flow rate of fuel stream in gas turbine (kg/h)=

1569600 kg/h – mass flow rate of fuel stream (kg/h).

The mass flow rate of the oxidant stream to be combustively reacted with the fuel stream containing 100 vol.% H_2S to produce 183 MW_e in the gas turbine was calculated to be 1450939 kg/h, and the mass flow rate of the oxidant stream to be combustively reacted with the fuel stream containing 100 vol.% CH_4 to produce 183 MW_e was calculated to be 1533365 kg/h.

The molar ratio of molecular oxygen to hydrogen sulfide in the combustion of the fuel stream containing 100 vol. % H_2S with the oxidant stream in the gas turbine to generate 183 MW_e was calculated to be greater than 1:1. The relative mass flow rates of the components of the oxidant stream were calculated based on the relative weight fractions of the components multiplied by the mass flow rate of the oxidant stream. The calculated input mass flow rates of the oxidant stream components are shown in Table 4.

TABLE 4

Component	Calculation	Input Mass Flow Rate (kg/h)
Oxygen (molecular)	1450939kg/h * 0.2315	335892
Nitrogen (molecular)	1450939kg/h * 0.7551	1095604
Argon	1450939kg/h * 0.0128	18572
Carbon Dioxide	1450939kg/h * 0.0006	871

The input molar flow rate of oxygen was calculated from the input mass flow rate of oxygen: [(335892 kg/h)*(1000 g/kg)]/32 g/m (O_2) = 10496625 m/h O_2, and the molar flow rate of hydrogen sulfide was calculated from the input mass flow rate of the fuel stream, where the fuel stream consisted only of hydrogen sulfide, [(118661 kg/h) * (1000 g/kg)]/34 g/m (H_2S) = 3490029 m/h H_2S. The ratio of molecular oxygen to hydrogen sulfide provided by the fuel stream containing hydrogen sulfide and the oxidant stream over the same period of time, therefore, was calculated to be 10496625/3490029 = 3.0.

A summary of the data and calculations for each of the fuel streams and the oxidant stream and production of power in the gas turbine is provided in Table 5.

TABLE 5

	Calculation 1	**Comparative Calculation 1**
Gas Turbine	GE 7FA	GE 7FA
Cycle Type	Simple Cycle	Simple Cycle
Fuel Composition, CH_4 (vol.%)	0	100
Fuel Composition, H_2S (vol.%)	100	0
Lower Heating Value, CH_4 (kJ/kg)	--	49862
Lower Heating Value, H_2S (kJ/kg)	15226	--
Heat Rate (kJ/kWh)	9873	9873
Total Gas Turbine Mass Flow Rate (kg/h)	1569600	1569600
Gas Turbine CH_4 Fuel Mass Flow Rate (kg/h)	0	36235
Gas Turbine H_2S Fuel Mass Flow Rate (kg/h)	118660	0
Oxidant Stream Mass Flow Rate (kg/h)	1450939	1533365
Molecular Oxygen Mass Flow Rate (kg/h)	335892	354974
Molecular Nitrogen Mass Flow Rate (kg/h)	1095604	1157844
Argon Mass Flow Rate (kg/h)	18572	19627
CO_2 Mass Flow Rate (kg/h)	871	920
Exhaust Temperature (°C)	604	604
Total Output Power (MW_e)	183	183

The calculations demonstrated that a fuel stream consisting of H_2S can be combusted in a simple cycle gas turbine to produce electrical power equivalent to the electrical power produced by combusting a natural gas stream consisting of methane in a similar manner. The fuel stream consisting of H_2S must be provided for combustion at a rate about 3 times greater than a fuel stream consisting of methane to produce an equivalent amount of power.

EXAMPLE 2

The electrical power produced by combusting selected fuel streams in a combined cycle gas-turbine/heat recovery steam generator system was calculated. For the calculations, the GE 7FA Simple Cycle gas turbine comprised the first cycle of the combined cycle power generation system and a heat recovery steam generator (HRSG) utilizing a GE S107FA steam turbine comprised the second cycle. The total electrical power produced in the combined cycle system is produced from expansion of a combusted gas stream in the gas turbine, as described above with respect to a single cycle system, and from expansion of a steam stream through a steam turbine, where the steam stream is produced by exchange of heat between the expanded combusted gas stream and water/steam in the HRSG. Calculations were made to determine the total electrical power produced by combustion of the selected fuel streams in the gas turbine, expansion of the resulting combusted gas stream through the gas turbine, exchanging heat between the expanded combusted gas stream and water and/or steam in the HRSG, and expansion of the resulting hot steam stream through a steam turbine in the HRSG. As shown above, the electrical power produced in the GE 7FA gas turbine is 183 MW_e, where the mass input flow rate of the fuel stream is adjusted to generate the electrical power produced by the gas turbine. The electrical power produced by the HRSG is dependent on the temperature of the expanded combusted gas stream exiting the gas turbine, the components, and relative quantity of these components, of the expanded combusted gas stream, and on the efficiency of the heat recovery cycle of the HRSG.

The temperature of the expanded combusted gas stream exiting the GE 7FA gas turbine produced from a fuel stream consisting of 100% CH_4 is provided in the specifications of the GE 7FA as 604°C in GE Publication GEH12985H. For purposes of the electrical power calculations, the exit temperature for each of the expanded combusted gas streams produced from the selected fuel streams was assumed to be 604°C.

The efficiency of the heat recovery cycle of the combined cycle system utilizing a GE 7FA gas turbine and an HRSG with a GE S107FA steam turbine was calculated by determining the thermal power (MW_t) that would be exchanged between an expanded combusted gas stream produced by combusting the fuel stream containing 100% CH_4 as described above in Example 1 and water/steam, then calculating the efficiency of the heat recovery cycle from the electrical power output specification of the heat recovery cycle of the system as published in GE Publication GEH12985H relative to the thermal power exchanged between the expanded combusted gas stream and the water/steam. The thermal power exchanged between the expanded combusted gas stream and water/steam in the HRSG was calculated from the temperature difference between the expanded combusted gas stream exiting the gas turbine and the flue gas stream exiting the HRSG, the mass flow rate of each of the components of the expanded combusted gas stream exiting the gas turbine, and the heat capacity of each of the components of the expanded combusted gas stream.

The temperature difference between the expanded combusted gas stream and water/steam in the HRSG and the flue gas stream is 484°C (484°K), assuming that the flue gas stream exits the HRSG at 120°C and the expanded combusted gas stream exits the gas turbine at 604°C as set forth in GE Publication GEH12985H. A temperature of 120°C is a typical flue gas stream temperature when the flue gas stream is formed by exchanging heat between a combusted gas stream and steam.

Calculation of the mass flow rate of each of the components of the expanded combusted gas stream produced from a fuel stream containing 100% CH_4 as the expanded combusted gas stream exits the gas turbine was dependent on determining the components of the expanded combusted gas stream and on calculation of the relative quantities of such components. The components and relative quantity of the components of the expanded combusted gas stream depends in turn on the components and relative quantities of such components of the fuel stream and the oxidant stream, and on the stoichiometry of the combustion reaction of the fuel stream and oxident stream. For the fuel stream containing 100% CH_4, the combustion reaction of CH_4 with air is $CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$. The components of the expanded combusted gas stream and the relative mass flow rate of these components in the expanded combusted gas stream, therefore, were determined from the components of air as the oxidant stream as set forth in Table 4 and their relative mass flow rates to the gas turbine as reacted with the CH_4 of the fuel stream at its mass flow rate to the gas turbine. Complete combustion and consumption of CH_4 was assumed since a substantial stoichiometric excess of oxygen is provided for combustion relative to CH_4. The mass flow rate of molecular oxygen in the expanded combusted gas stream was calculated according to formula (4):

(4)

$$\begin{gathered}
\textit{mass flow rate of molecular oxygen in expanded combusted gas stream (kg/h)} = \\
[\textit{mass flow rate of oxygen in oxidant stream (kg/h)}] - \\
\{[\frac{2*(\textit{molecular wt .of molecular oxygen } (\frac{g}{m}))}{\textit{molecular wt.methane } (\frac{g}{m})}] * \\
[\textit{mass flow rate of fuel stream of 100 vol.\% methane (kg/h)}]\}.
\end{gathered}$$

The mass flow rates of carbon dioxide and water vapor components in the expanded combusted gas stream were calculated according to formula (5):

(5)

$$\begin{gathered}
\textit{mass flow rate of component (kg/h)} = \\
[\textit{mass flow rate of component in oxidant stream (kg/h)}] + \\
\{[\frac{(\textit{moles of component produced per mole of methane}) * (\textit{molecular wt.of component } (\frac{g}{m}))}{\textit{molecular wt.methane } (\frac{g}{m})}] *
\end{gathered}$$

[mass flow rate of fuel stream of 100 vol.% methane (kg/h)]}.

The mass flow rate in the expanded combusted gas stream of components of the oxidant stream that are inert in the combustion reaction was assumed to be equivalent to the mass flow rate of those components in the oxidant stream fed to the gas turbine. Table 6 provides the calculated mass flow rates of components of the expanded combusted gas stream formed by combustion of the fuel stream containing 100 vol.% methane with a dry air oxidant stream in a GE 7FA gas turbine utilizing the fuel stream and oxidant stream mass flow rates set forth in Example 1 above.

TABLE 6

Combusted Gas Stream Component	**Calculated Mass Flow (kg/h)**
Oxygen	210032
Nitrogen	1157844
Argon	19627
Carbon Dioxide	100567
Water Vapor	81530

The thermal power contained in each component of the expanded combusted gas stream component was calculated according to formula (6):

(6)

$$\text{Thermal Power Provided by Component For Electrical Power Generation} = \frac{[(\text{heat capacity of component (kJ/kg)}) * (\text{mass flow rate of component (kg/h)} * (\Delta T_{\text{max expanded combusted gas stream } T - \text{flue gas stream } T})]}{3600000\ \text{kJ/MW}_t}.$$

The calculated thermal power of the components is shown in Table 7.

TABLE 7

Combusted Gas Stream Component	**Heat Capacity (kJ/(kg*K)**	**Mass Flow Rate (kg/h)**	**Temperature Change (°C)**	**Thermal Power MW$_t$**
Oxygen	0.92	210032	484	26.0
Nitrogen	1.04	1157844	484	161.9
Argon	0.52	19627	484	1.4

Carbon Dioxide	0.84	100567	484	11.4
Water Vapor	1.97	81530	484	21.6

The total thermal power contained in the expanded combusted gas stream available upon cooling the expanded combusted gas stream to 120°C was calculated to be 222.3 MW_t by summing the thermal power of the combusted gas stream components.

The efficiency of the HRSG system in converting the thermal power of the expanded combusted gas stream to electrical power was then calculated. According to General Electric Publication GEH 12985H, the second cycle of the combined cycle power generation system produces 86 MW_e electrical power utilizing a methane fuel stream. The efficiency of the second cycle of the combined cycle system, therefore, was calculated as $\frac{86\ MW\ (electrical)}{222.3\ MW\ (thermal)}$ $*100 = 38.7\%$.

Based on the calculated efficiency of the HRSG cycle in the combined cycle gas turbine/HRSG system, the electrical power capable of being produced by selected fuel streams containing H_2S and, optionally CH_4 and/or CO_2 was calculated. In particular, the electrical power capable of being produced in a combined cycle gas turbine/HRSG system by a fuel stream containing 100 vol% H_2S provided at a rate effective to produce 183 MW_e in the gas turbine cycle, as described in Example 1, was calculated. In addition, the electrical power capable of being produced in a combined cycle gas turbine/HRSG system by a fuel stream containing 50 vol.% H_2S and 50 vol.% CH_4, and by a fuel stream containing 33 vol.% H_2S, 33 vol.% CH_4, and 34 vol.% CO_2 was calculated. Mass flow rates of the selected fuel streams into the gas turbine were calculated in accordance with formula (1) above, where the lower heat capacity of the blended fuel streams was calculated in accordance with formula (7):

(7)

$$\textit{Lower Heating Value of Fuel Stream (kJ/kg)} = \\ [(\textit{wt. fraction } H_2S)*(\textit{lower heating value of } H_2S\ (kJ/kg))] + \\ [(\textit{wt. fraction } CH_4)*(\textit{lower heating value of } CH_4\ (kJ/kg)).$$

For each selected fuel stream, components of the expanded combusted gas stream produced by combustion of the fuel stream and air and their relative quantities were determined as set forth above with respect to the fuel stream containing 100 vol.% CH_4, except where H_2S is a component of the fuel stream the expanded combusted gas stream will contain sulfur dioxide and water vapor produced in accordance with the combustion reaction

$H_2S + 1.5\ O_2 \rightarrow SO_2 + H_2O$. In this case, the mass flow rate of molecular oxygen in the expanded combusted gas stream was adjusted to reflect consumption of molecular oxygen by reaction with H_2S by including the amount of oxygen consumed by reaction with H_2S in the calculation of the mass flow rate of molecular oxygen in the expanded combusted gas stream, where the amount of oxygen consumed by reaction with H_2S was calculated according to formula (8):

(8)

$$\text{molecular oxygen consumed by combustive reaction with } H_2S\ (kg/h) = [(\frac{1.5 * \text{molecular weight of molecular oxygen } (\frac{g}{m})}{\text{molecular weight of hydrogen sulfide} (\frac{g}{m})}) * \text{mass flow rate of hydrogen sulfide in fuel stream } (kg/h)].$$

The mass flow rate of sulfur dioxide in an expanded combusted gas stream produced by combusting a fuel stream containing H_2S, and optionally CH_4 and/or CO_2, was calculated according to formula (9):

(9) $$\text{mass flow rate of sulfur dioxide in expanded combusted gas stream } (kg/h) = [\frac{(\text{moles of } SO2 \text{ produced per mole of } H2S) * (mol.wt.SO2\ (\frac{g}{m}))}{mol.wt.H2S\ (\frac{g}{m})}] * [\text{mass flow rate of } H_2S \text{ in fuel stream } (kg/h)].$$

The mass flow rate of water vapor in an expanded combusted gas stream produced by combusting a fuel stream containing H_2S, and optionally CH_4 and/or CO_2, was calculated according to formula (10):

(10) $$\text{mass flow rate of water vapor in expanded combusted gas stream } (kg/h) = \{ [\frac{(\text{moles of } H2O \text{ produced per mole of } H2S) * (mol.wt.H2O\ (\frac{g}{m}))}{mol.wt.hydrogen\ H2S\ (\frac{g}{m})}] * [\text{mass flow rate of } H_2S \text{ in fuel stream } (kg/h)] \} + \{ [\frac{(\text{moles of } H2O \text{ produced per mole of } CH4) * (mol\ wt.H2O\ (\frac{g}{m}))}{mol.wt.CH4}] * [\text{mass flow rate of } CH_4 \text{ in fuel stream } (kg/h)] \}$$

The mass flow rate of carbon dioxide in an expanded combusted gas stream produced by combusting a fuel stream containing H_2S, and optionally CH_4 and/or CO_2, was calculated according to formula (11):

(11) *mass flow rate of carbon dioxide in expanded combusted gas stream (kg/h) =*

$$\left\{\left[\frac{(moles\ of\ CO2\ produced\ per\ mole\ of\ CH4)*(mol.wt.CO2\ (\frac{g}{m}))}{mol.wt.CH4\ (\frac{g}{m})}\right]* \right.$$

$$\left.\left[\textit{mass flow rate of CH4 in fuel stream (kg/h)}\right]\right\} +$$

$$\left[\textit{mass flow rate of } CO_2 \textit{ in fuel stream (kg/h)}\right] +$$

$$\left[\textit{mass flow rate of } CO_2 \textit{ in oxidant stream (kg/h)}\right]$$

The heat capacities of the components of expanded combusted gas stream produced by combustion of each selected fuel stream were used to calculate the thermal power contained in each component of the expanded combusted gas stream in accordance with formula (6), where the heat capacity of sulfur dioxide is 0.64. The electrical power capable of being produced from the selected fuel streams and the relevant data for calculating the electrical power is provided in Table 8.

TABLE 8

	Calculation 2	**Calculation 3**	**Calculation 4**	**Comparative Calculation 2**
Gas Turbine	GE 7FA	GE 7FA	GE 7FA	GE 7FA
Cycle Type	Combined	Combined	Combined	Combined
Fuel stream composition, CH_4 (Vol.%)	0	50	33	100
Fuel stream composition H_2S (Vol.%)	100	50	33	0
Fuel stream composition (CO_2) (Vol.%)	0	0	34	0
Lower Heating Value, CH_4 (kJ/kg)	--	49862	49862	49862
Lower Heating Value H_2S (kJ/kg)	15226	15226	15226	--
Lower Heating Value, Fuel Blend (kJ/kg)	--	26310	13799	--
Heat Rate (kJ/kWh)	9873	9873	9873	9873
Gas Turbine Mass Flow (kg/s)	436	436	436	436
Gas Turbine Power Generation (MW_e)	183	183	183	183
Expanded Combusted Gas Temperature (°C)	604	604	604	604
Gas Turbine Total Fuel Rate (kg/h)	118661	68673	130936	36235
Gas Turbine CH_4 Fuel Rate (kg/h)	0	21975	21975	36235
Gas Turbine H_2S Fuel Rate (kg/h)	118661	46698	46698	0
Gas Turbine CO_2 Fuel Rate (kg/h)	0	0	62263	0
Gas Turbine Oxidant (Dry Air) Stream Rate (kg/h)	1450939	1500927	1438664	1533365
Gas Turbine Molecular Oxygen Rate (Oxidant Stream) (kg/h)	335892	347465	333051	354974
Gas Turbine Nitrogen Rate (Oxidant Stream) (kg.h)	1095604	1133350	1086335	1157844
Gas Turbine Argon Rate (Oxidant Stream) (kg/h)	18572	19212	18415	19627
Gas Turbine Carbon Dioxide Rate (Oxidant Stream) (kg/h)	871	901	863	920
Molecular Oxygen Consumption (kg/h)	167521	153827	153827	144942
Expanded Combusted Gas Stream Carbon Dioxide (kg/h)	871	61333	123559	100567
Expanded Combusted Gas Stream Sulfur Dioxide (kg/h)	223361	87901	87901	0
Expanded Combusted Gas Stream Oxygen (kg/h)	168372	193637	179223	210032
Expanded Combusted Gas Stream Nitrogen (kg/h)	1095604	1133350	1086335	1157844
Expanded Combusted Gas Stream Argon (kg/h)	18572	19212	18415	19627
Expanded Combusted Gas Stream Water Vapor (kg/h)	62820	74167	74167	81530
Recoverable Thermal Power from Expanded Combusted Gas Stream 604°C to 120°C (MW_t)	211	218	217	222
Calculated HRSG Cycle Efficiency (%)	38.7	38.7	38.7	38.7
Flue Gas Stream Exhaust Temperature	120	120	120	120
HRSG Cycle Power (MW_e)	82	84	84	86
Total Output Power (MW_e)	265	267	267	269

The calculations demonstrated that a fuel stream consisting of H_2S, or containing 50 vol.% H_2S and 50 vol.% CH_4, or containing 33 vol.% H_2S, 33 vol.% CH_4, and 34 vol.% CO_2 can be combusted in a combined cycle gas turbine/HRSG to produce electrical power equivalent to the electrical power produced by combusting a natural gas stream consisting of methane in a similar manner. The fuel streams containing H_2S must be provided for combustion at rates greater than a fuel stream consisting of methane to produce an equivalent amount of power.

EXAMPLE 3

The electrical power produced by combusting selected fuel streams in a combined cycle gas-turbine/heat recovery steam generator system with supplemental firing to combust the oxygen in the expanded combusted gas stream of each combusted fuel stream was calculated. The fuel streams selected were the same as those selected in Example 2 except that each contained some quantity of hydrogen sulfide. No calculation of electrical power produced by supplemental firing of a fuel stream containing only methane was conducted since, as shown above, comparison of the electrical power produced in the gas turbine cycle versus the electrical power produced in the HRSG cycle indicates the gas turbine cycle produces more electrical power, therefore, the most efficient method for producing electrical power with a methane fuel stream would avoid using some of the fuel stream as fuel for supplemental firing. Supplemental firing, however, is preferred when utilizing a fuel stream containing hydrogen sulfide to combust all oxygen in the expanded combusted gas stream to prevent sulfur dioxide in the expanded combusted gas stream from reacting with oxygen and water to form sulfurous or sulfuric acid. For the calculations, the GE 7FA Simple Cycle gas turbine comprised the first cycle of the combined cycle power generation system and a supplemental firing unit in combination with a heat recovery steam generator (HRSG) utilizing a GE S107FA steam turbine comprised the second cycle.

The electrical power produced in the gas turbine cycle of the combined gas turbine/supplemental firing unit/HRSG using a GE 7FA gas turbine for each of the selected fuel streams is 183 MW_e, as discussed above.

The electrical power produced by the combined supplemental firing unit and HRSG was calculated by determining the thermal power contained in a second combusted gas stream produced by combusting all of the molecular oxygen in the expanded combusted gas stream with a selected supplemental fuel stream, calculating the efficiency of conversion of the thermal power in the second combusted gas stream to electrical power in the HRSG, and calculating the amount of electrical power produced in the HRSG from the thermal power contained in the second combusted gas stream and the efficiency of converting that thermal power to electrical power in the HRSG when the second combusted gas stream was cooled to 120°C in the HRSG. Each supplemental fuel stream was selected to have the same composition as the fuel stream used to produce power in the gas turbine cycle as set forth in Example 2 above.

The amount of thermal power in the second combusted gas stream available for conversion to electrical power in the HRSG system was calculated by adding the enthalpy of combustion produced by combusting the supplemental fuel stream and the expanded combusted gas stream in the supplemental firing unit and the thermal power contained in each component of the second combusted gas stream exchanged to produce steam in the HRSG, and subtracting the amount of thermal power required to raise the temperature of the expanded combusted gas stream from the temperature of the expanded combusted gas stream exiting the gas turbine (604°C) to a selected temperature. The amount of thermal power converted to electrical power in the second combusted gas stream, therefore, was calculated according to formula (12):

(12)

$$\text{Thermal power of second combusted stream converted to electrical power in HRSG} =$$
$$[(\text{Enthalpy of Combustion } (MW_t) + \text{Thermal Power Contained in Components of Second Combusted Gas Stream Exchanged in HRSG } (MW_t)) - \text{Thermal Power Required to Raise Expanded Combusted Gas Stream to Selected Temperature } (MW_t)].$$

The amount of electrical power produced by the combined cycle gas turbine/supplemental firing unit/HRSG system was calculated for each selected fuel stream using a second combusted gas stream temperature of 604°C (the temperature of the expanded combusted gas stream), 800°C, and 1000°C.

The enthalpy of combustion produced by combusting the supplemental fuel stream and the expanded combusted gas stream in the supplemental firing unit was calculated for each selected fuel stream according to formula (13):

(13)

$$\text{Enthalpy of Combustion } MW_t = \frac{\text{Mass Flow Rate of Supplemental Fuel Stream } \left(\frac{kg}{h}\right) * \text{Lower heating value of Supplemental Fuel Stream } \left(\frac{kJ}{kg}\right)}{3600000 \left(\frac{MWt}{\frac{kJ}{h}}\right)}$$

The mass flow rate of the supplemental fuel stream required to consume the oxygen present in the expanded combusted gas stream was calculated based on the mass flow rate of oxygen present in the expanded combusted gas stream as calculated above, the relative amounts of hydrogen sulfide and methane in the supplemental fuel stream, and on the molar ratios of molecular oxygen to hydrogen sulfide, and if present methane, as set forth in the reaction equations above. In particular, the mass flow rate of supplemental fuel stream needed to consume the oxygen present in the expanded combusted gas stream when the fuel stream contains 100 vol.% H_2S was calculated according to formula (14):

(14)

$$\text{mass flow rate of supplemental fuel stream to consume oxygen present in the expanded combusted gas stream (kg/h)} = \left(\frac{\text{molecular wt.}H2S\ (\frac{g}{m})}{1.5 * \text{molecular wt } O2\ (\frac{g}{m})}\right) * (\text{mass flow rate of molecular oxygen in expanded combusted gas stream (kg/h)}).$$

When the fuel stream contains 50 vol.% H_2S and 50 vol.% CH_4, the mass flow of the supplemental feed stream was calculated according to formula (15):

(15)

$$\text{mass flow of supplemental fuel stream to consume oxygen present in the expanded combusted gas stream (kg/h)} = \frac{\text{mass flow rate of molecular oxygen in expanded combusted gas stream } (\frac{kg}{h})}{\left\{\left[\left(\frac{2 * \text{mol.wt. of } O2\ (\frac{g}{m})}{\text{mol.wt. of } CH4\ (\frac{g}{m})}\right) * \text{wt fraction } CH4 \text{ in s.fuel stream}\right] + \left[\left(\frac{1.5 * \text{mol.wt. of } O2\ (\frac{g}{m})}{\text{mol.wt. of } H2S\ (\frac{g}{m})}\right) * \text{wt fraction of } H2S \text{ in s.fuel stream}\right]\right\}}$$

where the weight fraction of methane in the supplemental fuel stream is 0.32 and the weight fraction of hydrogen sulfide in the supplemental fuel stream is 0.68. When the supplemental fuel stream contains 33 vol.% H_2S, 33 vol.% CH_4, and 34 vol.% CO_2, the mass flow rate of the supplemental fuel stream needed to consume the oxygen present in the expanded combusted gas stream was calculated using formula (15) above where the weight fraction of methane in the fuel stream is 0.168 and the weight fraction of H_2S in the fuel stream is 0.357. The lower heating value of the supplemental fuel stream was calculated based on the lower heating values of H_2S and CH_4 and the relative weight ratios of these components in the supplemental fuel stream according to formula (16):

(16)

Lower Heating Value of Supplemental Fuel Stream =
*[(Lower Heating Value H_2S*Weight Fraction H_2S in Supplemental Fuel Stream) +*
*(Lower Heating Value CH_4 * Weight Fraction of CH_4 in Supplemental Fuel Stream)].*

Table 9 shows the calculated enthalpies of combustion of the selected supplemental fuel streams, the calculated mass flow rates of the selected fuel streams necessary to consume oxygen in the expanded combusted gas stream, and the lower heating values of the selected supplemental fuel streams for a combined cycle system producing 183 MW*e* in a GE 7FA gas turbine in a first cycle.

TABLE 9

	100% H_2S (vol.%)	50% H_2S/50% CH_4 (vol.%)	33% H_2S/33% CH_4/34% CO_2 (vol.%)
Mass flow rate fuel stream to gas turbine (from Table 8) to produce 183 MW*e* in gas turbine (kg/h)	118661	68673	1300936
Mass flow rate of oxidant stream to gas turbine (from Table 8) to produce 183 MW_e in gas turbine (kg/h)	1450939	1500927	1438664
Mass flow rate of molecular oxygen in expanded combusted gas stream (from Table 8) (kg/h)	168372	193637	179223
Wt. Fraction of CH_4 in supplemental fuel stream	0	0.32	0.17
Wt. Fraction of H_2S in supplemental fuel stream	1	0.68	0.35
Lower Heating Value of supplemental fuel stream (kJ/kg)	15226	26310	13799
Mass flow rate of supplemental fuel stream to supplemental firing unit to combust oxygen in expanded combusted gas stream (kg/h)	119263	86445	152553
Enthalpy of combustion of supplemental fuel in supplemental firing unit (MW_t)	504	632	585

The thermal power contained in the components of second combusted gas stream that is exchanged to produce steam in the HRSG was determined by calculating the thermal power contained in each of the components that is exchanged to produce steam in the HRSG and summing the results. The thermal power contained in each of the components of the second combusted gas stream that is exchanged to produce steam in the HRSG was calculated by determining the mass flow rate of each of the components of the second combusted gas stream produced by combustion of the expanded combusted gas stream and the selected supplemental fuel stream and multiplying the heat capacity for each component of the second combusted gas stream by its mass flow rate and the temperature difference of the second combusted gas stream and the flue gas stream exiting the HRSG at 120°C and dividing the result by 3600000 kj/MW_t. The thermal power contained in the second combusted gas stream may be expressed by formula (17):

(17)

$$\textit{Thermal Power Of the Second Combusted Gas Stream Exchanged to Produce Steam in HRSG} =$$

$$\Sigma\left[\frac{\text{Heat Capacity of Component X}\left(\frac{\text{kJ}}{\text{kg}}\right) * \text{Mass Flow Rate of Component X in 2d combusted gas stream}\left(\frac{\text{kg}}{\text{h}}\right) * \Delta T}{\frac{\text{MWt}}{3600000\left(\frac{\text{kJ}}{\text{kg}}\right)}}\right]$$

where X ranges from 1 to n, where n is the total number of components in the second combusted gas stream, and ΔT is the difference between the temperature of the second combusted gas stream (604°C, 800°C, or 1000°C depending on the selected temperature) and the temperature of the flue gas stream exiting the HRSG (120°C).

Calculation of the mass flow rate of the components of the second combusted gas stream was effected by 1) determining the components of the second combusted gas stream and 2) calculating of the relative quantities of such components. The components and relative quantity of the components of the second combusted gas stream depends on the components of the supplemental fuel stream and relative mass flow rates of such components (provided above in Table 9 for each of the selected fuels), on the components of the expanded combusted gas stream and the relative mass flow rates of such components (provided above in Table 8), and on the stoichiometry of the combustion reaction of the supplemental fuel stream and expanded combusted gas stream. Each of the supplemental fuel streams contains H_2S, where, as noted above, the combustion reaction of H_2S with molecular oxygen is $H_2S + 1.5O_2 \rightarrow SO_2 + H_2O$. For the supplemental fuel streams containing CH_4, the combustion reaction of CH_4 with molecular oxygen, as noted above, is $CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$. The components of the second combusted gas stream and the relative mass flow rates of these components in the second combusted gas stream, therefore, were determined from the components of the expanded combusted gas stream as set forth in Table 8 and their relative mass flow rates to the supplemental firing unit as reacted with the H_2S, and if present the CH_4, of the supplemental fuel stream at its mass flow rate to the supplemental firing unit. Complete combustion and consumption of H_2S and CH_4, if present in the supplemental fuel stream, and O_2 was assumed since the supplemental fuel stream is provided at a mass flow rate relative to the mass flow rate of oxygen in the expanded combusted gas stream to stoichiometrically react with all of the oxygen in the expanded combusted gas stream. The mass flow rate of sulfur dioxide in the second combusted gas stream was calculated according to formula (18):

(18)

$$\text{mass flow rate of sulfur dioxide in second combusted gas stream} =$$

$$\{[\frac{(\text{molecular wt .of } SO2\ (\frac{g}{m}))}{\text{molecular wt. } H2S(\frac{g}{m})}] * [\text{mass flow rate of hydrogen sulfide in supplemental fuel stream (kg/h)}]\}$$

$$+ \text{ mass flow rate of sulfur dioxde in expanded combusted gas stream.}$$

The mass flow rate of carbon dioxide in the second combusted gas stream was calculated according to formula (19):

(19)

$$\text{mass flow rate of carbon dioxide in second combusted gas stream} =$$

$$\{[\frac{(\text{molecular wt .of } CO2\ (\frac{g}{m}))}{\text{molecular wt. } CH4(\frac{g}{m})}] * [\text{mass flow rate of methane in supplemental fuel stream (kg/h)}]\}$$

$$+ \text{ mass flow rate of carbon dioxide in expanded combusted gas stream.}$$

The mass flow rate of water vapor in the second combusted gas stream was calculated according to formula (20):

(20)

$$\text{mass flow rate of water vapor in second combusted gas stream} =$$

$$\{[\left(\frac{\left(\text{molecular wt .of } H2O\ \left(\frac{g}{m}\right)\right)}{\text{molecular wt. } H2S\left(\frac{g}{m}\right)}\right) *$$

$$(\text{mass flow rate of supplemental fuel stream (kg/h)} * \text{wt. fraction of } H2S \text{ in supplemental fuel stream})] +$$

$$[(\frac{\text{molecular wt.of } H2O\ (\frac{g}{m})}{\text{molecular wt.of } CH4\ (\frac{g}{m})}) * (\text{mass flow rate of supplemental fuel stream (kg/h)} * \text{wt. fraction of } CH4)]\} +$$

$$\text{mass flow rate of water vapor in expanded combusted gas stream}$$

The mass flow rate in the second combusted gas stream of components of the expanded combusted gas stream that are inert in the combustion reaction in the supplemental firing unit was assumed to be equivalent to the mass flow rate of those components in the expanded combusted gas stream fed to the supplemental firing unit. Table 10 provides the calculated mass flow rates of components of the second combusted gas stream formed by combustion each of the selected supplemental fuel streams, the heat capacity of these components, the calculated thermal power of the components of the second combusted gas stream at temperatures of 604°C, 800°C, and 1000°C when cooled to produce a flue gas stream at 120°C, the enthalpy of combustion of each supplemental fuel stream in the supplemental firing unit, and the total thermal power for conversion to electrical power [(calculated thermal power of the components of the second combusted gas stream + enthalpy of combustion) – thermal power required to raise the temperature of the expanded combusted gas stream to selected temperature for heat exchange in the HRSG].

TABLE 10

	Calculation 2 (cont.)	Calculation 3 (cont)	Calculation 4 (cont)	Calculation 5	Calculation 6	Calculation 7	Calculation 8	Calculation 9	Calculation 10
$H_2S/CH_4/CO_2$ content of fuel stream and supplemental fuel stream (vol.%)	100/0/0	50/50/0	33/33/34	100/0/0	100/0/0	50/50/0	50/50/0	33/33/34	33/33/34
Max. Temp. of 2d combusted gas stream (ºC)	604	604	604	800	1000	800	1000	800	1000
Δ T in heat exchange of 2d combusted gas stream (ºC)	484	484	484	680	880	680	880	680	880
Mass Flow Rate SO_2 in 2d combusted gas stream (kg/h)	447857	198551	190315	447857	447857	198551	198551	190315	190315
SO_2 Heat Capacity (kJ/kg)	0.64	0.64	0.64	0.64	0.64	0.64	0.64	0.64	0.64
SO_2 Thermal Power Contribution in heat exchange of 2d combusted gas stream (MW_t)	38.5	17.1	16.4	54.1	70.1	24.0	31.1	23.0	29.8
Mass Flow Rate Water Vapor in 2d combusted gas stream (kg/h)	125960	167528	160578	125960	125960	167528	167528	160578	160578
Water Vapor Heat Capacity (kJ/kg)	1.97	1.97	1.97	1.97	1.97	1.97	1.97	1.97	1.97
Water Vapor Thermal Power Contribution in heat exchange of 2d combusted gas stream (MW_t)	33.4	44.4	42.5	46.9	60.7	62.3	80.7	59.8	77.3
Mass Flow Rate CO_2 in 2d Combusted Gas Stream (kg/h)	871	137405	193968	871	871	137405	137405	193968	193968
CO_2 Heat Capacity (kJ/kg)	0.84	0.84	0.84	0.84	0.84	0.84	0.84	0.84	0.84
CO_2 Thermal Power Contribution in heat exchange of 2d combusted gas stream (MW_t)	0.1	15.5	21.9	0.1	0.2	21.8	28.2	30.8	39.8
Mass Flow Rate of Nitrogen in 2d Combusted Gas Stream (kg/h)	1095604	1133350	1086335	1095604	1095604	1133350	1133350	1086335	1086335
Nitrogen Heat Capacity (kJ/kg)	1.04	1.04	1.04	1.04	1.04	1.04	1.04	1.04	1.04
Nitrogen Thermal Power Contribution in heat exchange of 2d combusted gas stream (MW_t)	153.2	158.5	151.9	215.2	278.6	222.7	288.1	213.4	276.2
Mass Flow Rate of Argon in heat exchange of 2d combusted gas stream (MW_t)	18572	19212	18415	18572	18572	19212	19212	18415	18415
Argon Heat Capacity (kJ/kg)	0.52	0.52	0.52	0.52	0.52	0.52	0.52	0.52	0.52
Argon Thermal Power Contribution in heat exchange of 2d combusted gas stream (MW_t)	1.3	1.3	1.3	1.8	2.4	1.9	2.4	1.8	2.4
Thermal Power Produced in Heat Exchange from components of 2d combusted gas stream (MW_t)	226.5	236.8	234.0	318.1	412.0	332.7	430.5	328.8	425.5
Enthalpy of Combustion of Supplemental Fuel Stream (MW_t)	504	632	585	504	504	632	632	585	585
Thermal Power Required to Raise 2d Combusted Gas Stream to its Max. Temp from 604ºC (MW_t)	0	0	0	(86)	(173)	(88)	(178)	(88)	(177)
Total Thermal Power available for conversion to electrical power in Second Cycle (MW_t)	730.5	868.8	819	736.1	743	876.7	884.5	825.8	833

The electrical power produced from the total thermal power in the heat recovery cycle of the combined cycle system was calculated by multiplying the total thermal power by the efficiency of the heat recovery cycle of the combined cycle system utilizing a GE S107FA steam turbine. The efficiency of the heat recovery cycle utilizing a GE S107FA at a temperature differential of 484°C was calculated above in Example 2. For the examples in which the maximum second combusted gas temperature is 800°C or 1000°C and the temperature differential between the maximum second combusted gas temperature and the flue gas temperature was 680°C or 880°C, the net efficiency of the heat recovery cycle was assumed to increase 1% for every 20°C difference between the maximum second combusted gas temperature and the temperature of the expanded combusted gas stream exiting the gas turbine (604°C), where the basis for this assumption is the National Coal Council Issue Paper 2009, Higher Efficiency Power Generation Reduces Emissions, J. M. Beér, p.2. The heat recovery efficiency for the selected fuel streams and supplemental fuel streams at the selected maximum second combustion gas stream temperatures, the calculated electrical power produced from the thermal power in the heat recovery cycle, and the total electrical power of the combined cycle system for the selected fuel streams is shown in Table 11.

TABLE 11

	Calculation 2 (continued)	**Calculation 3 (continued)**	**Calculation 4 (continued)**	**Calculation 5**	**Calculation 6**	**Calculation 7**	**Calculation 8**	**Calculation 9**	**Calculation 10**
$H_2S/CH_4/CO_2$ content of fuel stream and supplemental fuel stream (vol.%)	100/0/0	50/50/0	33/33/34	100/0/0	100/0/0	50/50/0	50/50/0	33/33/34	33/33/34
Total Thermal Power available for conversion to electrical power in the Heat Recovery Cycle (MW_t)	730.5	868.8	819	736.1	743	876.7	884.5	825.8	833
Efficiency of conversion of Thermal Power to Electrical Power in the Heat Recovery Cycle	0.387	0.387	0.387	0.487	0.587	0.487	0.587	0.487	0.587
Electrical Power from Heat Cycle (MW_e)	282.7	336.2	316.9	358.4	436.1	426.9	519.2	402.2	489.0
Electrical Power from Gas Turbine Cycle (MW_e)	183	183	183	183	183	183	183	183	183
Total Electrical Power of Combined Cycle System (MW_e)	465.7	519.2	499.9	541.4	619.1	609.9	702.2	585.2	672.0

The calculations demonstrated that a fuel stream consisting of H_2S, or containing 50 vol.% H_2S and 50 vol.% CH_4, or containing 33 vol.% H_2S, 33 vol.% CH_4, and 34 vol.% CO_2 can be combusted in a combined cycle gas turbine/HRSG with supplemental firing to produce substantial electrical power.

EXAMPLE 4

The electrical power produced by combusting selected fuel streams in a combined cycle gas-turbine/heat recovery steam generator system with supplemental firing and capturing electrical power from cooling the resulting flue gas to 3°C was calculated. The selected fuel streams were the fuel streams utilized in Example 3. The total electrical power produced when cooling the flue gas stream to 3°C is the electrical power calculated in Example 3 plus the additional electrical power produced from the thermal power available from the flue gas stream upon cooling the flue gas stream from 120°C to 3°C.

The calculations were performed based on using ammonia as a heat transfer fluid, where liquid ammonia at a temperature of about 0°C is contacted in an evaporator with the flue gas stream exiting the HRSG to cool the flue gas stream from 120°C to 3°C and to transfer thermal power from the flue gas stream to the ammonia, producing an evaporated ammonia heat transfer gas. The evaporated ammonia heat transfer gas is expanded in an isentropic expander to convert the thermal power contained therein to electrical power. The amount of thermal power available for conversion to electrical power produced by cooling the flue gas stream from 120°C to 3° was calculated according to formula (21):

(21)

$$\mathit{thermal\ power\ of\ flue\ gas\ stream\ (120^{\circ}C \rightarrow 3^{\circ}C)} =$$

$$\sum\{[(\text{Heat Capacity of Non} - \text{Water Component X (kJ/kg)} \ast \text{Mass Flow Rate of Non} -$$
$$\text{Water Component X in flue gas stream(kg/h)} \ast \Delta\text{T1}) \ast (\text{Heat Capacity of Water Vapor(kJ/kg)} \ast$$
$$\text{Mass Flow Rate of Water Vapor(kg/h)} \ast \Delta\text{T2(}^{\circ}\text{C)}\,) + (\text{Heat Capacity of Liquid Water(kJ/kg)} \ast$$
$$\text{Mass Flow Rate of Liquid Water(kg/h)} \ast \Delta\text{T3(}^{\circ}\text{C)}\,) + (\text{Mass Flow Rate of Water Vapor(kg/h)} \ast$$
$$\text{Latent Heat of Condensation of Water } (\tfrac{\text{kJ}}{\text{kg}}))]/(3600000\ ((\text{MWt/kJ})/\text{kg})\,)\}$$

where X ranges from 1 to n, where n is the total number of non-water components in the flue gas stream, ΔT1 is the difference between the temperature of the flue gas stream exiting the HRSG (120°C) and the temperature of the flue gas stream exiting the expander (3°C), ΔT2 is the difference between the temperature of the flue gas stream exiting the HRSG (120°C) and the temperature at which water condenses (100°C), ΔT3 is the difference between the temperature at which water condenses (100°C) and the temperature of the flue gas stream exiting the expander (3°C), and the latent heat of condensation of water is 2260 kJ/kg. The gaseous components of the flue gas stream, their heat capacities, and mass flow rates of these components are the same as the components of the second combusted gas stream, their heat capacities, and the mass flow rate of these components in the second combusted gas stream as set forth in Table 10. Water is converted to liquid upon cooling the flue gas stream from 120°C to 3°C: the mass flow rate of the liquid water is equivalent to the mass flow rate of water vapor in the flue gas stream but the heat capacity of liquid water is 4.18 kJ/kg. The electrical power produced by expanding the ammonia heat transfer gas in an isentropic expander was calculated by multiplying the calculated thermal power produced by cooling the flue gas stream from 120°C to 3°C by the isentropic turbine efficiency for producing electrical power from an ammonia heat transfer gas, where the isentropic turbine efficiency was assumed to be 0.3. The calculated thermal power produced by cooling the flue gas stream; the calculated electrical power produced from the thermal power produced by cooling the flue gas stream; and the total electrical power of the combined cycle system including electrical power produced by cooling the flue gas stream for the selected fuel streams is shown in Table 12.

TABLE 12

	Calculation 2 (continued)	**Calculation 3 (continued)**	**Calculation 4 (continued)**	**Calculation 5 (continued)**	**Calculation 6 (continued)**	**Calculation 7 (continued)**	**Calculation 8 (continued)**	**Calculation 9 (continued)**	**Calculation 10 (continued)**
$H_2S/CH_4/CO_2$ content of fuel stream and supplemental fuel stream (vol.%)	100/0/0	50/50/0	33/33/34	100/0/0	100/0/0	50/50/0	50/50/0	33/33/34	33/33/34
Total thermal power available for conversion to electrical power from cooling the flue gas stream from 120ºC to 3ºC (MW_t)	141	172	167	141	141	172	172	167	167
Assumed efficiency of conversion of thermal power to electrical power in the ammonia heat transfer gas isentropic expander	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3
Electrical Power from Cooling the Flue Gas Stream (MW_e)	42	52	50	42	42	52	52	50	50
Total Electrical Power of Combined Cycle System with Flue Gas Cooling (MW_e)	508	571	550	583	661	662	754	635	722

The calculations set forth in Table 12 demonstrated that considerable more electrical power may be generated by cooling the flue gas stream to capture the thermal power therein utilizing a non-aqueous heat transfer gas having a boiling point at least 50°C less than water and converting the thermal power from the flue gas stream to electrical power.

The present invention is well adapted to attain the ends and advantages mentioned as well as those that are inherent therein. The particular embodiments disclosed above are illustrative only, as the present invention may be modified and practiced in different but equivalent manners apparent to those skilled in the art having the benefit of the teachings herein. Furthermore, no limitations are intended to the details of construction or design herein shown, other than as described in the claims below. While compositions and methods are described in terms of “comprising,” “containing,” or “including” various components or steps, the compositions and methods can also “consist essentially of” or “consist of” the various components and steps. All numbers and ranges disclosed above may vary by some amount. Whenever a numerical range with a lower limit and an upper limit is disclosed, any number and any included range falling within the range is specifically disclosed. In particular, every range of values (of the form, “from about a to about b,” or, equivalently, “from approximately a to b,” or, equivalently, “from approximately a-b”) disclosed herein is to be understood to set forth every number and range encompassed within the broader range of values. Also, the terms in the claims have their plain, ordinary meaning unless otherwise explicitly and clearly defined by the patentee. Moreover, the indefinite articles “a” or “an”, as used in the claims, are defined herein to mean one or more than one of the element that it introduces.

CLAIMS

1. A method of producing power comprising combusting an oxidant stream and at least a portion of a fuel stream comprising one or more feed streams to generate a combusted gas stream containing thermal power, wherein the fuel stream has a hydrogen sulfide content of at least 30 mol%, at least 90 mass% of the components of the fuel stream are gaseous, the oxidant stream contains molecular oxygen, and the molar ratio of molecular oxygen to hydrogen sulfide is at least 1:1; and

generating electrical power from the thermal power of the combusted gas stream.

2. The method of claim 1 wherein the molar ratio of molecular oxygen to hydrogen sulfide is from 1.3:1 to 1.7:1.

3. The method of claim 1 wherein electrical power is generated from the thermal power of the combusted gas stream by exchanging sufficient heat between the combusted gas stream and a liquid aqueous stream to convert the liquid aqueous stream to steam and to cool the combusted gas stream, where the cooled combusted gas stream is a flue gas stream comprising sulfur dioxide; expanding the steam through an expander to generate mechanical power and an expanded steam stream; and, converting the mechanical power to electrical power.

4. The method of claim 3, further comprising the steps of exchanging heat between the expanded steam stream and a fluid having a boiling point at least 50°C lower than the liquid of the liquid aqueous stream at 0.101 MPa and having a latent heat of vaporization of at least 350 kJ/kg to condense water from the expanded steam stream and to form a heat transfer gas from the fluid, where the heat transfer gas contains thermal power transferred from the expanded steam stream by heat exchange with the fluid;

expanding the heat transfer gas through an expander to generate mechanical power and to form the fluid; and

generating electrical power from the mechanical power generated by expanding the heat transfer gas.

5. The method of claim 4 wherein a portion of the thermal power transferred from the expanded steam stream to the fluid is derived from the latent heat of condensation of water in the expanded steam stream.

6. The method of claim 4 wherein the fluid is selected from the group consisting of anhydrous ammonia, aqueous ammonia, anhydrous sulfur dioxide, carbon dioxide, and diethyl ether.

7. The method of claim 4 wherein the water condensed from the expanded steam stream has a temperature of at most 85°C and is provided as the liquid aqueous stream to cool the combusted gas stream.

8. The method of claim 3, wherein the flue gas stream further comprises steam and has a temperature of from greater than 100°C to 150°C, further comprising the steps of exchanging heat between the flue gas stream and a fluid having a boiling point at least 50°C lower than the liquid of the liquid aqueous stream at 0.101 MPa and having a latent heat of vaporization of at least 350 kJ/kg to cool the flue gas stream to a temperature of from greater than 0°C to 50°C and condense and separate water therefrom and to form a heat transfer gas from the fluid, where the heat transfer gas contains thermal power transferred from the flue gas stream to the fluid by heat exchange with the fluid;
expanding the heat transfer gas through an expander to generate mechanical power and to form the fluid; and
generating electrical power from the mechanical power generated by expanding the heat transfer gas.

9. The method of claim 8, wherein the fluid formed by expanding the heat transfer gas through an expander has a temperature of from -25°C to -100°C, further comprising the steps of exchanging heat between the fluid having a temperature of from -25°C to -100°C and a scrubbing solvent in which sulfur dioxide, carbon dioxide, or both may be dissolved to chill the scrubbing solvent to a temperature of from -100°C to 0°C and to raise the temperature of the fluid to at least 0°C without converting the fluid to a gas; and

contacting the flue gas stream from which water has been separated with the chilled scrubbing solvent to separate sulfur dioxide, carbon dioxide, or both from the flue gas stream into the scrubbing solvent.

10. The method of claim 9 further comprising the steps of heating the chilled scrubbing solvent to a temperature at which sulfur dioxide, carbon dioxide, or both separate from the scrubbing solvent, and separating sulfur dioxide, carbon dioxide, or both from the scrubbing solvent.

11. The method of claim 10 further comprising the step of exchanging heat between the scrubbing solvent from which sulfur dioxide, carbon dioxide, or both have been separated and the fluid having a temperature of from -25°C to -100°C to chill the scrubbing solvent to temperature of from -100°C to 0°C and to heat the fluid to a temperature of at least 0°C without converting the fluid to a gas.

12. The method of claim 1 wherein electrical power is generated from at least a portion of the thermal power of the combusted gas stream by expanding the combusted gas stream through an expander to generate mechanical power and an expanded combusted gas stream, and converting the mechanical power to electrical power.

13. The method of claim 12 wherein the molar ratio of molecular oxygen in the oxidant stream to hydrogen sulfide in the fuel stream is at least 2:1.

14. The method of claim 12 wherein electrical power is generated by exchanging heat between the expanded combusted gas stream and a liquid aqueous stream to convert the liquid aqueous stream to steam and to cool the expanded combusted gas stream, where the cooled expanded combusted gas stream is a flue gas stream comprising sulfur dioxide; expanding the steam through an expander to generate mechanical power and an expanded steam stream; and, converting the mechanical power to electrical power.

15. The method of claim 14, further comprising the steps of exchanging heat between the expanded steam stream and a fluid having a boiling point at least 50°C lower than the liquid of

the liquid aqueous stream at 0.101 MPa and having a latent heat of vaporization of at least 350 kJ/kg to condense water from the expanded steam stream and to form a heat transfer gas from the fluid, where the heat transfer gas contains thermal power transferred from the expanded steam stream by heat exchange with the fluid;
expanding the heat transfer gas through an expander to generate mechanical power and to form the fluid; and
generating electrical power from the mechanical power generated by expanding the heat transfer gas.

16. The method of claim 15 wherein the fluid is selected from the group consisting of anhydrous ammonia, aqueous ammonia, anhydrous sulfur dioxide, carbon dioxide, and diethyl ether.

17. The method of claim 15 wherein a portion of the thermal power transferred from the expanded steam stream to the fluid is derived from the latent heat of condensation of water in the expanded steam stream.

18. The method of claim 15 wherein the water condensed from the expanded steam stream has a temperature of at most 85ºC and is provided as the liquid aqueous stream to cool the expanded combusted gas stream.

19. The method of claim 14, wherein the flue gas stream further comprises steam and has a temperature of from greater than 100ºC to 150ºC, further comprising the steps of exchanging heat between the flue gas stream and a fluid having a boiling point at least 50ºC lower than the liquid of the liquid aqueous stream at 0.101 MPa and having a latent heat of vaporization of at least 350 kJ/kg to cool the flue gas stream to a temperature of from greater than 0ºC to 50ºC and condense and separate water therefrom and to form a heat transfer gas from the fluid, where the heat transfer gas contains thermal power transferred from the flue gas stream to the fluid by heat exchange with the fluid;
expanding the heat transfer gas through an expander to generate mechanical power and to form the fluid; and

generating electrical power from the mechanical power generated by expanding the heat transfer gas.

20. The method of claim 19, wherein the fluid formed by expanding the heat transfer gas through an expander has a temperature of from -25°C to -100°C, further comprising the steps of exchanging heat between the fluid having a temperature of from -25°C to -100°C and a scrubbing solvent in which sulfur dioxide, carbon dioxide, or both may be dissolved to chill the scrubbing solvent to a temperature of from -100°C to 0°C and to raise the temperature of the fluid to at least 0°C without converting the fluid to a gas; and
contacting the flue gas stream from which water has been separated with the chilled scrubbing solvent to separate sulfur dioxide, carbon dioxide, or both from the flue gas stream into the scrubbing solvent.

21. The method of claim 20 further comprising the steps of heating the chilled scrubbing solvent to a temperature at which sulfur dioxide, carbon dioxide, or both separate from the scrubbing solvent and separating sulfur dioxide, carbon dioxide, or both from the scrubbing solvent.

22. The method of claim 21 further comprising the step of exchanging heat between the scrubbing solvent from which sulfur dioxide, carbon dioxide, or both have been separated and the fluid having a temperature of from -25°C to -100°C to chill the scrubbing solvent to a temperature of from -100°C to 0°C and to heat the fluid to a temperature of at least 0°C without converting the fluid to a gas.

23. The method of claim 12 wherein the expanded combusted gas stream contains at least 5 vol.%, or at least 10 vol.% molecular oxygen, further comprising the steps of combusting a sulfur-containing fuel and the expanded combusted gas stream to generate a second combusted gas stream containing thermal power and generating electrical power from the thermal power of the second combusted gas stream.

24. The method of claim 23 further comprising the step of providing the sulfur-containing fuel for combustion in an amount effective to consume at least 99 mol% of the molecular oxygen in the expanded combusted gas stream.

25. The method of claim 23 wherein a portion of the fuel stream is provided as the sulfur-containing fuel.

26. The method of claim 23 wherein electrical power is generated by exchanging sufficient heat between the second combusted gas stream and a liquid aqueous stream to convert the liquid aqueous stream to steam and to cool the second combusted gas stream, where the cooled second combusted gas stream is a flue gas stream comprising sulfur dioxide; expanding the steam through an expander to generate mechanical power and an expanded steam stream; and converting the mechanical power to electrical power.

27. The method of claim 26, further comprising the steps of exchanging heat between the expanded steam stream and a fluid having a boiling point at least 50°C lower than the liquid of the liquid aqueous stream at 0.101 MPa and having a latent heat of vaporization of at least 350 kJ/kg to condense water from the expanded steam stream and to form a heat transfer gas from the fluid, where the heat transfer gas contains thermal power transferred from the expanded steam stream by heat exchange with the fluid;
expanding the heat transfer gas through an expander to generate mechanical power and to form the fluid; and
generating electrical power from the mechanical power generated by expanding the heat transfer gas.

28. The method of claim 27 wherein the fluid is selected from the group consisting of anhydrous ammonia, aqueous ammonia, anhydrous sulfur dioxide, carbon dioxide, and diethyl ether.

29. The method of claim 27 wherein a portion of the thermal power transferred from the expanded steam stream to the fluid is derived from the latent heat of condensation of water in the expanded steam stream.

30. The method of claim 27 wherein the water condensed from the expanded steam stream has a temperature of at most 85°C and is provided as the liquid aqueous stream to cool the second combusted gas stream.

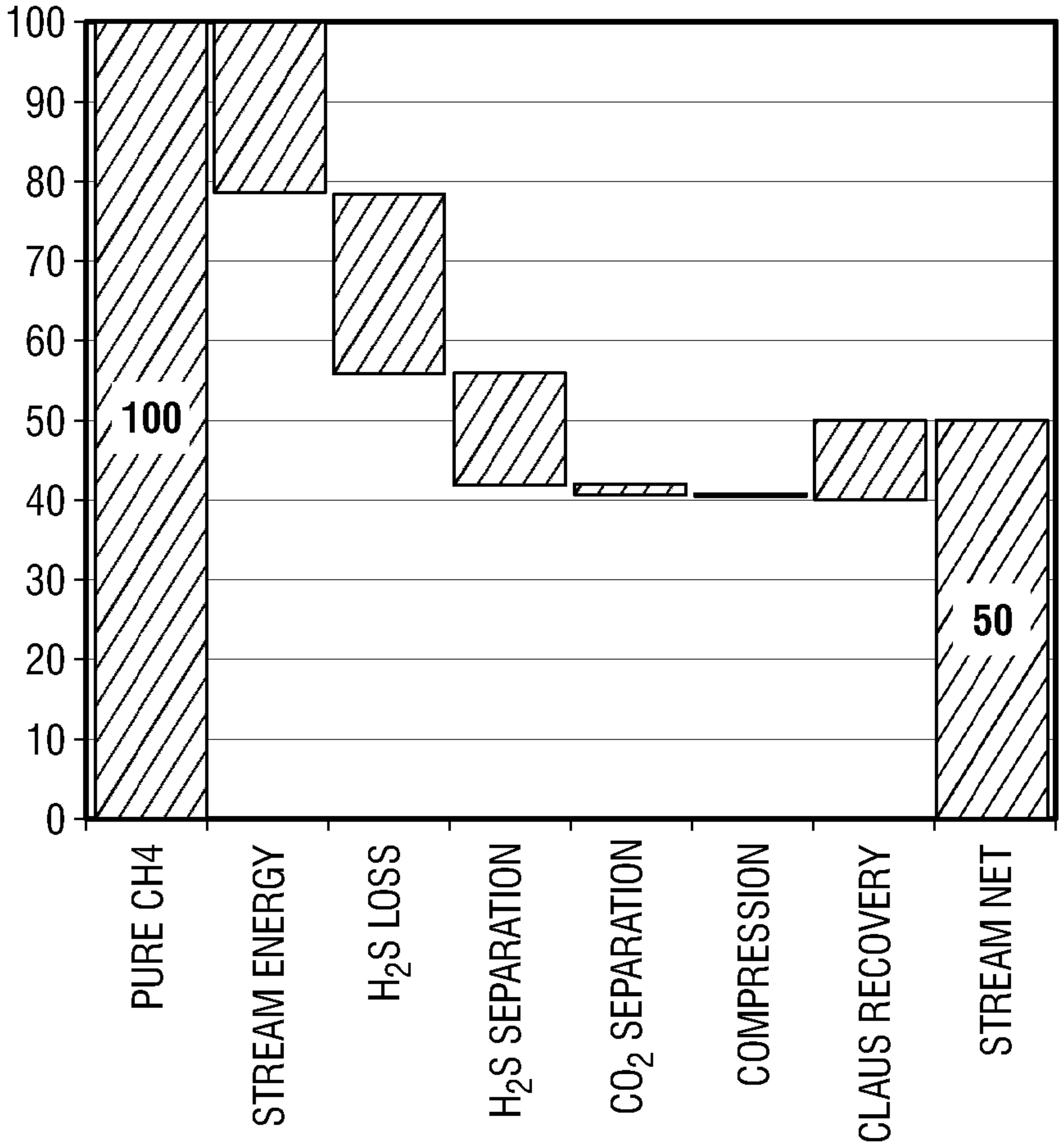
31. The method of claim 26, wherein the flue gas stream further comprises steam and has a temperature of from greater than 100°C to 150°C, further comprising the steps of exchanging heat between the flue gas stream and a fluid having a boiling point at least 50°C lower than the liquid of the liquid aqueous stream at 0.101 MPa and having a latent heat of vaporization of at least 350 kJ/kg to cool the flue gas stream to a temperature of from greater than 0°C to 50°C and condense and separate water therefrom and to form a heat transfer gas from the fluid, where the heat transfer gas contains thermal power transferred from the flue gas stream to the fluid by heat exchange with the fluid;
expanding the heat transfer gas through an expander to generate mechanical power and to form the fluid; and
generating electrical power from the mechanical power generated by expanding the heat transfer gas.

32. The method of claim 31 wherein the fluid is selected from the group consisting of anhydrous ammonia, aqueous ammonia, anhydrous sulfur dioxide, carbon dioxide, and diethyl ether.

33. The method of claim 31, wherein the fluid formed by expanding the heat transfer gas through an expander has a temperature of from -25°C to -100°C, further comprising the steps of exchanging heat between the fluid having a temperature of from -25°C to -100°C and a scrubbing solvent in which sulfur dioxide, carbon dioxide, or both may be dissolved to chill the scrubbing solvent to a temperature of from -100°C to 0°C and to raise the temperature of the fluid to at least 0°C without converting the fluid to a gas; and
contacting the flue gas stream from which water has been separated with the chilled scrubbing solvent to separate sulfur dioxide, carbon dioxide, or both from the flue gas stream into the scrubbing solvent.

34. The method of claim 33 further comprising the steps of heating the chilled scrubbing solvent to a temperature at which sulfur dioxide, carbon dioxide, or both separate from the scrubbing solvent, and, separating sulfur dioxide, carbon dioxide, or both from the scrubbing solvent.

35. The method of claim 34 further comprising the step of exchanging heat between the scrubbing solvent from which sulfur dioxide or carbon dioxide has been separated and the fluid having a temperature of from -25°C to -100°C to chill the scrubbing solvent to a temperature of -100°C to 0°C and to heat the fluid to a temperature of at least 0°C without converting the fluid to a gas.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CO_2 SEPARATION
COMPRESSION
CLAUS RECOVERY
STREAM NET

FIG. 1

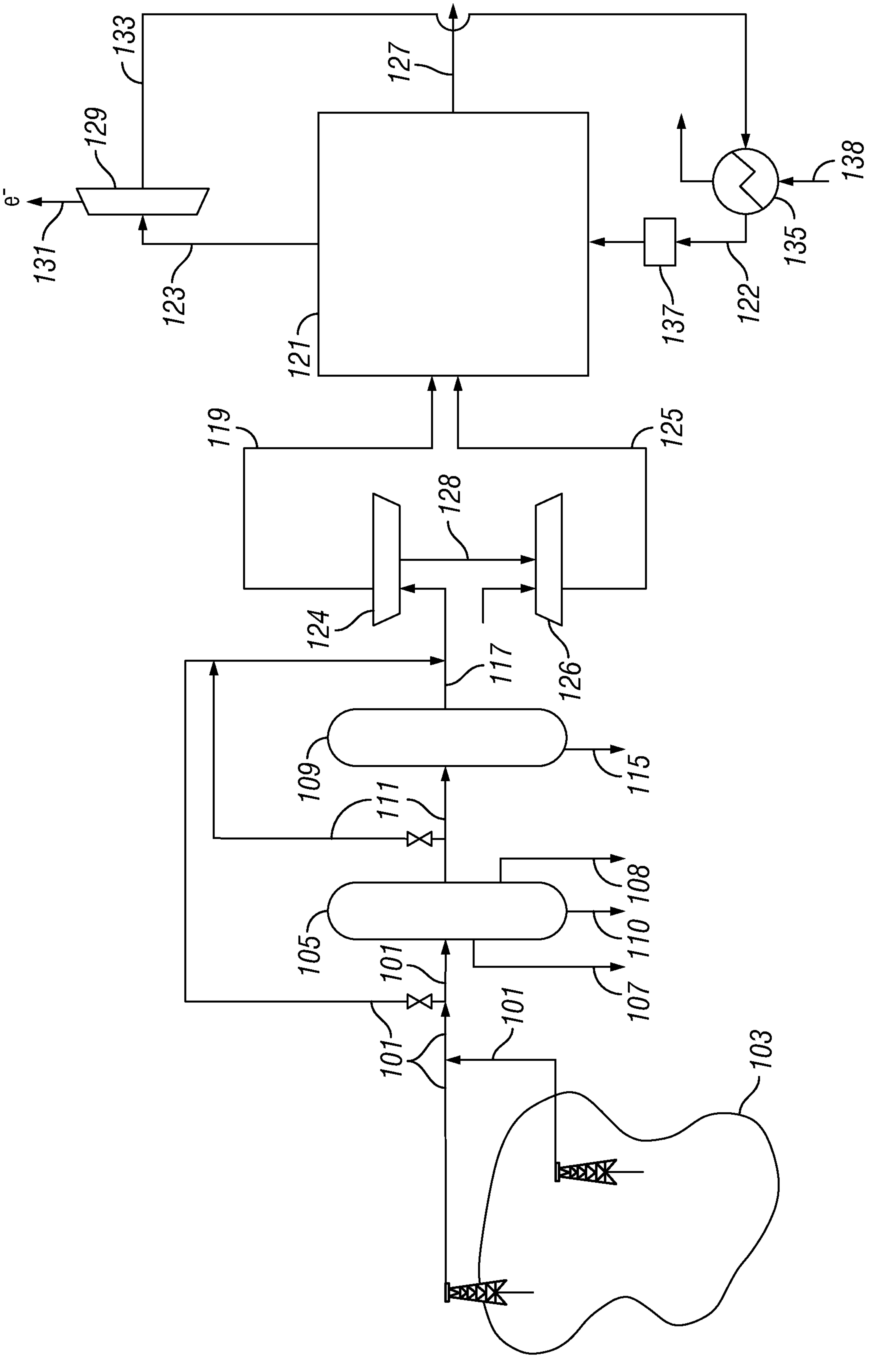
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FIG. 2

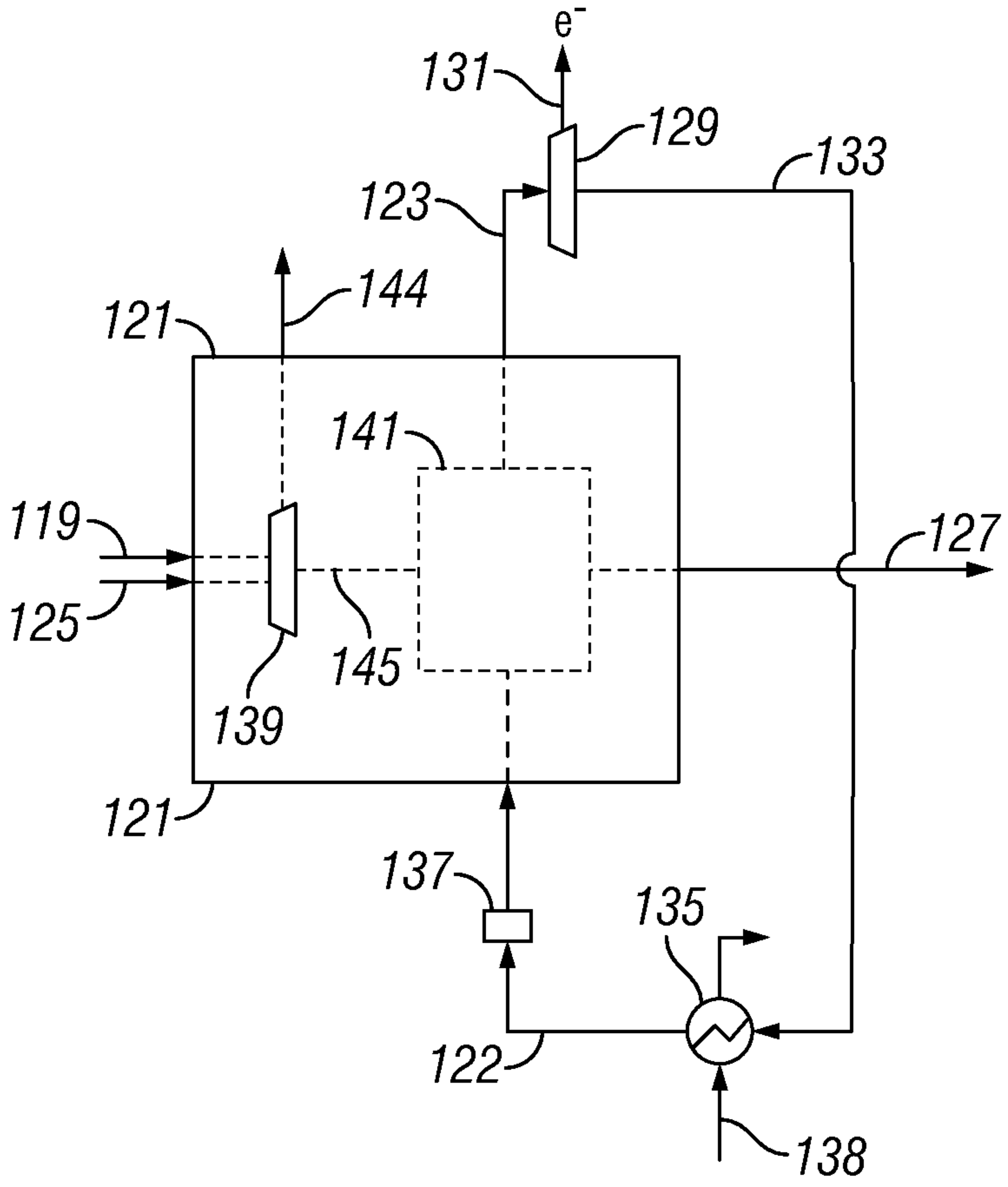
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FIG. 3

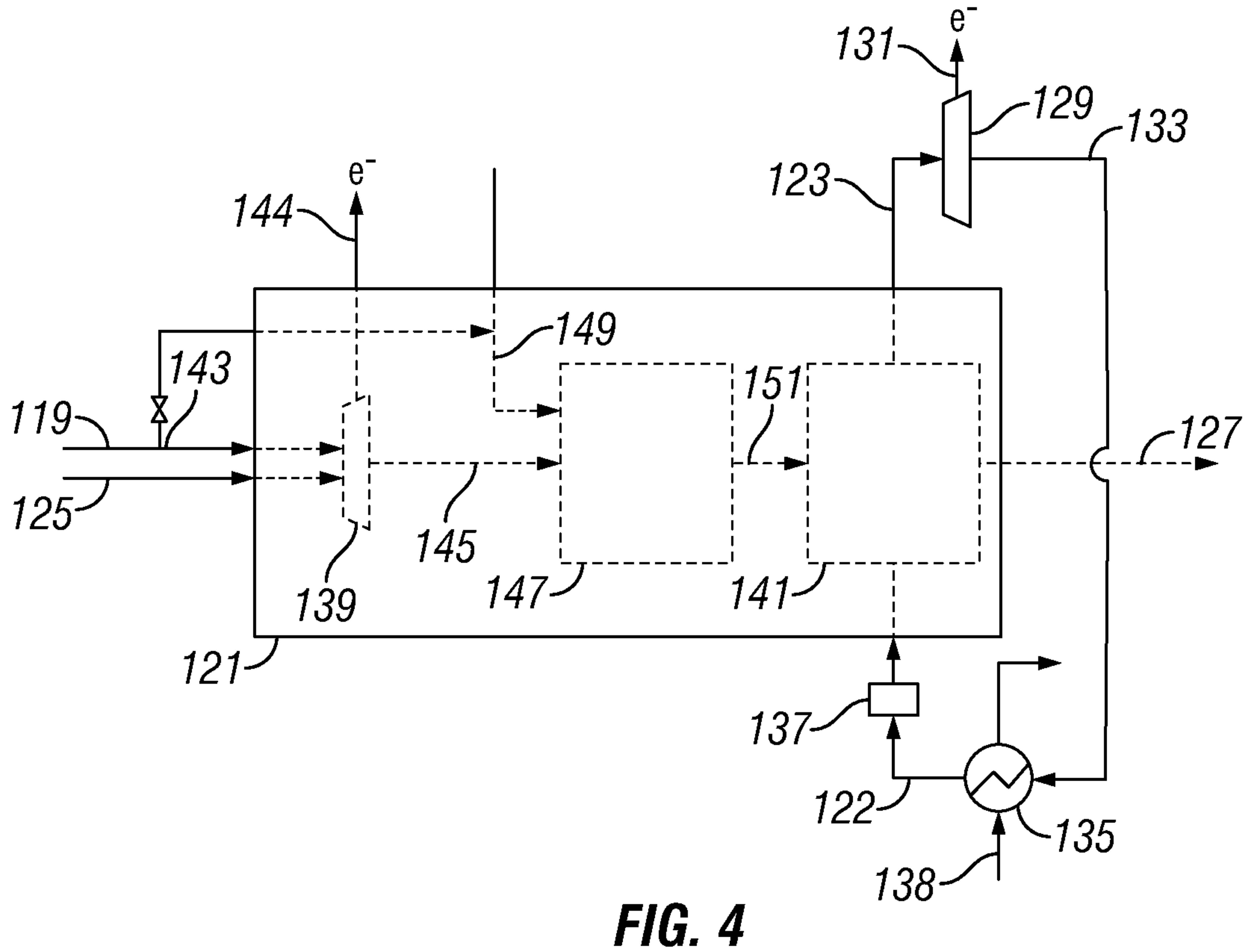
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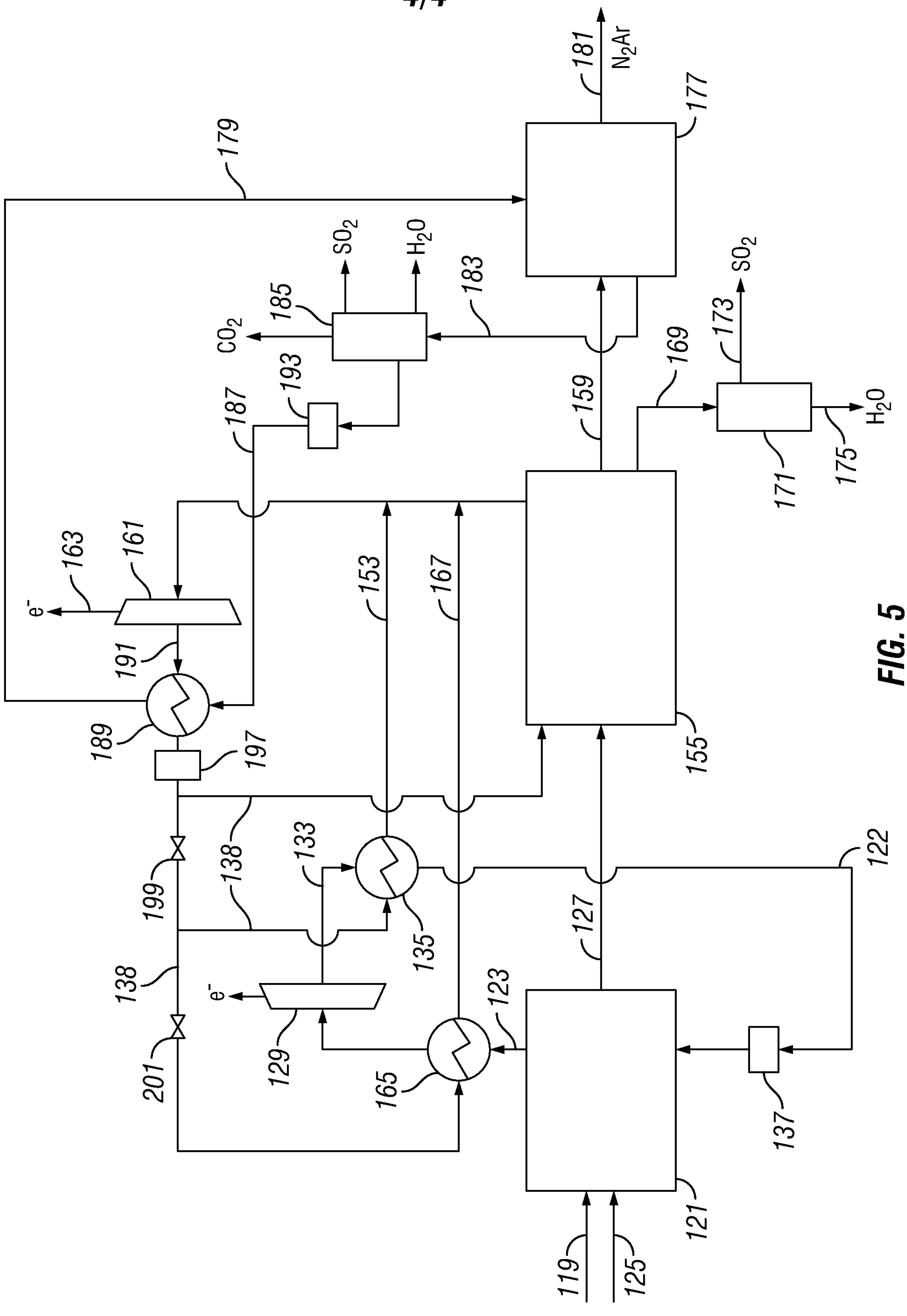
FIG. 4

4/4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163
161
e^-
153
167
191
189
197
155
138
133
199
135
127
122
138
e^-
123
129
201
165
137
121
119
125

FIG. 5

e-
163
189
191
161
138
201
199
197
179
e-
138
187
CO2
129
133
193
185
SO2
165
153
135
H2O
167
183
123
119
127
159
181
125
N2Ar
121
155
169
177
173
SO2
137
171
175
H2O
122

FIG. 5