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Hybrid Process Using a Membrane to Enrich Flue Gas CO₂ with a Solvent-Based Post-Combustion CO₂ Capture System

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(54) **HYBRID PROCESS USING A MEMBRANE TO ENRICH FLUE GAS CO₂ WITH A SOLVENT-BASED POST-COMBUSTION CO₂ CAPTURE SYSTEM**

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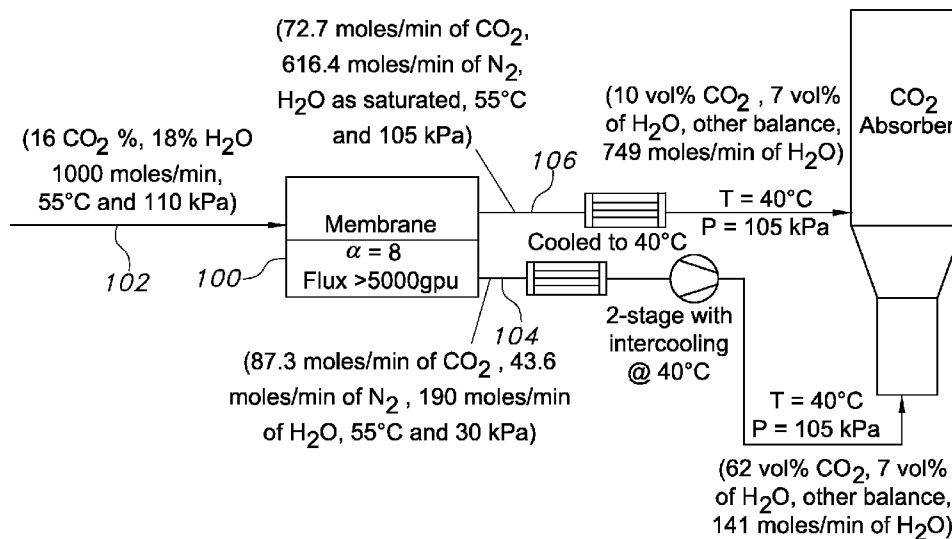
Primary Examiner — Jason M Greene

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(57) **ABSTRACT**

A process for recovery of CO₂ from a post-combustion gas includes pre-concentrating a CO₂ component of the post-combustion flue gas by passing the post-combustion gas through a CO₂-selective membrane module to provide a CO₂-enriched permeate stream and a CO₂-lean reject stream. Next, in a CO₂ absorber, both the CO₂-enriched permeate stream and CO₂ lean reject stream, fed to separate feed locations on the CO₂ absorber, are contacted with a scrubbing solvent to absorb CO₂ and provide a carbon-rich scrubbing solvent. Finally, absorbed CO₂ is stripped from the carbon-rich scrubbing solvent by a two-stage CO₂ stripping system. The CO₂-selective membrane may be a high flux, low pressure drop, low CO₂ selectivity membrane. The two stage stripping system includes a primary CO₂ stripping column for stripping CO₂ from the carbon-rich scrubbing solvent exiting the CO₂ absorber, and a secondary CO₂ stripping column for stripping CO₂ from a carbon-lean scrubbing solvent exiting the primary CO₂ stripping column. Apparatus for CO₂ removal from post-combustion gases in a pulverized coal power plant incorporating the described processes are described.

21 Claims, 5 Drawing Sheets



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10/06 (2013.01); **Y02C 10/10** (2013.01); **Y02E**
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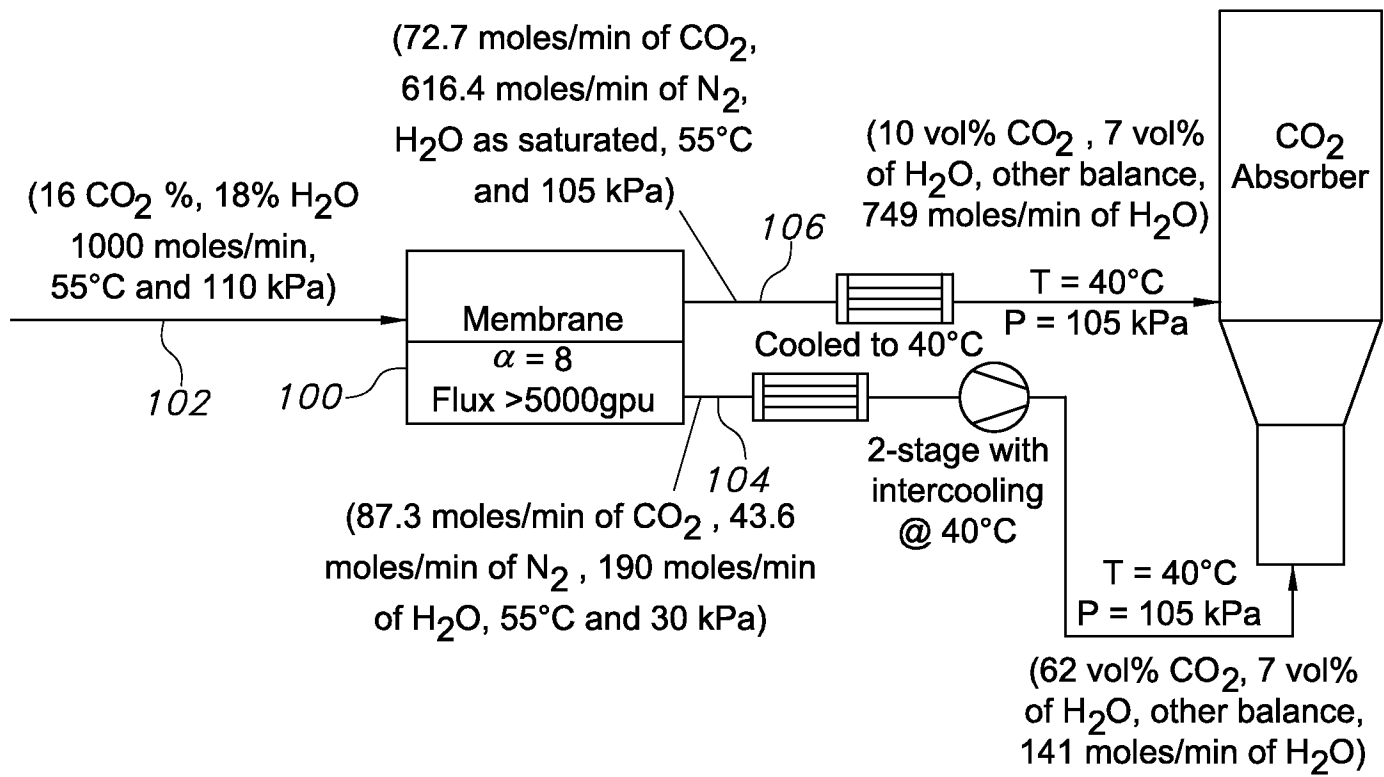


FIG. 1

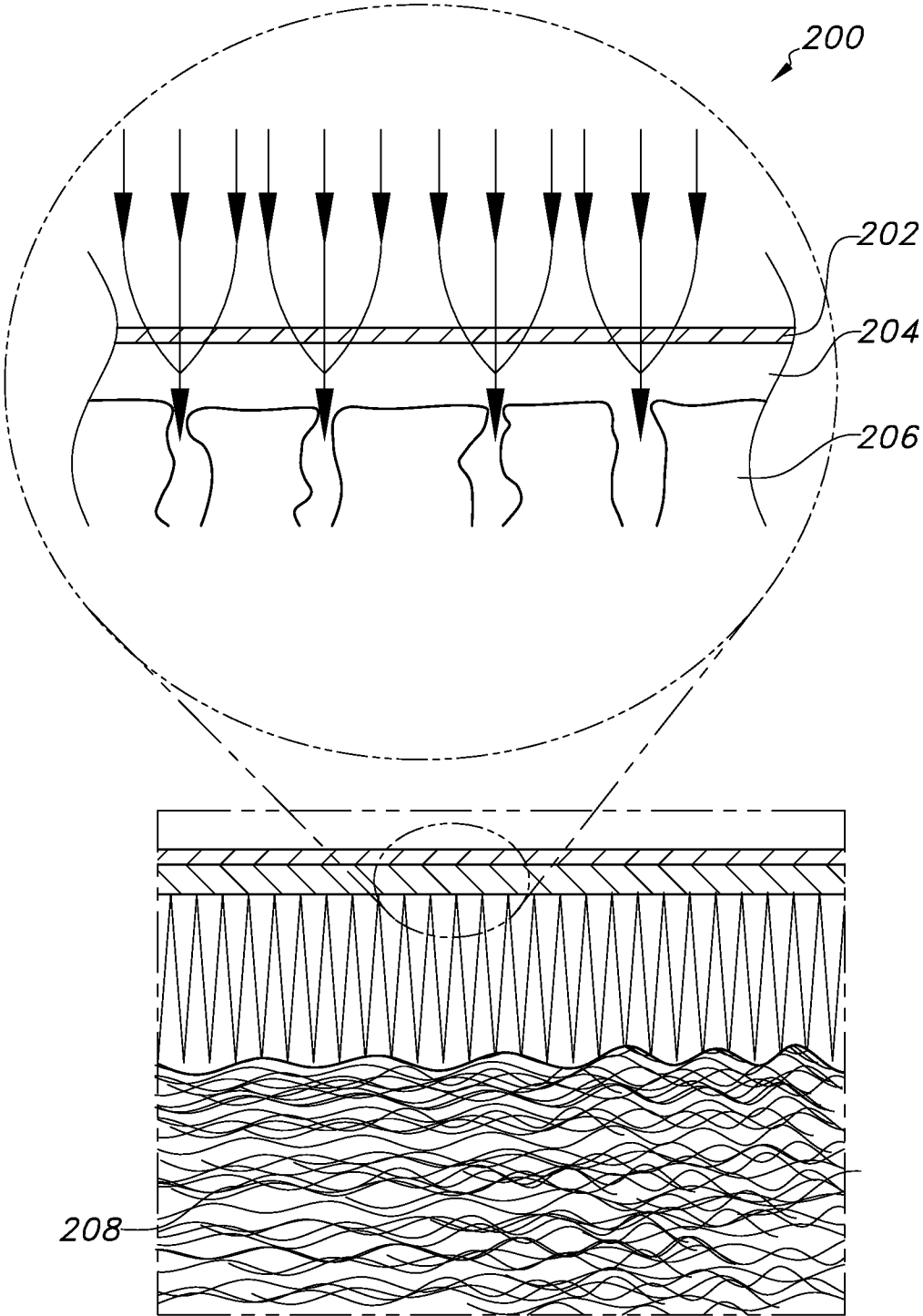


FIG. 2

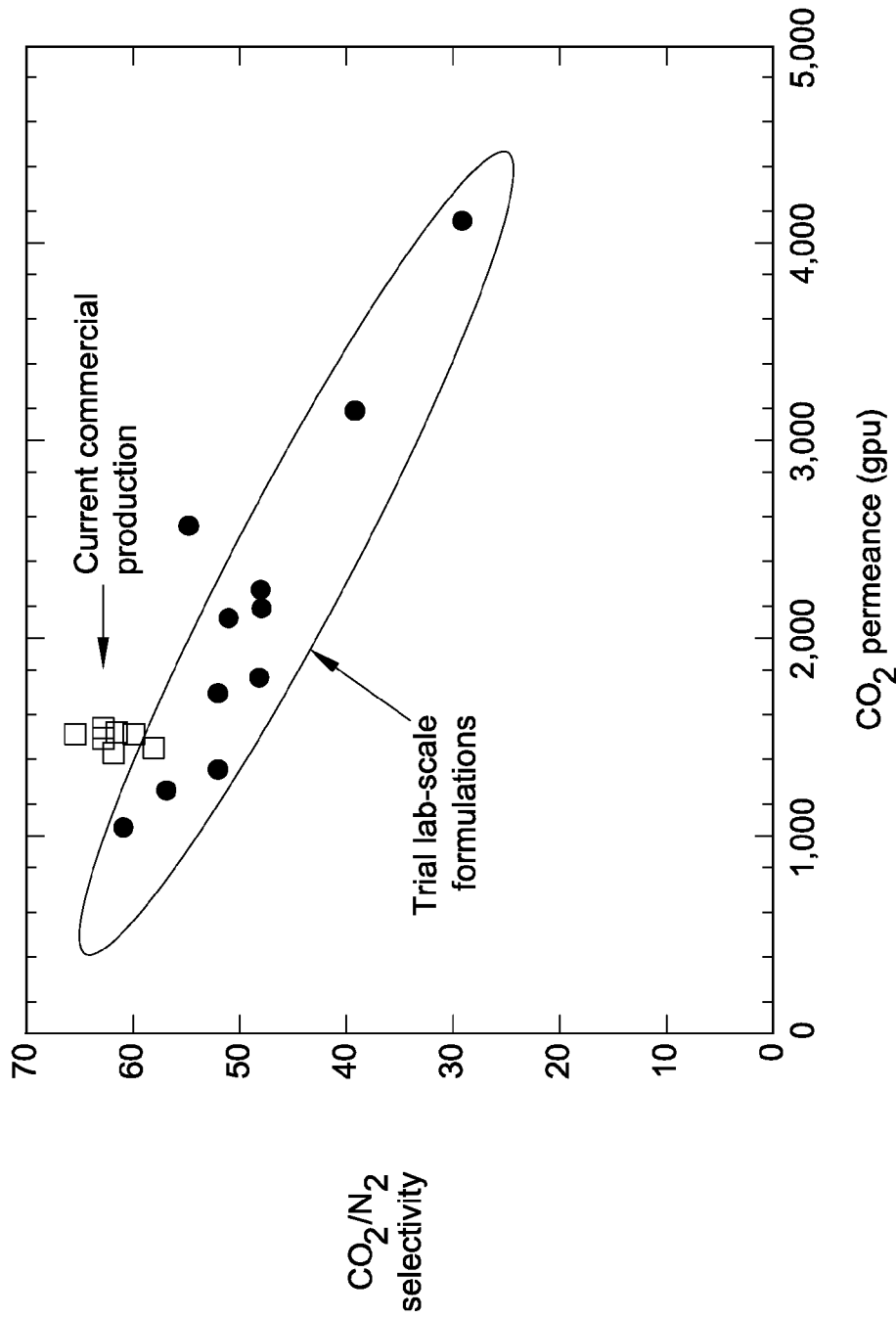


FIG. 3

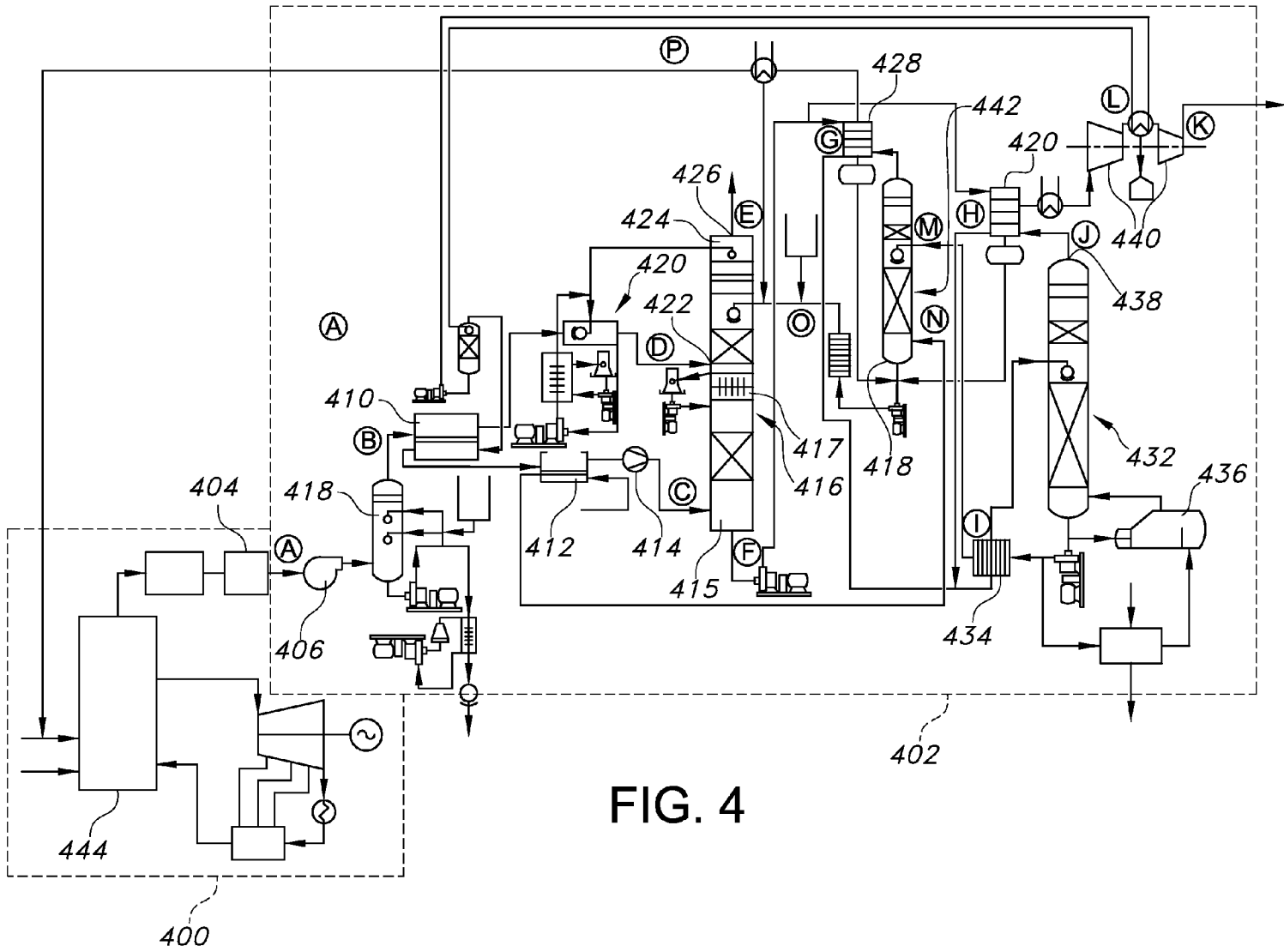


FIG. 4

The ambient conditions: Dry bulb Temperature = 35°C
and relative humidity, H = 60%
the absorber is operating at 40°C
the solvent working capacity is 2.5 mole/kg solution
the heat of the reaction is 27 kJ/gmole
stripper exiting temperature = 76.7°C

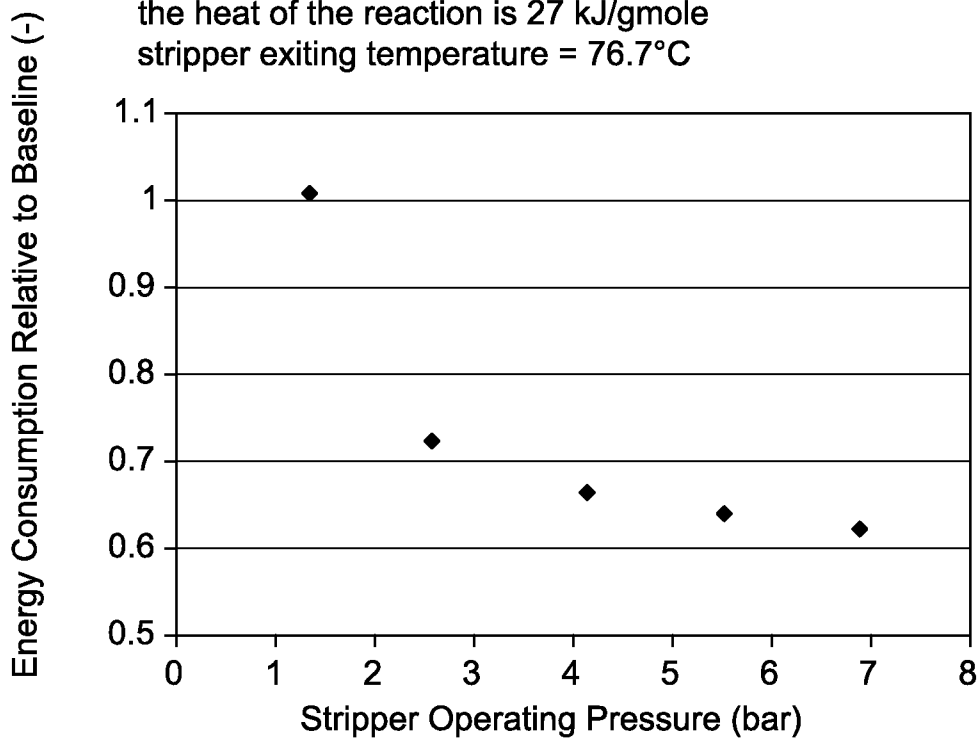


FIG. 5

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**HYBRID PROCESS USING A MEMBRANE TO
ENRICH FLUE GAS CO₂ WITH A
SOLVENT-BASED POST-COMBUSTION CO₂
CAPTURE SYSTEM**

This utility patent application claims the benefit of priority in U.S. Provisional Patent Application Ser. No. 61/924,373 filed on Jan. 7, 2014, the entirety of the disclosure of which is incorporated herein by reference.

TECHNICAL FIELD

The present disclosure relates to removal of CO₂ from post-combustion gases. In particular, the disclosure relates to a hybrid process for CO₂ removal including membrane-based pre-concentration of CO₂ in flue gas, followed by passage through a two-stage stripping unit for solvent regeneration. The described process finds utility in post-combustion carbon dioxide (CO₂) capture, such as from utility flue gases in coal-fired power plants and natural gas-fired power plant.

BACKGROUND OF THE INVENTION

In a typical coal-fired power plant, coal is burned in a boiler to make high temperature and pressure steam to drive a steam turbine and electricity generator. For the steam-side process, the initial superheated steam drives the high-pressure turbine first for power generation, before returning to the boiler for reheating to bring steam temperature back to over 540° C. (according to the steam cycle specifications and boiler design). The reheated steam then enters intermediate-pressure and low-pressure turbines to generate additional electricity. The steam pipe connecting the intermediate-pressure turbine exhaust and the low-pressure turbine inlet is called the cross-over section, where the steam for solvent regeneration in the CO₂ capture process is extracted. The exiting saturated steam from the low-pressure turbine is condensed in a water-to-steam heat exchanger (or condenser). The condensate is pressurized via feedwater pumps, heated via feedwater heaters and economizers, and fed back to the boiler to complete the steam cycle.

For the gas-side process, typically after combustion of the coal in low NOx burners the combustion gases leave the boiler and are treated by a NOx removal technology called Selective Catalytic Reduction (SCR). After the SCR treatment, the gas is further treated in a fly ash removal device, such as an electrostatic precipitator, to remove particulates. After this treatment, the gas is routed through an SO₂ removal device (SO₂ scrubber or polisher). At this point, the carbon capture process begins.

It is well-known in the art to use CO₂ absorbers/scrubbers including scrubbing solvents for capturing CO₂ from post-combustion gases, such as from utility flue gases. As an example, a conventional solvent often used is 30 weight percent monoethanolamine (MEA). Likewise, it is known to use solvents comprising piperazine, K₂CO₃/KHCO₃, NH₃, and others. In commercial use, the solvents must be regenerated, i.e. have absorbed CO₂ removed, for continued use in the systems. All of these processes incur costs in the form of at least energy costs and capital costs for equipment to accomplish them. An industry goal is to provide processes and systems for carbon capture and recovery, such as in commercial scale power plants, with as minimal an energy and capital cost as possible, to provide power at as low a cost as possible.

To address the aforementioned and other issues, and to meet the U.S. Department of Energy's goals of 90% CO₂ capture with 95% CO₂ purity at a cost of no more than \$40/

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tonne of CO₂ captured, a hybrid process is described to achieve high percent CO₂ capture from low driving force coal post-combustion flue gas, CO₂ purity for compression, and low process energy requirement. The described process incorporates membrane-based CO₂ enrichment of post-combustion gases, coupled with a heat-integrated aqueous CO₂ capture system and a two-stage solvent regeneration process to provide a desired CO₂ purity from utility flue gases.

SUMMARY OF THE INVENTION

In accordance with the purposes and advantages of the present invention as described herein, in one aspect of the present disclosure a process for removal and recovery of CO₂ from a post-combustion gas such as a boiler or a gas turbine exhaust flue gas is described. The process includes steps of pre-concentrating a CO₂ component of the boiler exhaust gas by passing the flue gas through a CO₂-selective membrane module to provide a CO₂-enriched post-combustion gas and a CO₂-lean stream gas. Next, in a CO₂ absorber, the CO₂-enriched post-combustion gas is fed to a CO₂ absorber bottom and contacted with a scrubbing solvent to absorb CO₂ and provide a carbon-rich scrubbing solvent. The CO₂-lean stream is fed to the CO₂ absorber at an intermediate height to achieve 90% overall CO₂ removal. Then, absorbed CO₂ is stripped from the carbon-rich scrubbing solvent by a two-stage CO₂ stripping system. The process may further include evaporating at least a portion of a water as carrier vapor for permeate stream before the step of pre-concentrating the post-combustion gas CO₂ component. In embodiments, a CO₂-rich permeate stream is passed from the CO₂-selective membrane module to a bottom portion of the CO₂ absorber, and a CO₂-lean reject stream is crossed from the CO₂-selective membrane module to the CO₂ absorber at an appropriate point above the CO₂-rich permeate stream feed point of the CO₂ absorber. The CO₂-membrane module may be a high flux, low pressure drop, low CO₂ selectivity membrane.

The two stage stripping system includes a primary CO₂ stripping column for stripping CO₂ from the carbon-rich scrubbing solvent exiting the CO₂ absorber and a secondary CO₂ stripping column for stripping CO₂ from a carbon-lean scrubbing solvent exiting the primary CO₂ stripping column. In an embodiment, the secondary CO₂ stripping column is an ambient pressure air-swept, packed column secondary CO₂ stripping column. The process may include a step of passing a CO₂-enriched air source exiting the secondary CO₂ stripping column to a preheater and therefrom for use as a boiler combustion air for a pulverized coal boiler of a pulverized coal plant.

In an embodiment, the CO₂ absorber includes a bottom packing portion with a first cross-sectional dimension, and a top packing portion with a second cross-sectional dimension that is different from the first cross-sectional dimension. One or more intercoolers may be disposed on the absorber. The location of the intercooler(s) may be between the bottom packing portion and the top packing portion and/or at an individual section. In another embodiment, the first cross-sectional dimension is approximately 1/3 that of the second cross-sectional dimension.

In another aspect, a CO₂ capture and recovery system for recovery of CO₂ from a post-combustion gas is described, including systems and devices for accomplishing the process as described above. Incorporation of the described CO₂ capture and recovery system into a pulverized coal-fired power plant is described.

In the following description there are shown and described several different embodiments, simply by way of illustration

of some of the modes best suited to carry out the invention. As it will be realized, the described subject matter is capable of other different embodiments and its several details are capable of modification in various, obvious aspects all without departing from the invention. Accordingly, the drawings and descriptions will be regarded as illustrative in nature and not as restrictive.

BRIEF DESCRIPTION OF THE DRAWINGS

The accompanying drawings incorporated herein and forming a part of the specification, illustrate several aspects of the present invention and together with the description serve to explain certain principles of the invention. In the drawings:

FIG. 1 schematically shows a CO₂-selective membrane module according to the present disclosure;

FIG. 2 shows an exemplary membrane for use in the module of FIG. 1;

FIG. 3 graphically depicts CO₂/N₂ selectivity of the membrane of FIG. 2;

FIG. 4 schematically depicts integration of a CO₂ stripping and recovery system according to the present disclosure into a pulverized coal-fired power plant; and

FIG. 5 graphically depicts impact of CO₂ stripper operating parameters on energy consumption for solvent regeneration.

Reference will now be made in detail to the present preferred embodiment of the invention, examples of which are illustrated in the accompanying drawings.

DETAILED DESCRIPTION OF THE INVENTION

To solve the foregoing problems, at a high level the present disclosure is directed to a process for capturing CO₂ from low driving force post-combustion flue gases including an initial step of CO₂ enrichment and a subsequent heat integrated aqueous CO₂ capture system utilizing a solvent to upgrade the CO₂ purity. With reference to FIG. 1, a high flux, low pressure drop and low selectivity membrane **100** is used to split the flue gas **102** exiting an SO₂ polisher (not shown) into two streams. Such membranes are known in the art. The first stream is a CO₂-enriched permeate stream **104** comprising approximately 14% of the total flue gas, and containing approximately 60% CO₂ after removal of water vapor at 40° C. The second stream is a CO₂-lean reject stream **106** (approximately 76% of the total flue gas flow rate) containing approximately 10% CO₂ after water vapor removal at 40° C. CO₂ permeability through the membrane **100** is enhanced by providing low-quality energy available from a CO₂ stripper overhead condenser and/or from CO₂ compression intercoolers, to produce water vapor diluting the membrane permeate stream to reduce the CO₂ partial pressure and maintain a reasonable driving force at relatively low pressure vacuum conditions.

In one embodiment (see FIG. 2), a modified membrane **200** developed by MTR was used. This membrane comprised a CO₂-selective layer **202**, a gutter layer **204**, a porous support membrane **206**, and a non-woven paper support **208**. As shown in FIG. 3, such membranes exhibit decreasing selectivity as permeance increases. In laboratory trials evaluating pure CO₂ and N₂ at ambient temperature, the highest permeance sample had a CO₂ permeance of 4100 gpu and a CO₂/N₂ selectivity of 29.

As briefly described above, an effective method for removing CO₂ from post-combustion gases is use of a scrubbing solvent to absorb acidic CO₂. Exemplary solvents for use in this process include primary, secondary, and tertiary amines,

although many others are known including piperazine and the stripper disclosed in the present assignee's own U.S. Published Patent Appl. No. 2012/0082604 to Liu et al. for Contaminant-tolerant Solvent and Stripping Chemical and Process for Using Same for Carbon Capture from Combustion Gases, the entirety of the disclosure of which is incorporated herein by reference. Commonly, following removal of CO₂ from post-combustion gases, the solvent is regenerated for re-use by removing CO₂ therefrom by means of a single stage stripper.

In a next aspect of the present disclosure, a two-stage stripping unit for solvent regeneration was provided. As will be described in greater detail, the unit includes an added air-based second stage stripper, inserted between a conventional rich-lean crossover heat exchanger and a lean solution temperature polishing heat exchanger. Advantageously, the outlet stream from the air-based second stage stripper, containing a low concentration of CO₂, is used as combustion air for a pulverized coal-fired boiler (see description of FIG. 4 below), enriching the flue gas with CO₂ (with approximately 0.7% de-rating of boiler efficiency).

In an embodiment (see FIG. 4), the described CO₂-enriching membrane and two-stage stripping unit are integrated into an existing power plant **400**. The CO₂ capture and compression system of the present disclosure are generally referenced as block **402**. Summarizing, the CO₂ capture and compression system **402** includes at least a pre-treatment tower, a membrane module (see FIG. 1) coupled with a water evaporator, a moisture separator, a vacuum pump, a packed column CO₂ absorber, two packed bed CO₂ strippers with one reboiler and reclaimer, and various heat exchangers, pumps, filtration devices, and a multistage compressor with intercoolers.

In more detail, after passage of post-combustion gas through an SO₂ polisher **404**, flue gas enters a booster fan **406** to overcome downstream pressure drop. At this point (A) the flue gas is saturated with water at a temperature of around 55° C., water content of 17 vol. %, and CO₂ concentration of approximately 15.8 vol. % of the total wet gas stream. The flue gas then enters a counter-flow pretreatment tower **408** which uses a dilute caustic solution for further SO₂ and other contaminant removal. Tower **408** polishes the flue gas to <5 ppm SO₂, which minimizes solvent degradation and potential membrane fouling. The spent solution is cooled, crystallizing a portion of the sulfur product that is then removed via filtration.

The SO₂-polished flue gas then enters (see point B) a CO₂ membrane pre-concentration module **410** (in the depicted embodiment having a selectivity of approximately 8 and a permeance of ≥5000 gpu) to produce two streams as described above.

In pilot testing, the effect of pre-concentration of CO₂ by a membrane as described was evaluated. Specifically, a pre-concentration membrane provided by MTR was evaluated (see Table 1), using a stripper operated at 45 psi to eliminate any effect of discrepancy of thermal compression from various carbon loadings obtained at the CO₂ absorber. Various liquid/gas ratios were tested to ensure a capture efficiency of approximately 80% or greater. As indicated with similar carbon/nitrogen ratios in a lean solution returning to the CO₂ absorber by varying heat flux at a bottom of the CO₂ stripper, supplying a high CO₂ inlet concentration to the CO₂ absorber reduced energy consumption required for solvent regeneration. The energy needed for regeneration with a pre-concentration membrane was found to be approximately 30% less than that obtained from 14 vol. % CO₂ inlet concentration without the membrane installed.

TABLE 1

Summary of Carbon Loadings and Energy of Regeneration for Various Experiments.							
Expt	Feed CO ₂ (%)	Perm. CO ₂ (%)	Residue CO ₂ (%)	L/G (wt/wt)	Stripper bottom. temp. (° C.)	% Capture	Energy Btu/lb CO ₂
M4	14	—	9	2.9	136	80	1007
M5	14	25	—	3.0	136	79	956
M6	14	29	—	3.2	136	81	1074
R1	14	—	—	4.8	133	79	1468
R2	14	—	—	5.0	132	75	1445

Using the total pressure (P_{tot}) in the stripper of 45 psi (310 kPa) and the partial pressure of steam (P_{H_2O}) at the stripper exhaust temperature, the partial pressure of CO₂ (P_{CO_2}) was estimated as $(P_{tot})=(P_{H_2O})+(P_{CO_2})$. As shown in Table 2, a reduction of steam requirement was observed by inclusion of the pre-concentration membrane.

TABLE 2

Partial Pressure of H ₂ O and CO ₂ at Stripper Exhaust.				
Expt	Stripper Exhaust Temp. (° C.)	Steam Partial Pressure (P_{H_2O}) kPa	CO ₂ Partial Pressure (P_{CO_2}) kPa	P_{CO_2}/P_{H_2O}
With membrane	105	121	189	1.56
Without	107	130	180	1.38

In the depicted embodiment the membrane module **410** is operated at 55° C., and at slightly above ambient pressure at the reject side, and at a vacuum of 30 kPa at the permeate side. The permeate passes through a moisture separator or condenser **412**, a compressor **414**, and then enters a bottom **415** of a CO₂ absorber **416** (see point C). At this point, the permeate stream consists of approximately 60 vol. % CO₂, 31 vol. % N₂+O₂+Ar, trace SO₂ and halogens, and the balance of water vapor. In the depicted embodiment, moisture separator **412** is a membrane separator of known design. In such embodiments, a portion of the combustion air is used as a carrying gas at the permeate side of moisture separator **412**. The moisture-rich air is then fed to a secondary air-based CO₂ stripper **418** (see point N) for CO₂ stripping as described below.

The reject stream proceeds through a direct water contactor **420** (in the depicted embodiment being a water spray system with liquid collection at a bottom thereof) and enters the CO₂ absorber **416** at an intermediate portion **422** (see point D). This is unlike conventional CO₂ absorber configurations including only one feed point for flue gases. The reject stream includes 10 vol. % CO₂ prior to entering the CO₂ absorber **416**.

In an embodiment, the cross-sectional dimension of the bottom packing of CO₂ absorber **416** is approximately 1/3 of the cross-sectional dimension of the top packing, to balance the gas-side pressure drop and mass transfer requirements. An intermediate heat exchanger (cooler) **417** is interposed between the top and bottom portions of CO₂ absorber **416**. CO₂-depleted flue gas is treated in a water-wash section **424** disposed at a top of the CO₂ absorber, using water from the direct water contactor **420** to remove residual solvent. At an exhaust point **426** of CO₂ absorber **416** (see point E), the flue gas is saturated with water at approximately 42° C.

After gaseous CO₂ is transferred to the liquid phase, the carbon-rich solution exits the CO₂ absorber (see point F), is pressurized, and is sent to a heat recovery unit **428** (see point

G) and then to an overhead condenser **430** of a primary CO₂ stripper **432** (see point H). Then, the carbon rich solution proceeds to a lean/rich heat exchanger **434** (see point I). Then, the solution proceeds to the primary CO₂ stripper **432**, which is a conventional stripper of known design. This portion of the process requires external energy to drive a steam reboiler **436**. At an exit point **438** of the primary stripper **432** (see point J), the gas stream consists primarily of CO₂ (60 vol. %) and water vapor (40 vol. %) at a pressure of approximately 3.7 bar and a temperature of approximately 110° C.

After exiting the heat recovery unit **420**, the CO₂ enriched gas stream is pressurized to approximately 153 bar with inter-cooling for downstream utilization or sequestration (see point K). Sensible heat rejected from intercoolers **440** (see point L) is used for water evaporation to increase CO₂ permeance driving force through CO₂ membrane pre-concentration module **410** as described above.

Carbon-lean solution exiting the primary stripper **432** is sent to a lean/rich heat exchanger **434** (see point I) where the heat is recovered with the carbon-rich solution. Next, the carbon-lean stream (temperature approximately 90° C.) is sent to a top (see point M) of an ambient pressure air-swept, packed column secondary CO₂ stripper **442** to further reduce the carbon load. An air-swept secondary CO₂ stripper **442** was selected due to the discovery that this system improved CO₂ stripping from solvent compared to conventional systems at like operating temperatures. As shown in Table 3, under laboratory conditions, at the same column operating temperature the CO₂ concentration in the evolved stream from the air-based stripper increased with increased solvent solution carbon loading compared to conventional systems. Likewise, at a same solvent carbon loading, high operating temperatures provided a high stream CO₂ concentration.

TABLE 3

Impact of Stripping Conditions on Carbon Release (30 wt % MEA solvent).		
Temperature (° C.)	Carbon Loading (mol C/kg soln.)	% CO ₂ in evolved gas
80	1.90	2.43
	2.00	2.59
	1.80	3.55
	1.86	4.35
	1.90	5.06
95	1.93	5.08
	1.89	5.92
	1.95	6.40

Finally, the carbon-lean stream was cooled to approximately 40° C. by power plant cooling water, and recycled to the CO₂ absorber **416** (see point O). Water-saturated air used as carrying gas for CO₂ stripping is obtained from the membrane moisture separator **412** described above. The CO₂ enriched air exiting the secondary CO₂ stripper **442** (approximately 2-4% CO₂ content) is fed to an air preheater and sent to the pulverized coal boiler **444** of the pulverized coal plant **400** as boiler combustion air.

Advantageously, the processes and systems of the present disclosure provide significant advantages over conventional CO₂ recovery technology. According to the Gibbs free energy equations (Gibbs-Helmholtz equation and Clausius-Clapeyron relation), the CO₂ partial pressure at a top of a CO₂ stripper can be expressed as:

$$P_{CO_2} = P_{CO_2,scrub}^* \times e^{\left[\frac{\Delta h_{abs,CO_2}}{R} \times \left(\frac{1}{T_{ref}} - \frac{1}{T} \right) \right]}$$

where $P_{CO_2,scrub}^*$ is the CO_2 equilibrium pressure in the carbon rich solution with a carbon loading at the bottom of a CO_2 absorber with a temperature of T_{ref} . For a given solvent and a temperature, $P_{CO_2,scrub}^*$ directly corresponds to the carbon loading (Carbon/Nitrogen, C/N ratio) resulting from the mass transfer M inside the CO_2 absorber and guided by the equation of $M=K_G \cdot A \cdot \Delta P_{CO_2}$ (driving force defined by differential CO_2 pressure between gas and liquid phase). With an increased CO_2 driving force, a higher C/N ratio is achieved at a bottom of the CO_2 absorber, resulting in a high CO_2 partial pressure at a top of the CO_2 stripper at a desired temperature. This reduces the energy required to regenerate the same amount of CO_2 from a stripping solvent.

As shown in Table 4, compared to conventional equipment and a benchmark solvent (30 wt % monoethanolamine), the presently disclosed processes and systems provide a significant advantage over conventional configurations including CO_2 absorbers with a single feeding point. At an identical flue gas temperature (40° C.) and lean solution temperature (40° C.) fed into the CO_2 scrubber and a constant flue gas flowrate, the presently described technology from process simulation provided an 8% increase in carbon loading, a 19% reduction in liquid recirculation, and a 15% reduction in energy consumption while maintaining a same carbon loading of the lean solution and a 90% overall CO_2 capture efficiency.

TABLE 4

Comparison of CO_2 absorber with a single flue gas feed point to a CO_2 absorber with two flue gas feed points.		
	Single flue gas feed point (normalized to unity)	Dual flue gas feed point
Carbon loading	1	1.08
Solvent recirculation	1	0.81
Energy for regeneration	1	0.85

In turn, the heat integration processes and systems described above provide significant advantages in reduction of energy requirements for recovery of CO_2 compared to conventional systems. In particular, the use of an air-based secondary stripper **442** as described above is advantageous. The extra lean solvent produced from solvent regeneration by the presently described process (compared to conventional systems using a single CO_2 stripper for solvent regeneration) allow use of smaller CO_2 absorbers and CO_2 strippers, reducing capital costs. Further reduction of carbon loading in the lean solvent provides a higher free amine concentration (higher pH) and lower liquid CO_2 partial pressure at a top of the CO_2 absorber. The recycling of CO_2 from the secondary stripper (as combustion air for a pulverized coal boiler, see description above) yields a higher CO_2 concentration (15-20%) compared to conventional systems. Lower liquid CO_2 partial pressure and higher gas phase CO_2 concentration increases the driving force for CO_2 diffusion through a liquid/gas reaction film, resulting in a higher mass transfer coefficient. Still more, reaction kinetics are enhanced by the higher free amine concentration in the upper part of the CO_2 absorber.

Still more, steam increases water vapor concentration in the permeate-side of CO_2 -selective membrane module **410** that is generated by heat rejection from compressor intercoolers. This dilutes the permeance to reduce the vacuum requirement for achieving a reasonable driving force between feed and permeate sides of the CO_2 -selective membrane, allowing reducing membrane surface area (and associated cost) for a given permeance flux. Moreover, the steam saturates (wets) the portion of combustion air in module **412** used for the secondary CO_2 stripper **442**, reducing energy demand for water evaporation (temperature increase from bottom to top along the secondary CO_2 stripper). In turn, selection of low selectivity/ambient pressure operation membranes for use in the CO_2 -selective membrane module **410** allows selection of cost-effective materials for the membrane.

Still yet more, as described above a CO_2 absorber **416** with staged cross-sectional areas is provided, including an intermediate cooler **417**. Unlike conventional absorbers, the CO_2 absorber of the present disclosure includes a bottom packing having a smaller cross-sectional dimension than that of the top packing. This configuration takes advantage of a low flue gas flow rate (14% of total flue gas) while maintaining a reasonable flue gas velocity to minimize gas-side mass transfer resistance and to reduce a flooding tendency caused by a high liquid/gas ratio. In one embodiment, the present CO_2 absorber bottom portion cross-sectional dimension is approximately $\frac{1}{3}$ that of the top portion cross-sectional dimension. By use of the described CO_2 absorber including staged cross-sectional areas of this ratio, the liquid/gas ratio is increased 7.7 times in the bottom section of packing to enhance gas-liquid contact and reshape the vapor-liquid-equilibrium operating curve in an interior of the CO_2 absorber, providing maximum benefit from high inlet CO_2 concentration (CO_2 gas pressure).

In another feature, the primary CO_2 stripper **432** is operated at an elevated pressure (approximately 3.7 bar), reducing energy consumption for CO_2 compression by 30% compared to conventional systems. As shown in FIG. 5, energy consumption is significantly reduced as stripper operating pressure increases. As the stripper pressure increases, the mole ratio of CO_2/H_2O (vapor) at the stripper outlet is increased at a given temperature, leading to overall reduction in energy consumption. The operating pressure (3.7 bar) was found to maximize the energy benefit while minimizing capital investments and solvent degradation (solvent degradation is known to increase with increasing temperature, but energy savings for solvent regeneration by increasing stripper operating pressure reach a point of diminishing returns).

Numerous other advantages accrue from the presently described processes and systems. Combining membrane pre-concentration of flue gas CO_2 and use of a two-stage stripping system as described provided a CO_2 mass-transfer flux of energy on average 2.5 times faster than that of conventional systems, with further benefits deriving from the CO_2 absorber including staged cross-sectional dimensions as described. The high mass-transfer flux resulted in the ability to provide a smaller CO_2 absorber to capture the same amount of CO_2 from a flue gas stream. CO_2 absorber size reductions of up to 50% compared to conventional absorbers are anticipated compared to conventional systems/solvents.

Further, the described heat integration processes are anticipated to reduce plant cooling tower heat rejection duty (the amount of heat processes by the cooling tower) by 12-19%, reducing capital cost for the cooling tower and auxiliary equipment by approximately 15%.

The foregoing description has been presented for purposes of illustration and description. It is not intended to be exhaus-

tive or to limit the invention to the precise form disclosed. Obvious modifications or variations are possible in light of the above teachings. The embodiments were chosen and described to provide the best illustration of the principles of the invention and its practical application to thereby enable one of ordinary skill in the art to utilize the invention in various embodiments and with various modifications as are suited to the particular use contemplated. All such modifications and variations are within the scope of the invention as determined by the appended claims when interpreted in accordance with the breadth to which they are fairly, legally and equitably entitled. In turn, the drawings and preferred embodiments do not and are not intended to limit the ordinary meaning of the claims in their fair and broad interpretation in any way when interpreted in accordance with the breadth to which they are fairly, legally and equitably entitled.

What is claimed:

1. A process for removal and recovery of CO₂ from a post-combustion flue gas, comprising:

pre-concentrating a CO₂ component of the post-combustion flue gas by passing the post-combustion flue gas through a CO₂-selective membrane module to provide a CO₂-enriched permeate stream and a CO₂-lean reject stream;

in a CO₂ absorber, contacting the CO₂-enriched permeate stream and the CO₂-lean reject stream with a scrubbing solvent to absorb CO₂ and provide a carbon-rich scrubbing solvent; and

stripping absorbed CO₂ from the carbon-rich scrubbing solvent by a two-stage CO₂ stripping system.

2. The process of claim 1, further including evaporating at least a portion of a water component of a post-combustion flue gas condensate for use as a carrier vapor for a permeate side of the CO₂-selective membrane module.

3. The process of claim 1, further including passing a CO₂-rich permeate stream from the CO₂-selective membrane module to a feed point disposed at a bottom portion of the CO₂ absorber and passing the CO₂-lean reject stream from the CO₂-selective membrane module to an intermediate feed point above the CO₂ absorber bottom portion feed point.

4. The process of claim 1, further including providing a CO₂-membrane module comprising a high flux, low pressure drop, low CO₂ selectivity membrane.

5. The process of claim 1, further including providing a two stage stripping system comprising:

a primary CO₂ stripping column for stripping CO₂ from the carbon-rich scrubbing solvent exiting the CO₂ absorber; and

a secondary CO₂ stripping column for stripping CO₂ from a carbon-lean scrubbing solvent exiting the primary CO₂ stripping column.

6. The process of claim 5, including providing an ambient pressure air-swept, packed column secondary CO₂ stripping column.

7. The process of claim 1, including providing a CO₂ absorber having a bottom packing portion with a first cross-sectional dimension, and a top packing portion with a second cross-sectional dimension that is different from the first cross-sectional dimension.

8. The process of claim 1, including providing one or more intercoolers at intermediate positions of the CO₂ absorber.

9. The process of claim 7, wherein the first cross-sectional dimension is approximately $\frac{1}{3}$ that of the second cross-sectional dimension.

10. The process of claim 6, including passing a CO₂-enriched air source exiting the secondary CO₂ stripping column to a preheater and therefrom as a boiler combustion air for a pulverized coal boiler of a pulverized coal plant.

11. A CO₂ capture and recovery system for recovery of CO₂ from a post-combustion gas, comprising:

a CO₂-selective membrane module for pre-concentrating a CO₂ component of an SO₂-depleted post-combustion gas to provide a CO₂-enriched permeate stream and a CO₂-lean reject stream;

a CO₂ absorber for providing a carbon-rich scrubbing solvent by contacting the CO₂-enriched permeate stream and the CO₂-lean reject stream with a scrubbing solvent to absorb CO₂; and

a two-stage CO₂ stripping system for stripping absorbed CO₂ from the carbon-rich scrubbing solvent.

12. The system of claim 11, further including a moisture separator or condenser for removing at least a portion of a water vapor component of the CO₂-enriched permeate stream.

13. The system of claim 11, wherein the CO₂-enriched permeate stream from the CO₂-selective membrane module is passed to a feed point disposed at a bottom of the CO₂ absorber and the CO₂-lean reject stream from the CO₂-selective membrane module is passed to an intermediate feed point above the CO₂-absorber bottom feed point.

14. The system of claim 11, wherein the CO₂-membrane module comprises a high flux, low pressure drop, low CO₂-selectivity membrane.

15. The system of claim 11, wherein the two stage stripping system comprises:

a primary CO₂ stripping column for stripping CO₂ from the carbon-rich scrubbing solvent exiting the CO₂ absorber; and

a secondary CO₂ stripping column for stripping CO₂ from a carbon-lean scrubbing solvent exiting the primary CO₂ stripping column.

16. The system of claim 15, wherein the secondary CO₂ stripping column is an ambient pressure air-swept, packed column CO₂ stripping column.

17. The system of claim 15, wherein the CO₂ absorber includes a bottom packing portion with a first cross-sectional dimension and a top packing portion with a second cross-sectional dimension that is different from the first cross-sectional dimension.

18. The system of claim 11, wherein the CO₂ absorber further includes one or more intercoolers disposed at intermediate positions of the CO₂ absorber.

19. The system of claim 17, wherein the first cross-sectional dimension is approximately $\frac{1}{3}$ that of the second cross-sectional dimension.

20. The system of claim 16, wherein a CO₂-enriched air source exiting the secondary CO₂ stripping column is passed to a preheater and therefrom to a pulverized coal boiler of a pulverized coal power plant as a boiler combustion air.

21. A pulverized coal-fired power plant comprising the CO₂ capture and recovery system of claim 11.

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