

(43) International Publication Date
22 November 2007 (22.11.2007)

(10) International Publication Number
WO 2007/133595 A2

- (51) **International Patent Classification:**
C12P 21/04 (2006.01) *C07K 14/405* (2006.01)
G01N 33/00 (2006.01)

(21) **International Application Number:**
PCT/US2007/011234

(22) **International Filing Date:** 8 May 2007 (08.05.2007)

(25) **Filing Language:** English

(26) **Publication Language:** English

(30) **Priority Data:**
60/798,489 8 May 2006 (08.05.2006) US

(71) **Applicant (for all designated States except US):** **THE BOARD OF TRUSTEES OF THE UNIVERSITY OF ILLINOIS** [US/US]; 506 South Wright Street, 352 Henry Administration Building, Urbana, IL 61801 (US).

(72) **Inventors; and**

(75) **Inventors/Applicants (for US only):** **CHEN, Shiaguo** [CN/US]; 2111 Emerald Dr., Champaign, IL 61822 (US). **LU, Yonggi** [CN/US]; 804 Brighton Dr., Urbana, IL 61801 (US). **ROSTAM-ABADI, Massoud** [IR/US]; 1108 Sterling Dr., Champaign, 61821 (US).

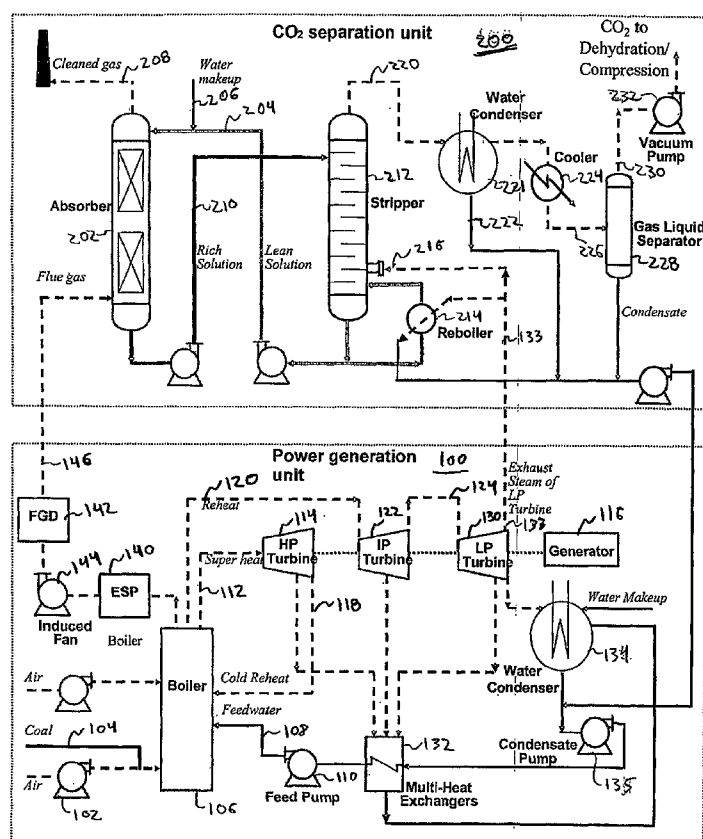
(74) **Agents:** **RANSON, Arik, B.** et al.; Greer, Burns & Crain, Ltd., 300 S. Wacker Drive, Suite 2500, Chicago, IL 60606 (US).

(81) **Designated States (unless otherwise indicated, for every kind of national protection available):** AE, AG, AL, AM, AT, AU, AZ, BA, BB, BG, BH, BR, BW, BY, BZ, CA, CH, CN, CO, CR, CU, CZ, DE, DK, DM, DZ, EC, EE, EG, ES, FI, GB, GD, GE, GH, GM, GT, HN, HR, HU, ID, IL, IN, IS, JP, KE, KG, KM, KN, KP, KR, KZ, LA, LC, LK, LR, LS, LT, LU, LY, MA, MD, MG, MK, MN, MW, MX, MY, MZ, NA, NG, NI, NO, NZ, OM, PG, PH, PL, PT, RO, RS, RU, SC, SD, SE, SG, SK, SL, SM, SV, SY, TJ, TM, TN, TR, TT, TZ, UA, UG, US, UZ, VC, VN, ZA, ZM, ZW.

(84) **Designated States (unless otherwise indicated, for every kind of regional protection available):** ARIPO (BW, GH, GM, KE, LS, MW, MZ, NA, SD, SL, SZ, TZ, UG, ZM, ZW), Eurasian (AM, AZ, BY, KG, KZ, MD, RU, TJ, TM), European (AT, BE, BG, CH, CY, CZ, DE, DK, EE, ES, FI,

[Continued on next page]

- (54) Title:** INTEGRATED VACUUM ABSORPTION STEAM CYCLE GAS SEPARATION



(57) Abstract: Methods and systems for separating a targeted gas from a gas stream emitted from a power plant. The gas stream is brought into contact with an absorption solution to preferentially absorb the targeted gas to be separated from the gas stream so that an absorbed gas is present within the absorption solution. This provides a gas-rich solution, which is introduced into a stripper. Low pressure exhaust steam from a low pressure steam turbine of the power plant is injected into the stripper with the gas-rich solution. The absorbed gas from the gas-rich solution is stripped in the stripper using the injected low pressure steam to provide a gas stream containing the targeted gas. The stripper is at or near vacuum. Water vapor in a gas stream from the stripper is condensed in a condenser operating at a pressure lower than the stripper to concentrate the targeted gas. Condensed water is separated from the concentrated targeted gas.



FR, GB, GR, HU, IE, IS, IT, LT, LU, LV, MC, MT, NL, PL,
PT, RO, SE, SI, SK, TR), OAPI (BF, BJ, CF, CG, CI, CM,
GA, GN, GQ, GW, ML, MR, NE, SN, TD, TG).

For two-letter codes and other abbreviations, refer to the "Guidance Notes on Codes and Abbreviations" appearing at the beginning of each regular issue of the PCT Gazette.

Published:

- *without international search report and to be republished upon receipt of that report*

INTEGRATED VACUUM ABSORPTION STEAM CYCLE GAS SEPARATION

PRIORITY CLAIM

This application claims the benefit of U.S. Provisional
5 Application Serial No. 60/798,489, filed May 8, 2006, under 35 U.S.C. § 119.

STATEMENT OF GOVERNMENT INTEREST

This invention was made with Government assistance under
Department of Energy Grant No. DE-FC26-03NT41994. The Government has
certain rights in this invention.

10

BACKGROUND ART

A field of the invention is gas treatment. A particular exemplary
application of the invention is to flue gas treatment, such as from flue gases of
power plants. The invention is particularly applicable to treatment of gas
streams.

15

Conventional coal-firing plants use a steam cycle for power
generation. Steam cycles are also used in newer power plants, such as
Integrated Gasification Combined Cycle (IGCC) power plants or Natural Gas
Combined Cycle (NGCC) power plants. In a conventional steam cycle power
plant, only about one-third of the heat energy contained in the fossil fuel is
20 converted to electricity. This leaves more than half of the energy wasted in the
water cooling towers. The water cooling tower transfers all of the waste heat in
the low pressure exhaust steam (1.2 psia, 107.8 °F) to the environment through
the vaporization of cooling water.

25

On the other hand, CO₂ separation from post-combustion flue gas
requires a large amount of heat (in the form of steam). For example, the
current state-of-the-art monoethanol amine (MEA) absorption process requires
about 4MJ of total heat for every kilogram of CO₂ captured. The steam that
carries the heat has a pressure of around 60 psia. An absorption process that

uses steam at a much lower pressure will have a significant advantage over the MEA-based absorption process.

Conventional CO₂ separation processes are energy intensive. The required energy is provided by steam in the power plant. The extraction of steam from the steam turbine reduces the electricity generation (output). For the same amount of steam, the higher the pressure (and thus the temperature) of the extracted steam is, the higher the loss of the electricity will be. State-of-the-art amine-based absorption processes reduce electricity output by about 30%. For example, the installation of an MEA-based process in a steam cycle power plant will reduce its net electricity output from 500 MW to around 350 MW.

This energy cost, and the associated financial cost, inhibits the adoption of CO₂ separation. With conventional MEA-based absorption processes, the cost of CO₂ separation from post-combustion flue gases takes more than 80% of the total CO₂ sequestration cost. High separation cost is the main reason for the delaying of CO₂ emission control. High CO₂ separation cost is also hindering the acceptance of CO₂ based Enhanced Oil Recovery (EOR) technology.

CO₂ emission control from power plants is not regulated at this time in the United States. There are several thousand power plants in the United States alone. If CO₂ emission control becomes mandatory, an efficient and cost-effective process will be required. If voluntary reductions of emissions are to be encouraged, the cost of emission control must also be reduced. Currently, the CO₂ emission level in the United States is about 20% above the level of 1990. A goal defined by the Kyoto protocol is that the United States needs to control around 20% of CO₂ emission, which is about 12 Billion metric tons annually, equivalent to 300 of 500MW power plants. Acceptance and adoption of CO₂ separation, whether voluntary or by treaty or regulation, requires an improved CO₂ separation process.

DISCLOSURE OF THE INVENTION

According to exemplary embodiments of the present invention, methods and systems are provided for separating a targeted gas from a gas stream emitted from a power plant. In a preferred method, the gas stream is brought into contact with an absorption solution to preferentially absorb the targeted gas to be separated from the gas stream so that an absorbed gas is present within the absorption solution. This provides a gas-rich solution, which is introduced into a stripper.

Low pressure exhaust steam from a low pressure steam turbine of the power plant is injected into the stripper with the gas-rich solution. The absorbed gas from the gas-rich solution is stripped in the stripper using the injected low pressure steam to provide a gas stream containing the targeted gas. The stripper is at or near vacuum. Water vapor in a gas stream from the stripper is condensed in a condenser operating at a pressure lower than the stripper to concentrate the targeted gas. Condensed water is separated from the concentrated targeted gas.

BRIEF DESCRIPTION OF THE DRAWINGS

FIG. 1 is a schematic diagram of an embodiment of the invention, showing an exemplary power plant and an integrated absorption-based CO₂ separation process;

FIG. 2 is a schematic diagram of a membrane-based integrated CO₂ separation process, according to an alternative embodiment of the present invention;

FIG. 3 shows an exemplary relationship between stripper vacuum degree and electricity loss;

FIG. 4 is a table showing main operating conditions used in a simulated CO₂ removal system and process, according to an embodiment of the present invention;

FIG. 5 shows results of energy and mass balance calculations for a reference pulverized coal (PC) power plant;

FIG. 6 shows energy distributions in the PC power plant of FIG. 5;

5 FIG. 7 shows steam properties during expansion in IP and LP turbines;

FIG. 8 shows mass and energy balance calculations for a simulated monoethanolamine (MEA) system and process;

10 FIG. 9 shows electricity output and use in a power plant with and without the MEA system shown in FIG. 8;

FIG. 10 shows CO₂ and water partial pressures over a 20%wt K₂CO₃ aqueous solution, according to an embodiment of the present invention;

15 FIG. 11 shows mass and energy balances for an exemplary vacuum carbonate process, according to an embodiment of the present invention;

FIG. 12 shows electricity loss at different stripping pressures;

FIG. 13 is a table showing electricity use and output in a PC power plant, PC+MEA plant, and a PC+vacuum carbonate system according to an embodiment of the present invention;

20 FIG. 14 shows the effect on liquid-to-gas ratio (L/G ratio) at different CO₂ loadings;

FIG. 15 shows the effect on steam extraction loss at different CO₂ loadings;

25 FIG. 16 shows the effect on steam extraction loss at different L/G ratios; and

FIG. 17 shows cost comparisons for a simulated vacuum-carbonate process according to an embodiment of the present invention, and for an MEA process.

BEST MODE OF CARRYING OUT THE INVENTION

An integrated CO₂ sequestration scheme includes three general steps: CO₂ capture, transportation, and permanent storage. A vast majority of the fossil fuel-fired power plants, which are the most likely point sources for CO₂ capture, generally produce diluted CO₂ gas streams. Capture of CO₂ from these sources is the most expensive step, consuming more than 70% of the total sequestration cost. Reducing the CO₂ capture cost would clearly have the most significant impact on the economic performance of a sequestration process.

Currently, MEA-based absorption processes are considered the best available option for post-combustion CO₂ capture. However, the cost for CO₂ removal from the post-combustion flue gas of a pulverized coal (PC) power plant with a state-of-the-art MEA process typically ranges from \$40 to \$60 per ton of CO₂ avoided. Such a high cost may be economically unacceptable. The electricity and heat used for the MEA process has been estimated to contribute to about half of the total cost, and the O&M cost and capital cost each are estimated to contribute about a quarter. Thus, reducing the energy consumption of the CO₂ capture process goes a large way to lowering the total cost.

J. Ciferno, P. Dipietro, and T. Tarka, "An Economic Scoping Study for CO₂ Capture Using Aqueous Ammonia", US DOE/NETL Report, February 2005, showed that the large energy consumption of the MEA process reduces electricity output of a PC power plant by about 28%. Breakdowns of the electricity loss revealed that about 65% is due to the steam extracted from the power plant and used for heat in the stripper reboiler. CO₂ compression prior to pipeline transportation accounted for another 25% of the loss, and gas blowers and liquid pumps required in the MEA process are responsible for the remaining 10%.

Since the energy use for CO₂ compression is unavoidable for an absorption process, minimizing the reboiler heat duty and/or lowering the quality of the extracted steam is the most viable way to improve the process. The reboiler heat duty in an absorption process mainly depends on the reaction

heat (heat of absorption) and stripping heat. The sensible heat can be mostly recovered through heat exchange in the absorption-desorption loop. For the MEA process, the reaction and stripping heats are 825 and 800 Btu/lb CO₂, respectively. To reduce the heat duty of the reboiler, an absorption solution
5 with lower heat of reaction and stripping should be used.

One should note that heat energy should be described by both its quantity and quality. Heat energy at a higher temperature will contain more exergy (exergy is a measure of the work potential at a certain state, relative to the reference state, defined as $T_0 = 77\text{ }^{\circ}\text{F}$, $P_0 = 14.7\text{ psia}$) compared with heat at
10 a lower temperature. If the heat energy consumed in the CO₂ absorption process is of low quality (i.e., its T and P are close to T_0 and P_0), or even waste heat (whose T and P are virtually T_0 and P_0), then the total cost of the CO₂ capture process will be low even though the amount of heat required is large. As a typical MEA process operates at 200 to 250 $^{\circ}\text{F}$ (1-2 atm) for CO₂
15 stripping, to maintain a required temperature difference, which is the driving force for heat transfer, the steam used in the reboiler of an MEA process needs to be extracted from the power plant steam cycle at a pressure of about 60 psia (saturation temperature of 293 $^{\circ}\text{F}$). The high temperature requirement increases the electricity loss due to steam extraction.

20 Embodiments of the present invention provide a uniquely integrated CO₂ capture/separation process for post-combustion flue gases of power plants with steam cycle. An organically integrated system according to embodiments of the invention, which integrates two processes, can save large amounts of energy and significantly reduce the CO₂ separation cost. In
25 exemplary embodiments, a vacuum absorption process and the steam cycle are integrated efficiently so that the abundantly available, low quality exhaust steam in the steam cycle can be used in the CO₂ separation process. More particularly, in embodiments of the invention, low quality exhaust steam in the steam cycle is used for the stripping of CO₂ in a vacuum absorption-based CO₂
30 separation process. A lower stripping temperature can make possible the use of

the heat of lower quality steam, and the temperature in the stripper can be low at a low pressure; e.g., vacuum.

Preferred systems and methods of the invention combine a vacuum absorption process with a power plant steam cycle in a way to take advantage of the low-pressure exhaust steam in the steam cycle. In preferred embodiments of the invention, the electricity output of a 500MW power plant that adopts the CO₂ separation will be reduced to around 400MW, a savings of around 50MW compared to state-of-the-art amine-based CO₂ separation processes. Preliminary economic analysis reveals that the CO₂ avoidance cost of this process will be around \$33/tonne, compared with \$55/tonne for an MEA process.

Preferred embodiment systems and methods of the invention, which integrate a vacuum absorption-based CO₂ separation process and a steam cycle in a power plant, take advantage of the low-pressure (LP) exhaust steam (1.2 to 14.7 psia) abundantly available in the steam cycle. This significantly reduces electricity loss due to the installation of the exemplary CO₂ separation process. In embodiments of the invention, electricity loss can be substantially minimized to its theoretical limit. Additionally, compared to conventional systems, some heat exchangers can be eliminated due to the fact that absorption and desorption processes are operated at about the same temperature.

Embodiments of the invention can substantially reduce the CO₂ separation cost, e.g., a reduction of half of the current state-of-the-art MEA-based absorption processes and a 40% cost reduction. Such significantly reduced electricity loss can encourage the adoption of a CO₂ separation process.

Referring now to the drawings, FIG. 1 is a schematic diagram of an embodiment of the invention for absorption-based CO₂ separation processes. In the conventional CO₂ separation processes, the steam cycle 100 and the CO₂ separation process 200 are isolated. By "isolated", it is meant that there is no or little materials exchange between the two systems. The steam cycle may be, as nonlimiting examples, a pulverized coal (PC) firing power plant, an

integrated gasification combined cycle power plant, or a natural gas combined cycle power plant.

Within the steam cycle (power generation unit) 100, for example, air 102 and coal 104, such as pulverized coal (PC), are used to heat a boiler 106.

5 Within the boiler 106, feedwater 108 is fed via a feed pump 110 and heated to produce superheated steam 112. The superheated steam 112 is fed to a high-pressure (HP) turbine 114, which in turn drives a generator 116. Exhaust steam 118 from the HP turbine is reheated by the boiler 106, and the reheated steam 120 is fed to an intermediate pressure (IP) turbine 122, also driving the
10 generator 116. Exhaust steam from the IP turbine 124 in turn is fed to a low pressure (LP) turbine 130, which, again, drives the generator 116.

Additional exhaust steam from the HP, IP, and LP turbines 114, 122, 130 is fed to multi-heat exchangers 132. Some exhaust steam from the LP turbine 130 enters a water condenser 134, and the resulting water is fed into the
15 multi-heat exchangers 132 using a condensate pump 135. Thus, in the steam cycle, the low pressure exhaust steam 133 from the LP steam turbine 114 can go to the condenser 134 to be condensed. The condensed water is then recycled to generate steam in the boiler 106, and the cycle goes on. The operating condition in the condenser 134 is about 1.2 psia and 107.8 °F. The
20 lower the operating temperature (thus the pressure, since it is a saturated steam) is, the higher the thermal efficiency of the power generation system 100 will be. However, the lowest operating temperature is limited by the temperature of the cooling water, which depends on water sources and the season as well. In the summer, the operating temperature will be higher than in the winter.

25 Flue gas from the boiler 106 enters an electrostatic precipitator (ESP) 140 and then a flue gas desulfurization (FGD) 142 via an induced fan 144. Cleaned flue gas 146 containing carbon dioxide (CO₂) exits to the CO₂ separation unit 200.

Within the CO₂ separation unit 200, the cleaned flue gas 146
30 enters an absorber 202, into which is fed a lean solution 204 (along with water makeup 206) for absorbing acid gas such as CO₂. The absorber 202 may

include, for example, any suitable absorption device, such as, but not limited to, an absorption tower, that allows the flue gas stream 146 to be brought into contact with the absorption solution 204 to preferentially absorb the targeted gas (e.g., CO₂) to be separated from the main gas. After absorption, the
5 cleaned gas 208 exits, while the CO₂-rich solution 210 is introduced to a stripper 212 for CO₂ stripping. A nonlimiting example of the stripper 212 is a stripping tower.

In the conventional absorption process, the stripping steam is generated in a reboiler at the bottom of the stripper. The steam extracted from
10 the steam turbine provides the heat to vaporize the CO₂-rich solution at one side of the reboiler and itself is condensed at the other side of the reboiler. Since the heat is supplied through the reboiler, a temperature difference (ΔT) between the steam side and solution side is always required. This ΔT will impact the quality of the extracted steam. A ΔT of 20 °C could lead to a 3%
15 thermal efficiency loss. Further, the (huge) reboiler requires a large capital investment.

In the CO₂ separation unit 200, the stripper 212 may also be heated by a stripper reboiler 214, which may be heated, for example, using part of the exhaust steam 133 from the LP turbine 130. Preferably, a heat
20 exchanger is used to transfer heat from the LP exhaust steam 133 to the absorption solution in the reboiler 214.

Further, in exemplary embodiments of the present invention, the low-pressure exhaust steam from the lower pressure steam turbine 130 will be injected 216 directly into the CO₂ stripping tower 212 to strip CO₂ before it
25 goes to the water condenser 134. The direct introduction of the steam 133 into the stripper 212 will be able to reduce or eliminate the ΔT and reduce the reboiler 214 size. Generally, the “low pressure” turbine from which the LP exhaust steam 133 refers to the lowest energy steam available before it goes to the condenser 134, provided that this steam is within an acceptable temperature
30 range to operate within the stripper 212. In preferred embodiments, the LP exhaust steam 133 has a pressure between 1.2 and 14.7 psia, with a temperature

between 40 and 100 °C. Most preferably, the pressure is between 3 and 7 psia, with a temperature between 60 and 85 °C.

The exemplary gas stream 220 from the top of the stripper 212 is a mixture of CO₂ and steam with a CO₂/steam mole ratio of about 1:3. When
5 the mixture 220 passes through a water condenser 221, the water is condensed, and the CO₂ is concentrated. The condenser 221 operates at a pressure lower than that of the stripper 212. The condensed water will be recycled 222 to the power generation system 100 after necessary water treatment. Alternatively, additional water treatment may be omitted. The concentrated CO₂ stream from
10 the condenser 221 is cooled 224, and additional water 226 is condensed. The condensed water 226 is then separated in a gas-liquid separator 228, and the pure CO₂ 230 is finally drawn from the separator to a vacuum pump 232 to be compressed to a desired pressure.

FIG. 2 describes another embodiment of the invention, where like
15 parts are represented with like numerals. In FIG. 2, a membrane CO₂ separation process 300 is integrated with the steam cycle 100. In this system 300, a membrane module 302, such as but not limited to a membrane contractor or an integrated membrane system, replaces the absorber 202 and the stripper 212 used in the CO₂ separation unit 200. Particularly, within the
20 membrane module 302, a CO₂ absorption solvent (the CO₂ lean solution) 304 is in contact with flue gas at the retentate side to absorb CO₂ from gas phase and produces the CO₂ rich solution. The CO₂ rich solution carries CO₂ from the high pressure side (retentate side) to the low pressure side (permeate) through a porous membrane. The low pressure exhaust steam 133 from the LP turbine
25 130 is used to strip the permeate side of the membrane module 302. Some of the CO₂ rich solution may be left at the retentate side and is recycled through a pump 306. A liquid seal 308 is used to block the desorbed CO₂ rich gas from the permeate side. Different membranes, such as ultrafiltration membranes and nanofiltration membranes, can be used for the membrane module 302.

30 To allow the LP exhaust steam 133 to be used to strip CO₂ within the stripper 212 or the membrane module 302, one or more significant features

of the present invention may be employed. One such feature is that the stripper 212 or the low pressure (permeate) side of the membrane module should operate at a pressure lower than the LP 133 pressure. This allows, for example, the use of absorbents such as potassium carbonate or absorbents with similar properties, and, as will be explained below with respect to exemplary
5 embodiments of the invention, significantly reduces electricity requirements.

In preferred embodiments, the stripper 212 is a vacuum stripper operating at absolute pressures between 2 to 6 psia and corresponding temperatures between 50 to 70 °C. For the membrane module 302, preferably
10 the permeate side operates under vacuum at pressure between 1.2 to 14.7 psia, with a preferred pressure between 2 and 6 psia. The vacuum degree is a significant operating parameter in exemplary embodiments of the invention. Higher vacuum will allow the use of lower pressure steam and thus reduce electricity loss. However, the high vacuum will lead also to a larger gas
15 volume and result in larger stripper and vacuum pump sizes. FIG. 3 shows an exemplary relationship between vacuum degree and electricity loss. In preferred embodiments of the present invention, the condenser 221, which automatically contains a vacuum, is used as a “free” vacuum to reduce energy. In certain embodiments, the condenser 221 in the CO₂ separation unit 200 may
20 be the same part or part of the condenser 134 in the power plant 100. More particularly, in preferred embodiments, the vacuum of the stripper 212 is partially obtained by condensing water vapor in the gas stream from the stripper.

As an additional consideration, during operation of the absorber
25 202 (and the retentate side of the membrane module 302), a solvent should be chosen that can absorb the targeted gas (e.g., CO₂), yet allow the CO₂ to be released under vacuum pressure (or substantially vacuum pressure) within the stripper 212 (or permeate side of the membrane module). For example, different CO₂ absorption solutions, such as aqueous solutions of K₂CO₃,
30 Na₂CO₃, K₃PO₄, Na₃PO₄, amino-acid salts, and other salts, may be used as long as the absorption solution can adsorb the targeted gas (e.g., CO₂) at or near the

flue gas conditions, and as long as the absorbed CO₂ in the solutions can be stripped by the exhaust steam (1.2-10 psia) from the low pressure steam turbine 130 of the power generation unit 100. Since the steam extracted is saturated steam, the mentioned pressure range corresponding to an operating temperature and pressure of the stripping tower should be comparable with the conditions of the exhaust steam between 40 to 80 °C.

Further considerations for reducing energy include increasing the rate at which the CO₂ is absorbed in the solvent within the absorber 202. For example, by requiring less contact time between the gas and the solution, the size of the absorption column can be reduced. This may be done, for example, by adding a promoter to the solvent. Exemplary promoters include, but are not limited to, piperazine, monoethanolamine, and other organic base or inorganic base chemicals. Yet another consideration may be to reduce the energy needed to evaporate the water by adding another water soluble additive to the solution to reduce the vapor pressure of the solution. Exemplary additives include, but are not limited to, a water-soluble chemical such as ethylene glycol, or a salt such as NaCl or K₂SO₄.

The principle of the current invention can also be used for other gas separation applications. For example, an absorption-based oxygen separation process may use the exhaust steam for oxygen stripping, where the gas stream is air or other oxygen-containing gas mixture. In this case, the absorption solution would be an oxygen-absorbing solution that can absorb O₂ from air (for example) at a temperature close to the stripping temperature and can be stripped out at desirable stripper conditions, such as those set out above.

Experiment

The energy performance of an exemplary method according to the present invention, referred to as a vacuum-carbonate process, for CO₂ capture is evaluated applied to a 528 MW (gross) pulverized coal-fired power plant. Steady state process simulations were performed to compare the mass and energy balance of the vacuum carbonate process and a conventional MEA

process. The quality of steam heat required for the vacuum process and the associated electricity loss were significantly less than those for the MEA process.

The exemplary vacuum-carbonate process according to an embodiment of the present invention employs a potassium carbonate (K_2CO_3) aqueous solution for CO_2 absorption. Due to its weaker alkalinity, the heat of absorption of the CO_2/K_2CO_3 system (262 Btu/lb) is much less than that of the CO_2/MEA system (825 Btu/lb). This reduction in the heat of absorption will reduce the reboiler heat duty. In addition, while the absorption takes place at flue gas conditions similar to MEA processes, the stripper is operated under vacuum. Because of the reduced pressure, the CO_2 -rich solution boils at a lower temperature than that near the atmospheric pressure. This allows low quality steam to be used as a heat source and can result in reduced electricity loss. Both the low heat of absorption and desorption under vacuum conditions close to the exhaust steam condition make the vacuum-carbonate process an attractive technology.

A process simulation software, CHEMCAD (version 5.3.7), developed by Chemstations Inc., was used to perform steady state process simulations for a reference PC power plant, with an MEA process, and with an exemplary vacuum-carbonate process. FIG. 4 shows the major operating conditions employed in these simulated processes. Other process parameters, such as stripping temperature, heat duty of the reboiler, and steam extraction pressure, were derived from process simulations.

The reference plant is a standard air-blown PC power plant with a 528 MWe gross output. A typical Illinois coal was selected as the fuel. This coal has a moisture content of 6.08%, an ash content of 8.90%, and an HHV of 12,475 Btu/lb. The coal was expressed in an equivalent molecular formula as $C_{164}H_{134}O_{11}N_3S_2$ on a dry ash-free basis. The coal combustion in the boiler was modeled by the Gibbs free energy reactor module, which determines the equilibrium state through the minimization of the Gibbs free energy of the reaction system. The expander module was used to simulate the steam turbines.

The expansion was assumed as adiabatic expansion. The heat exchanger module was applied to heat exchanges in the boiler.

The CO₂ absorption and stripping columns of the vacuum-carbonate process were modeled by the rigorous multi-stage vapor-liquid equilibrium module (SCDS). The vapor-liquid equilibrium (VLE) calculation was selected in the SCDS module in this experiment. Fifteen stages were assumed for both the adsorption and the stripping columns. While the level of CO₂ removal was fixed in absorption, the liquid-to-gas ratio (L/G) was adjustable by varying the stage equilibrium efficiency. The thermodynamic equilibrium constant values (K values) were calculated using the electrolyte Non-Randomness Two-Liquid (NRTL) model for the liquid phase and the ideal vapor model for the gas phase. The binary interaction parameters for the species in the H₂O-CO₂-K₂CO₃ solution were adopted from the CHEMCAD database. The enthalpy and entropy modeling was based on the latent heat method. The chemical reaction equilibrium constants and transport properties of the system, both of which are temperature-dependent, were based on the regression equations incorporated in CHEMCAD.

The SCDS module was also used for modeling the MEA process. The procedure of the VLE calculation was similar to that for the vacuum-carbonate process. The thermodynamic equilibrium K values were calculated by the amine K-value model, which is based on the Soave-Redlich-Kwong (SRK) equation of state. The chemical reaction equilibrium constants involved in the CO₂-MEA aqueous solution were built into the CHEMCAD package. The enthalpies of chemical components were estimated by the amine enthalpy model, which employs the SRK equation for the vapor enthalpy and the method of Crynes and Maddox for the liquid enthalpy.

Results

Reference PC power plant

The results of the energy and mass balance calculation are shown in FIG. 5. The plant operating at full capacity demanded a heat input of 4,499

MMBtu/hr (180 ton/hr as-received coal). Complete combustion of coal was assumed in the simulation. The flue gas exited from the boiler at a flow rate of 5.04 million scfh, composed of 14.6%vol of CO₂, 75.7% of N₂, 2.7% of O₂, 6.8% water vapor, 1,774 ppm SO₂, 325 ppm NO, and 5.2 g/m³ fly ash. Most of the ash and SO₂ were then removed in the ESP and FGD devices. The clean gas leaving the FGD had a slightly increased volume (5.34 million scfh) due to water evaporation in the FGD absorber. This gas stream contained 13.9%vol of CO₂, 71.4% of N₂, 2.5% of O₂, 12.2% water vapor, 34 ppm SO₂, and 31 ppm NO. The flue gas was blown to provide the pressure drop of 2 psia required by the CO₂ capture process downstream. However, the associated electricity use was counted into the CO₂ process discussed below.

In the steam cycle, the main steam (1000 °F, 2415 psia) was fed to the HP turbine. The cold steam exiting from the HP turbine (635 °F, 604 psia) was reheated to 1000 °F (545 psia) and went to the IP and the LP turbines in series. The exhaust steam from the LP turbine (108 °F, 1.2 psia) was condensed in the water condenser and then fed back to the boiler after being preheated through seven consecutive heat exchangers. For simplicity and illustrative purposes, the calculated results for these heat exchangers are combined in FIG. 5. Through the steam cycle, a gross electricity output of 527.6 MW was generated, of which 150.6 MW was from the HP turbine, 124.7 MW was from the IP, and 252.3 MW was from the LP. However, due to the in-plant electricity use by equipment such as pumps, blowers, and fans, the net electricity output decreased to 492.9 MW. In accordance, the net thermal efficiency, i.e., the conversion of the total heat input to the net electricity output, was 37.4%.

FIG. 6 summarizes the heat distribution of the power generation system derived from the above mass and energy balance calculations. The result shows that only 41% of the total heat input was converted to usable energy forms, particularly 40% to electricity through the turbines and 1% to work through the feed-water pump. A large amount of the heat (47%) was wasted in the exhaust steam from the LP turbine, since the low temperature of

the steam (108 °F), close to the ambient, made it non-usable in the current plant. This heat was released in the exhaust steam condenser and finally dissipated into the atmosphere through the heat exchange between the cooling water and the air.

5 Since the heat contained in steam can vary in quality, it is meaningful to see how such quality changes with the steam conditions. FIG. 7 shows the change of steam properties and its capability of generating electricity when it passes through the IP and LP turbines. Here, an adiabatic expansion process was assumed, and the expander had an efficiency of 88.5%. During the
10 expansion, the capability of the steam to generate electricity decreases logarithmically with the decrease of steam pressure. At the IP inlet (1000 °F, 545.4 psia) about 32% of the total heat contained in the steam can be converted to electricity, while little or no such heat can be converted to electricity at the outlet of the LP turbine (107.8 °F, 1.2 psia). However, according to
15 embodiments of the present invention, the large amount of heat contained in the exhaust steam may be used as an energy source for a CO₂ separation process if appropriate stripping conditions in a CO₂ absorption process are designed. Examples of such conditions are explained below with regard to the exemplary vacuum-carbonate process.

20 MEA process

The simulation of the MEA process was based on the flue gas leaving the wet FGD from the reference plant. Since the flue gas was cooled to 129 °F through the FGD absorber, no further cooling was assumed before the flue gas entered the MEA process. Again, the inlet flue gas contained
25 13.9%vol CO₂, equivalent to a partial pressure of 2.3 psia.

The results of the process mass and energy balance calculation are shown in FIG. 8. 90% of the CO₂ in the flue gas was removed in the absorption column. The cleaned flue gas contained 74.1%vol N₂, 21.7% water vapor, 2.6% O₂, and 1.6% CO₂. MEA regeneration from the rich solution
30 required a heat duty of about 1,400 MMBtu/hr in the stripper reboiler, which is

equivalent to 1,680 Btu/lb CO₂ removed. The gas stream from the stripper contained 58.7%vol CO₂ and 41.3% H₂O. The stream was cooled to 104 °F to condense water vapor, and the CO₂ was concentrated (95.7%vol CO₂ and 4.3% H₂O).

5 The steam used for MEA regeneration was extracted from the power plant. The steam condition was determined based on the required temperature for the stripper. As most of the heat from steam was stored as the latent heat that would be released during condensation, the saturation temperature of the steam was kept above the solution temperature at the bottom
10 of the stripping tower. In addition, a temperature difference of 40-50 °F is typical for a stripper reboiler in the MEA process. As a result, steam with a saturation temperature of about 290 °F was needed for the reboiler. The corresponding saturation pressure of the steam was about 60 psia (FIG. 7). Since the steam extracted from the exit of the IP turbine was at 175 psia, it was
15 therefore expanded to 60 psia with a temperature of 475 °F. This steam was further de-superheated prior to entering the reboiler. Part of the electricity was recovered during the steam expansion. The steam was then used to heat the MEA solution and was condensed itself in the reboiler. The condensate from the reboiler was returned to the feed water system in the power plant steam
20 cycle. In this simulation, it was assumed that no steam was directly introduced to the stripper to strip CO₂.

 The extraction of steam resulted in a large reduction in electricity generation. With more than half the total steam extracted from the exit of the IP turbine, the gross electricity generation was reduced by 89.4 MW (power
25 recovery was considered). Another large electricity loss was caused by CO₂ compression. The assumed four stage inter-stage cooling compressor consumed a total of 35.4 MW of electricity to compress all the captured CO₂ from 25 to 1200 psia. In addition, 10.0 MW was required for the flue gas blower to overcome a pressure drop of 2 psia in the MEA absorber, and 1.8 MW for the
30 MEA solution pumps. However, the other auxiliary power use in the PC+MEA plant was slightly less than the reference plant. This is because the steam

extraction demanded by the MEA process reduced the loading of the exhaust steam condenser and cooling tower in the power plant, which in turn reduced the electricity usage. The results of the electricity use/loss are summarized in FIG. 9. Overall, the net electricity output in the PC+MEA power plant was
5 reduced by 134 MW, equivalent to a loss of 27.2% of the net electricity output from the reference PC plant.

Vacuum carbonate process

The CO₂ and water partial pressures in the vapor over a 20%wt K₂CO₃ solution at equilibrium were calculated and are shown in FIG. 10. The
10 predicted CO₂ partial pressure agreed well with the experimental data at 158 and 194 °F reported by Tosh, J.S., Field, J.H., Benson, H.E., Haynes, W.P., Equilibrium Study of the System Potassium Carbonate, Potassium Bicarbonate, Carbon Dioxide, and Water, United States Bureau of Mines, Report of Investments, No. 5484, 1959, while the predicted water vapor pressure only
15 slightly deviated upward at high CO₂ loadings. The equilibrium data showed that at a temperature range of 104-194 °F, the total pressure of CO₂ and water vapor changed from about 1-9 psia when CO₂ loading in the solution was kept at 1% wt at the exit of the stripper.

When the absorption was operated around the atmospheric
20 pressure, the CO₂ partial pressure decreased from about 2.3 to 0.23 psia to achieve 90% removal efficiency. Assuming that the gas phase CO₂ and the absorption solution were at equilibrium at the bottom of the column, then the loading of CO₂ in the CO₂ rich solution ranged from 2.6%wt to 4.2%wt in the temperature range of 194-104 °F (FIG. 10). Compared with the MEA solution,
25 the K₂CO₃ solution has a lower CO₂ absorption capacity, and a higher liquid flow rate is needed to achieve the same CO₂ removal.

FIG. 11 shows operation of an exemplary vacuum-absorption process according to this exemplary embodiment of the present invention. The low-pressure steam used in the stripper was extracted from the LP turbine,
30 typically at pressure ranging from 3-9 psia. These pressures are much lower

than the steam used in the MEA process, where the steam pressure was 60 psia. In the exemplary vacuum-carbonate process, most of the steam is directly introduced into the stripper for stripping and heating. Only a small part of the steam (e.g., about 1/3) was used for indirect heating in the reboiler. This is quite different from the MEA process, for example, where the majority of the steam is introduced to the reboiler at the bottom of the stripper for indirect heating.

As a nonlimiting example, mass and heat balances for the vacuum process operated at 3 psia (at the top of the stripper) were listed in FIG. 11. The calculations were based on a 20% K_2CO_3 solution and 1%wt of CO_2 loading in the lean solution. The liquid to gas mass ratio (L/G) used in this example was 1.2 times the minimum L/G ratio, $(L/G)_{min}$. This ratio (7.92) roughly doubled that of the MEA process (3.86). The results showed that about 1.6 million lb/hr steam were consumed for CO_2 regeneration, of which 62% was directly introduced into the stripper, and 38% went to the reboiler. A temperature difference (ΔT) of 20 °F was adopted for the heat transfer in the reboiler, since the lower heat duty (627 MMBtu/hr) of the vacuum process gives some flexibility for the reboiler. For comparison, the heat duty of the MEA process is 1,400 MMBtu/hr.

The gas stream from the top of the stripper was a mixture of 24.1%vol CO_2 and 75.9%vol water vapor. The majority of the water vapor in the stream was condensed in the water condenser, and the condensate was mixed with the condensate from the reboiler and returned to the steam cycle of the power plant. After the primary condensation, the gas stream included 91.7%vol CO_2 and 8.3% water vapor. The temperature of the mixed condensate was about 82 °F, a temperature close to the exhaust steam condenser in the power plant. In an exemplary real (not simulated) process, the condensate from the CO_2 /water vapor stream, containing 388 ppm CO_2 in water, may need treatment to meet the water quality requirement of the steam cycle. Exemplary treatments will be apparent to those of ordinary skill in the art.

In the above calculations, the amounts of steam used in the stripper and reboiler were optimized to maintain a water balance in the absorption-desorption loop. Introduction of excessive steam into the stripper would result in water accumulation in the process. However, indirect heating through the reboiler would reduce the thermal efficiency of the steam usage, since a temperature difference (ΔT) is necessary. For example, an increase of ΔT to 40 °F could lead to an additional 4 MW electricity loss compared to the ΔT of 20 °F used in this calculation. Due to this fact, as much of the steam as possible was directly introduced to the stripper to reduce or eliminate the ΔT and reduce the reboiler heat duty.

Based on the results shown in FIG. 12, the thermal efficiencies of the power plants equipped with the vacuum carbonate process and the MEA process were compared. The results of the comparison are shown in FIG. 13. The electricity needed for compressing the CO₂ to 1200 psia was also included. The different electricity use for CO₂ compression between the two processes was due to the different starting pressures (14.7 and 25 psia, respectively). The steam extraction loss in the vacuum carbonate process was as low as about 38 MWe, of which 52% was due to the stripping steam and 48% due to the reboiler steam. However, in the vacuum carbonate process, a significant amount of electricity was consumed by the vacuum pump to pressurize the CO₂ stream from the stripping pressure to one atmospheric pressure (14.7 psia). The exemplary vacuum carbonate process also required large pumps to circulate the absorption solution due to its higher L/G ratio. The thermal efficiencies of the two processes were 29.6% and 27.2%, respectively. Overall, about 104 MWe in total was consumed for CO₂ capture in the vacuum carbonate process, compared to 137 MWe in the MEA process, an energy saving of about 24%.

All of the above calculations were conducted at a stripping pressure of 3 psia, a CO₂ lean loading of 1.0%wt, and an L/G ratio of 1.2 (L/G)_{min}. Since these three parameters are significant operating parameters for

the vacuum carbonate process, their impacts on energy performance were investigated in more detail.

Effect of stripping pressure

The effect of the stripping pressure on steam extraction loss and electricity use of the vacuum pump, the two most energy-consuming components in the process, were examined for a stripping pressure range of 2-8 psia. The steam extraction loss was calculated based on the amount of steam expanded from the extraction pressure to 1.2 psia. The electricity use by the vacuum pump was calculated based on the process to compress the CO₂ stream to one atmospheric pressure. Since the absorber and vacuum stripper were operated at similar temperatures (closer at a lower stripping pressure), the heat duty of the heat exchanger for the CO₂ lean and rich solutions was significantly reduced compared to the MEA process. The ΔT between the lean solution inlet and rich solution outlet was thus selected as 10, 15, and 20 °F for the process simulations at stripping pressures of 3, 4, and 8 psia, respectively, and no lean-rich solution heater exchanger was required at 2 psia.

The process simulations showed that the steam extraction loss increased with increasing stripping pressure to a power of 0.59, while the electricity use of the vacuum pump decreased with increasing operating pressure to a power of -1.07 (FIG. 12). Overall, the total electricity loss increased with increasing stripping pressure to a power of 0.29. The total electricity loss changed from 50 to 74 MW as the stripping pressure increased from 2-8 psia at an L/G ratio of 1.2 (L/G)_{min}. This corresponds to about 56% - 82% of the steam extraction loss of the MEA process.

Only considering the energy consumption for the selected operating pressure range, it is expected that a lower stripping pressure will allow the use of lower quality steam, which will result in a higher thermal efficiency of the power plant. On the other hand, the lowest operating pressure has a boundary, since the steam condensation is limited by the cooling water temperature, which depends on the water sources and the season as well. The

lowest stripping pressure is also limited by the fact that a lower stripping pressure will lead to a larger stripper, which in turn increases the capital cost. An optimal stripping pressure should thus exist, at which pressure the best economic performance can be achieved.

5 Effect of CO₂ loading in the lean solution

CO₂ loading in the lean solution affects the equilibrium vapor pressure of CO₂ in the gas phase, and thus impacts the minimum liquid-to-gas ratio, $(L/G)_{\min}$, in the absorption column. A lower loading of CO₂ in the lean solution favors CO₂ absorption, resulting in a reduced $(L/G)_{\min}$, and thus the
10 real L/G ratio, to achieve the same level of CO₂ absorption (FIG. 14). The CO₂ loading in the rich solution only changed slightly with decreasing CO₂ loadings in the lean solution.

The mass and energy balance calculation showed that a lower lean CO₂ loading at the stripper bottom shifted the temperature profile upward
15 in the column compared to a higher CO₂ loading. For example, at a stripping pressure of 3 psia, the temperatures at the bottom and top of the column increased by about 3 °F and 10 °F, respectively, when the lean CO₂ loading varied from 1.5 to 0.5%. Therefore, more heat was required to provide for greater water vaporization and the greater heat duty of the reboiler. Although
20 the L/G ratio decreased with decreasing lean CO₂ loading, the resulting reduction of sensible heat was relatively not significant compared to the increased requirement for water latent heat. As a result, a decrease of CO₂ loading from 1.5% to 0.5% almost doubled the steam extraction loss (FIG. 15). While a lower CO₂ loading is not economic, it can also be noted that a CO₂
25 loading above 1.5% would not greatly improve the energy use performance of the process.

Effect of L/G ratio

The L/G ratio affects both the treated gas purity and the heat requirement of the absorption process. In any chemical absorption process, the
30 L/G ratio typically ranges from 1.1-1.5 times the $(L/G)_{\min}$. The L/G ratios

examined in this experiment were 1.0, 1.2, and 1.5 times $(L/G)_{\min}$. The impact on the heat requirement of the stripper is shown in FIG. 16.

At 90% CO₂ removal and a 1% CO₂ lean loading in solution, the CO₂ rich loading at the stripper inlet decreased from 3.8% to 2.8% as the L/G increased from 1.0 to 1.5 times $(L/G)_{\min}$. The temperature at the stripper bottom was related only to the CO₂ lean loading in the equilibrium calculation, and thus remained the same. A lower CO₂ loading in the rich solution required a slightly higher temperature in the stripping column to drive out the CO₂ from solution. Therefore, more water vaporization was expected at a higher L/G ratio. On the other hand, a higher L/G ratio required more sensible heat for heating the solution. However, as mentioned above, the sensible heat contributed less than 20% of the total heat requirement for operations at 1% CO₂ loading, and therefore was not considered significant. Combining the increased heat requirement for water vaporization and solution heating, the electricity loss due to steam use increased by about 45% for a change in the L/G ratio from 1.0 to 1.5 times $(L/G)_{\min}$.

Preliminary cost estimation

A preliminary cost comparison was made between the MEA process and the vacuum-carbonate process (FIG. 17). The IECM model (e.g., as described in Singh, D., Croiset, E., Douglas, P.L., Douglas, M.A., Techno-Economic Study of CO₂ Capture from An Existing Coal-Fired Power Plant: MEA Scrubbing vs. O₂/CO₂ Recycle Combustion, Energy Conversion and Management 2003, 44, 3073-3091) was employed to estimate the cost of the MEA process, based on default financial assumptions built into the model and the process parameters input from this experiment. The resulting breakdown costs for the MEA process were comparable to those previously reported (e.g., see Rao, A.B., Rubin, E.S, Identifying Cost-Effective CO₂ Control Levels from Amine-Based CO₂ Capture Systems, Ind. Eng. Chem. Res. 2006, 45, 2421-2429). The cost analysis of the exemplary vacuum-carbonate process was based on a pressure of 3 psia at the top of the stripper, a 20wt% K₂CO₃ solution,

a lean loading of 1% CO₂ in the solution, and an L/G ratio of 1.2 times (L/G)_{min}. For a preliminary estimation, the capital cost of the vacuum process was assumed 50% greater than the MEA process, allowing for a larger gas volume under a vacuum condition and the additional expenditure for the vacuum pump.

5 The fixed O&M cost was assumed to be proportional to the capital cost. The energy cost in the vacuum carbonate process was obtained by comparing the electricity loss/use (104 MW) to that of the MEA process (137 MW). Because there are no degradation/polymerization problems in the process, the carbonate solvent is only lost through stoichiometric reaction with SO₂. The resulting
10 replacement cost of solvent was estimated based on a mass-based price of 1/6 of the cost of MEA. The other variable cost in the vacuum carbonate process was the use of corrosion inhibitor, which was assumed as 20% of the solvent cost. The cost results showed that a significant amount of the cost for the vacuum carbonate process was contributed by the capital investment. Due to
15 its lower energy requirement and lower solvent loss, the variable costs for the vacuum process were nearly half those of the MEA process. As a result, the CO₂ avoidance cost of the vacuum carbonate process was 18% less than the MEA process. Additional information on kinetic-based equipment sizing would be useful to obtain more detailed cost estimates.

20 An improved process for the amine-based CO₂ capture process was developed by Mitsubishi Heavy Industries and the Kansai Electric Power Company. This process employs the KS-1 solvent, an undisclosed amine with steric hindrance. The energy use for CO₂ capture (not including CO₂ compression) was reported to be 195 kWh/ton CO₂. The total cost is reduced
25 by 20% compared to the MEA process. It can be seen that the techno-economic performance of the exemplary vacuum-carbonate process in this simulation is comparable to the amine-based process where KS-1 is used.

Conclusions

30 About half of the steam generated in the power plant at a typical operating condition is extracted by the exemplary vacuum-carbonate process,

among which 60% is directly introduced into the stripper. Because of this steam extraction, the steam condenser in the reference PC power plant will have excess capacity (about 40%). When the vacuum-carbonate process is retrofitted to an existing power plant, the excess capacity of the existing
5 condenser can be used to condense the CO₂/water stream from the stripper. Since the operating conditions are similar for the two condensation steps, the exhaust steam condenser in the existing power plant may be used with minimal modifications. If the vacuum-carbonate process is used in a new power plant, then more than half of the exhaust steam condenser can be eliminated, and
10 instead a CO₂/water condenser should be added. The resulting plant integrates the exemplary vacuum-carbonate absorption process and the power plant steam cycle. Such integration is not possible for an MEA process.

The potassium carbonate solution used in this modeling experiment has a slower chemical reaction rate with CO₂ than the MEA
15 solution. At a flue gas temperature of around 122-140 °F (downstream of a wet FGD), the chemical reaction rate could be slow and become the rate-limiting step for the absorption process. As a result, the absorber for the vacuum process should be larger than the absorber for the MEA process. To reduce the absorber size, developing a good promoter/activator is a significant factor for a
20 preferred vacuum-carbonate process. Many promoters have been tested and some have been commercialized. Nonlimiting examples may be found in Kohl, A.L., Riesenfeld, F.C., Gas Purification (Fourth Edition), Gulf Publishing Company, Houston, Texas, 1985.

Some earlier studies achieved major improvements for the hot
25 potassium process. For example, addition of sodium or potassium arsenite as an activator in K₂CO₃ solution has been shown to increase the rate of CO₂ absorption and desorption by about 4 times, resulting in appreciable improvements in regeneration heat, equipment size, and treated gas purity. Another example is the Catacarb potassium carbonate process, where an
30 undisclosed additive was used to catalyze absorption and desorption of the acid gases, especially of CO₂. Recently, Cullinane and Rochelle demonstrated that

a 20 wt% K_2CO_3 solution promoted with 0.6 M piperazine (PZ) could achieve absorption rates comparable to those in 5M MEA at temperatures of 40-80 °C. More concentrated solvents, such as 3.6 M K^+ /1.8M PZ, had absorption rates a factor of 1.5 higher than 5M MEA. Thus, it is possible to promote the absorption rate of the carbonate solution to a magnitude comparable to that of an MEA solvent.

Another potential issue is that vacuum in the stripper will significantly increase the gas stream volume especially when the vacuum is high. At 3 psia, the total CO_2 volume will be more than 600,000 acfm for a 500 MW power plant. At a 5m/s gas velocity, the total cross section area at the inlet of the vacuum pump will be about 600 ft². Clearly, too high a vacuum will make the capital cost of the vacuum pump high.

The concentration of the carbonate solution is limited by $KHCO_3$ dissolution. Higher K_2CO_3 concentration in solution is possible based on the typical operating conditions employed in this process. For example, at 130 °F and 50% conversion of K_2CO_3 to $KHCO_3$, the K_2CO_3 concentration in solution can reach about 35%. A higher K_2CO_3 concentration in solution will increase the working capacity (difference of lean and rich CO_2 loadings) and decrease the L/G requirement. If a 30% K_2CO_3 solution is used, the energy use and cost will be further reduced, compared to the 20% K_2CO_3 solution used in the above process simulation and cost analysis.

In sum, process simulations showed that, when an exemplary vacuum-carbonate process is operated at a stripping pressure of 3 psia, a 1.0% CO_2 lean loading and an L/G at 1.2 (L/G)_{min}, the electricity loss due to steam extraction is only about 42% of that in the MEA process. The total electricity loss of the vacuum-carbonate process, including the CO_2 compression, is about 76% of that of the MEA process.

The impacts on the energy performance of the vacuum-carbonate process were investigated for three process parameters: stripping pressure, CO_2 loading, and L/G ratio. Increasing the stripping pressure resulted in an increase of steam usage and quality, and a decrease of electricity use for the vacuum

pump. Overall, as stripping pressure rose from 2-8 psia, the total electricity loss increased at a power of 0.22. A decrease of CO₂ lean loading from 1.5% to 0.5% almost doubled the steam extraction loss. However, a CO₂ lean loading higher than 1.0% did not significantly improve the energy performance
5 of the process. Increasing the L/G ratio from 1.0-1.5 times the $(L/G)_{min}$ increased the electricity loss due to steam extraction by 45%.

The preliminary cost analysis showed that the vacuum-carbonate process is economically attractive for post-combustion CO₂ capture with an 18% reduction of the CO₂ avoidance cost compared to the MEA process.

10 While various embodiments of the present invention have been shown and described, it should be understood that other modifications, substitutions, and alternatives are apparent to one of ordinary skill in the art. Such modifications, substitutions, and alternatives can be made without departing from the spirit and scope of the invention, which should be
15 determined from the appended claims.

Various features of the invention are set forth in the appended claims.

CLAIMS

1. A method for separating a targeted gas from a gas stream emitted from a power plant, the method comprising:

bringing the gas stream into contact with an absorption solution
5 to preferentially absorb the targeted gas to be separated from the gas stream so that an absorbed gas is present within the absorption solution, providing a gas-rich solution;

introducing the gas-rich solution into a stripper;

injecting low pressure exhaust steam from a low pressure steam
10 turbine of the power plant into the stripper with the gas-rich solution;

stripping the absorbed gas from the gas-rich solution in the stripper using the injected low pressure steam within the stripper to provide a gas stream containing the targeted gas, wherein the stripper is at or near vacuum;

15 condensing water vapor in a gas stream from the stripper in a condenser operating at a pressure lower than the stripper to concentrate the targeted gas;

separating condensed water from the concentrated targeted gas.

20 2. The method of claim 1, wherein the stripper during said stripping is at a pressure between 1.2 and 14.7 psia, and a temperature $T=40\sim 100^{\circ}\text{C}$.

3. The method of claim 2, wherein the low pressure exhaust
25 steam is between 1.2 and 14.7 psia, and a temperature $T=40\sim 100^{\circ}\text{C}$.

4. The method of claim 1, wherein the stripper during said stripping is at a pressure between 3 and 7 psia, and a temperature $T=60\sim 85^{\circ}\text{C}$, and wherein the low pressure exhaust steam is at a pressure between 3 to 7 psia
30 and a temperature $T=60\sim 85^{\circ}\text{C}$.

5. The method of claim 1, wherein the low pressure turbine feeds into a condenser in the power plant, and wherein the low pressure steam is obtained from the low pressure turbine for said injecting before the low pressure steam reaches the condenser in the power plant.

5

6. The method of claim 1, wherein said injecting comprises injecting the low pressure exhaust steam directly into the stripper.

7. The method of claim 1, wherein the power plant includes a plurality of turbines leading to a condenser in the power plant, and wherein the low pressure turbine is the lowest pressure turbine before the condenser in the power plant.

8. The method of claim 1, further comprising:
15 using a heat exchanger to transfer heat from the low pressure steam to the absorption solution in a reboiler at the bottom of the stripper.

9. The method of claim 1, further comprising:
recycling the condensed water to a power generation system of
20 the power plant.

10. The method of claim 1, further comprising:
before said separating, cooling the concentrated gas stream from the condenser and condensing additional water in the gas stream;

25

11. The method of claim 1, wherein said separating the condensed water comprises using a gas-liquid separator.

12. The method of claim 1, further comprising:
30 using a vacuum pump to draw the concentrated gas from said separating to a compressor;

compressing the concentrated gas to a selected pressure.

13. The method of claim 1, wherein the targeted gas is CO₂, and the gas stream is flue gas from the power plant.

5

14. The method of claim 13, wherein the absorption solution comprises an aqueous solution including at least one salt.

15. The method of claim 14, wherein the absorption solution
10 comprises an aqueous solution including at least one of K₂CO₃, Na₂CO₃, K₃PO₄, Na₃PO₄, and amino-acid salts.

16. The method of claim 1, wherein the targeted gas is O₂, and the gas stream is an oxygen-containing gas mixture.

15

17. The method of claim 1, wherein said bringing comprises bringing the gas stream into at least one of a membrane contactor and an integrated membrane system, wherein the absorption solution permeates through a membrane side, and a permeate side operates substantially under
20 vacuum.

18. The method of claim 1, wherein the power plant comprises at least one of a coal-firing power plant, an integrated gasification combined cycle power plant, and a natural gas combined cycle power plant.

25

19. The method of claim 1, further comprising:
bringing the stripper at or near vacuum at least partially by condensing water vapor in the gas stream from the stripper.

30

20. The method of claim 1, further comprising:

adding a promoter to the absorption solution to accelerate an absorption rate.

21. The method of claim 20, wherein the promoter comprises
5 at least one of piperazine, monoethanolamine, an organic base chemical, and an inorganic base chemical.

22. The method of claim 1, further comprising:
adding a chemical to the absorption solution to lower a vapor
10 pressure of the absorption solution.

23. The method of claim 22, wherein the added chemical comprises at least one of a water soluble chemical and a salt.

15 24. A system for separating a targeted gas from a gas stream emitted from a power plant, the method comprising:

an absorber in fluid communication with the power plant for bringing the gas stream into contact with an absorption solution to preferentially absorb the targeted gas to be separated from the gas stream so
20 that an absorbed gas is present within the absorption solution, providing a gas-rich solution;

a stripper in fluid communication with said absorber that receives the gas-rich solution and strips the absorbed gas from the gas-rich solution to provide a gas stream containing the targeted gas via an outlet, said stripper
25 being in fluid communication with a low pressure steam turbine of the power plant to receive injected low pressure steam;

a vacuum source for providing a pressure within the stripper that is at or near vacuum;

a stripper reboiler for providing heat to said stripper;
30 a condenser in fluid communication with the outlet of the stripper for condensing water vapor in the gas stream from the stripper, said condenser

being configured to operate at a pressure lower than the stripper to concentrate the targeted gas;

a gas-liquid separator in fluid communication with said condenser for separating condensed water from the concentrated targeted gas.

5

25. The system of claim 24, wherein said vacuum source is configured to provide a pressure within said stripper that is between 1.2 and 14.7 psia.

10

26. The system of claim 24, wherein the low pressure steam turbine feeds into said stripper before it feeds into a condenser in the power plant.

15

27. The system of claim 24, wherein the low pressure steam turbine is coupled to said stripper such that the low pressure exhaust steam is injected directly into the stripper.

20

28. The system of claim 24, wherein the power plant includes a plurality of turbines leading to a condenser in the power plant, and wherein the low pressure turbine is the lowest pressure turbine before the condenser in the power plant.

25

29. The system of claim 24, further comprising:
a heat exchanger that transfers heat from the low pressure steam to the absorption solution in said stripper reboiler.

30

30. The system of claim 24, further comprising:
a cooler disposed between said condenser and said gas-liquid separator for cooling the concentrated gas stream from the condenser and condensing additional water in the gas stream.

31. The system of claim 24, further comprising:
a compressor for compressing the concentrated gas to a selected
pressure;

5 a vacuum pump that draws the concentrated gas from said gas-
liquid separator to said compressor.

32. The system of claim 24, wherein the targeted gas is CO₂,
and the gas stream is flue gas from the power plant.

10 33. The system of claim 32, wherein the absorption solution
comprises an aqueous solution including at least one salt.

34. The system of claim 24, wherein the targeted gas is O₂,
and the gas stream is an oxygen-containing gas mixture.

15 35. The system of claim 24, wherein said absorber comprises
at least one of a membrane contactor and an integrated membrane system
wherein the absorption solution passes from a retentate side to a permeate side,
wherein said stripper comprises the permeate side, and wherein the permeate
20 side operates substantially under vacuum.

36. The system of claim 24, wherein the power plant
comprises at least one of a coal-firing power plant, an integrated gasification
combined cycle power plant, and a natural gas combined cycle power plant.

25 37. The system of claim 24, wherein said vacuum source
comprises said condenser, wherein said condenser brings said stripper at or
near vacuum at least partially by condensing water vapor in the gas stream
from said stripper.

30

38. The system of claim 24, wherein the absorption solution includes a promoter to accelerate an absorption rate.

39. The system of claim 24, wherein the absorption solution
5 includes at least one chemical for lowering a vapor pressure of the absorption solution.

40. A gas separation method for separating a targeted gas from a gas stream of a power plant, the method comprising:
10 within an absorption device, bringing the gas stream into contact with an absorption solution to preferentially absorb the targeted gas;
stripping the absorbed gas in the absorption solution within a vacuum stripper operating at absolute pressures between 1.2 to 14.7 psi and corresponding temperatures between 40 to 100 °C;
15 obtaining a low pressure exhaust steam from a lower pressure steam turbine of the power plant operating at pressures between 1.2 to 14.7 psi (T= 40~100 °C);
injecting the obtained low pressure exhaust steam directly into the stripper;
20 using the injected low pressure steam at pressures between 1.2-14.7psi (T= 40~100 °C) to strip the gas in the stripper;
transferring heat from the low pressure steam to the absorption solution in a reboiler at the bottom of the stripper using a heat exchanger;
condensing the water vapor in the gas stream from the stripper in
25 a condenser that operates at a pressure lower than the stripper to provide condensed water and concentrate the gas;
recycling the condensed water, with or without further treatment, to a power generation system of the power plant;
cooling the concentrated gas stream from the condenser and
30 condensing additional water in the gas stream;
separating the condensed water in a gas-liquid separator;

using a vacuum pump to draw the concentrated gas from the gas-liquid separator to a compressor; and
compressing the concentrated gas to a desired pressure.

1/12

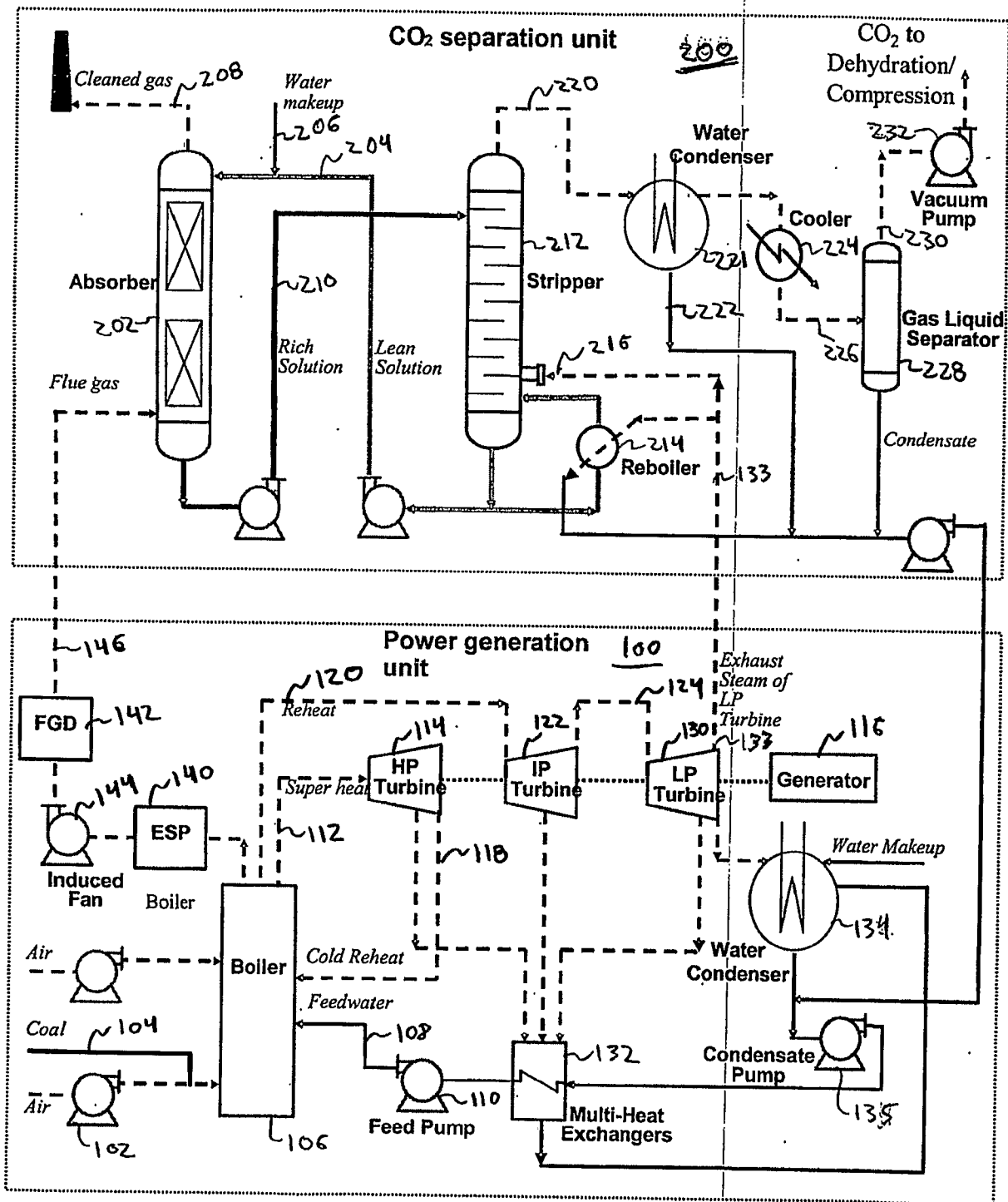


FIG. 1

2/12

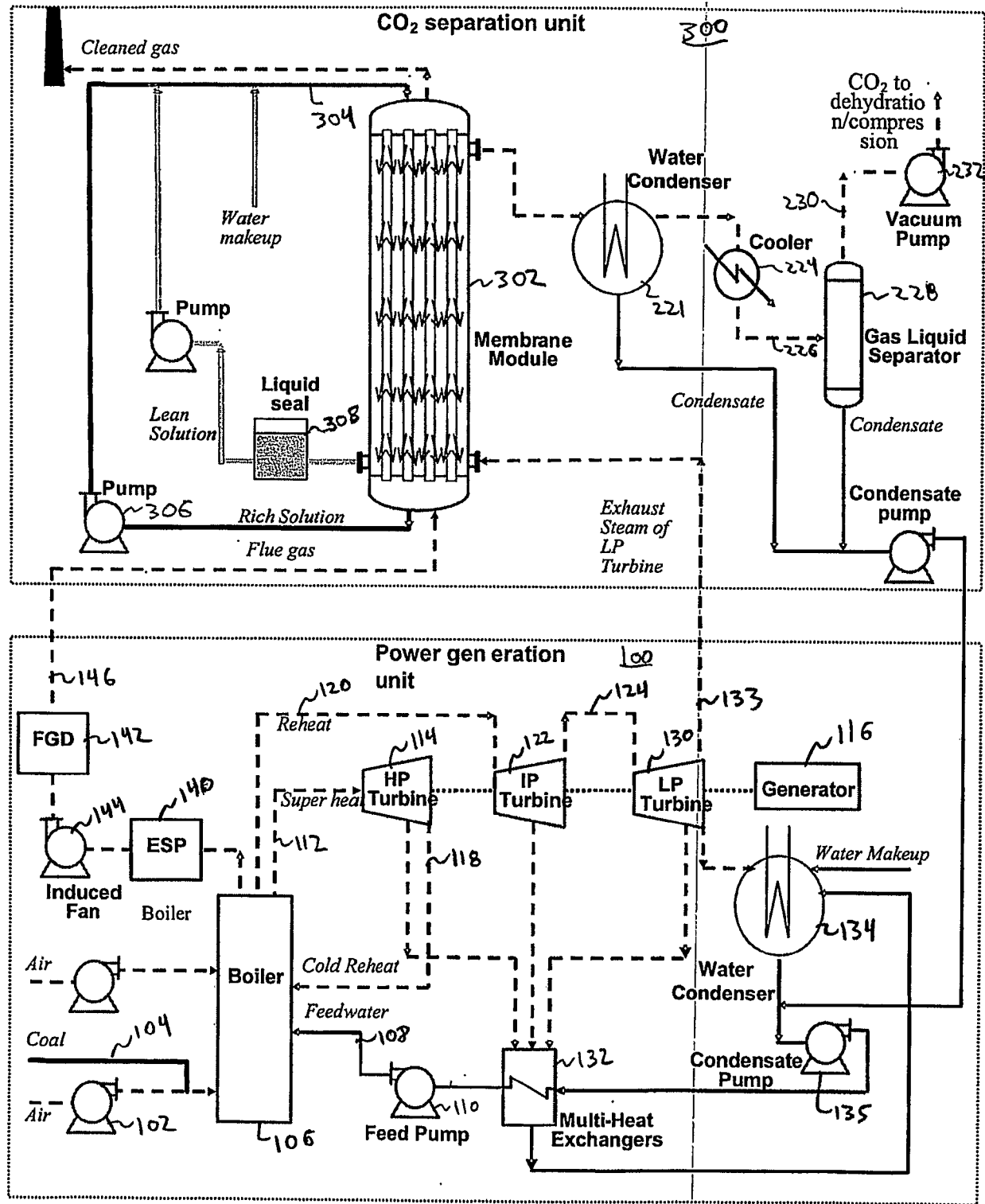


FIG. 2

3/12

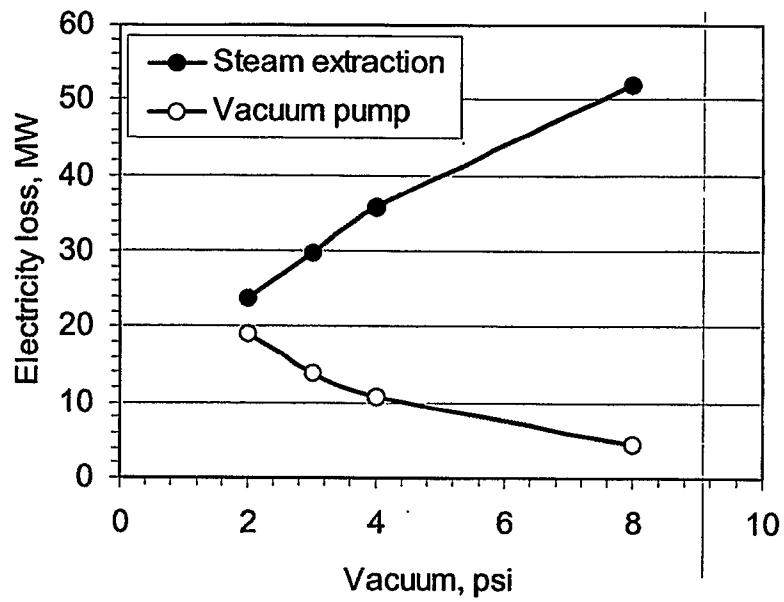


FIG. 3

PC power plant		
Gross output, MWe	528	
Unit type	Sub-critical PC	
Coal	IL bituminous	
Excess Air, %	15%	
Gas temp. exiting air preheater, °F	295	
Main steam condition	2,415 psia/1000 °F	
Reheat steam condition	545 psia/1000 °F	
Turbine efficiency, %	88.5%	
SO ₂ removal, %	98% (Limestone Forced Oxidation)	
Particle removal, %	99% (ESP)	
CO ₂ capture and compression		
	MEA process	K ₂ CO ₃ process
Solvent concentration, %wt	30% MEA	20% K ₂ CO ₃
Pressure drop in absorber, psia	2	2
Temp. of flue gas entering absorber, °F	129	129
Temp. of solvent entering absorber, °F	104	122
Liquid to gas ratio, lb/lb	3.86	1, 1.2, 1.5 (L/G) _{min}
Lean CO ₂ loading, %wt	5.5%	0.5%, 1%, 1.5%
Pressure at the top of stripper, psia	25	2 ,4, 6, 8
Pressure drop in stripper, psia	2	1
Vacuum pump efficiency, %	-	85%
CO ₂ removal, %	90%	90%
Compressor efficiency, %	82% (4-stage, inter-stage cooling)	
Inter-stage cooling temperature, °F	104	
Compression end pressure, psia	1,200	

FIG. 4

Greer Burns & Crain, Ltd.
Reference No.: 1201.75142PCT
Sheet 5 of 12

(Steven P. Fallon)
(312) 360-0080

5/12

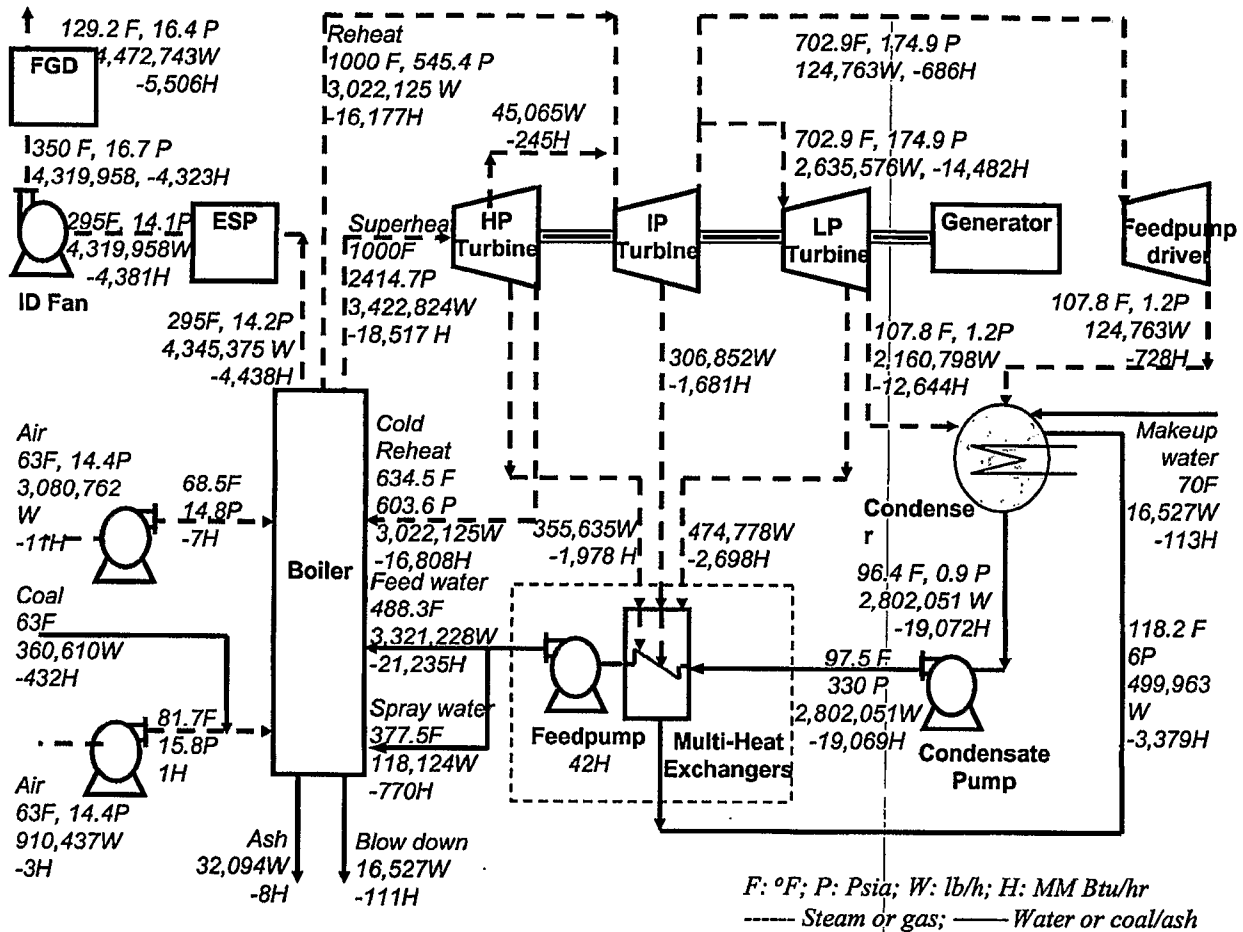


FIG. 5

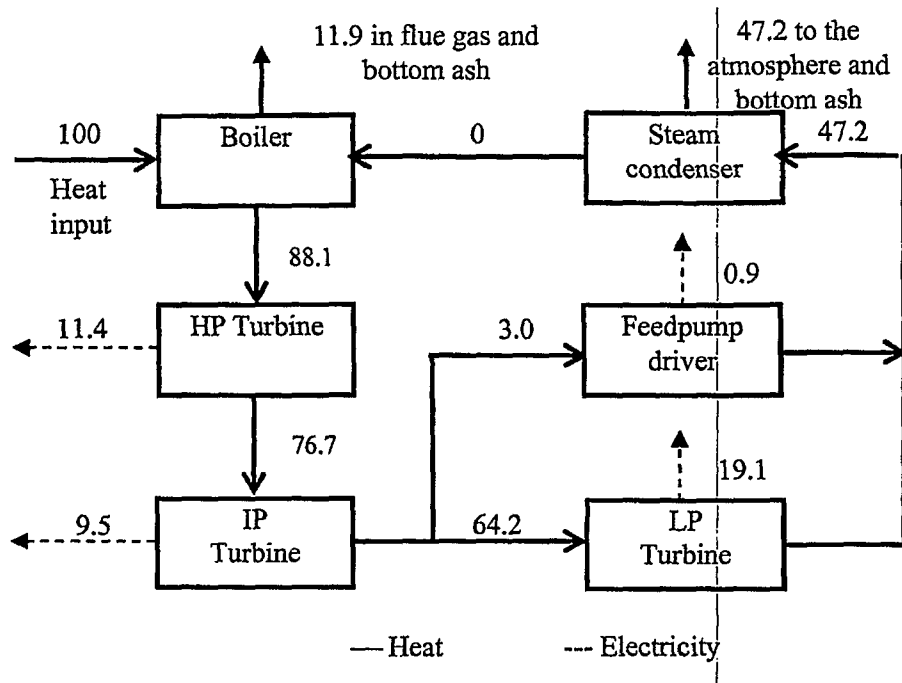


FIG. 6

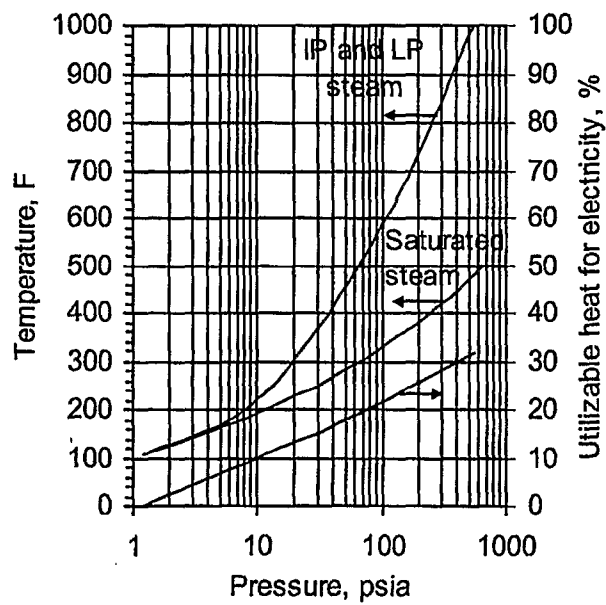


FIG. 7

7/12

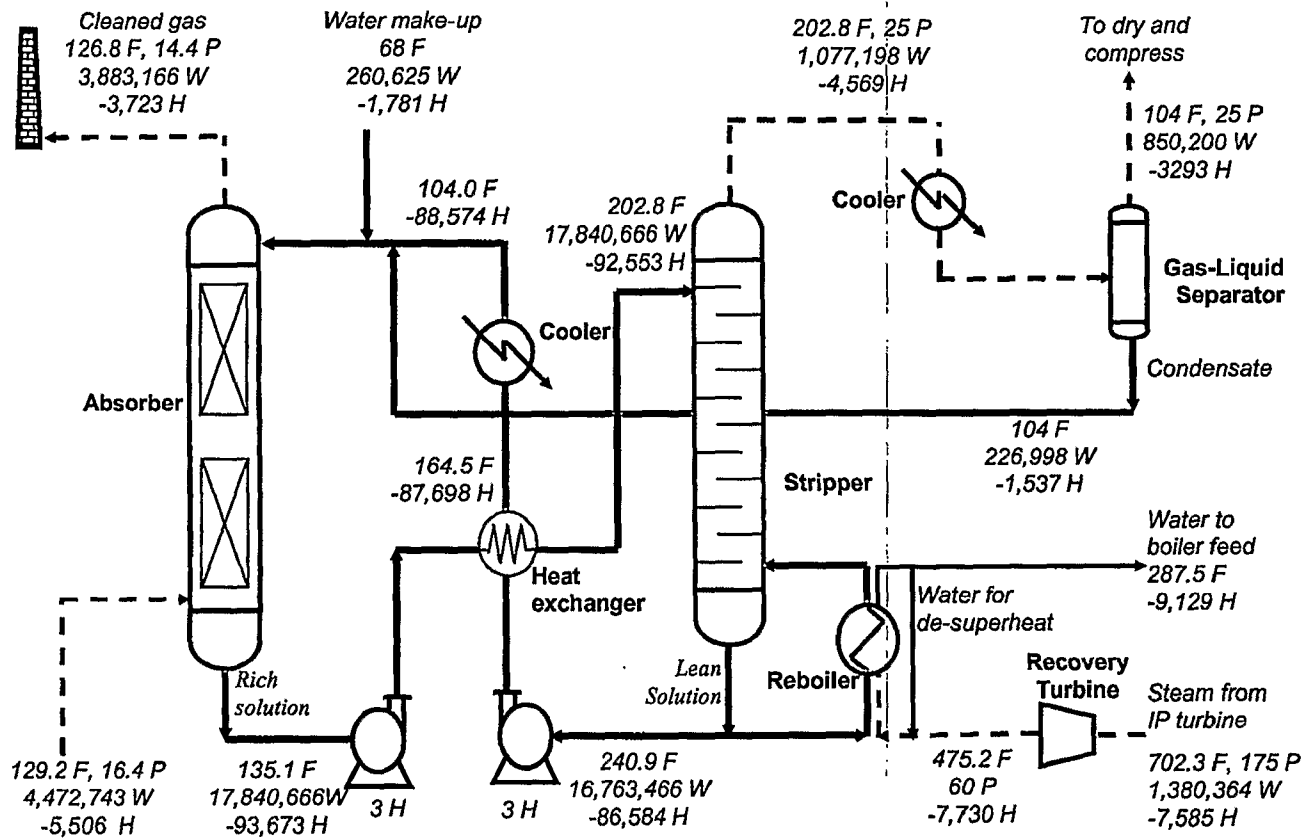


FIG. 8

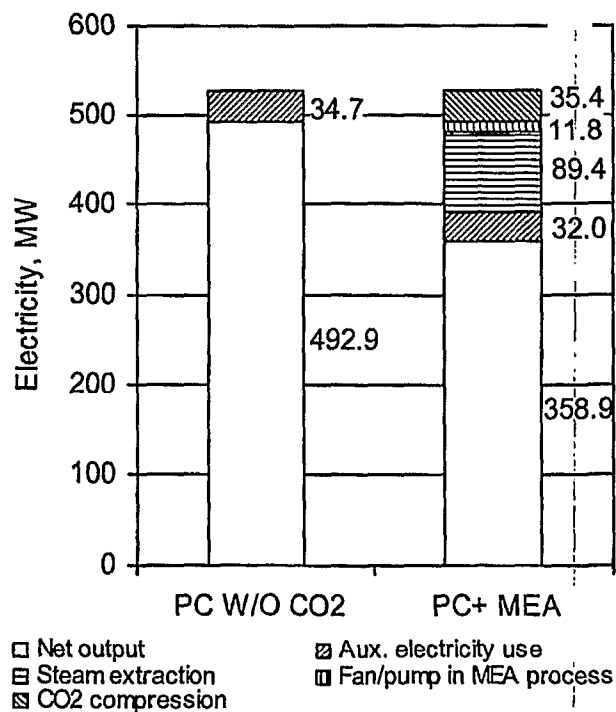


FIG. 9

8/12

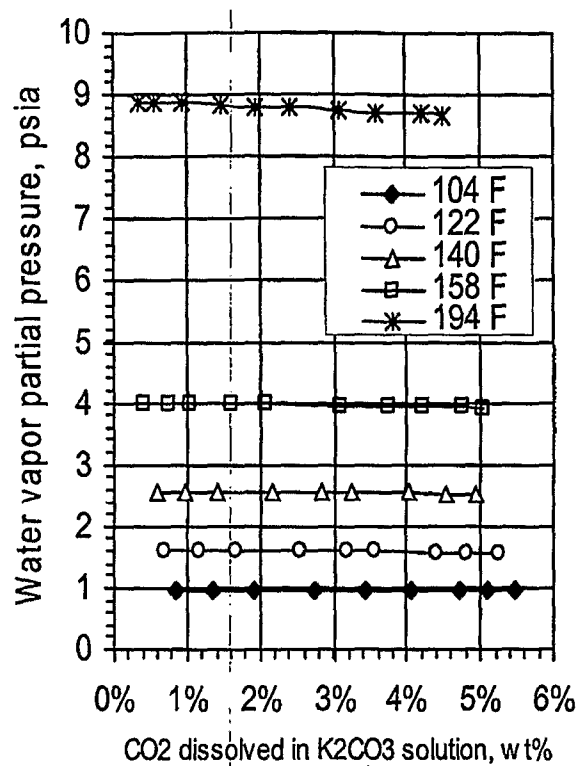
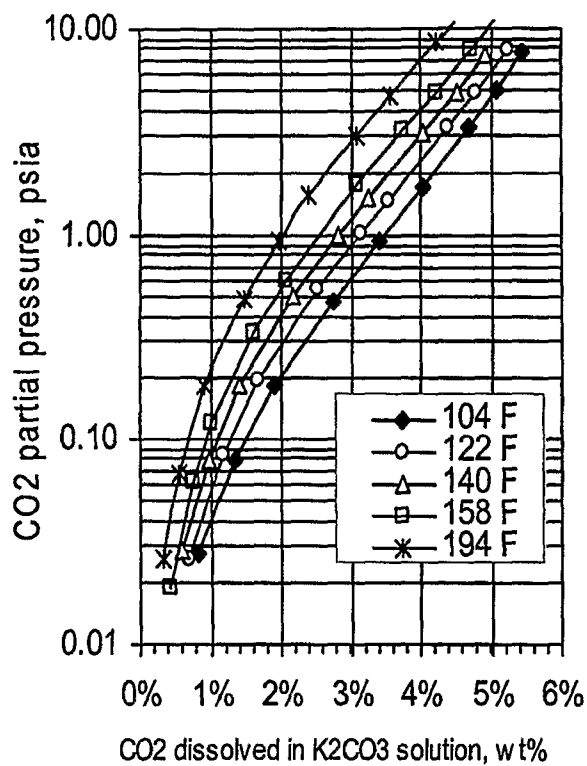


FIG. 10

9/12

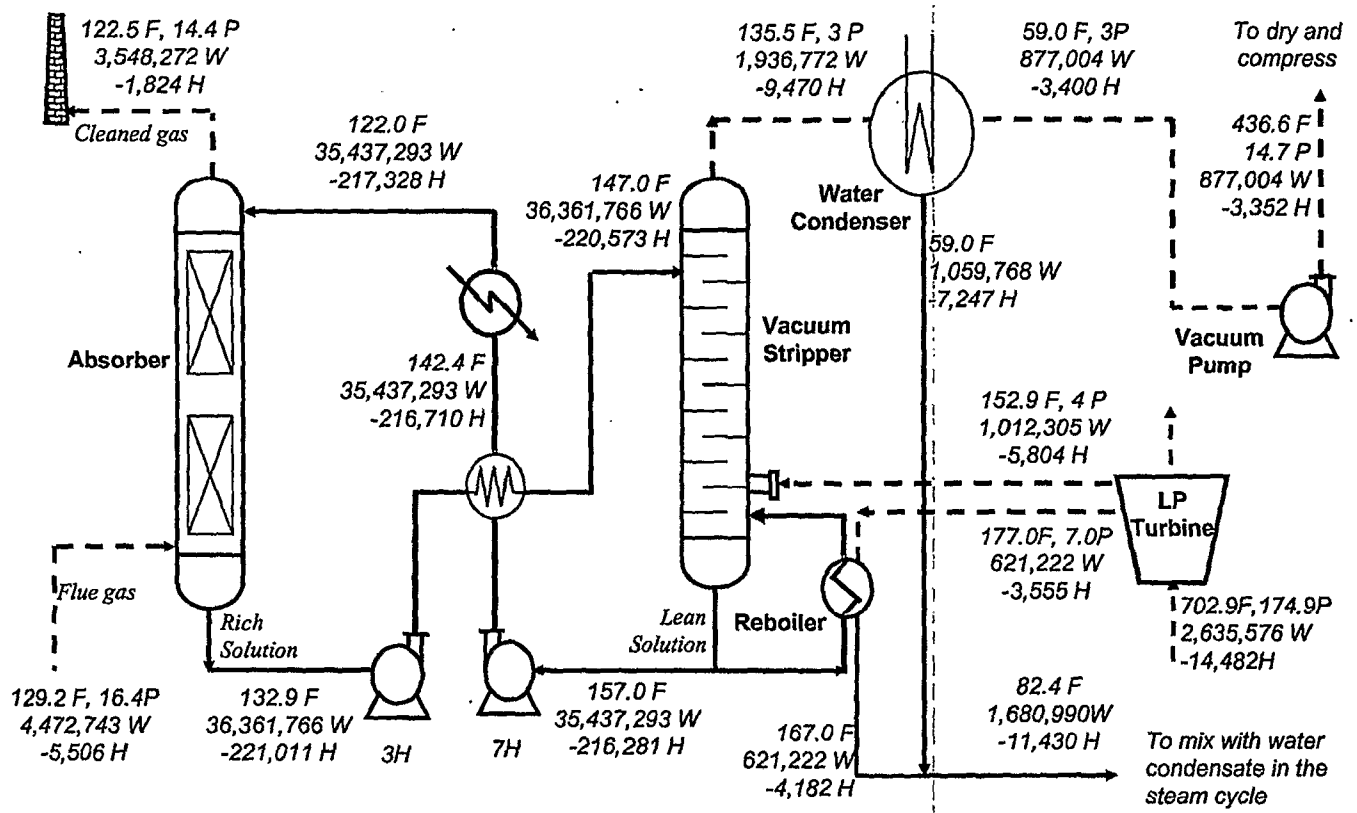


FIG. 11

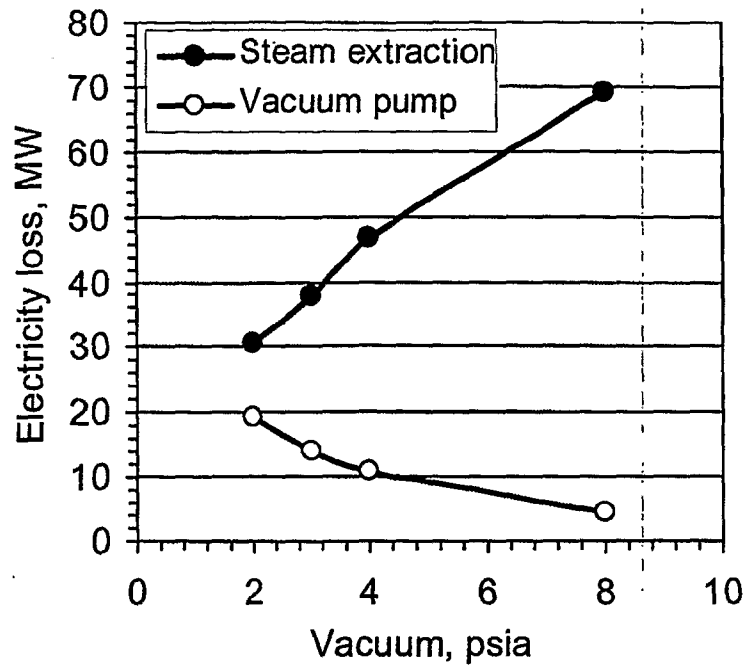


FIG. 12

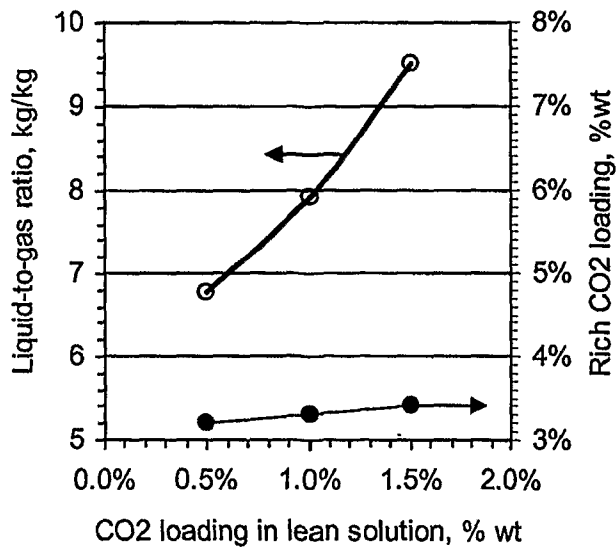


FIG. 14

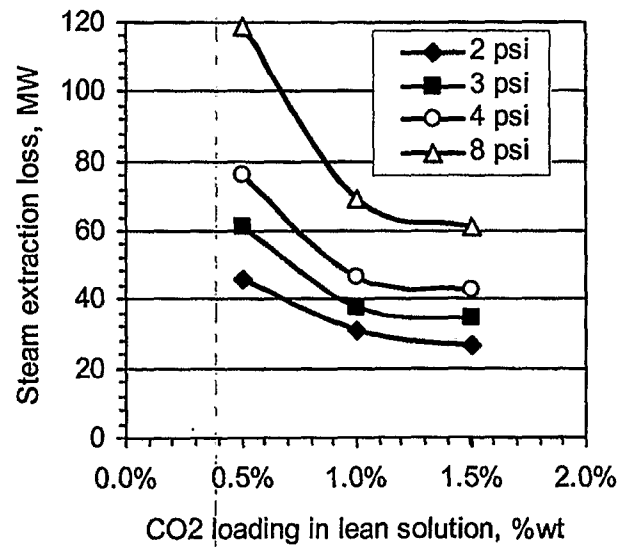


FIG. 15

11/12

Electricity, MWe	PC	PC+ Vacuum	PC+ MEA
CO ₂ capture process:			
Steam extraction loss	0	37.81	89.43
Gas blower	0	10.03	10.03
Liquid pump	0	2.81	1.79
Vacuum pump	0	13.96	0
CO ₂ compression	0	39.65	35.42
Aux. energy use in power plant	34.74	33.27	32.00
Net electricity output	492.86	390.07	358.93

FIG. 13

	MEA process		Vacuum process	
Annualized cost of CO ₂ capture	Cost, M\$/y	Share, %	Cost, M\$/y	Share, %
Capital cost	24.30	30.98%	36.45	51.38%
Fixed O&M	4.97	6.34%	7.46	10.52%
Variable O&M-energy	32.56	41.51%	24.99	35.23%
Variable O&M -sorbent	11.26	14.35%	1.70	2.39%
Variable O&M -others	5.35	6.82%	0.34	0.48%
Subtotal	78.44	100.00%	70.94	100.00%
Net electricity, MW	358.93		390.07	
Loading factor, %	75%		75%	
Annualized cost, c/kWh	3.33		2.77	
Reference CO ₂ emissions, lb/kWh	1.88		1.88	
CO ₂ emissions with control, lb/kWh	0.26		0.24	
CO ₂ avoidance cost, \$/ton	41.00		33.69	

12/12

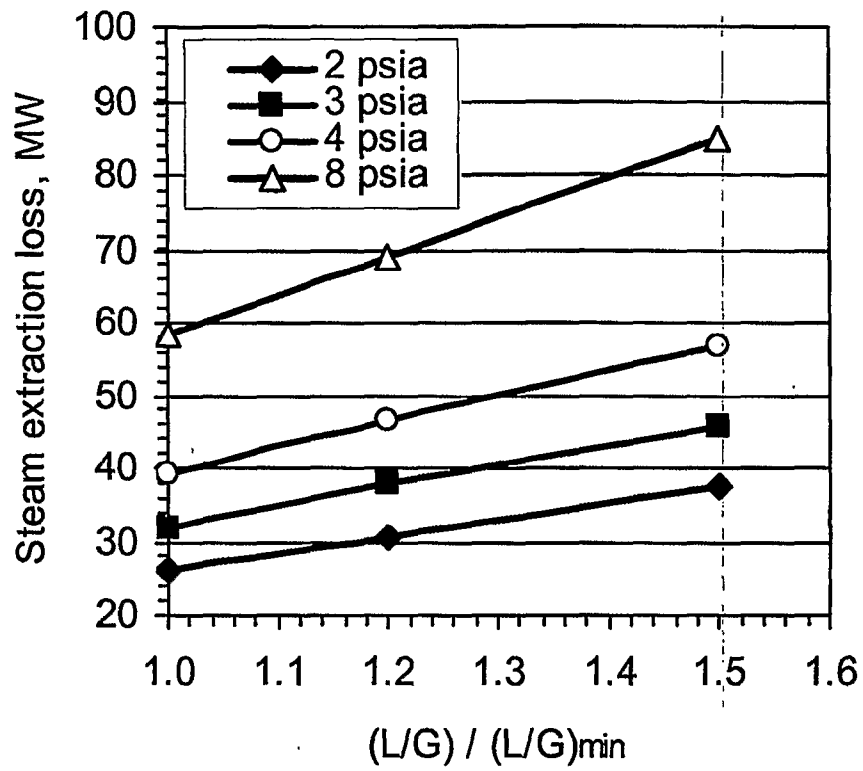


FIG. 16