



Industria y Comercio

SUPERINTENDENCIA

Delegatura de propiedad Industrial

División de Nuevas Creaciones

PCT
SOLICITUD

PATENTE DE INVENCION

21. EXPEDIENTE No.

05-103104

54. TÍTULO Regeneración de una solución acuosa de un proceso de absorción de gas ácido mediante vaporización instantánea y separación con vapor en cuerpos múltiples

51. CLASIFICACIÓN INTERNACIONAL B01D 53/14

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22. BOGOTÁ, D. C., Octubre 10/05

FORMULARIO ÚNICO DE SOLICITUD DE PATENTE
PCT
ENTRADA EN FASE NACIONAL

1) SOLICITUD DE:			
☒ Patente de invención		☐ Patente de Modelo de Utilidad	
☐ Capítulo I.		☒ Capítulo II.	
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5) PCT (86) Solicitud Internacional No. PCT/US2004/006580 Fecha: MARZO 4, 2004 Publicación Internacional No. WO 2004/080573 A1 Fecha: Sept. 23, 2004			
6) Título (54) REGENERACIÓN DE UNA SOLUCIÓN ACUOSA DE UN PROCESO DE ABSORCIÓN DE GAS ÁCIDO MEDIANTE VAPORIZACIÓN INSTANTÁNEA Y SEPARACIÓN CON VAPOR EN CUERPOS MÚLTIPLES.			
7) Clasificación Internacional: B01D 53/14			
8) PRIORIDAD		33) País de Origen	(32) Fecha
SI ☒ NO ☐		Estados Unidos	Marzo 10, 2003
			(31) Número de Solicitud 60/453,279
9) Comprobante de pago N°: 05-76,952 Fecha: Octubre 5, 2005			
Instrucciones para el diligenciamiento del presente formulario. Ver reverso de esta página			

SUPERINDUSTRIA Y COMERCIO
Radicación : 85181104 00000000 Folios: 128
Fecha (AM): 2005-10-10 15:19:20
Trámite : 011 PCT-PATENT 379 FASE NACI 111 PRESENTAC
Dependencia: 2020 DIVISION DE NUEVAS CREACIONES

SUPERINTENDENCIA DE INDUSTRIA Y COMERCIO

NIT : 800176089-2

RECIBO OFICIAL DE CAJA : 05 - 76,952
FECHA : OCTUBRE 5 DE 2005

***** CONSIGNACION *****

DEPOSITANTE	TIPO PAGO	BANCO	CUENTA	No. PAGO	FECHA PGO	Vr. PAGO
CLARKE MODET Y CO. COLOMBIA	CONSIGNACION	BANCO POPULAR	050-00110-6	40241679	05/10/2005	9,375,000.00

***** CONCEPTO *****

CANT.	RENTISTICO	CONCEPTO	TOTAL CONCEPTO
1	50005-01-01 SOLICITUDES	1 TRAMITES DE SOL. DE PATENTE DE IN	443,000.00
		TOTAL :	\$ 443,000.00

SOM: CUATROCIENTOS CUARENTA Y TRES MIL PESOS

RESPONSABLE : _____

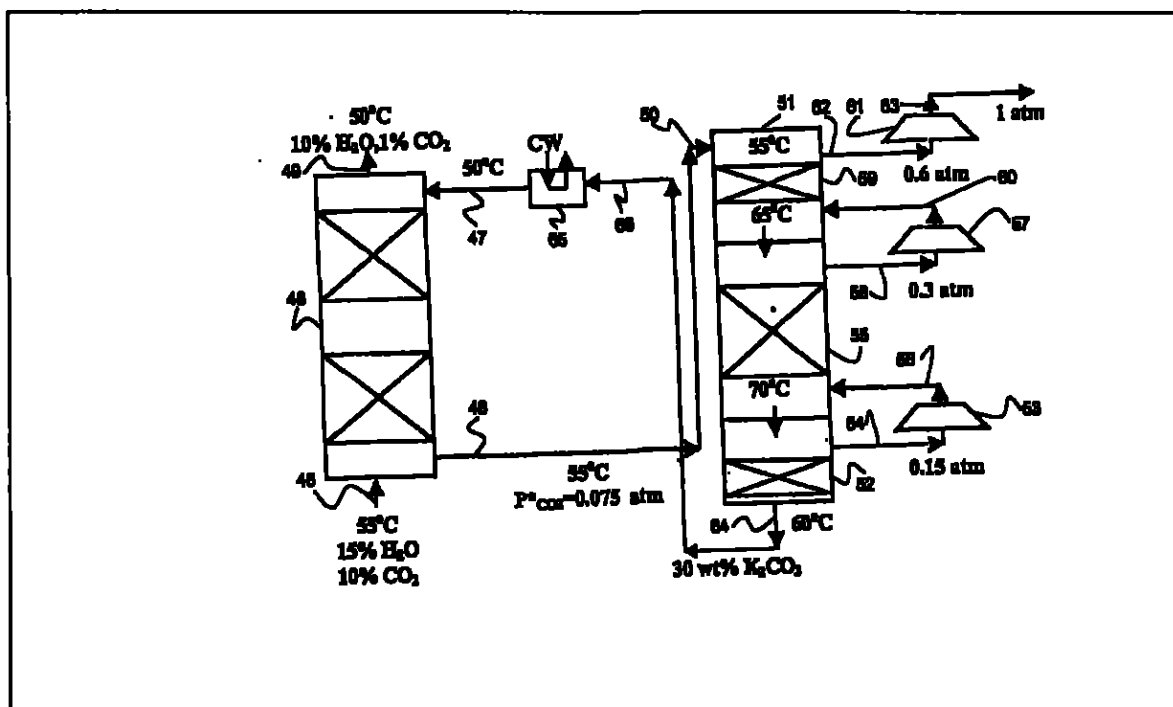
RECIBO DE CAJA APLICADO AL EXPEDIENTE No. _____

10)

ANEXOS

- ☒ Comprobante de pago de la tasa de presentación de la solicitud
☐ Comprobante de pago por reivindicación de prioridad
☐ Documento que acredite la existencia y representación legal cuando el solicitante sea persona Jurídica
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☒ Otros:
1. Reporte internacional de búsqueda
 2. Reporte patente Nr. 4.780.115 de Oct. 25, 1998
 3. Reporte patente Nr. 5,435,977 de Julio 25, 1995
 4. Reporte Patente Nr. 4,160,810 de Julio 10, 1979
 5. Reporte Patente Nr. 1.484.050
 6. Reporte Patente Nr. 4,258,019 de Marzo 24, 1981
 7. Notificación de la transmisión de la copia de la solicitud de publicación

11) FIGURA CARACTERÍSTICA:



12) Solicitud concesión de la patente,

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FIRMA: _____

C.C. 41.654.882 de Bogota TP 20824 CSJ

Instrucciones para el diligenciamiento del presente formulario. Ver reverso de esta pagina



(51) International Patent Classification: B01D 53/14

(21) International Application Number:
PCT/US2004/006380

(22) International Filing Date: 4 March 2004 (04.03.2004)

(25) Filing Language: English

(26) Publication Language: English

(30) Priority Data:
60453,279 10 March 2003 (10.03.2003) US

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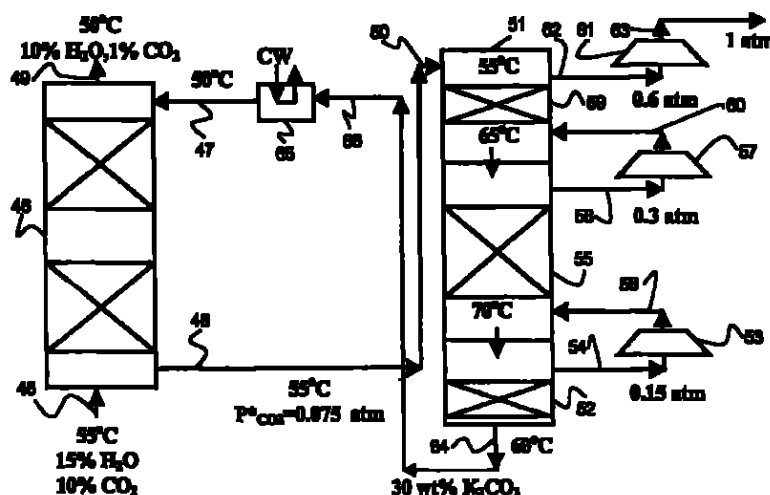
(81) Designated States (unless otherwise indicated, for every
kind of national protection available): AE, AG, AL, AM,
AT, AU, AZ, BA, BB, BG, BR, BW, BY, BZ, CA, CH, CN,
CO, CR, CU, CZ, DE, DK, DM, DZ, EC, EE, EG, ES, FI,
GB, GD, GH, GM, GR, GU, HK, HU, ID, IL, IN, IS, JP, KE,
KG, KP, KR, KZ, LC, LK,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

(84) Designated States (unless otherwise indicated, for every
kind of regional protection available): ARIPO (BW, GH,
GM, KE, LS, MW, MZ, SD, SL, SZ, TZ, UG, ZM, ZW),
Eurasian (AM, AZ, BY, KG, KZ, MD, RU, TJ, TM), Euro-
pean (AT, BE, BG, CH, CY, CZ, DE, DK, EE, ES, FI, FR,
GB, GR, HU, IE, IT, LU, MC, NL, PL, PT, RO, SE, SI, SK,
TR), OAPI (BF, BJ, CF, CG, CI, CM, GA, GN, GQ, GW,
ML, MR, NE, SN, TD, TO).

Published:
— with international search report

[Continued on next page]

(54) Title: REGENERATION OF AN AQUEOUS SOLUTION FROM AN ACID GAS ABSORPTION PROCESS BY MULTI-
STAGE FLASHING AND STRIPPING



(57) Abstract: An acid gas such as carbon dioxide, hydrogen sulfide, or a mixture thereof is removed from gaseous streams using aqueous absorption and stripping processes. By replacing the conventional stripper used to regenerate the aqueous solvent and capture the acid gas with a multipressure stripper (51) that combines acid gas compression with stripping, less energy is consumed. The multipressure stripper is a multistage flash (52, 55, 59) in which the total vapor flow from each stage is compressed and fed to the bottom of the previous flash stage at a higher pressure. In this process, the heat in the water content of the vapor exiting each stage is utilized at a higher pressure in the previous stage. The described stripping process generates the acid gas at a higher pressure without operating the stripper at a higher temperature, thereby reducing the energy consumption of the system.

6 4

WO 2004/080573 A1



— before the expiration of the time limit for amending the claims and to be republished in the event of receipt of amendments

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REGENERATION OF AN AQUEOUS SOLUTION FROM AN ACID GAS ABSORPTION PROCESS BY MULTISTAGE FLASHING AND STRIPPING

5 This application claims priority to, and incorporates by reference, U.S. Provisional Patent Application Serial No. 60/453,279 filed March 10, 2003. The present invention relates generally to the removal and capture of acid gases such as carbon dioxide, hydrogen sulfide and mixtures thereof from gases containing same through aqueous absorption and stripping processes. More particularly, it provides methods for reducing the energy consumption of such absorption and
10 stripping processes.

 A common viewpoint held by a significant segment of the environmental community is that carbon dioxide released into the air plays a major role in global climate change. Thus, global climate change initiatives such as the Kyoto Protocol have identified the curtailment of carbon dioxide releases from fossil fuel combustion and other point sources as a primary means
15 of reducing global climate change. Extensive programs already in place are beginning to demonstrate the economic and technical feasibility of sequestering carbon dioxide by approaches such as injection in underground reservoirs (see, Bergman, P.D. et al, "Disposal of Power Plant CO₂ in Depleted Oil and Gas Reservoirs in Texas," presented at the Third International Conference on Carbon Dioxide Removal, Cambridge, MA, Sept 9-11, 1996) and disposal in the
20 deep ocean (Fujioke, Y. et al., "Cost Comparison of Various CO₂ Ocean Disposal Options," presented at the Third International Conference on Carbon Dioxide Removal, Cambridge, MA, Sept 9-11, 1996).

 One method of curtailing carbon dioxide releases in the industrial arena involves removing carbon dioxide from combustion gases and other gases. Carbon dioxide is emitted in
25 large quantities from fuel combustion by mobile and stationary sources. Carbon dioxide capture/sequestration will be most effective if applied to large stationary sources. The largest single sources of carbon dioxide are conventional coal-fired power plants. These sources represent 30 to 40% of the carbon dioxide emissions in the United States. Technology developed for such sources should also be applicable to CO₂ capture from gas and oil fired
30 boilers, combined cycle power plants, coal gasification, and hydrogen plants. Absorption/stripping is primarily a tail-end technology and is therefore suitable for both existing and new boilers. Specifically, it can be used with existing coal-fired boilers, especially if they already have scrubbers for SO₂-abatement.

The use of absorption and stripping processes with aqueous solvents such as alkanolamines and promoted potassium carbonate is a known, effective technology for the removal and capture of carbon dioxide from flue gas, natural gas, hydrogen, synthesis gas, and other gases. U.S. Patents 4,477,419 and 4,152,217, each of which is incorporated herein by reference, describe aspects of this technology. Alkanolamine absorption/stripping is one proven and effective technology for carbon dioxide capture from gas. The first generation of this technology uses aqueous solutions of monoethanolamine (MEA). Advances in this technology have provided other alkanolamine solvents for carbon dioxide treating in various industries. Monoethanolamine (MEA), diethanolamine (DEA), and the hindered amine AMP are used alone in an aqueous solution. Typical solvent blends include a methyldiethanolamine (MDEA) solution promoted by piperazine or other secondary amines. Also, potassium carbonate solvents are commonly promoted by DEA or other reactive amines.

Gas absorption is a process in which soluble components of a gas mixture are dissolved in a liquid. Stripping is essentially the inverse of absorption, as it involves the transfer of volatile components from a liquid mixture into a gas. In a typical carbon dioxide removal process, absorption is used to remove carbon dioxide from a combustion gas, and stripping is subsequently used to regenerate the solvent and capture the carbon dioxide contained in the solvent. Once carbon dioxide is removed from combustion gases and other gases, it can be captured and compressed for use in a number of applications, including sequestration, production of methanol, and tertiary oil recovery.

The conventional method of using absorption/stripping processes to remove carbon dioxide from gaseous streams is described in U.S. Patent 4,384,875, which is incorporated herein by reference. In the absorption stage, the gas to be treated, containing the carbon dioxide to be removed, is placed in contact, in an absorption column, with the chosen absorbent under conditions of pressure and temperature such that the absorbent solution removes virtually all the carbon dioxide. The purified gas emerges at the top of the absorption column and, if necessary, it is then directed towards a scrubber employing sodium hydroxide, in which the last traces of carbon dioxide are removed. At the bottom of the absorption column, the absorbent solution containing carbon dioxide (also called "rich solvent") is drawn off and subjected to a stripping process to free it of the carbon dioxide and regenerate its absorbent properties.

To effect the regeneration of the absorbent solution, the rich solvent drawn off from the bottom of the absorption column is introduced into the upper half of a stripping column, and the rich solvent is maintained at its boiling point under pressure in this column. The heat necessary for maintaining the boiling point is furnished by reboiling the absorbent solution contained in the stripping column. The reboiling process is effectuated by indirect heat exchange between part of

the solution to be regenerated located in the lower half of the stripping column and a hot fluid at appropriate temperature, generally saturated water vapor. In the course of regeneration, the carbon dioxide contained in the rich solvent to be regenerated maintained at its boiling point is released and stripped by the vapors of the absorbent solution. Vapor containing the stripped carbon dioxide emerges at the top of the stripping column and is passed through a condenser system which returns to the stripping column the liquid phase resulting from the condensation of the vapors of the absorbent solution which pass out of the stripping column with the gaseous carbon dioxide. At the bottom of the stripping column, the hot regenerated absorbent solution (also called "lean solvent") is drawn off and recycled to the absorption column after having used part of the heat content of the solution to heat, by indirect heat exchange, the rich solvent to be regenerated, before its introduction into the stripping column.

In simple absorption/stripping as it is typically practiced in the field, aqueous rich solvent is regenerated at 100-120°C in a simple, countercurrent, reboiled stripper operated at a single pressure, which is usually 1-2 atm. The rich solvent feed is preheated by cross-exchange with hot lean solvent product to within 5-30°C of the stripper bottoms. The overhead vapor is cooled to condense water, which is returned as reflux to the countercurrent stripper. When used for carbon dioxide sequestration and other applications, the product carbon dioxide is compressed to 100-150 atm.

A major problem with the existing absorption/stripping process described above is that it is very energy intensive, and this is largely because the heat required for the heat reboiler is significant. In application on a coal-fired power plant, the required heat of such a process can reduce net power production by as much as 15 to 30%. (Herzog, H., E. Drake, & E. Adams, "CO₂ Capture, Reuse, and Storage Technologies for Mitigating Global Climate Change," final rept, DOE Order No. DE-AF22-96PC01257, 1997). Therefore, it is important to maximize energy efficiency in the design and operation of these systems. The primary method for enhancing energy efficiency is the recovery of useful heat from the overhead condenser, as the overhead vapor can contain one to ten moles of water vapor for every mole of carbon dioxide. Common forms of heat recovery currently practiced include vapor recompression and multieffect strippers. In vapor recompression, the overhead vapor is compressed by a factor of two to ten and then exchanged with the bottoms liquid to provide heat for the reboiler. With multieffect strippers, two or more strippers are operated in parallel, but each stripper is operated at a significantly different pressure. The vapor from a higher pressure stripper is used to heat the reboiler of a lower pressure stripper in a cascade arrangement. Unfortunately, both of these configurations result in a loss of energy in the required heat exchanger.

As noted in U.S. Patent 4,152,217, several attempts have been made to reduce the overall cost associated with the regeneration of absorbent liquid streams. By devising a system in which the spent absorbent/lean absorbent heat exchanger, the overhead cooler-condenser, the reflux drum, and the reflux pump, ordinarily constructed and used with a conventional stripper or regenerator, could be eliminated, the patentee in U.S. Patent 3,690,861 sought to reduce capital investment costs. While capital investment was considerably reduced in the disclosed process, no consideration was given to how the elimination of heat exchangers would affect overall heat requirements of the system. As described in U.S. Patent 4,152,217, it has been found that elimination of heat exchangers increases the overall heat requirements of the system. Thus, while initial capital expenditures are considerably lessened, long term operating expenses, especially in view of rising energy costs, would be higher in an absorption/stripping process that eliminates heat exchangers.

Other patents have disclosed various methods for improving the cost effectiveness of carbon dioxide removal systems by reducing the energy requirements of such systems. One such patent, U.S. Patent 4,553,984, which is incorporated herein by reference, discloses a method in which the rich solvent laden with carbon dioxide is regenerated without the use of a stripping column simply by flashing in one or more flash stages. The disclosed method is said to substantially reduce both capital costs and energy costs. However effective at reducing costs such a technology may be, it may be ineffective for applications which require the captured carbon dioxide to be produced at a higher pressure for use in sequestration, production of methanol, tertiary oil recovery, or other applications. At the very least, such a technology may require significant capital expenditure for the addition of a compressor to enable the captured carbon dioxide to be compressed for use in applications that require higher pressure carbon dioxide.

In light of the above, it would be advantageous to provide for technology in which carbon dioxide can be removed from combustion gases and other gases by an absorption/stripping process that is significantly more energy efficient than the processes currently practiced. The ideal system would generate carbon dioxide at a higher pressure without operating the stripper at a greater temperature. Such a system would, in turn, significantly reduce the energy cost associated with systems in which the carbon dioxide must be compressed for sequestration, production of methanol, tertiary oil recovery, or other applications.

Any problems or shortcomings enumerated in the foregoing are not intended to be exhaustive but rather are among many that tend to impair the effectiveness of previously known techniques. Other noteworthy problems may also exist; however, those presented above should

be sufficient to demonstrate that apparatus and methods appearing in the art have not been altogether satisfactory and that a need exists for the techniques disclosed herein.

The present invention overcomes deficiencies in the prior art by providing, among other things, a method for using aqueous absorption and stripping processes to remove an acid gas from gaseous streams in a manner that generates said acid gas at a higher pressure while consuming less energy than the existing technology. The method involves replacing the conventional single-pressure stripper used to regenerate the aqueous solvent and capture the carbon dioxide with a multipressure stripper that combines acid gas compression with stripping. By generating the acid gas at a higher pressure without operating the stripper at a greater temperature, the method reduces the energy consumption of systems in which the carbon dioxide must be compressed for sequestration, production of methanol, tertiary oil recovery, or other applications.

In one embodiment, the present invention concerns an improved method for removing an acid gas from a gaseous stream containing same using aqueous absorption and stripping equipment, the improvement comprising:

- (a) passing the acid gas-rich solvent stream exiting the absorbing equipment through a multipressure stripper in which the acid gas-rich solvent stream passes through multiple flash stages such that the vapor obtained by flashing the acid gas-rich solvent stream in each flash stage strips the acid gas from the acid gas-rich solvent stream, resulting in an acid gas-rich gaseous stream exiting the multipressure stripper and an acid gas-lean solvent stream exiting the multipressure stripper; wherein the multipressure stripper is operated at multiple pressure levels such that the vapor obtained in each flash stage is compressed in a compressor and fed to the previous flash stage at a higher pressure; and
- (b) recycling the acid gas-lean solvent stream exiting the multipressure stripper in step (a) back to the absorption equipment at least once.

In another embodiment of the present invention, each stage of the multipressure stripper from the stripping process is a countercurrent contactor.

In another embodiment of the present invention, one or more of the stages of the multipressure stripper from the stripping process is a cocurrent contactor.

In another embodiment of the present invention, one or more of the stages of the multipressure stripper from the stripping process is a mixed contactor.

In another embodiment of the present invention, additional heat is supplied at one or more of the stages of the stripping column.

5 In another embodiment of the present invention, the invention involves a stripping equipment for stripping an acid gas from an acid gas-rich solvent stream. This equipment includes a stripping column containing multiple flash stages. The equipment also includes a compressor connected between each of the flash stages such that vapor produced in each flash stage is compressed and fed to the previous flash stage at a higher pressure.

In another embodiment of the present invention, each stage of the stripping column from the stripping equipment is a countercurrent contactor.

10 In another embodiment of the present invention, one or more of the stages of the stripping column from the stripping equipment is a cocurrent contactor.

In another embodiment of the present invention, one or more of the stages of the stripping column from the stripping equipment is a mixed contactor.

15 In another embodiment of the present invention, additional heat is supplied at one or more of the stages of the stripping column.

In the present invention, in an absorption process stage, the gaseous stream is contacted with an aqueous solvent (such as an aqueous amine, an aqueous alkanolamine or mixtures thereof, or an amine promoted aqueous potassium carbonate) in an absorption equipment such that the acid gas in the gaseous stream is transferred from the gaseous stream to the solvent, resulting in a purified gaseous stream exiting the absorption equipment and an acid gas-rich solvent stream exiting the absorption equipment. In a stripping stage, the acid gas-rich solvent stream exiting the absorption equipment is passed through a multipressure stripper in which the acid-gas-rich solvent stream passes through multiple flash stages such that the vapor obtained by flashing the acid gas-rich solvent stream in each flash stage strips the acid gas from the acid gas-rich solvent stream, resulting in an acid gas-rich gaseous stream exiting the multipressure stripper and an acid gas-lean solvent stream exiting the multipressure stripper. In the stripping stage, the multipressure stripper is operated at multiple pressure levels such that the vapor obtained in each flash stage is compressed in a compressor and fed to the previous flash stage at a higher pressure. In the final step of the method of the present invention, the acid gas-lean solvent stream exiting the multipressure stripper is recycled back to the absorption equipment.

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Conveniently, the gaseous stream to be treated by the present method is comprised of flue gas, natural gas, hydrogen gas, or synthesis gas.

As used in the specification, "a" or "an" may mean one or more. As used in the claim(s), when used in conjunction with the word "comprising", the words "a" or "an" mean one or more than one unless explicitly stated otherwise. As used herein "another" may mean at least a second or more.

5 Other objects, features, and advantages of the present invention will become apparent from the following detailed description. It should be understood, however, that the detailed description and the specific examples, while indicating illustrative embodiments of the invention, are given by way of illustration only, since various changes and modifications within the spirit and scope of the invention will become apparent to those skilled in the art from this detailed
10 description.

The following drawings form part of the present specification and are included to further demonstrate certain aspects of the present invention. The invention may be better understood by reference to one or more of these drawings in combination with the detailed description of specific embodiments presented herein.

15 FIG. 1 is a schematic drawing of CO₂ capture by potassium carbonate absorption/stripping utilizing one possible vapor recompression for energy integration.

FIG. 2 is a schematic drawing of CO₂ capture by potassium carbonate absorption/stripping utilizing conventional turbine exhaust steam for energy integration.

20 FIG. 3 is a schematic drawing of CO₂ capture by potassium carbonate absorption/stripping utilizing multipressure stripping for energy integration according to embodiments of the present invention.

FIG. 4 is a schematic diagram of the multipressure stripping process according to embodiments of the present invention.

25 According to the present invention it is possible to minimize the heat energy requirements associated with the removal and capture of carbon dioxide from a gaseous stream by aqueous absorbent and stripping processes. This is attained by combining aqueous absorption with multipressure stripping. Because the present invention results in the captured acid gas being in a compressed state, the invention may be most beneficial in systems requiring compressed acid gas for applications such as sequestration, production of methanol, and tertiary
30 oil recovery.

The present invention utilizes aspects of the process of aqueous absorption, as described in U.S. Patent 6,139,605, which is incorporated herein by reference. Through this process, an

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acid gas such as carbon dioxide, hydrogen sulfide, or a mixture thereof is removed from gaseous streams by contacting the gaseous stream with a liquid absorbent that absorbs the acid gas. As industrial applications expected to benefit from the present invention require the concentration of the acid gas to be reduced to a very low level, the absorbent is generally selected to be one that reacts with the acid gas. Examples of absorbent liquids suitable for use in the present invention for the absorption of the acid gas include, but are not limited to, amine promoted aqueous potassium carbonate and aqueous solutions of amines and alkanolamines. Non-limiting examples of alkanolamines suitable for use in the present invention are monoethanolamine (MEA), diethanolamine (DEA), and methyl diethanolamine (MDEA).

The absorption step may be carried out by contacting the gaseous stream at a relatively low temperature but at an elevated pressure in an absorption column with a stream of the absorbent liquid, referred to at this stage as "lean solvent," flowing counter-current to the gaseous stream. As described in U.S. Patent 4,384,875, the purified gaseous stream emerges from the top of the absorber, while the absorbent liquid containing the acid gas, referred to at this stage as "acid gas-rich solvent," emerges from the bottom of the absorber. To capture the acid gas and regenerate the absorbent liquid so it can be recirculated back to the top of the absorber column as lean solvent, the rich solvent is treated with a stripping process.

In conventional absorption/stripping, the aqueous solvent is regenerated in a simple, countercurrent, reboiled stripper operated at a single pressure. To carry out the stripping process in the present invention, a multipressure stripper is employed. The multipressure stripper integrates the acid gas compression with stripping. The stripper itself is a multistage flash in which the total vapor flow from each stage is compressed and fed to the previous flash stage at a higher pressure. In this process, the heat in the water content of the vapor exiting each stage is utilized at a higher pressure in the previous stage. This is significant because the overhead vapor can contain one to ten moles of water vapor for every mole of the acid gas. The described stripping process generates the acid gas at a higher pressure without operating the stripper at a higher temperature, thereby reducing the energy consumption of the system.

The multipressure stripping process of the present invention may be undesirable in certain distillation and stripping applications because compressors are currently expensive to the point of being cost prohibitive. However, in applications where a compressor is already required to achieve a desired end result, the multipressure stripping process can be implemented with little additional cost.

The present invention finds application in a vast array of fields. Because it integrates compression with stripping to produce a compressed acid gas product, the invention may be particularly attractive in systems that require compression of the product acid gas for

applications such as sequestration, production of methanol, and tertiary oil recovery. Such systems will already have a compressor incorporated into their current configuration, and thus it will not be necessary to expend the significant capital necessary to purchase a compressor for use in the present invention.

5 In addition, as environmental concerns magnify, industry may be forced to further explore and adopt various technologies that effectively reduce acid gas (such as, for example, carbon dioxide) emissions in an energy-efficient manner. In such a scenario, the present invention will be highly desirable to an even greater spectrum of industry due to its energy-efficient method of operation.

10 The following non-limiting examples are included to demonstrate specific embodiments of the invention. It should be appreciated by those of skill in the art that the techniques disclosed in the examples which follow represent techniques discovered by the inventor to function well in the practice of the invention, and thus can be considered to constitute specific modes for its practice. However, those of skill in the art should, in light of the present disclosure, appreciate
15 that many changes can be made in the specific embodiments which are disclosed and still obtain a like or similar result without departing from the spirit and scope of the invention.

FIG. 1, FIG. 2, and FIG. 3 show the same carbon dioxide capture process utilizing three different modes of energy integration. The solvent used in the example process is an amine promoted aqueous potassium carbonate, although other solvents known in the art may be used as well. FIG. 1 and FIG. 2 show processes utilizing known modes of energy integration, whereas
20 FIG. 3 shows a process utilizing multipressure stripping for energy integration according to embodiments of the present invention.

Comparative Example 1

FIG. 1 shows a process utilizing one possible vapor recompression for energy integration.
25 Referring to FIG. 1, a gas which contains a high concentration of CO₂, for example a natural gas or a fine gas, is passed, via line 1, into the bottom of absorption column 2 and flows up through column 2 counter-current to absorbent flowing down the column. At the same time, lean absorption solvent, in this example a promoted potassium carbonate, is passed via line 3 to the upper end of the absorption column and semi-lean absorption solvent is fed via line 4 to a
30 location intermediate the upper and lower ends of column 2. The absorption solvent, which is fed counter-current to the gas, becomes laden with CO₂, and the rich absorption solvent is taken off at the bottom of the absorption column via line 5. The washed gas is taken off at the top of the absorption column via line 6.

Thus, the gas to be treated is scrubbed against semi-lean absorbent near the middle and lower end of absorption column 2 and against lean absorbent near the top of absorption column 2 to absorb most of the specified gas while passing most of the other gases in the gas to be treated out of the absorption step. The bulk of the absorption takes place near the bottom of column 2 by using semi-lean absorbent. Only the last traces of the CO₂ are removed from the gas in the top section of absorption column 2.

The CO₂-rich absorbent solvent is then fed, via line 7, to the upper end of a vacuum stripping column 8, where the rich absorbent flows down column 8. Heat is provided to the stripper by compressing the stripper overhead exiting column 8 via line 9 in compressors 10 and 11 and condensing its contained water vapor against a recirculating stream from the stripper bottoms in condensers 12 and 13. The recirculating stream from the stripper bottoms exits column 8 via line 15, and a portion of it is fed through condensers 12 and 13 via line 16. The recirculating stream is fed back to column 8 via line 14. Compressed gas containing the CO₂ stripped from the rich absorbent liquid exits the system via line 21. The heat provided to the stripper is used to generate the vapor necessary for stripping the CO₂ from the rich absorbent. The vapor generated from heating the rich absorbent rises up through the stripping column 8 countercurrent to the rich absorbent liquid flowing down column 8. As the vapor rises to the top of the stripping column, it strips the CO₂ from the rich absorbent. The vapor rising from the stripper bottom is mainly water vapor since aqueous solvents generally have a low volatility. At the top of column 8, the vapor carrying the CO₂ stripped from the rich absorbent is passed via line 9 through compressor 10 as described previously.

Semi-lean absorbent liquid is withdrawn from a location intermediate the lower and upper ends of stripping column 8 via line 17 and is recycled to a location intermediate the upper and lower ends of absorption column 2. Thus, the semi-lean absorbent withdrawn via line 17 is fed to heat exchanger 18, where it is cooled with cooling water to a temperature and is then fed to absorption column 2 via line 4. A portion of the lean absorbent liquid taken from the bottom of stripping column 8 via line 15 is recycled to the upper end of absorption column 2. Thus, some of the lean absorbent withdrawn via line 15 is fed to heat exchanger 20 via line 19, where it is cooled with cooling water to a temperature and is then fed to absorption column 2 via line 3.

Comparative Example 2

FIG. 2 shows a process utilizing conventional turbine exhaust steam for energy integration. Referring to FIG. 2, a gas which contains a high concentration of CO₂, for example a natural gas or a flue gas, is passed, via line 22, into the bottom of absorption column 23 and flows up through column 23 counter-current to absorbent flowing down the column. At the

same time, lean absorption solvent, in this example a promoted potassium carbonate, is passed via line 24 to the upper end of the absorption column and semi-lean absorption solvent is fed via line 25 to a location intermediate the upper and lower ends of column 23. The absorption solvent, which is fed counter-current to the gas, becomes laden with CO₂, and the rich absorption solvent is taken off at the bottom of the absorption column via line 26. The washed gas is taken off at the top of the absorption column via line 27.

Thus, the gas to be treated is scrubbed against semi-lean absorbent near the middle and lower end of absorption column 23 and against lean absorbent near the top of absorption column 23 to absorb most of the specified gas while passing most of the other gases in the gas to be treated out of the absorption step. The bulk of the absorption takes place near the bottom of column 23 by using semi-lean absorbent. Only the last traces of the CO₂ are removed from the gas in the top section of absorption column 23.

The CO₂-rich absorbent solvent is then fed, via line 28, to the upper end of a vacuum stripping column 29, where the rich absorbent flows down column 29. Heat is provided to the stripper by using exhaust steam from steam turbine 30. High pressure steam is fed to turbine 30 via line 36, and the turbine produces some form of work, such as electricity. The exhaust steam from turbine 30 is fed to condenser 31 via line 32, where it is condensed against a recirculating stream from the stripper bottoms. The recirculating stream from the stripper bottoms exits column 29 via line 33, and a portion of it is fed through condenser 31 via line 34. The recirculating stream is fed back to column 29 via line 35. The heat provided to the stripper is used to generate the vapor necessary for stripping the CO₂ from the rich absorbent. The vapor generated from heating the rich absorbent rises up through the stripping column 29 countercurrent to the rich absorbent liquid flowing down column 29. As the vapor rises to the top of the stripping column, it strips the CO₂ from the rich absorbent. The vapor rising from the stripper bottom is mainly water vapor since aqueous solvents generally have a low volatility. At the top of column 29, the vapor carrying the CO₂ stripped from the rich absorbent is passed via line 36 through condenser 37, where heat exchange with cooling water causes most of the water in the vapor to condense out. The remaining vapor, which is high in CO₂ concentration, is fed to compressor 38 via line 39. The compressed CO₂-rich gas exits the system via line 44.

Semi-lean absorbent liquid is withdrawn from a location intermediate the lower and upper ends of stripping column 29 via line 40 and is recycled to a location intermediate the upper and lower ends of absorption column 23. Thus, the semi-lean absorbent withdrawn via line 40 is fed to heat exchanger 41, where it is cooled with cooling water to a temperature and is then fed to absorption column 23 via line 25. A portion of the lean absorbent liquid taken from the bottom of stripping column 29 via line 33 is recycled to the upper end of absorption column 23.

Thus, some of the lean absorbent withdrawn via line 33 is fed to heat exchanger 43 via line 42, where it is cooled with cooling water to a temperature and is then fed to absorption column 23 via line 24.

Example 1

5 FIG. 3 shows a process utilizing multipressure stripping for energy integration according to embodiments of the present invention. Referring to FIG. 3, a gas which contains a high concentration of CO₂, for example a natural gas or a flue gas, is passed, via line 45, into the bottom of absorption column 46 and flows up through column 46 counter-current to absorbent flowing down the column. At the same time, lean absorption solvent, in this example a promoted potassium carbonate, is passed via line 47 to the upper end of the absorption column.
10 The absorption solvent, which is fed counter-current to the gas, becomes laden with CO₂, and the rich absorption solvent is taken off at the bottom of the absorption column via line 48. The washed gas is taken off at the top of the absorption column via line 49.

 The CO₂-rich absorbent solvent is then fed, via line 50, to the upper end of a
15 multipressure stripping column 51, where the rich absorbent flows down column 51. The multipressure stripping column is a multistage flash, whereby the lowest flash stage 52 provides stripping steam by flashing the absorption solvent. The stripping steam then contacts the rich absorbent liquid in each flash stage and strips the CO₂ from the rich absorbent liquid. The total vapor flow produced in lowest stage 52 is fed through compressor 53 via line 54. The
20 compressed vapor exiting compressor 53 is fed to the bottom of intermediate flash stage 55 via line 56. The total vapor flow produced in intermediate flash stage 55 is fed through compressor 57 via line 58. The compressed vapor exiting compressor 57 is fed to the bottom of top flash stage 59 via line 60. The total vapor flow produced in top flash stage 59 is fed through compressor 61 via line 62. The CO₂-rich compressed vapor exiting compressor 62 exits the
25 system via line 63.

 The lean absorbent liquid taken from the bottom of multipressure stripping column 51 via line 64 is recycled to the upper end of absorption column 46. Thus, some of the lean absorbent withdrawn via line 64 is fed to heat exchanger 65 via line 66, where it is cooled with cooling water to a temperature and is then fed to absorption column 46 via line 47.

30 Using known vapor recompression as shown in FIG. 1, heat will be provided to the stripper by compressing the stripper overhead and condensing its contained water vapor against a recirculating stream from the stripper bottoms. The compression work will also be recovered as heat in the stripper bottoms. This feature replaces steam with equivalent power. Vapor recompression may require an axial flow compressor with high volume capacity for water vapor.

The capital cost of this compressor may be significant. It will be comparable in size to the air compressor of a large gas turbine for a power plant of the equivalent size, but the shaft power requirement will be an order of magnitude smaller because of the low density of the vacuum stream.

5 Using conventional turbine exhaust steam as shown in FIG. 2, a steam turbine could produce electricity and could provide the additional work required to compress the CO₂ from 0.15 atm to 1 atm. The capital cost of compressors would be significantly reduced relative to that for vapor recompression. However, the capital cost of a steam turbine may be significant.

10 Compared to the conventional modes of energy integration shown in FIG. 1 and FIG. 2, the multipressure stripping mode of energy integration of the present invention shown in FIG. 3 offers the minimum energy configuration with the minimum amount of heat exchange. The bottom stage provides stripping steam by flashing the solvent. The top stage recovers heat and drives the overall stripper to a greater temperature and pressure. All of the energy is provided as work, either from off-peak electricity or from steam turbines.

15 Example 2

A schematic diagram of a multipressure stripping process according to embodiments of the present disclosure is shown in FIG. 4. Referring to FIG. 4, CO₂-rich absorbent solvent is fed, via line 87, through cross exchanger 88. The solvent is then fed, via line 67, through cross
20 exchanger 68. The rich absorbent is then fed, via line 69, to the top stage 70 of a multipressure stripping column, where the rich absorbent flows down the column via lines 71 and 73. The multipressure stripping column is a multistage flash, whereby the lowest flash stage 74 provides stripping steam by flashing the absorption solvent. The stripping steam then contacts the rich absorbent liquid in each flash stage and strips the CO₂ from the rich absorbent liquid. The total
25 vapor flow produced in lowest stage 74 is fed through compressor 76 via line 75. The compressed vapor exiting compressor 76 is fed to the bottom of intermediate flash stage 72 via line 77. The total vapor flow produced in intermediate flash stage 72 is fed through compressor 79 via line 78. The compressed vapor exiting compressor 79 is fed to the bottom of top flash stage 70 via line 80. The CO₂-rich compressed vapor exiting top flash stage 70 exits the system
30 via line 81.

A portion of the lean absorbent liquid exiting lowest flash stage 74 via line 82 is recycled to lowest flash stage 74. Thus, some of the lean absorbent withdrawn via line 82 is fed to cross exchanger 84 via line 83, where it is then fed to lowest flash stage 74 via line 85. The remainder

of the lean absorbent liquid exiting lowest flash stage 74 via line 82 that is not recycled to lowest flash stage 74 is fed to cross exchanger 88 via line 86. The lean absorbent liquid exiting cross exchanger 88 is then recycled back to the absorption stage of the process via line 89.

5 The stripper configuration depicted in FIG. 4 was simulated assuming equilibrium contacting in a spreadsheet simulation with a hypothetical solvent. The nonlinear equilibrium of the system was represented by the simple equation:

$$\ln P_{CO_2} = a + B \cdot \ln g - \Delta H/RT$$

10 The constant B was selected to give the specified capacity ($m = \text{gmol/kg water}$) for CO_2 absorption at the nominal absorber temperature, T. The heat of absorption, ΔH , was also varied to simulate different solvents; a value of 22 kcal/gmol is typical of monoethanolamine (MEA); a value of 15 kcal/gmol would represent an alternative solvent. The heat rate, Q, was provided at a maximum temperature of 120°C with a 10°C driving force. The equivalent work of the heat was calculated from the Carnot equation:

$$\text{Equivalent Work of Steam} = Q (T_{\text{atm}} - 313)/T_{\text{atm}}$$

15 The approach to equilibrium at the rich and lean end of the stripper was taken to be 33% (typical of MEA systems) and 75% (possible with reactive alternative solvent). Table 1 shows the results of the aforementioned spreadsheet simulation.

22

TABLE 1. Simulated Energy Requirements for Multipressure Stripper

ΔH (kcal/gmol)	Approach (%)	Q (kcal/gmol)	Compressor Work (kcal/gmol)	Equivalent Work of Steam (kcal/gmol)
15	33	32	2.2	8.9
15	75	28.6	1.4	7.3
22	33	34	1.5	7.5
22	75	34	1.0	6.8
MEA simple	33 rich, 5 lean	50	2.0	12

Parameters: 10% CO₂, 40° Absorber, 90% removal, compression to 8 atm CO₂
0.5 m capacity, 5°C cross-x approach, 4 contact stages (3 compressor stages)

As shown in Table 1, the multipressure stripper requires 6.8 to 8.9 kcal equivalent work/mole CO₂ removed. This is 25 to 40% less than the conventional MEA stripper (12 kcal/gmol). The actual compressor work to get a CO₂ product at 8 atm is 1.0 to 2.2 kcal/gmol. Even though the multipressure stripper requires some compression for water vapor, the compressor work required is not significantly different from the simple MEA stripper because much of the CO₂ is produced at higher pressure.

The net heat requirement for the multipressure stripper (29-34 kcal/gmol CO₂) is approximately the heat of absorption of the CO₂ plus 5°C sensible heat to drive the cross exchanger 88. It is significantly less than the net heat required (40-50 kcal/gmol) for the state-of-the-art MEA technology. Therefore, the compressor capital cost and power requirement may not be significantly different from the conventional MEA stripper, but the heat requirement may be significantly reduced. A closer approach to equilibrium in the absorber reduces the equivalent total work by 10 to 20%. The solvent with the lower heat of absorption appears to be 10 to 15 % better.

While the present invention may be adaptable to various modifications and alternative forms, specific embodiments have been shown by way of example and described herein. However, it should be understood that the present invention is not intended to be limited to the particular forms disclosed. Rather, it is to cover all modifications, equivalents, and alternatives falling within the spirit and scope of the invention as defined by the appended claims. Moreover, the different aspects of the disclosed equipment and methods may be utilized in various combinations and/or independently. Thus the invention is not limited to only those combinations shown herein, but rather may include other combinations, as well.

CLAIMS

1. An improved method for removing an acid gas from a gaseous stream containing same using aqueous absorption and stripping equipment, the improvement comprising:
 - 5 (a) passing the acid gas-rich solvent stream exiting the absorbing equipment through a multipressure stripper in which the acid gas-rich solvent stream passes through multiple flash stages such that the vapor obtained by flashing the acid gas-rich solvent stream in each flash stage strips the acid gas from the acid gas-rich solvent stream, resulting in an acid gas-rich gaseous stream exiting the multipressure stripper and an acid gas-lean solvent stream exiting the
 - 10 multipressure stripper; wherein the multipressure stripper is operated at multiple pressure levels such that the vapor obtained in each flash stage is compressed in a compressor and fed to the previous flash stage at a higher pressure; and
 - (b) recycling the acid gas-lean solvent stream exiting the multipressure stripper in step (a) back to the absorption equipment at least once.
- 15 2. The method of claim 1, wherein each stage of the multipressure stripper is a countercurrent contactor.
3. The method of claim 1, wherein one or more of the stages of the multipressure stripper is a cocurrent contactor.
4. The method of claim 1, wherein one or more of the stages of the multipressure stripper is
 - 20 a mixed contactor.
5. The method of any one of claims 1 to 4, wherein the gaseous stream comprises flue gas, natural gas, hydrogen gas, or synthesis gas.
6. The method of any one of claims 1 to 4, wherein the acid gas comprises carbon dioxide.
7. The method of any one of the claims 1 to 4, wherein the acid gas comprises hydrogen
 - 25 sulfide.
8. The method of any one of claims 1 to 4, wherein the solvent is an aqueous amine, an aqueous alkanolamine, or mixtures thereof.
9. The method of any one of claims 1 to 4, wherein the solvent is an amine promoted aqueous potassium carbonate.
- 30 10. The method of any one of claims 1 to 4, wherein additional heat is supplied at one or more of the stages of the multipressure stripper.
11. An equipment for stripping an acid gas from an acid gas-rich solvent stream, comprising:
 - (a) a stripping column comprising multiple flash stages; and
 - 35 (b) at least one compressor connected between each of the flash stages such that vapor produced in each flash stage is compressed and fed to the previous flash stage at a higher pressure.

12. The equipment of claim 11, wherein each stage of the stripping column is a countercurrent contactor.
13. The equipment of claim 11, wherein one or more of the stages of the stripping column is a cocurrent contactor.
- 5 14. The equipment of claim 11, wherein one or more of the stages of the stripping column is a mixed contactor.
15. The equipment of claim 11, wherein additional heat is supplied at one or more of the stages of the multipressure stripper column.

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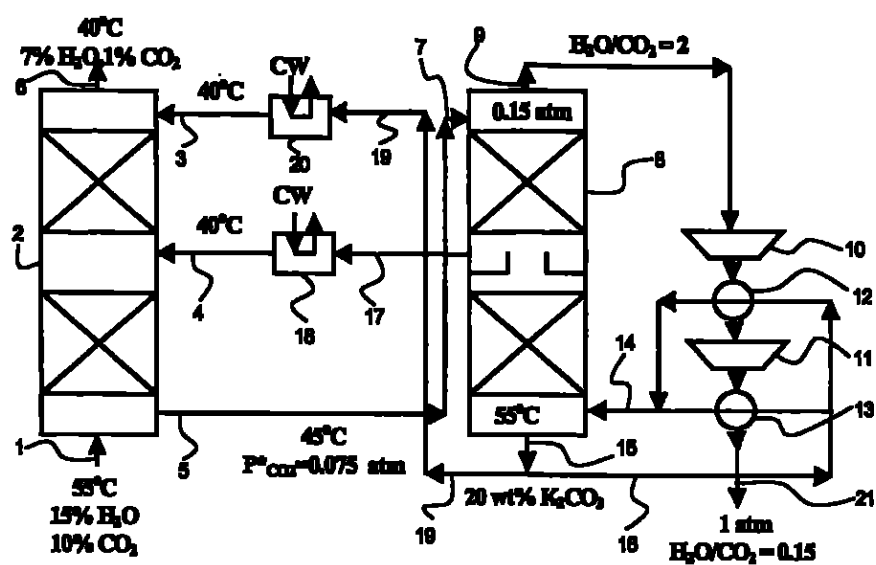


FIG. 1

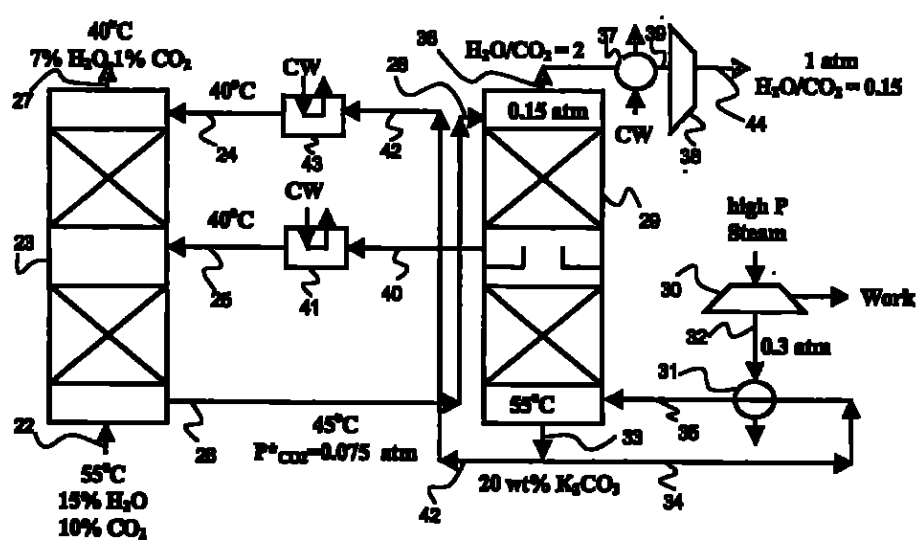


FIG. 2

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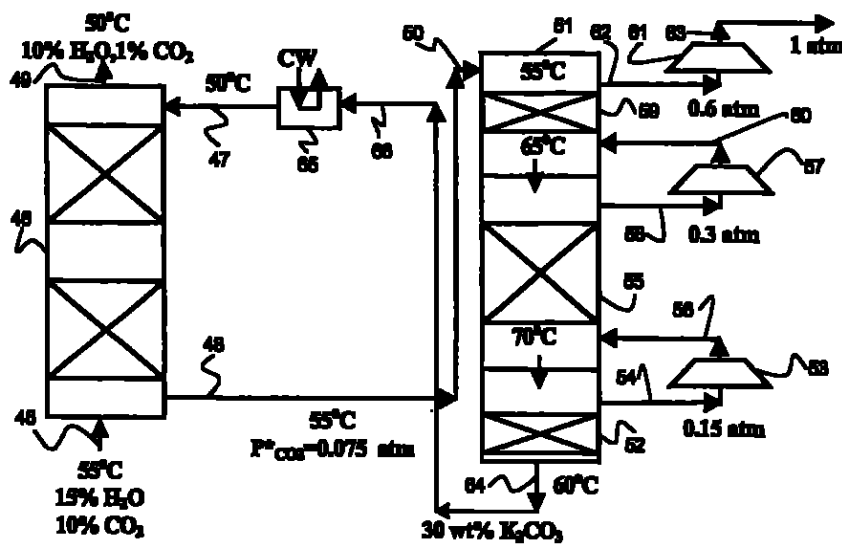


FIG. 3

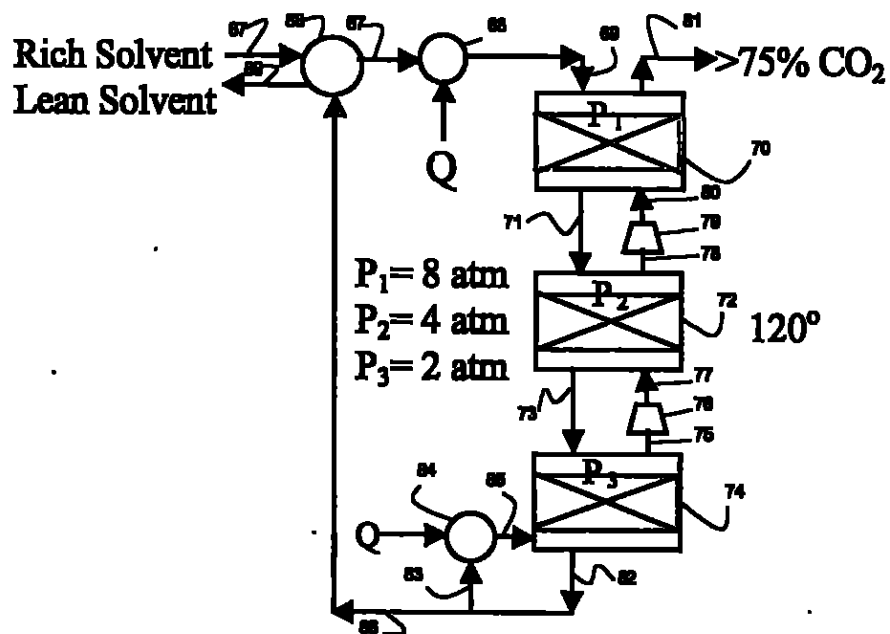


FIG. 4

REGENERACIÓN DE UNA SOLUCIÓN ACUOSA DE UN PROCESO DE
ABSORCIÓN DE GAS ÁCIDO MEDIANTE VAPORIZACIÓN
INSTANTÁNEA Y SEPARACIÓN CON VAPOR EN CUERPOS
MÚLTIPLES

5

DESCRIPCIÓN

Esta solicitud reclama prioridad a la solicitud de patente
estadounidense provisional número de serie 60/453,279, presentada
10 el 10 de marzo de 2003, y la incorpora aquí como referencia. La
presente invención se refiere en general a la extracción y captura de
gases ácidos tales como dióxido de carbono, sulfuro de hidrógeno y
sus mezclas, de gases que los contienen, mediante procesos de
absorción acuosa y separación con vapor. Más particularmente,
15 proporciona métodos para reducir el consumo de energía de estos
procesos de absorción y separación con vapor.

Un punto de vista común sostenido por un segmento
significativo de la comunidad ambiental es que el dióxido de carbono
liberado en el aire juega un papel principal en el cambio climático
20 global. Por eso, las iniciativas de cambio del clima global tales
como el Protocolo de Kyoto, han identificado la reducción de
emisiones de dióxido de carbono de la combustión de combustible
fósil y de otras fuentes puntuales como un medio primordial para
reducir el cambio climático global. Programas extensivos ya en
25 funcionamiento están comenzando a demostrar la factibilidad

económica y técnica de eliminar el dióxido de carbono mediante enfoques tales como inyección en depósitos subterráneos (véase Bergman, P.D. y coinvestigadores, "disposición de CO₂ de plantas de energía en depósitos de petróleo y de gas agotados en Texas",
5 presentado en la Tercera Conferencia Internacional para la Eliminación de Dióxido de Carbono, Cambridge, MA, 9 al 11 de septiembre de 1996) y su disposición en el océano profundo (Fujioka, Y., y coinvestigadores, "Comparación de Costos de Diversas Opciones para la Disposición de CO₂ en el Océano", presentada en
10 la Tercera Conferencia Internacional para la Eliminación de Dióxido de Carbono, Cambridge, MA, 9 al 11 de septiembre de 1996).

Un método para reducir las emisiones de dióxido de carbono en el campo industrial involucra extraer dióxido de carbono de gases de combustión y de otros gases. El dióxido de carbono se
15 emite en grandes cantidades a partir de la combustión de combustible por fuentes móviles y estacionarias. La captura o eliminación de dióxido de carbono será más efectiva si se aplica a grandes fuentes estacionarias. Las fuentes simples más grandes de dióxido de carbono son plantas de energía convencionales que
20 funcionan con carbón. Estas fuentes representan de 30 a 40 por ciento de las emisiones de dióxido de carbono en Estados Unidos. La tecnología desarrollada para estas fuentes debe ser aplicable también a la captura de CO₂ de calentadores de gas y de petróleo, plantas de energía de ciclos combinados, gasificación de carbón, y
25 plantas de hidrógeno. La absorción/separación con vapor es

principalmente una tecnología que está al final de la cola, y por consiguiente es apropiada para los calentadores existentes y los nuevos. Específicamente, puede usarse con los calentadores existentes que funcionan con carbón, especialmente si ya tienen lavadores de gases para la eliminación de SO_2 .

El uso de procesos de absorción y de separación con vapor con solventes acuosos tales como alcanolaminas y carbonato de potasio promovido, es una tecnología efectiva y conocida para la extracción y captura de dióxido de carbono del gas de combustión, gas natural, hidrógeno, gas de síntesis, y otros gases. Las patentes estadounidenses números 4,477,419 y 4,152,217, cada una de las cuales está incorporada a la presente solicitud mediante referencia, describen aspectos de esta tecnología. La absorción con alcanolamina/separación con vapor, es una tecnología probada y efectiva para la captura de dióxido de carbono de un gas. La primera generación de esta tecnología usa soluciones acuosas de monoetanolamina (MEA). Los avances en esta tecnología han proporcionado otros solventes de alcanolamina para el tratamiento del dióxido de carbono en diversas industrias. La monoetanolamina (MEA), la dietanolamina (DEA) y la amina impedida AMP se usan solas en una solución acuosa. Las mezclas de solvente típicas incluyen una solución de metildietanolamina (MDEA) promovida por piperazina u otras aminas secundarias. También, los solventes de carbonato de potasio son promovidos comúnmente por DEA u otras aminas reactivas.

La absorción de gas es un proceso en el cual los componentes solubles de una mezcla de gas se disuelven en un líquido. La separación con vapor esencialmente es el inverso de la absorción, ya que involucra la transferencia de componentes volátiles de una mezcla líquida a un gas. En un proceso típico de extracción de dióxido de carbono, se usa la absorción para extraer el dióxido de carbono o de un gas de combustión, y la separación con vapor se usa a continuación para regenerar el solvente y capturar el dióxido de carbono contenido en el solvente. Una vez que el dióxido de carbono se elimina de los gases de combustión y de otros gases, puede ser capturado y comprimido para su uso en una cantidad de aplicaciones, incluyendo eliminación, producción de metanol, y recuperación terciaria de aceite.

El método convencional para usar los procesos de absorción/separación con vapor para extraer dióxido de carbono de corrientes gaseosas, se describe en la patente estadounidense número 4,384,875, la cual está incorporada a la presente solicitud mediante referencia. En el cuerpo de absorción, el gas que va a ser tratado, que contiene dióxido de carbono que va a ser extraído, se pone en contacto, en una columna de absorción, con el absorbente elegido bajo condiciones de presión y temperatura tales que la solución absorbente extrae virtualmente todo el dióxido de carbono. El gas purificado sube a la parte superior de la columna de absorción, y si es necesario, se dirige luego hacia un lavador de gases que emplea hidróxido de sodio, en el cual se eliminan las

últimas trazas de dióxido de carbono. En el fondo de la columna de absorción, la solución absorbente que contiene dióxido de carbono (también denominado "solvente rico") se extrae por destilación, y se somete a un proceso de separación con vapor para liberarlo del dióxido de carbono y regenerar sus propiedades absorbentes.

Para efectuar la regeneración de la solución absorbente, el solvente rico extraído por destilación del fondo de la columna de absorción, se introduce en la mitad superior de una columna de separación con vapor, y el solvente rico se mantiene en su punto de ebullición bajo presión en esta columna. El calor necesario para mantener el punto de ebullición se proporciona vaporizando la solución absorbente contenida en la columna de separación con vapor. El proceso de evaporación se efectúa por intercambio de calor indirecto entre parte de la solución que va a ser regenerada, localizada en la mitad inferior de la columna de separación con vapor, y un fluido caliente a temperatura apropiada, generalmente vapor de agua saturado. En el curso de la regeneración, el dióxido de carbono contenido en el solvente rico que va a ser regenerado mantenido en su punto de ebullición, se libera y se separa mediante los vapores de la solución absorbente. El vapor que contiene el dióxido de carbono separado sube a la parte superior de la columna de separación con vapor, y se hace pasar a través de un sistema condensador, que devuelve a la columna de separación con vapor la fase líquida resultante de la condensación de los vapores de la solución absorbente que se extrae de la columna de separación por

vapor con el dióxido de carbono gaseoso. En el fondo de la columna de separación con vapor, la solución absorbente regenerada por calor (también llamada "solvente empobrecido"), se extrae por destilación y se recicla hacia la columna de absorción después de haber usado parte del contenido de calor de la solución para calentar, por intercambio de calor indirecto, el solvente rico que va a ser regenerado, antes de su introducción en la columna de separación con vapor.

En la absorción /separación con vapor simple, tal como se practica comúnmente en la materia, el solvente acuoso rico se regenera a 100 - 120 °C en un separador con vapor simple, contracorriente, de evaporación, que funciona a una sola presión, la cual usualmente es de 1 a 2 atm. El solvente rico alimentado se precalienta por intercambio cruzado con el producto solvente empobrecido caliente a entre 5 y 30 °C de los residuos del separador. El vapor saliente de la parte alta se enfría para condensar el agua, la cual es devuelta como reflujo hacia el separador con vapor en contracorriente. Cuando se usa para la eliminación del dióxido de carbono y para otras aplicaciones, el dióxido de carbono producido se comprime a 100 - 150 atm.

Un problema principal con el proceso existente de absorción/separación con vapor descrito anteriormente es que es de uso intensivo de energía, y ésta es mayor porque el calor requerido por el evaporador es significativo. En aplicación en una planta de energía que funciona con carbón, el calor requerido por este tipo de

proceso puede reducir la producción de energía neta en tanto como 15 a 30 % (Herzog, H., E. Drake, y E. Adams, "CO₂, Tecnologías para Captura, Reutilización y Almacenamiento para Mitigar el Cambio Climático Global", rep. final, Orden DOE No. DE-AF22-5 96PC01257, 1997). En consecuencia, es importante aumentar al máximo la eficiencia energética en el diseño y operación de estos sistemas. El método primordial para mejorar la eficiencia energética es la recuperación de calor útil del condensador superior, ya que el vapor saliente de la parte alta puede contener de uno a diez moles de vapor de agua por cada mol de dióxido de carbono. Las formas comunes de recuperación de calor que se practican actualmente incluyen recompresión de vapor y separadores de efecto múltiple. En la recompresión de vapor, el vapor saliente de la parte superior se comprime en un factor de dos a diez y luego se intercambia con el líquido del fondo para proporcionar calor al evaporador. En el caso de los separadores con vapor de efecto múltiple, dos o más separadores funcionan en paralelo, pero cada separador se usa para calentar el evaporador de un separador con vapor de menor presión en un arreglo de cascada. Desafortunadamente, ambas 15 configuraciones dan como resultado una pérdida de energía en el intercambiador de calor requerido.

Como se señala en la patente estadounidense número 4,152,217, se han hecho varios intentos para reducir el costo general asociado con la regeneración de corrientes de líquido absorbente. 25 Ideando un sistema en el cual puedan eliminarse el intercambiador

de calor absorbente gastado/ absorbente empobrecido, el enfriador condensador superior, el tambor de reflujo, y la bomba de reflujo, contruidos y usados ordinariamente con un separador de vapor o con un regenerador convencionales, el titular de la patente, en la
5 patente estadounidense número 3,690,861 buscó reducir los costos de inversión de capital. Si bien la inversión de capital se redujo considerablemente en el proceso descrito, no se dio consideración a cómo afectaría la eliminación de los intercambiadores de calor a los requerimientos totales de calor del sistema. Tal como se describe en
10 la patente estadounidense número 4,152,217, se ha encontrado que la eliminación de los intercambiadores de calor aumenta los requerimientos totales de calor del sistema. Por eso, si bien los gastos de capital inicial disminuyen considerablemente, los gastos de operación a largo plazo, especialmente en vista de los costos en
15 aumento de la energía, serán más altos en un proceso de absorción/separación con vapor que elimine los intercambiadores de calor.

Otras patentes han descrito diversos métodos para mejorar la efectividad de costo de los sistemas de extracción de
20 dióxido de carbono reduciendo los requerimientos de energía de estos sistemas. Una de estas patentes, la patente estadounidense número 4,553,984, la cual está incorporada a la presente solicitud mediante referencia, describe un método en el cual el solvente rico cargado con dióxido de carbono es regenerado sin el uso de una
25 columna de separación con vapor, simplemente vaporizando

instantáneamente en uno o más cuerpos de vaporización instantánea. Se dice que el método descrito reduce sustancialmente tanto los costos de capital como los costos de energía. A pesar de lo efectiva para reducir costos que puede ser esta tecnología, puede ser
5 inefectiva para aplicaciones que requieran que el dióxido de carbono capturado sea producido a una presión más alta para su uso en eliminación, producción de metanol, recuperación de aceite terciario y otras aplicaciones. Al final, este tipo de tecnología puede requerir gastos de capital significativos para la adición de un compresor que
10 permita comprimir el dióxido de carbono capturado para su uso en aplicaciones que requieran dióxido de carbono con mayor presión.

A la luz de lo anterior, sería ventajoso proporcionar una tecnología en la cual el dióxido de carbono pueda ser extraído de los gases de combustión y de otros gases por medio de un proceso de
15 absorción/separación con vapor que sea significativamente más eficiente desde el punto de vista energético que los procesos practicados actualmente. El sistema ideal generaría dióxido de carbono a una presión más alta sin que el separador por vapor tenga que funcionar a una temperatura más alta. Este sistema, a su vez,
20 reduciría significativamente el costo de energía asociado con los sistemas en los cuales el dióxido de carbono debe ser comprimido para eliminación, producción de metanol, recuperación de aceite terciario, u otras aplicaciones.

Cualesquiera problemas o defectos enumerados en lo
25 anterior no pretenden ser exhaustivos, sino que están entre muchos

de los que tienden a obstaculizar la efectividad de las técnicas previamente conocidas. También pueden existir otros problemas notables, aquellos que se presentan antes deben ser suficientes para demostrar que los aparatos y métodos divulgados en la técnica no han sido enteramente satisfactorios, y que existe una necesidad de las técnicas que aquí se describen.

La presente invención soluciona deficiencias en la técnica anterior proporcionando, entre otras cosas, un método para usar procesos de absorción acuosa y de separación con vapor para extraer un gas ácido de corrientes gaseosas en una forma que genera dicho gas ácido a una presión más alta, y al mismo tiempo consume menos energía que la tecnología existente. El método involucra reemplazar el separador con vapor de una sola presión, que se usa para regenerar el solvente acuoso y capturar el dióxido de carbono, con un separador con vapor a presión múltiple que combina la compresión del gas ácido y la separación con vapor. Al generar el gas ácido a una presión más alta sin hacer funcionar el separador a una temperatura mayor, el método reduce el consumo de energía de los sistemas en los cuales el dióxido de carbono debe ser comprimido para el eliminación, producción de metanol, recuperación de aceite terciario u otras aplicaciones.

En una modalidad, la presente invención se refiere a un método mejorado para eliminar un gas ácido de una corriente gaseosa que lo contiene usando equipo para absorción acuosa y para separación con vapor, la mejora comprende:

(a) hacer pasar la corriente de solvente rico en gas ácido fuera del equipo de absorción, mediante un separador con vapor de presión múltiple, en el cual la corriente de solvente rico en gas ácido pasa a través de múltiples cuerpos de vaporización instantánea, de tal forma que el vapor obtenido por vaporización instantánea de la corriente de solvente rico en gas ácido en cada cuerpo de vaporización instantánea, separa el gas ácido de la corriente de solvente rico en gas ácido, dando como resultado una corriente gaseosa rica en gas ácido que sale del separador con vapor de presión múltiple y una corriente de solvente empobrecida en gas ácido sale del separador de presión múltiple; en donde el separador de presión múltiple funciona a niveles de presión múltiples, de tal forma que el vapor obtenido en cada cuerpo de vaporización instantánea se comprime en un compresor y alimenta el cuerpo de vaporización instantánea previo a una presión más alta; y

(b) reciclar la corriente de solvente empobrecida en gas ácido que sale del separador de presión múltiple en el paso (a) de regreso al equipo de absorción al menos una vez.

En otra modalidad de la presente invención, cada cuerpo del separador con vapor de presión múltiple del proceso de separación con vapor, es un contactor contracorriente.

En otra modalidad de la presente invención, cada cuerpo del separador con vapor de presión múltiple del proceso de separación con vapor, es un contactor co-corriente.

En otra modalidad de la presente invención, una o más de

las cuerpos del separador con vapor de presión múltiple del proceso de separación con vapor es un contactor mixto.

En otra modalidad de la presente invención, se suministra calor adicional a uno o más de los cuerpos de la columna de
5 separación con vapor.

En otra modalidad de la presente invención, la invención involucra un equipo para la separación con vapor de un gas ácido, de una corriente de solvente rica en gas ácido. Este equipo incluye un compresor conectado entre cada uno de los cuerpos de vaporización
10 instantánea de tal manera que el vapor producido en cada cuerpo de vaporización instantánea se comprima y alimente a el cuerpo de vaporización instantánea previo, a una presión mayor.

En otra modalidad de la presente invención, cada cuerpo de la columna de separación con vapor del equipo de separación con
15 vapor, es un contactor contracorriente.

En otra modalidad de la presente invención, uno o más de los cuerpos de la columna de separación con vapor del equipo de separación con vapor, es un contactor co-corriente.

En otra modalidad de la presente invención, uno o más de
20 los cuerpos de la columna de separación con vapor del equipo de separación con vapor, es un contactor mixto.

En otra modalidad de la presente invención, se suministra calor adicional a uno o más de los cuerpos de la columna de separación por vapor.

25 En la presente invención, en una etapa del proceso de

absorción, la corriente gaseosa se pone en contacto con un solvente acuoso (tal como una amina acuosa, una alcanolamina acuosa o mezclas de ellas, o un carbonato de potasio acuoso promovido) en un equipo de absorción tal, que el gas ácido en la corriente gaseosa se transfiera desde la corriente gaseosa hacia el solvente, dando como resultado una corriente gaseosa purificada que sale del equipo de absorción y una corriente de solvente rico en gas ácido que sale del equipo de absorción. En un cuerpo de separación con vapor, la corriente de solvente rica en gas ácido que sale del equipo de absorción, pasa a través de un separador con presión múltiple, en el cual la corriente de solvente rico en gas ácido pasa a través de múltiples cuerpos de vaporización instantánea de tal forma que el vapor obtenido por vaporización instantánea de la corriente de solvente rico en gas ácido en cada cuerpo de vaporización instantánea, separe el gas ácido de la corriente de solvente rico en gas ácido, dando como resultado una corriente gaseosa rica en gas ácido que sale del separador con vapor de presión múltiple, y una corriente de solvente empobrecida en gas ácido que sale del separador con vapor de presión múltiple. En el cuerpo de separación con vapor, el separador con presión múltiple funciona con niveles de presión múltiples, de tal forma que el vapor obtenido en cada cuerpo de vaporización instantánea es comprimido en un compresor y alimenta al cuerpo de vaporización instantánea previo, a una presión mayor. En el paso final del método de la presente invención, la corriente de solvente empobrecida en gas ácido que sale del

separador de presión múltiple, es reciclada hacia el equipo de absorción.

Convenientemente, la corriente gaseosa que va a ser tratada por el presente método, está constituida por gas de
5 combustión, gas natural, gas hidrógeno, o gas de síntesis.

Como se usa en la especificación, "un" puede significar uno o más. Como se usa en las reivindicaciones, cuando se usa en conjunto con la palabra "comprende" la palabra "un" significa uno o más de uno, a menos que se especifique explícitamente otra cosa.
10 Como se usa aquí, "otro" puede significar al menos un segundo o más.

Otros objetos, características y ventajas de la presente invención serán evidentes a partir de la siguiente descripción detallada. Se debe entender, sin embargo, que la descripción
15 detallada y los ejemplos específicos, si bien indican modalidades ilustrativas de la invención, se dan a manera de ilustración solamente, dado que serán evidentes diversos cambios y modificaciones dentro del espíritu y el alcance de la invención, para los conocedores de la materia, a partir de esta descripción detallada.

20 Los siguientes dibujos forman parte de la presente especificación y están incluidos para demostrar adicionalmente ciertos aspectos de la presente invención. La invención será mejor comprendida mediante referencia a uno o más de estos dibujos en combinación con la descripción detallada de modalidades específicas
25 presentadas aquí.

La figura 1 es un dibujo esquemático de captura de CO_2 por medio de absorción con carbonato de potasio/separación con vapor, que utiliza una posible recompresión de vapor para la integración de energía.

5 La figura 2 es un dibujo esquemático de captura de CO_2 por medio de absorción con carbonato de potasio/separación con vapor que utiliza agotamiento de vapor de turbina convencional para la integración de energía.

La figura 3 es un dibujo esquemático de captura de CO_2
10 por medio de absorción con carbonato de potasio/separación con vapor, que utiliza separación por vapor a presión múltiple para la integración de energía, de acuerdo con modalidades de la presente invención.

La figura 4 es un diagrama esquemático del proceso de
15 separación con vapor a presión múltiple, de acuerdo con modalidades de la presente invención.

De acuerdo con la presente invención, es posible reducir al mínimo los requerimientos de energía calorífica asociados con la extracción y captura de dióxido de carbono de una corriente gaseosa
20 mediante procesos de absorción acuosa y de separación con vapor. Esto se logra combinando la absorción acuosa y la separación con vapor a presión múltiple. Debido a que la presente invención da como resultado el gas ácido capturado en un estado comprimido, la invención puede ser más beneficiosa en los sistemas que requieren
25 gas ácido comprimido para aplicaciones tales como eliminación,

producción de metanol y recuperación de aceite terciario.

La presente invención utiliza aspectos del proceso de absorción acuosa, tal como se describe en la patente estadounidense número 6,139.605, la cual está incorporada a la presente solicitud
5 mediante referencia. A través de este proceso, se extrae un gas ácido tal como dióxido de carbono, sulfuro de hidrógeno, o una mezcla de ellos, de corrientes gaseosas, poniendo en contacto la corriente gaseosa con un absorbente líquido que absorbe el gas ácido. Como las aplicaciones industriales que se espera se
10 beneficien a partir de la presente invención requieren que la concentración del gas ácido esté reducida a un nivel muy bajo, el absorbente generalmente se selecciona para que sea uno que reaccione con el gas ácido. Los ejemplos de líquidos absorbentes apropiados para su uso en la presente invención para la absorción
15 del gas ácido incluyen carbonato de potasio acuoso promovido por amina, y soluciones acuosas de aminas y alcanolaminas, pero no están limitados a ellos. Los ejemplos no limitantes de alcanolaminas apropiadas para su uso en la presente invención son monoetanolamina (MEA), dietanolamina (DEA) y metildietanolamina
20 (MDEA).

El paso de absorción puede llevarse a cabo poniendo en contacto la corriente gaseosa a una temperatura relativamente baja, pero a una presión elevada en una columna de absorción con una corriente del líquido absorbente, denominada en esta etapa como
25 "solvente empobrecido", que fluye contra la corriente con respecto a

la corriente gaseosa. Como se describe en la patente estadounidense número 4,384,875, la corriente gaseosa purificada surge de la parte superior del absorbedor, mientras que el líquido absorbente que contiene el gas ácido, denominado en esta etapa 5 "solvente rico en gas ácido", surge de los residuos del absorbedor. Para capturar el gas ácido y regenerar el líquido absorbente de manera que pueda recircular de regreso hacia la parte superior de la columna de absorbedor como solvente empobrecido, el solvente rico se trata con un proceso de separación con vapor.

10 En la absorción/separación con vapor convencionales, el solvente acuoso se regenera en un separador evaporador simple, contracorriente, que funciona a una sola presión. Para llevar a cabo el proceso de separación por vapor de la presente invención, se emplea un separador de presión múltiple. El separador con vapor de 15 presión múltiple integra la compresión de gas ácido con la separación. El separador por vapor por sí solo es un vaporizador instantáneo de cuerpos múltiples en el cual el flujo total de vapor de cada cuerpo se comprime y alimenta a el cuerpo de vaporización instantánea previo a una presión más alta. En este proceso, se 20 utiliza el calor en el contenido de agua del vapor que sale de cada cuerpo, a una presión más alta en el cuerpo previo. Esto es significativo debido a que el vapor que sale de la parte alta puede contener de uno a diez moles de vapor de agua por cada mol del gas ácido. El proceso de separación con vapor descrito genera el gas 25 ácido a una presión más alta sin que el separador con vapor funcione

a una temperatura más alta, reduciendo con ello el consumo de energía del sistema.

El proceso de separación con vapor de presión múltiple de la presente invención puede ser indeseable en ciertas aplicaciones de destilación y de separación con vapor, debido a que los compresores actualmente son costosos hasta el punto de ser prohibitivos en costos. Sin embargo, en aplicaciones en donde ya se requiere un compresor para lograr un resultado final deseado, el proceso de separación por vapor a presión múltiple puede implementarse con muy poco costo adicional.

La presente invención encuentra aplicación en un vasto conjunto de campos. Debido a que integra la compresión con la separación con vapor para producir un producto de gas ácido comprimido, la invención puede ser particularmente atractiva en sistemas que requieren compresión del gas ácido producido para aplicaciones tales como eliminación, producción de metanol y recuperación de aceite terciario. Estos sistemas ya tendrán un compresor incorporado en su configuración actual, y por eso no será necesario gastar el capital significativo que se requiere para comprar un compresor con el fin de usarlo en la presente invención.

Además, como las preocupaciones ambientales aumentan, la industria puede ser obligada a explorar adicionalmente y a adoptar diversas tecnologías que reduzcan efectivamente las emisiones de gas ácido (tal como, por ejemplo, dióxido de carbono) en una forma eficiente energéticamente. En este escenario, la presente invención

sería altamente deseable para un espectro aún mayor de la industria, debido a su método de operación eficiente energéticamente.

Los siguientes ejemplos no limitantes se incluyen para demostrar modalidades específicas de la invención. Los
5 conocedores de la materia se darán cuenta de que las técnicas descritas en los ejemplos que siguen representan técnicas descubiertas por el inventor que funcionan bien en la práctica de la invención, y por ello puede considerarse que constituyen formas específicas para su práctica. Sin embargo, los conocedores de la
10 materia se darán cuenta, a la luz de la presente descripción, que se puede hacer muchos cambios en las modalidades específicas que están descritas y todavía obtener un resultado igual o similar sin apartarse del espíritu y del alcance de la invención.

La figura 1, la figura 2 y la figura 3, muestran el mismo
15 proceso para la captura de dióxido de carbono utilizando tres modos diferentes de integración de energía. El solvente utilizado en el proceso de ejemplo es un carbonato de potasio acuoso promovido por amina, a pesar de que también puede usarse otros solventes conocidos en la materia. La figura 1 y la figura 2 muestran procesos
20 que utilizan modos conocidos de integración de energía, de acuerdo con modalidades de la presente invención.

EJEMPLO COMPARATIVO 1

25 La figura 1 muestra un proceso que utiliza una posible

recompresión de vapor para la integración de energía. Con referencia a la figura 1, un gas que contiene una alta concentración de CO_2 , por ejemplo un gas natural o un gas de combustión, se hace pasar, por la línea 1, al fondo de la columna de absorción 2, y fluye a través de la columna 2 contra la corriente con respecto al flujo absorbente hacia abajo de la columna. Al mismo tiempo, el solvente de absorción empobrecido, en este ejemplo un carbonato de potasio promovido, se hace pasar por la línea 3 hacia el extremo superior de la columna de absorción y el solvente de absorción semi-empobrecido se alimenta por la línea 4 hacia una ubicación intermedia entre los extremos superior e inferior de la columna 2. El solvente de absorción, que se alimenta contra corriente con respecto al gas, se hace empobrecido en CO_2 , y el solvente de absorción rico se extrae al fondo de la columna de absorción mediante la línea 5. El gas lavado se extrae en la parte superior de la columna de absorción por medio de la línea 6.

Así, el gas que va a ser tratado se lava contra el absorbente semi-agotado cerca del extremo medio e inferior de la columna de absorción 2 y contra el absorbente empobrecido cerca de la parte superior de la columna de absorción 2 para absorber la mayor parte del gas especificado, mientras pasa la mayor parte de los otros gases al gas que va a ser tratado fuera del paso de absorción. El flujo de absorción tiene lugar cerca del fondo de la columna 2 usando absorbente semi-empobrecido. Solamente las últimas trazas del CO_2 se extraen del gas en la sección superior de

la columna de absorción 2.

El solvente absorbente rico en CO_2 se alimenta entonces, por la línea 7, al extremo superior de una columna de separación al vacío 8, en donde el absorbente rico fluye hacia abajo en la columna 8. Se proporciona calor al separador comprimiendo el vapor saliente de la parte alta de la columna 8, por medio de la línea en los compresores 10 y 11 y condensando su vapor de agua contenida contra una corriente de recirculación de los residuos del separador en los condensadores 12 y 13. La corriente recirculante de los residuos del separador sale por la columna 8 mediante la línea 15, y una porción de él se alimenta a través de los condensadores 12 y 13 por la línea 16. La corriente recirculante se retroalimenta a la columna 8 por medio de la línea 14. El gas comprimido que contiene el CO_2 separado con vapor del líquido absorbente rico sale del sistema por la línea 21. El calor proporcionado al separador se usa para generar el vapor necesario para separar por vapor el CO_2 del absorbente rico. El vapor generado del calentamiento del absorbente rico se eleva a través de la columna de separación con vapor 8 contra la corriente con respecto al líquido absorbente que fluye hacia abajo de la columna 8. A medida que el vapor sube hacia la parte superior de la columna de separación con vapor, separa el CO_2 del absorbente rico. El vapor que se eleva desde el fondo del separador principalmente es vapor de agua, dado que los solventes acuosos generalmente tienen una baja volatilidad. En la parte superior de la columna 8, el vapor que porta el CO_2 separado del

absorbente rico se hace pasar por la línea 9 por medio del compresor 10, tal como se describió anteriormente.

El líquido absorbente semi-empobrecido que sale por la línea 15, se alimenta al intercambiador de calor 20 por la línea 19, en donde se enfría con agua de enfriamiento hasta una temperatura y luego se alimenta a la columna de absorción 2 por medio de la línea 3.

EJEMPLO COMPARATIVO 2

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La figura 2 muestra un proceso que utiliza vapor agotado convencional de turbina para la integración de energía. Con referencia a la figura 2, se hace pasar un gas que contiene una alta concentración de CO₂, por ejemplo un gas natural o un gas de combustión, por la línea 22, al fondo de la columna de absorción 23 y fluye hacia arriba a través de la columna 23 contra la corriente con respecto al absorbente que fluye hacia abajo de la columna. Al mismo tiempo, el solvente de absorción empobrecido, en este ejemplo un carbonato de potasio promovido, se hace pasar por la línea 24 hacia el extremo superior de la columna de absorción y el solvente de absorción semi-empobrecido se alimenta por la línea 25 a una ubicación intermedia entre los extremos superior e inferior de la columna 23. El solvente de absorción, que se alimenta contra la corriente con respecto al gas, se hace cargado con CO₂, y el solvente de absorción rico se saca al fondo de la columna de

absorción por la línea 26. El gas lavado se saca en la parte superior de la columna de absorción por la línea 27.

Así, el gas que va a ser tratado se lava contra el absorbente semi-empobrecido cerca del medio y el extremo inferior de la columna de absorción 23, y contra el absorbente cerca de la columna de absorción superior 23, para absorber la mayor parte del gas especificado, mientras se hace pasar la mayoría de los otros gases al gas que va a ser tratado fuera del paso de absorción. El volumen de absorción tiene lugar cerca del fondo de la columna 23 usando absorbente semi-empobrecido. Solamente las últimas trazas del CO₂ se extraen del gas en la sección superior de la columna de absorción 23.

El solvente absorbente rico en CO₂ se alimenta entonces, por la línea 28, al extremo superior de una columna separadora al vacío 29, en donde el absorbente rico fluye hacia abajo de la columna 29. Se proporciona calor al separador usando vapor agotado de la turbina de vapor 30. Se alimenta vapor a alta presión a la turbina 30 por la línea 36, y la turbina produce alguna forma de trabajo, tal como electricidad. El vapor agotado de la turbina 30 se alimenta al condensador 31 por la línea 32, en donde se condensa contra una corriente recirculante de los residuos del separador. La corriente recirculante de los residuos del separador sale de la columna 29 por la línea 33, y una porción de ella se alimenta a través del condensador 31 por la línea 34. La corriente recirculante se retroalimenta a la columna 29 por la línea 35. El calor

proporcionado al separador se usa para generar el vapor necesario para separar el CO₂ del absorbente rico. El vapor generado del calentamiento del absorbente rico aumenta a través de la columna de separación 29 contracorriente con respecto al flujo de líquido absorbente rico hacia abajo de la columna 29. A medida que el vapor sube hacia la parte superior de la columna de separación con vapor, separa el CO₂ del absorbente rico. El vapor que se eleva desde el fondo del separador principalmente es vapor de agua, dado que los solventes acuosos generalmente tienen una baja volatilidad.

5 En la parte superior de la columna 29, el vapor que arrastra el CO₂ separado del absorbente rico se hace pasar por la línea 36 a través del condensador 37, en donde el intercambio de calor con agua de enfriamiento ocasiona que la mayor parte del agua en el vapor se condense. El vapor restante, que es alto en concentración de CO₂,

10 se alimenta al compresor 38 por la línea 39. El gas comprimido rico en CO₂ sale del sistema por la línea 44.

El líquido absorbente semi-empobrecido se extrae de una ubicación intermedia entre los extremos inferior y superior de la columna de separación 29 por la línea 40 y se recicla a una ubicación intermedia entre los extremos superior e inferior de la columna de absorción 23. Así, el absorbente semi-empobrecido extraído por la línea 40 se alimenta al intercambiador de calor 41, en donde se enfría con agua de enfriamiento a una temperatura y luego se alimenta a la columna de absorción 23 por la línea 25. Una parte

20 del líquido absorbente empobrecido tomado del fondo de la columna

25

de separación 29 por la línea 33, se recicla hacia el extremo superior de la columna de absorción 23. Así, algo del absorbente empobrecido extraído por la línea 33 se alimenta al intercambiador de calor 43 por la línea 42, en donde se enfría con agua de enfriamiento a una temperatura y luego se alimenta en la columna de absorción 23 por la línea 24.

EJEMPLO 1

10 La figura 3 muestra un proceso que utiliza separación con vapor a múltiples presiones para la integración de energía de acuerdo con modalidades de la presente invención. Con referencia a la figura 3, se hace pasar un gas que contiene una alta concentración de CO₂, por ejemplo un gas natural o un gas de
15 combustión, por la línea 45, al fondo de la columna de absorción 46 y fluye hacia arriba a través de la columna 46 contra la corriente con respecto al absorbente que fluye hacia abajo de la columna. Al mismo tiempo, el solvente de absorción empobrecido, en este ejemplo un carbonato de potasio promovido, se hace pasar por la
20 línea 47 al extremo superior de la columna de absorción. El solvente de absorción, que se alimenta contra corriente con respecto al gas, se hace empobrecido en CO₂, y el solvente de absorción rico se extrae en el fondo de la columna de absorción por la línea 48. El gas lavado se extrae en la parte superior de la columna de absorción
25 por la línea 49.

Luego se alimenta el solvente rico en CO_2 , por la línea 50, hacia el extremo superior de una columna de separación de presión múltiple 51, en donde el absorbente rico fluye hacia abajo de la columna 51. La columna de separación de presión múltiple es un vaporizador instantáneo de cuerpos múltiples, en el que el cuerpo más bajo de vaporización instantánea 52 proporciona vapor de separación por vaporización instantánea del solvente de absorción. El vapor separador luego hace contacto con el líquido absorbente rico en cada cuerpo de vaporización instantánea, y separa el CO_2 del líquido absorbente rico. El flujo total de vapor producido en el cuerpo inferior 52, se alimenta al compresor 53 mediante la línea 54. El vapor comprimido que sale del compresor 53 se alimenta al fondo del cuerpo intermedio de vaporización instantánea 55 por la línea 56. El flujo de vapor total producido en el cuerpo intermedio de vaporización instantánea 55, se alimenta al compresor 57 por la línea 58. El vapor comprimido que sale del compresor 57 se alimenta al fondo del cuerpo superior de vaporización instantánea 59 por la línea 60. El flujo total de vapor producido en el cuerpo de vaporización instantánea 59 se alimenta al compresor 61 por la línea 62. El vapor comprimido rico en CO_2 que sale del compresor 62 sale del sistema por la línea 63.

El líquido absorbente empobrecido tomado del fondo de la columna de separación de presión múltiple 51 por la línea 64, se recicla hacia el extremo superior de la columna de absorción 46. Así, algo del absorbente extraído por la línea 64 se alimenta al

intercambiador de calor 65 por la línea 66, en donde se enfría con agua de enfriamiento hasta una temperatura y luego se alimenta a la columna de absorción 46 por la línea 47.

Usando la recompresión de vapor conocida como lo muestra la figura 1, se proporcionará calor al separador comprimiendo el vapor saliente de la parte alta del separador y condensando su contenido de vapor de agua contra una corriente recirculante de los residuos del separador. El trabajo de compresión también será recuperado como calor en los residuos del separador. Esta característica reemplaza el vapor con energía equivalente. La recompresión del vapor puede requerir un compresor de flujo axial con alta capacidad de volumen para el vapor de agua. El costo de capital de este compresor puede ser significativo. Será comparable en tamaño con el compresor de aire de una turbina de gas grande para una planta de energía de tamaño equivalente, pero la potencia en el eje requerida puede tener un orden de magnitud más pequeño, debido a la baja densidad de la corriente al vacío.

Usando vapor agotado de turbina convencional como lo muestra la figura 2, una turbina a vapor podría producir electricidad y podría proporcionar el trabajo adicional requerido para comprimir el CO₂ de 0.15 atm a 1 atm. El costo de capital de los compresores sería reducido significativamente con relación al de la recompresión de vapor. Sin embargo, el costo de capital de una turbina de vapor puede ser significativo.

Comparado con las formas de integración de energía

convencionales que se muestran en la figura 1 y en la figura 2, el modo de integración de energía por separación con vapor a presión múltiple de la presente invención, mostrado en la figura 3, ofrece la configuración de energía mínima con la cantidad mínima de intercambio de calor. El cuerpo del fondo proporciona vapor para separación vaporizando instantáneamente el solvente. El cuerpo superior recupera calor y conduce el vapor saliente de la parte alta del separador a una temperatura y presión mayores. Toda la energía se proporciona como trabajo, ya sea de electricidad fuera de pico o de turbinas de vapor.

EJEMPLO 2

En la figura 4 se muestra un diagrama esquemático de un proceso de separación con vapor de presión múltiple de acuerdo con modalidades de la presente descripción. Con referencia a la figura 4, se alimenta solvente absorbente rico en CO₂ por la línea 87, a través del intercambiador de calor 88. Luego se alimenta el solvente, por la línea 67, a través del intercambiador cruzado 68. Luego se alimenta el absorbente rico, por la línea 69, al cuerpo superior 70 de una columna de separación con vapor de presión múltiple, en donde el absorbente rico fluye hacia abajo de la columna por las líneas 71 y 73. La columna de separación de presión múltiple es un vaporizador instantáneo de cuerpos múltiples, en el que el cuerpo inferior del vaporizador instantáneo 74 proporciona vapor de

separación, vaporizando instantáneamente el solvente de absorción. El vapor de separación luego hace contacto con el líquido absorbente en cada cuerpo del vaporizador instantáneo, y separa el CO₂ del líquido absorbente rico. El flujo total de vapor producido en el cuerpo inferior 74 se alimenta al compresor 76 por la línea 75. El vapor comprimido que sale del compresor 76 se alimenta al fondo del cuerpo del vaporizador instantáneo intermedio 72 por la línea 77. El flujo de vapor total producido en el cuerpo del vaporizador instantáneo intermedio 72 se alimenta al compresor 79 por la línea 78. El vapor comprimido que sale del compresor 79 se alimenta al fondo del cuerpo superior del vaporizador instantáneo 70 por la línea 80. El vapor comprimido rico en CO₂ que sale del cuerpo del vaporizador instantáneo 70 sale del sistema por la línea 81.

Una porción del líquido absorbente empobrecido que sale del cuerpo del vaporizador instantáneo 74 por la línea 82 se recicla hacia el cuerpo inferior del vaporizador instantáneo. Así, algo del absorbente empobrecido extraído por la línea 82 se alimenta al intercambiador cruzado 84 por la línea 83, en donde se alimenta entonces al cuerpo inferior del vaporizador instantáneo 74 por la línea 85. El remanente del líquido absorbente empobrecido que sale del cuerpo inferior del vaporizador instantáneo 74 por la línea 82 que no se recicla hacia el cuerpo inferior del vaporizador instantáneo 74 se alimenta al intercambiador cruzado 88 por la línea 86. El líquido absorbente empobrecido que sale del intercambiador cruzado 88 se recicla de regreso hacia la etapa de absorción del proceso, por la

línea 89.

La configuración del separador con vapor ilustrada en la figura 4 fue simulada asumiendo el equilibrio de contacto en una simulación en hoja de cálculo con un solvente hipotético. El equilibrio no lineal del sistema se representó mediante la ecuación simple:

$$\ln P_{CO_2} = a + B \cdot Idg - \Delta H/RT$$

La constante B se seleccionó para dar la capacidad especificada ($m = \text{gmol/kg de agua}$) para la absorción de CO_2 a la temperatura nominal del absorbedor, T. El calor de absorción, ΔH , también se hizo variar para simular diferentes solventes; un valor de 22 kcal/gmol es típico de la monoetanolamina (MEA); un valor de 15 kcal/gmol representaría un solvente alternativo. La tasa de calor, Q, se proporcionó a una temperatura máxima de 120 °C con una fuerza de accionamiento de 10 °C. El trabajo equivalente del calor se calculó a partir de la ecuación Carnot:

$$\text{Trabajo equivalente de Vapor} = Q(T_{atm} - 313)/T_{atm}$$

La aproximación al equilibrio en el extremo rico y en el empobrecido del separador se tomó como 33% (típico de los sistemas con MEA) y 75% (posible con solvente alternativo reactivo). La tabla 1 muestra los resultados de la simulación en hoja de cálculo

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antes mencionada.

**TABLA 1. REQUERIMIENTOS DE ENERGÍA SIMULADOS PARA
SEPARADOR CON VAPOR DE PRESIÓN MÚLTIPLE**

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ΔH (kcal/gmol)	Aproximación (%)	Q (kcal/gmol)	Trabajo de compresor (kcal/gmol)	Trabajo equivalente de vapor (kcal/gmol)
15	33	32	2.2	8.9
15	75	28.6	1.4	7.3
22	33	34	1.5	7.5
22	75	34	1.0	6.8
MEA simple	33 rico, 5 empobrecido	50	2.0	12

Parámetros: 10% CO₂, 40% absorbedor, 90% extracción, compresión a 8 atm de CO₂, 0.5 m de capacidad, 5°C aproximación cruzada-x, 4 etapas de contacto (3 etapas de compresor)

20

Tal como lo muestra la Tabla 1, el separador con vapor de presión múltiple requiere de 6.8 a 8.9 kcal de trabajo equivalente/mol de CO₂ extraído. Esto es de 25 a 40 % menos que el separador convencional con MEA (12 kcal/gmol). El compresor actual trabaja para obtener un producto de CO₂ a 8 atm que tenga de

25

1.0 a 2.2 kcal/gmol. Aún a pesar de que el separador con vapor de presión múltiple requiere algo de compresión para el vapor de agua, el trabajo del compresor requerido no es significativamente diferente del separador con MEA simple debido a que mucho del CO₂ se produce a presión más alta.

El requerimiento de calor neto del separador con vapor de presión múltiple (29-34 kcal/mol de CO₂) es aproximadamente el calor de absorción del CO₂ mas 5 °C de calor sensible para accionar el intercambiador cruzado 88. Es significativamente menor que el calor neto requerido (40-50 kcal/gmol) para el estado del arte de la tecnología MEA. Por lo tanto, el costo de capital del compresor y el requerimiento de energía pueden no ser significativamente diferentes del separador convencional con MEA, pero el requerimiento de calor puede ser reducido significativamente. Una aproximación más cercana al equilibrio en el absorbedor reduce el trabajo equivalente total en 10 a 20 %. El solvente con el menor calor de absorción parece ser de 10 a 15% mejor.

Si bien la presente invención puede ser adaptable a diversas modificaciones y formas alternativas, se ha mostrado y descrito modalidades específicas a manera de ejemplo en la presente solicitud. Sin embargo, debe entenderse que la presente invención no pretende estar limitada a las formas particulares descritas. En lugar de eso, pretende cubrir todas las modificaciones, equivalentes y alternativas que caigan dentro del espíritu y alcance de la invención tal como se definen por medio de las reivindicaciones

anexas. Más aún, los diferentes aspectos del equipo y métodos descritos pueden utilizarse en diversas combinaciones y/o independientemente. Así, la invención no está limitada solamente a aquellas combinaciones que se muestran aquí, sino que también
5 puede incluir otras combinaciones.

REIVINDICACIONES

1. Un método mejorado para extraer un gas ácido de una corriente gaseosa que lo contiene usando equipo para absorción acuosa y separación con vapor, la mejora comprende:

(a) hacer pasar la corriente de solvente rico en gas ácido fuera del equipo de absorción, mediante un separador con vapor de presión múltiple, en el cual la corriente de solvente rico en gas ácido pasa a través de múltiples cuerpos de vaporización instantánea, de tal forma que el vapor obtenido por vaporización instantánea de la corriente de solvente rico en gas ácido en cada cuerpo de vaporización instantánea, separa el gas ácido de la corriente de solvente rico en gas ácido, dando como resultado una corriente gaseosa rica en gas ácido que sale del separador con vapor de presión múltiple y una corriente de solvente empobrecida en gas ácido sale del separador de presión múltiple; en donde el separador de presión múltiple funciona con niveles de presión múltiples, de tal forma que el vapor obtenido en cada cuerpo de vaporización instantánea se comprime en un compresor y alimenta el cuerpo de vaporización instantánea previo a una presión más alta; y

(b) reciclar la corriente de solvente empobrecida en gas ácido que sale del separador de presión múltiple en el paso (a) de regreso al equipo de absorción, al menos una vez.

2. El método de la reivindicación 1, caracterizado además porque cada cuerpo del separador de presión múltiple es un

contactor contracorriente.

3. El método de la reivindicación 1, caracterizado además porque uno o más de los cuerpos del separador de presión múltiple es un contactor co-corriente.

5 4. El método de la reivindicación 1, caracterizado además porque uno o más de los cuerpos del separador de presión múltiple es un contactor mixto.

5. El método de cualquiera de las reivindicaciones 1 a 4, caracterizado además porque la corriente gaseosa comprende gas de
10 combustión, gas natural, gas hidrógeno, o gas de síntesis.

6. El método de cualquiera de las reivindicaciones 1 a 4, caracterizado además porque el gas ácido comprende dióxido de carbono.

7. El método de cualquiera de las reivindicaciones 1 a 4,
15 caracterizado además porque el gas ácido comprende sulfuro de hidrógeno.

8. El método de cualquiera de las reivindicaciones 1 a 4, caracterizado además porque el solvente es una amina acuosa, una alcanolamina acuosa, o mezclas de ellas.

20 9. El método de cualquiera de las reivindicaciones 1 a 4, caracterizado además porque el solvente es un carbonato de potasio acuoso promovido por amina.

10. El método de cualquiera de las reivindicaciones 1 a 4, caracterizado además porque se suministra calor adicional a uno o
25 más de los cuerpos del separador con vapor de presión múltiple.

11. Un equipo para separar con vapor un gas ácido de una corriente solvente rica en gas ácido, que incluye:

(a) una columna de separación con vapor con múltiples cuerpos de vaporización instantánea; y

5 (b) al menos un compresor conectado entre cada uno de los cuerpos del vaporizador instantáneo, de tal forma que el vapor producido en cada cuerpo del vaporizador instantáneo se comprima y se alimente al cuerpo del vaporizador instantáneo previo, a una presión mayor.

10 12. El equipo de la reivindicación 11, caracterizado además porque cada cuerpo de la columna de separación con vapor es un contactor contracorriente.

13. El equipo de la reivindicación 11, caracterizado además porque uno o más de los cuerpos de la columna de
15 separación con vapor es un contactor co-corriente.

14. El equipo de la reivindicación 11, caracterizado además porque uno o más de los cuerpos de la columna de separación con vapor es un contactor mixto

15. El equipo de la reivindicación 11, caracterizado
20 además porque se suministra calor adicional a uno o más de los cuerpos de la columna del separador con vapor de presión múltiple.

RESUMEN

Un gas ácido tal como dióxido de carbono, sulfuro de hidrógeno, o una mezcla de ellos se saca de corrientes gaseosas usando procesos de absorción acuosa y de separación con vapor. Al reemplazar el separador con vapor convencional que se usa para regenerar el solvente acuoso y capturar el gas ácido con un separador de presión múltiple (51) que combina la compresión del gas ácido con la separación, se consume menos energía. El separador con vapor de presión múltiple es un vaporizador instantáneo de cuerpos múltiples (52, 55, 59) en el cual el flujo total de vapor de cada cuerpo se comprime y se alimenta al fondo de el cuerpo de vaporización instantánea previa a una presión más alta. En este proceso, se utiliza el calor del contenido de agua del vapor que sale de cada cuerpo a una presión más alta en el cuerpo previo. El proceso de separación con vapor descrito, genera el gas ácido a una presión más alta sin que el separador de extracción funcione a una temperatura más alta, reduciendo por ello el consumo de energía del sistema.

1/2

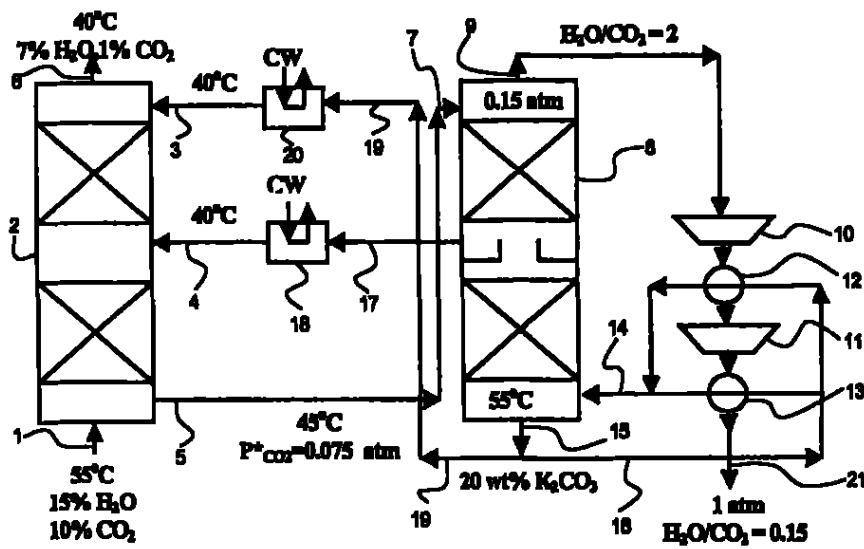


FIG. 1

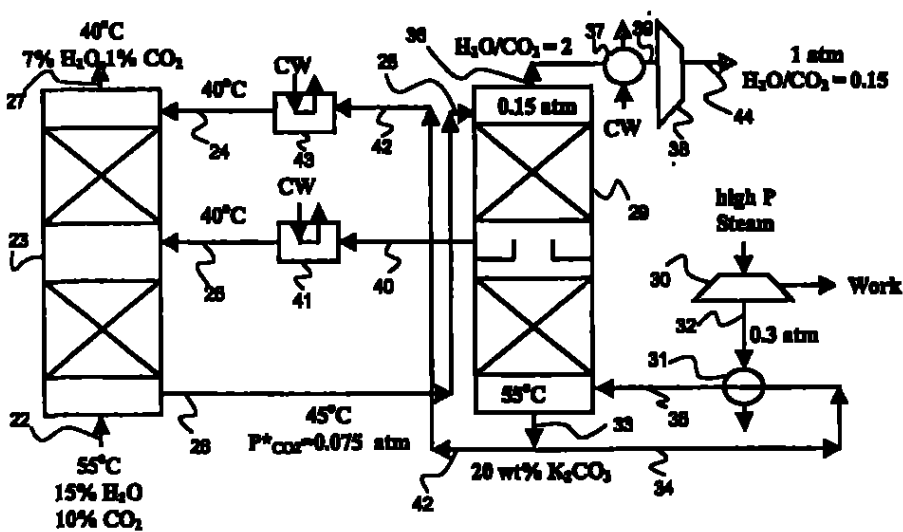


FIG. 2

2/2

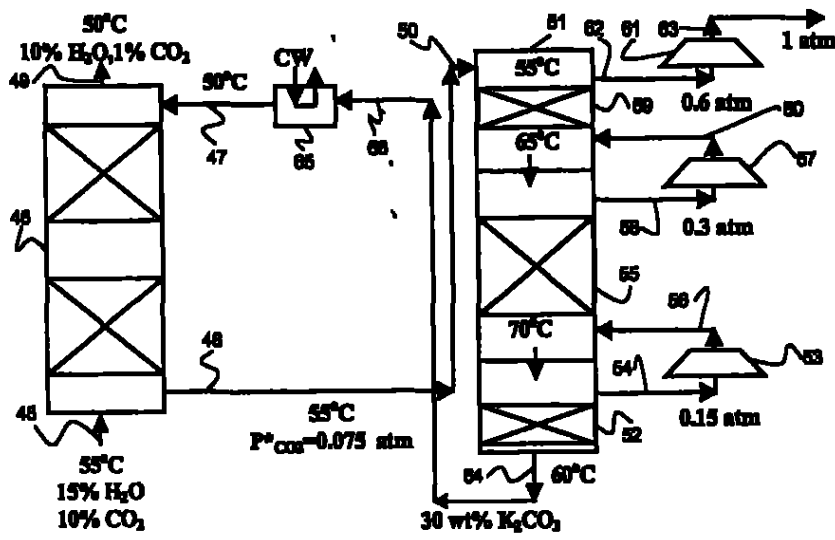


FIG. 3

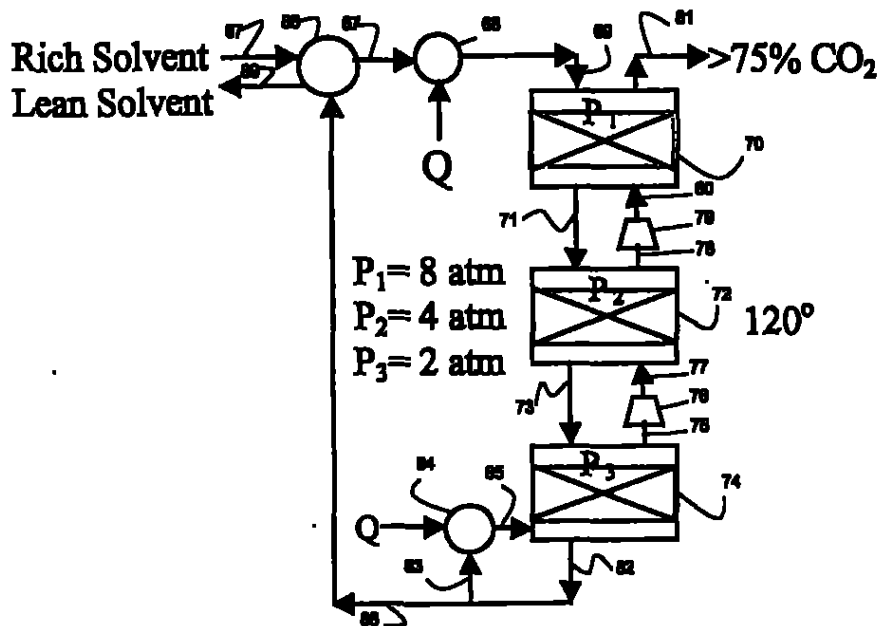


FIG. 4

PATENT COOPERATION TREATY

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From the INTERNATIONAL SEARCHING AUTHORITY

To:

FULBRIGHT & JAWORSKI L.L.P.
Attn. Barrett, Michael C.
600 Congress Avenue, Suite 2400
Austin, TX 78701
UNITED STATES OF AMERICA

NOTIFICATION OF TRANSMITTAL OF
THE INTERNATIONAL SEARCH REPORT AND
THE WRITTEN OPINION OF THE INTERNATIONAL
SEARCHING AUTHORITY, OR THE DECLARATION

*Article 19 Amendments due 07-26-04
Response to Written Opinion due 01-10-05*

(PCT Rule 44.1)

AUG 12 2004

Date of mailing
(day/month/year)

26/07/2004

Applicant's or agent's file reference

UTFB:718W0

Attorney(s): *LTP/mew/mcb*

FOR FURTHER ACTION

See paragraphs 1 and 4 below

International application No.

PCT/US2004/006580

International filing date
(day/month/year)

04/03/2004

Applicant

BOARD OF REGENTS, THE UNIVERSITY OF TEXAS SYSTEM

1. ☒ The applicant is hereby notified that the international search report and the written opinion of the International Searching Authority have been established and are transmitted herewith.

Filing of amendments and statement under Article 19:

The applicant is entitled, if he so wishes, to amend the claims of the International Application (see Rule 46):

When? The time limit for filing such amendments is normally 2 months from the date of transmittal of the International Search Report; however, for more details, see the notes on the accompanying sheet.

Where? Directly to the International Bureau of WIPO, 34 chemin des Colombettes
1211 Geneva 20, Switzerland, Facsimile No.: (41-22) 740.14.35

For more detailed instructions, see the notes on the accompanying sheet.

2. ☐ The applicant is hereby notified that no international search report will be established and that the declaration under Article 17(2)(a) to that effect and the written opinion of the International Searching Authority are transmitted herewith.
3. ☐ With regard to the protest against payment of (an) additional fee(s) under Rule 40.2, the applicant is notified that:

- ☐ the protest together with the decision thereon has been transmitted to the International Bureau together with the applicant's request to forward the texts of both the protest and the decision thereon to the designated Offices.
- ☐ no decision has been made yet on the protest; the applicant will be notified as soon as a decision is made.

4. Reminders

Shortly after the expiration of 18 months from the priority date, the international application will be published by the International Bureau. If the applicant wishes to avoid or postpone publication, a notice of withdrawal of the international application, or of the priority claim, must reach the International Bureau as provided in Rules 90bis.1 and 90bis.3, respectively, before the completion of the technical preparations for international publication.

The applicant may submit comments on an informal basis on the written opinion of the International Searching Authority to the International Bureau. The International Bureau will send a copy of such comments to all designated Offices unless an international preliminary examination report has been or is to be established. These comments would also be made available to the public but not before the expiration of 30 months from the priority date.

Within 19 months from the priority date, but only in respect of some designated Offices, a demand for international preliminary examination must be filed if the applicant wishes to postpone the entry into the national phase until 30 months from the priority date (in some Offices even later); otherwise, the applicant must, within 20 months from the priority date, perform the prescribed acts for entry into the national phase before those designated Offices.

In respect of other designated Offices, the time limit of 30 months (or later) will apply even if no demand is filed within 19 months.

See the Annex to Form PCT/IB/301 and, for details about the applicable time limits, Office by Office, see the *PCT Applicant's Guide*, Volume II, National Chapters and the WIPO Internet site.

Name and mailing address of the International Searching Authority



European Patent Office, P.B. 5818 Patentaan 2
NL-2280 HV Rijswijk
Tel. (+31-70) 340-2040, Tx. 31 651 epo nl,
Fax: (+31-70) 340-3016

Authorized officer

Véronique Baillou

These Notes are intended to give the basic instructions concerning the filing of amendments under article 19. The Notes are based on the requirements of the Patent Cooperation Treaty, the Regulations and the Administrative Instructions under that Treaty. In case of discrepancy between these Notes and those requirements, the latter are applicable. For more detailed information, see also the PCT Applicant's Guide, a publication of WIPO.

In these Notes, "Article", "Rule", and "Section" refer to the provisions of the PCT, the PCT Regulations and the PCT Administrative Instructions respectively.

INSTRUCTIONS CONCERNING AMENDMENTS UNDER ARTICLE 19

The applicant has, after having received the international search report, one opportunity to amend the claims of the international application. It should however be emphasized that, since all parts of the international application (claims, description and drawings) may be amended during the international preliminary examination procedure, there is usually no need to file amendments of the claims under Article 19 except where, e.g. the applicant wants the letter to be published for the purposes of provisional protection or has another reason for amending the claims before international publication. Furthermore, it should be emphasized that provisional protection is available in some States only.

What parts of the international application may be amended?

Under Article 19, only the claims may be amended.

During the international phase, the claims may also be amended (or further amended) under Article 34 before the International Preliminary Examining Authority. The description and drawings may only be amended under Article 34 before the International Examining Authority.

Upon entry into the national phase, all parts of the international application may be amended under Article 28 or, where applicable, Article 41.

When?

Within 2 months from the date of transmittal of the international search report or 16 months from the priority date, whichever time limit expires later. It should be noted, however, that the amendments will be considered as having been received on time if they are received by the International Bureau after the expiration of the applicable time limit but before the completion of the technical preparations for international publication (Rule 46.1).

Where not to file the amendments?

The amendments may only be filed with the International Bureau and not with the receiving Office or the International Searching Authority (Rule 46.2).

Where a demand for international preliminary examination has been/is filed, see below.

How?

Either by cancelling one or more entire claims, by adding one or more new claims or by amending the text of one or more of the claims as filed.

A replacement sheet must be submitted for each sheet of the claims which, on account of an amendment or amendments, differs from the sheet originally filed.

All the claims appearing on a replacement sheet must be numbered in Arabic numerals. Where a claim is cancelled, no renumbering of the other claims is required. In all cases where claims are renumbered, they must be renumbered consecutively (Administrative Instructions, Section 205(b)).

The amendments must be made in the language in which the international application is to be published.

What documents must/may accompany the amendments?

Letter (Section 205(b)):

The amendments must be submitted with a letter.

The letter will not be published with the international application and the amended claims. It should not be confused with the "Statement under Article 19(1)" (see below, under "Statement under Article 19(1)").

The letter must be in English or French, at the choice of the applicant. However, if the language of the international application is English, the letter must be in English; if the language of the international application is French, the letter must be in French.

The letter must indicate the differences between the claims as filed and the claims as amended. It must, in particular, indicate, in connection with each claim appearing in the international application (it being understood that identical indications concerning several claims may be grouped), whether

- (i) the claim is unchanged;
- (ii) the claim is cancelled;
- (iii) the claim is new;
- (iv) the claim replaces one or more claims as filed;
- (v) the claim is the result of the division of a claim as filed.

The following examples illustrate the manner in which amendments must be explained in the accompanying letter:

1. [Where originally there were 48 claims and after amendment of some claims there are 51]:
"Claims 1 to 29, 31, 32, 34, 35, 37 to 48 replaced by amended claims bearing the same numbers; claims 30, 33 and 36 unchanged; new claims 49 to 51 added."
2. [Where originally there were 15 claims and after amendment of all claims there are 11]:
"Claims 1 to 15 replaced by amended claims 1 to 11."
3. [Where originally there were 14 claims and the amendments consist in cancelling some claims and in adding new claims]:
"Claims 1 to 6 and 14 unchanged; claims 7 to 13 cancelled; new claims 15, 16 and 17 added." or
"Claims 7 to 13 cancelled; new claims 15, 16 and 17 added; all other claims unchanged."
4. [Where various kinds of amendments are made]:
"Claims 1-10 unchanged; claims 11 to 13, 16 and 19 cancelled; claims 14, 15 and 16 replaced by amended claim 14; claim 17 subdivided into amended claims 15, 16 and 17; new claims 20 and 21 added."

"Statement under article 19(1)" (Rule 48.4)

The amendments may be accompanied by a statement explaining the amendments and indicating any impact that such amendments might have on the description and the drawings (which cannot be amended under Article 19(1)).

The statement will be published with the international application and the amended claims.

It must be in the language in which the international application is to be published.

It must be brief, not exceeding 500 words if in English or if translated into English.

It should not be confused with and does not replace the letter indicating the differences between the claims as filed and as amended. It must be filed on a separate sheet and must be identified as such by a heading, preferably by using the words "Statement under Article 19(1)."

It may not contain any disparaging comments on the international search report or the relevance of citations contained in that report. Reference to citations, relevant to a given claim, contained in the international search report may be made only in connection with an amendment of that claim.

Consequence if a demand for international preliminary examination has already been filed

If, at the time of filing any amendments under Article 19, a demand for international preliminary examination has already been submitted, the applicant must preferably, at the same time of filing the amendments with the International Bureau, also file a copy of such amendments with the International Preliminary Examining Authority (see Rule 62.2(a), first sentence).

Consequence with regard to translation of the international application for entry into the national phase

The applicant's attention is drawn to the fact that, where upon entry into the national phase, a translation of the claims as amended under Article 19 may have to be furnished to the designated/elected Offices, instead of, or in addition to, the translation of the claims as filed.

For further details on the requirements of each designated/elected Office, see Volume II of the PCT Applicant's Guide.

PATENT COOPERATION TREATY

PCT

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INTERNATIONAL SEARCH REPORT

(PCT Article 18 and Rules 43 and 44)

Applicant's or agent's file reference UTFB: 718WO	FOR FURTHER ACTION see Form PCT/ISA/220 as well as, where applicable, item 5 below.	
International application No. PCT/US2004/006580	International filing date (day/month/year) 04/03/2004	(Earliest) Priority Date (day/month/year) 10/03/2003
Applicant BOARD OF REGENTS, THE UNIVERSITY OF TEXAS SYSTEM		

This International Search Report has been prepared by this International Searching Authority and is transmitted to the applicant according to Article 18. A copy is being transmitted to the International Bureau.

This International Search Report consists of a total of 5 sheets.

☒ It is also accompanied by a copy of each prior art document cited in this report.

1. Basis of the report

a. With regard to the language, the international search was carried out on the basis of the international application in the language in which it was filed, unless otherwise indicated under this item.

☐ The international search was carried out on the basis of a translation of the international application furnished to this Authority (Rule 23.1(b)).

b. ☐ With regard to any nucleotide and/or amino acid sequence disclosed in the international application, see Box No. I.

2. ☐ Certain claims were found unsearchable (See Box II).

3. ☐ Unity of invention is lacking (see Box III).

4. With regard to the title,

☐ the text is approved as submitted by the applicant.

☒ the text has been established by this Authority to read as follows:

REGENERATION OF AN AQUEOUS SOLUTION FROM AN ACID GAS ABSORPTION PROCESS BY MULTISTAGE FLASHING AND STRIPPING

5. With regard to the abstract,

☐ the text is approved as submitted by the applicant.

☒ the text has been established, according to Rule 36.2(b), by this Authority as it appears in Box No. IV. The applicant may, within one month from the date of mailing of this international search report, submit comments to this Authority.

6. With regards to the drawings,

a. the figure of the drawings to be published with the abstract is Figure No. 3

☐ as suggested by the applicant.

☒ as selected by this Authority, because the applicant failed to suggest a figure.

☐ as selected by this Authority, because this figure better characterizes the invention.

b. ☐ none of the figures is to be published with the abstract.

INTERNATIONAL SEARCH REPORT

International application No.

6b

PCT/US2004/006580

Box No. IV Text of the abstract (Continuation of Item 5 of the first sheet)

An acid gas such as carbon dioxide, hydrogen sulfide, or a mixture thereof is removed from gaseous streams using aqueous absorption and stripping processes. By replacing the conventional stripper used to regenerate the aqueous solvent and capture the acid gas with a multipressure stripper (51) that combines acid gas compression with stripping, less energy is consumed. The multipressure stripper is a multistage flash (52,55,59) in which the total vapor flow from each stage is compressed and fed to the bottom of the previous flash stage at a higher pressure. In this process, the heat in the water content of the vapor exiting each stage is utilized at a higher pressure in the previous stage. The described stripping process generates the acid gas at a higher pressure without operating the stripper at a higher temperature, thereby reducing the energy consumption of the system.

A. CLASSIFICATION OF SUBJECT MATTER
IPC 7 B01D53/14

According to International Patent Classification (IPC) or to both national classification and IPC

B. FIELDS SEARCHEDMinimum documentation searched (classification system followed by classification symbols)
IPC 7 B01D

Documentation searched other than minimum documentation to the extent that such documents are included in the fields searched

Electronic data base consulted during the International search (name of data base and, where practical, search terms used)

EPO-Internal

C. DOCUMENTS CONSIDERED TO BE RELEVANT

Category *	Citation of document, with indication, where appropriate, of the relevant passages	Relevant to claim No.
X	US 4 780 115 A (RANKE GERHARD) 25 October 1988 (1988-10-25)	1-7
Y	figure 4	8-15
Y	US 5 435 977 A (CHAO I-MEEN) 25 July 1995 (1995-07-25)	10-15
Y	column 1, line 12 - line 27; figures 1,3,4	
Y	US 4 160 810 A (BENSON HOMER E ET AL) 10 July 1979 (1979-07-10)	8,9
Y	column 1, line 7 - line 55; figures 1,2	
Y	GB 1 484 050 A (HECKE F VAN) 24 August 1977 (1977-08-24)	8,9
A	page 1, line 9 - line 52; figures 1-4	1,2,5,6, 10-15
	----- -/-	

☒ Further documents are listed in the continuation of box C.☒ Patent family members are listed in annex.

* Special categories of cited documents:

- "A" document defining the general state of the art which is not considered to be of particular relevance
- "E" earlier document but published on or after the International filing date
- "L" document which may throw doubts on priority claim(s) or which is cited to establish the publication date of another citation or other special reason (as specified)
- "O" document referring to an oral disclosure, use, exhibition or other means
- "P" document published prior to the International filing date but later than the priority date claimed

- "T" later document published after the International filing date or priority date and not in conflict with the application but cited to understand the principle or theory underlying the invention
- "X" document of particular relevance; the claimed invention cannot be considered novel or cannot be considered to involve an inventive step when the document is taken alone
- "Y" document of particular relevance; the claimed invention cannot be considered to involve an inventive step when the document is combined with one or more other such documents, such combination being obvious to a person skilled in the art.
- "&" document member of the same patent family

Date of the actual completion of the International search

14 July 2004

Date of mailing of the International search report

26/07/2004

Name and mailing address of the ISA

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Authorized officer

de Blasio, A

INTERNATIONAL SEARCH REPORT

International Application No
PCT/US2004/006580

10

C.(Continuation) DOCUMENTS CONSIDERED TO BE RELEVANT

Category *	Citation of document, with indication, where appropriate, of the relevant passages	Relevant to claim No.
A	US 4 258 019 A (DOERGES ALEXANDER ET AL) 24 March 1981 (1981-03-24) figure 3 -----	1,2,5-7, 10,11

INTERNATIONAL SEARCH REPORT

Information on patent family members

International Application No
PCT/US2004/006580

Patent document cited in search report		Publication date	Patent family member(s)	Publication date
US 4780115	A	25-10-1988	DE 3613110 A1 DD 256079 A5 EP 0249702 A2	22-10-1987 27-04-1988 23-12-1987
US 5435977	A	25-07-1995	NONE	
US 4160810	A	10-07-1979	BE 49 T1 CA 1106153 A1 DE 2952855 A1 EP 0004043 A1 FR 2446125 A1 GB 2041348 A ,B IN 150688 A1 IT 1148284 B JP 1367456 C JP 54155172 A JP 61035890 B MX 150006 A NL 190981 B NL 7915016 T	28-03-1980 04-08-1981 31-07-1980 19-09-1979 08-08-1980 10-09-1980 20-11-1982 26-11-1986 26-02-1987 06-12-1979 15-08-1986 28-02-1984 01-07-1994 29-04-1980
GB 1484050	A	24-08-1977	AT 369421 B AT 979075 A AU 503541 B2 BE 836929 A1 BR 7508536 A CA 1068638 A1 DE 2557531 A1 DK 588975 A ES 443792 A1 FR 2295781 A1 IE 41981 B1 IN 144785 A1 IT 1051858 B JP 1237307 C JP 51089876 A JP 59007487 B NZ 179638 A PH 12514 A ZA 7507939 A	27-12-1982 15-05-1982 06-09-1979 22-06-1976 24-08-1976 24-12-1979 08-07-1976 25-06-1976 16-04-1977 23-07-1976 07-05-1980 08-07-1978 20-05-1981 31-10-1984 06-08-1976 18-02-1984 18-12-1978 20-04-1979 24-11-1976
US 4258019	A	24-03-1981	DE 2743753 A1 GB 2004850 A ,B JP 1473312 C JP 54051997 A JP 63017877 B NL 7807194 A ,B,	12-04-1979 11-04-1979 27-12-1988 24-04-1979 15-04-1988 02-04-1979

PATENT COOPERATION TREATY

72

From the
INTERNATIONAL SEARCHING AUTHORITY

PCT

WRITTEN OPINION OF THE INTERNATIONAL SEARCHING AUTHORITY (PCT Rule 43bis.1)

To:

see form PCT/ISA/220

Date of mailing
(day/month/year) see form PCT/ISA/210 (second sheet)

Applicant's or agent's file reference
see form PCT/ISA/220

FOR FURTHER ACTION
See paragraph 2 below

International application No.
PCT/US2004/006580

International filing date (day/month/year)
04.03.2004

Priority date (day/month/year)
10.03.2003

International Patent Classification (IPC) or both national classification and IPC
B01D53/14

Applicant
BOARD OF REGENTS, THE UNIVERSITY OF TEXAS SYSTEM

1. This opinion contains indications relating to the following items:

- ☒ Box No. I Basis of the opinion
- ☒ Box No. II Priority
- ☐ Box No. III Non-establishment of opinion with regard to novelty, inventive step and industrial applicability
- ☐ Box No. IV Lack of unity of invention
- ☒ Box No. V Reasoned statement under Rule 43bis.1(a)(i) with regard to novelty, inventive step or industrial applicability; citations and explanations supporting such statement
- ☐ Box No. VI Certain documents cited
- ☐ Box No. VII Certain defects in the international application
- ☐ Box No. VIII Certain observations on the international application

2. FURTHER ACTION

If a demand for international preliminary examination is made, this opinion will usually be considered to be a written opinion of the International Preliminary Examining Authority ("IPEA"). However, this does not apply where the applicant chooses an Authority other than this one to be the IPEA and the chosen IPEA has notified the International Bureau under Rule 66.1bis(b) that written opinions of this International Searching Authority will not be so considered.

If this opinion is, as provided above, considered to be a written opinion of the IPEA, the applicant is invited to submit to the IPEA a written reply together, where appropriate, with amendments, before the expiration of three months from the date of mailing of Form PCT/ISA/220 or before the expiration of 22 months from the priority date, whichever expires later.

For further options, see Form PCT/ISA/220.

3. For further details, see notes to Form PCT/ISA/220.

Name and mailing address of the ISA:



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Authorized Officer

de Biasio, A

Telephone No. +49 89 2399-8627



**WRITTEN OPINION OF THE
INTERNATIONAL SEARCHING AUTHORITY**International application No.
PCT/US2004/006580

Box No. I Basis of the opinion

1. With regard to the language, this opinion has been established on the basis of the international application in the language in which it was filed, unless otherwise indicated under this item.
☐ This opinion has been established on the basis of a translation from the original language into the following language , which is the language of a translation furnished for the purposes of international search (under Rules 12.3 and 23.1(b)).
2. With regard to any nucleotide and/or amino acid sequence disclosed in the international application and necessary to the claimed invention, this opinion has been established on the basis of:
 - a. type of material:
☐ a sequence listing
☐ table(s) related to the sequence listing
 - b. format of material:
☐ in written format
☐ in computer readable form
 - c. time of filing/furnishing:
☐ contained in the international application as filed.
☐ filed together with the international application in computer readable form.
☐ furnished subsequently to this Authority for the purposes of search.
3. ☐ In addition, in the case that more than one version or copy of a sequence listing and/or table relating thereto has been filed or furnished, the required statements that the information in the subsequent or additional copies is identical to that in the application as filed or does not go beyond the application as filed, as appropriate, were furnished.
4. Additional comments:

**WRITTEN OPINION OF THE
INTERNATIONAL SEARCHING AUTHORITY**

International application No.
PCT/US2004/006580

Box No. II Priority

1. ☒ The following document has not been furnished:

☒ copy of the earlier application whose priority has been claimed (Rule 43bis.1 and 66.7(a)).

☐ translation of the earlier application whose priority has been claimed (Rule 43bis.1 and 66.7(b)).

Consequently it has not been possible to consider the validity of the priority claim. This opinion has nevertheless been established on the assumption that the relevant date is the claimed priority date.

2. ☐ This opinion has been established as if no priority had been claimed due to the fact that the priority claim has been found invalid (Rules 43bis.1 and 64.1). Thus for the purposes of this opinion, the international filing date indicated above is considered to be the relevant date.

3. Additional observations, if necessary:

Box No. V Reasoned statement under Rule 43bis.1(a)(I) with regard to novelty, inventive step or industrial applicability; citations and explanations supporting such statement

1. Statement

Novelty (N)	Yes: Claims	8-15
	No: Claims	1-7
Inventive step (IS)	Yes: Claims	-
	No: Claims	1-15
Industrial applicability (IA)	Yes: Claims	1-15
	No: Claims	-

2. Citations and explanations

see separate sheet

Preliminary Remark:

1. Claim 1 has been interpreted in such a way that the last technical features of item (a) does not apply to the first flash stage, as it should make no sense to feed its vapour to a previous stage that does not exist. Consequently, the vapour obtained from the first stage is not necessarily compressed, which should correspond to the embodiment of figure 4 of the present application. A narrower interpretation of claim 1, i.e. vapour of stage 1 is necessarily compressed, would lead to a contradiction with the embodiment of figure 4 (Art. 6 PCT).
2. Similarly, from the embodiment of figure 3 it does not seem to be necessary that the last flash stage also comprises stripping. In figure 3, there is no stripping current in the lowest flashing stage 52. In other words, would it be necessary for each flashing stage of the process of claim 1 to include a stripping current, then the embodiment of figure 3 would not fall within the definition given in claim 1 (Art. 6 PCT).

Re Item V

Reasoned statement with regard to novelty, inventive step or industrial applicability; citations and explanations supporting such statement

Reference is made to the following document/s/:

D1: US-A-4.780.115

D2: US-A-5.435.977

- 1.1 The subject-matter of claim 1 does not seem to fulfill the requirements of Art. 33(2) PCT:

US-A-4.780.115 (D1) discloses (cf figure 4) a method for removing an acid gas from a gaseous stream containing said acid gas by using an aqueous absorption solution (col. 2) and stripping equipment, said method comprising:

- (a) passing the acid-gas-rich solvent stream (5) exiting the absorbing equipment through a multipressure stripper in which the acid-gas-rich solvent stream passes through multiple flash stages (drums 7, 10, 14) such that the vapour obtained by flashing the acid-gas-rich solvent stream in each flash stage strips the acid gas from the acid-gas-rich solvent stream, resulting in an acid-gas-rich gaseous stream exiting the multipressure stripper and an acid-gas-

lean solvent stream exiting the multipressure stripper; wherein the multipressure stripper is operated at multiple pressure levels such that the vapour obtained in each flash stage (with exception of the first one) is compressed in a compressor (6a, 9) and fed to the previous flash stage at a higher pressure, and.

- (b) recycling the acid-gas-lean solvent stream exiting the multipressure stripper in step (a) back to the absorption equipment (2) at least once.

All technical features of claim 1 appear to be known from D1 which is thus novelty destroying (Art. 33(2) PCT).

2. The equipment disclosed in figure 4 of D1 differs from that defined in present claim 11 only in that the single flash stages are comprised in a stripping column.

US-A-5.435.977 (D2) discloses a similar method using a stripping column including in its lower portion two flash stages. The teaching of D2 is to place flash stages and stripper within one single column in order to cut the investment and operating expenses (col. 2, lines 30-49). It would thus be obvious for the skilled person starting from D1 to apply the teaching of D2 and thus obtain an equipment as defined in present claim 11 (Art. 33(3) PCT).

3. Dependent claims 2-10 and 12-15 do not contain any features which, in combination with the features of any claim to which they refer, meet the requirements of the PCT in respect of novelty and/or inventive step, see all documents cited in the search report and the corresponding passages. As far as the additional technical features of dependent claims 3, 4, 13, 14 are concerned, they merely appear to be straightforward possibilities from which the skilled person would select, in accordance with circumstances, without the exercise of inventive skill.

[54] PROCESS FOR REMOVING UNDESIRABLE COMPONENTS FROM GASES

[75] Inventor: Gerhard Ranke, Poecking, Fed. Rep. of Germany

[73] Assignee: Linde Aktiengesellschaft, Wiesbaden, Fed. Rep. of Germany

[21] Appl. No.: 39,874

[22] Filed: Apr. 20, 1987

[30] Foreign Application Priority Data

Apr. 18, 1986 [DE] Fed. Rep. of Germany 3613110

[51] Int. Cl.⁴ B01D 47/00

[52] U.S. Cl. 55/43; 55/44;

55/48; 55/68; 55/73; 62/17

[58] Field of Search 55/68, 73, 43, 48, 46, 55/44; 62/17

[56] References Cited

U.S. PATENT DOCUMENTS

3,899,312 8/1975 Krus 55/73

4,324,567 4/1982 Ranke et al. 55/43

FOREIGN PATENT DOCUMENTS

0053424 6/1982 European Pat. Off. 55/43

119610 5/1976 German Democratic Rep. .

147622 4/1981 German Democratic Rep. 55/73
286622 6/1929 United Kingdom 55/73

OTHER PUBLICATIONS

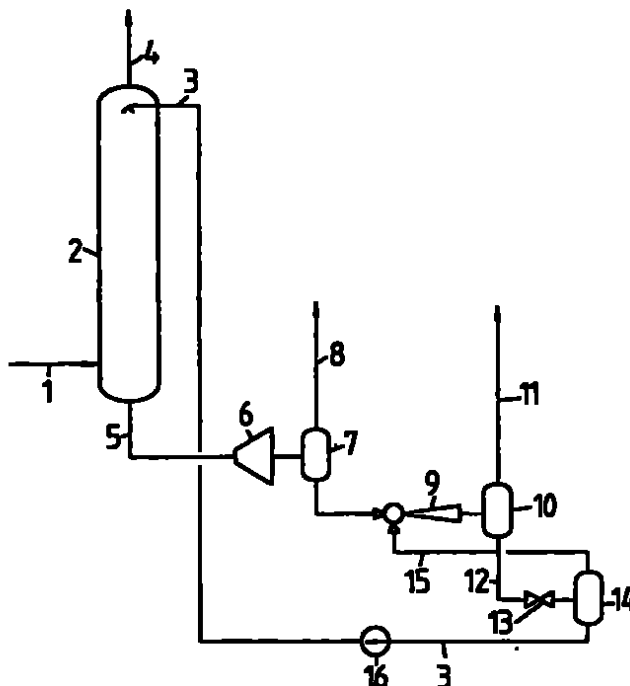
Ullmanns Encyklopaedie der technischen Chemie, 4th Edition, vol. 3, pp. 107-110 and p. 172.

Primary Examiner—Bernard Novick
Attorney, Agent, or Firm—Millen & White

[57] ABSTRACT

In a process for the regeneration of a scrubbing medium which has been utilized in a scrubbing operation and therein loaded with scrubbed-out components, the scrubbing medium is expanded and regenerated and hence can be reused in the scrubbing operation. In order to be able to subject expansion gases obtained during the expansion to an after-treatment under elevated pressure, or to be able to discharge these expansion gases under a higher pressure, the loaded scrubbing medium is expanded in at least two stages. Expansion in one stage is performed by means of a liquid-jet compressor and the expansion gases of a subsequent expansion stage are brought to an elevated pressure by means of the liquid-jet compressor.

19 Claims, 4 Drawing Sheets



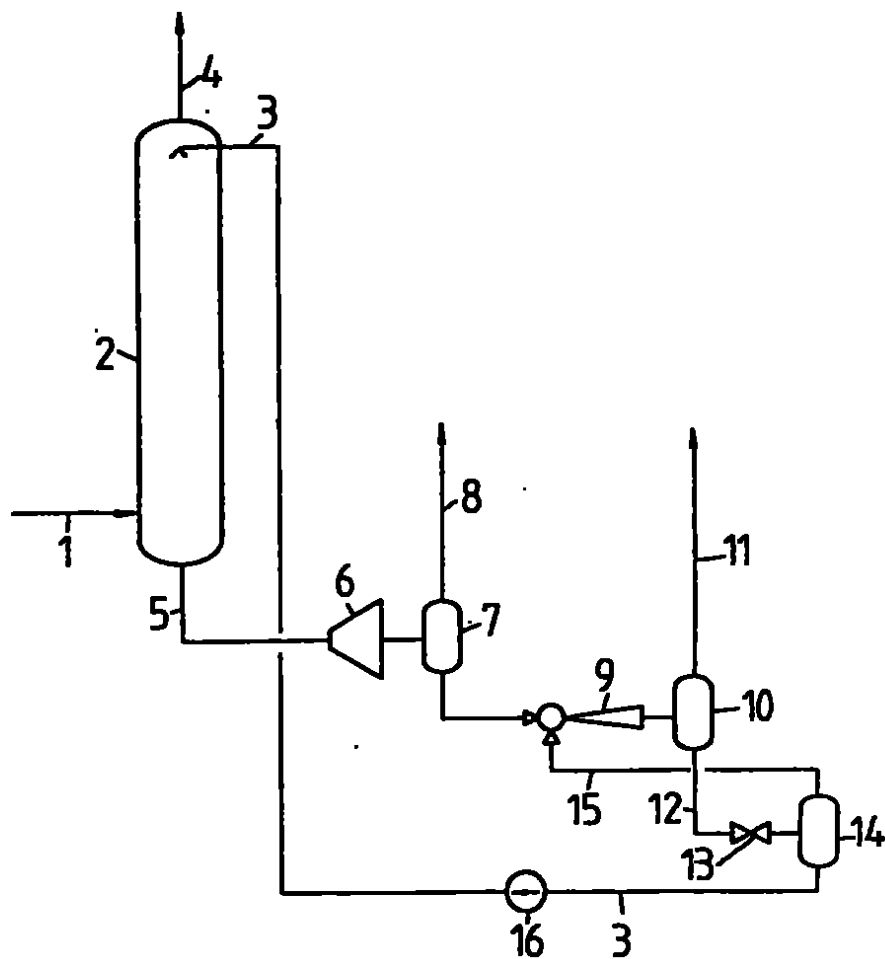


Fig.1

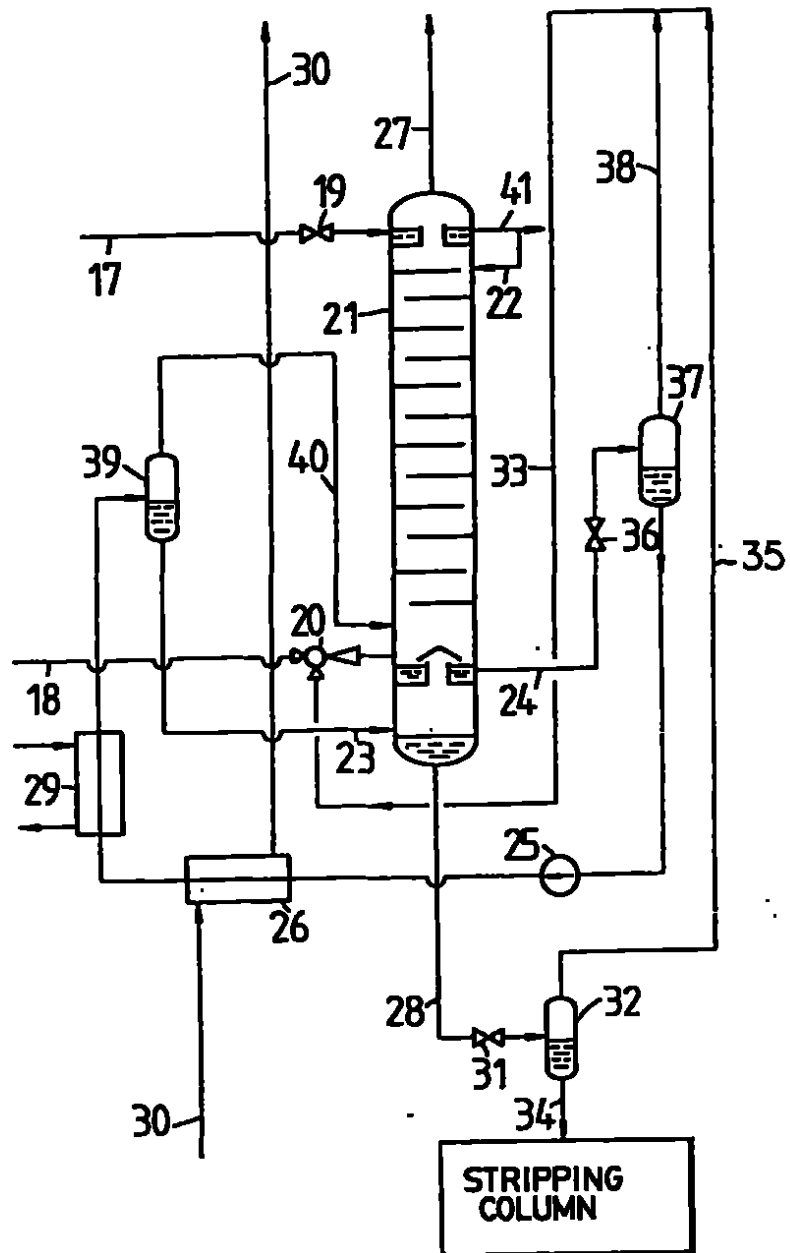


Fig. 2

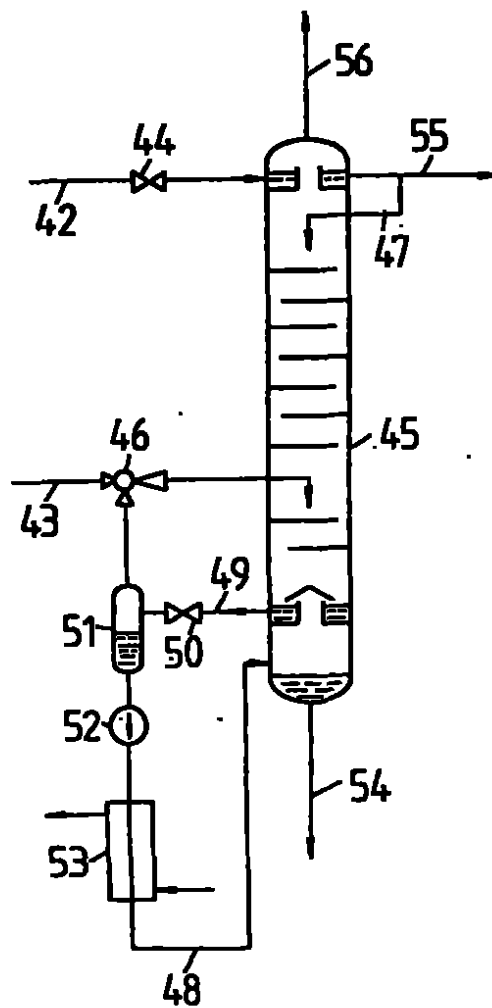


Fig.3

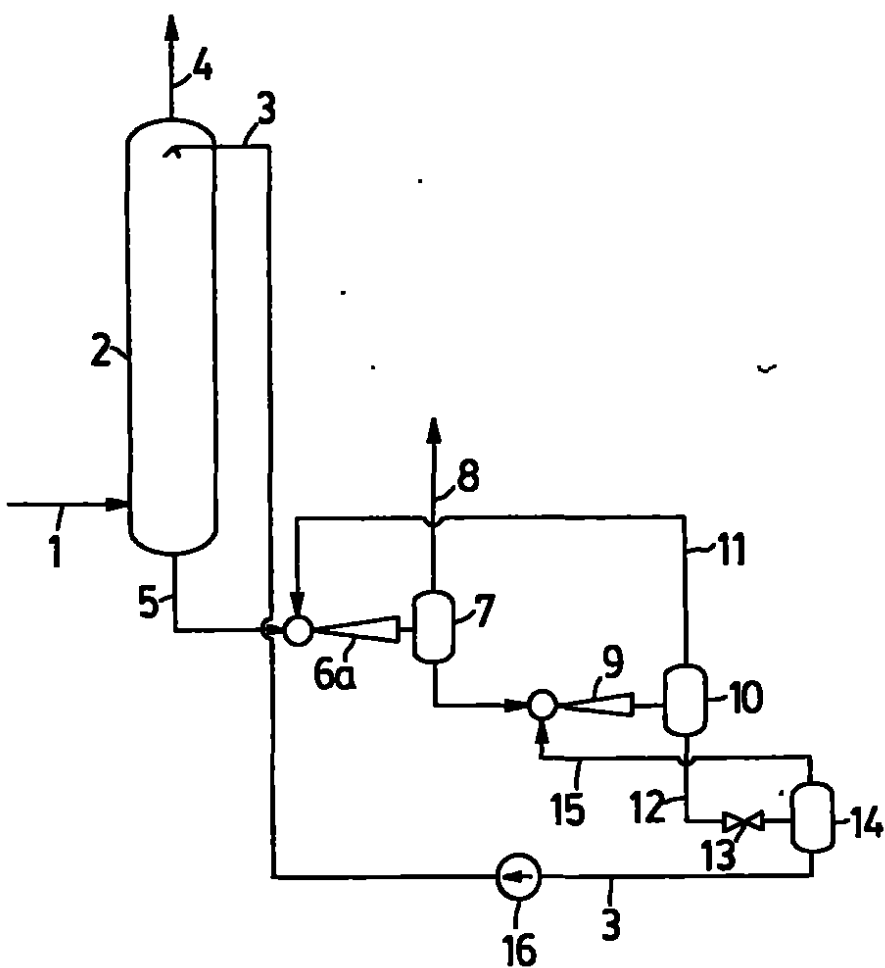


Fig. 4

PROCESS FOR REMOVING UNDESIRABLE COMPONENTS FROM GASES

BACKGROUND OF THE INVENTION

This invention relates to a process for the regeneration of a scrubbing medium which has been utilized in a pressurized scrubbing operation and therein loaded with scrubbed-out components, this scrubbing medium being expanded, regenerated, and then reused in the scrubbing operation.

In scrubbing operations, throttle valves are usually employed for expanding the loaded scrubbing medium from the scrubbing column to a lower pressure. Expansion turbines are utilized only in cases where large liquid quantities are involved for purposes of energy recovery. The use of expansion turbines, in spite of energy recovery, has several drawbacks.

The cost of turbines is very high so that they are economical only if the amount of recoverable energy is potentially large.

If the loaded scrubbing medium is extensively degasified, difficulties are encountered in the turbine due to an unfavorable liquid/gas volume ratio. On account of the fact that expansion turbines are prone to trouble, it is necessary either to insert a second unit in parallel as a standby or to install a throttle valve.

If an expansion turbine is coupled with the scrubbing medium pump, a motor must also be connected to the pump to provide make-up pump energy as required, and in order to provide the entire pump energy in case of failure of the turbine or during startup of the plant.

Based on the aforementioned disadvantages, it is customary, for example in acid gas scrubbing operations, to utilize expansion turbines only in large installations (i.e., in case of liquid quantities of several 100 m³/h) and with relatively large pressure drops. On the other hand, there is interest in providing for gas compression in many scrubbing operations—especially in physical scrubbing processes for acid gas removal—wherein the liquid quantities or pressure drops do not justify the employment of an expensive turbine. For example, gas compression becomes of interest when expansion gases are to be recycled into the crude gas or utilized as a product for further treatment or when the scrubbing medium is to be subjected to a vacuum in order to lower the gas load in the expanded scrubbing medium.

Customarily, compressors are utilized for such compression needs. Besides the energy consumed for compression, an additional disadvantage associated with the employment of compressors is that the gas is heated in the compressor. For this reason, in most cases a cooler is also required for recooling the compressed gas. If the compressed gas is to be further treated in the scrubbing operation, which is very often the case, especially with H₂S-containing gases, refrigeration losses are imposed by this compression, and these losses must be covered by external refrigeration.

Therefore, East German Pat. No. 119,610 teaches one to effect compression of expansion gases, obtained under vacuum, by means of a jet compressor, also called a jet ejector. In this process, gases obtained during an intermediate expansion, which have a higher pressure than the gases to be compressed, are utilized as the propulsion media for the jet compressor.

The essential advantage of this process resides in that no external energy is needed for compression, and, at the same time, there is no heating for the exhausted

expansion gases. However, the method can only be employed if a relatively large amount of expansion gas, at a higher pressure, is already available and, at the same time, if all gases can be discharged without being under pressure. A further drawback involved is the additional installation of expansion tanks which serve for obtaining propulsion gases for the jet compressors.

SUMMARY OF THE INVENTION

An object of one aspect of the present invention is to provide a process wherein expansion gases obtained in an expansion step can be further treated under elevated pressure or can be discharged at a higher pressure.

Upon further study of the specification and appended claims, further objects and advantages of this invention will become apparent to those skilled in the art.

These objectives have been attained according to this invention by expanding the loaded scrubbing medium in at least two stages, performing the expansion in one stage by means of a liquid-jet ejector or liquid-jet compressor, and then bringing the expansion gases from a subsequent expansion stage to an elevated pressure by means of the liquid-jet compressor.

It has been found in the context of this invention that, besides gas, it is also possible to use the loaded scrubbing medium as a liquid propulsion medium in a liquid-jet compressor whereby a substantially larger amount of gas can be compressed than, for example, in the process according to East German Pat. No. 119,610 which employs a gaseous propulsion medium.

Since the loaded scrubbing medium is used as the propulsion medium rather than expansion gases, the liquid-jet compressor herein operates even if, during pressure lowering, no gas or only very little gas is released, and thus can operate in cases where the method known from the East German patent cannot be utilized.

In some instances, especially in cases where concomitantly dissolved inert gases are to be obtained under maximally high pressure, it is advantageous to expand the loaded scrubbing medium in a two-stage procedure so that the gases can be recycled to the pressure scrubbing step with a minimum expenditure of energy. The concomitantly dissolved inert gases are released during the first expansion and then the scrubbing medium is further expanded in a second expansion wherein it is employed as a propulsion medium for a liquid-jet compressor. Thus, initially the loaded scrubbing medium is expanded to an intermediate pressure, for example in an expansion turbine, and the liberated gaseous phase is recycled to a point upstream of the pressure scrubbing stage. The partially expanded scrubbing medium is then fed to a liquid-jet compressor for further expansion. It is clear that also in this case of a two-stage pressure lowering operation, the compression energy produced by such a pressure lowering operation can be utilized for raising the pressure of expansion gases obtained in subsequent process steps.

According to another advantageous embodiment for the process according to this invention, the loaded scrubbing medium can be expanded stepwise by means of several series-connected liquid-jet compressors. At the same time, the gases released in a subsequent expansion stage can herein be compressed in a preceding expansion stage. The advantage of this mode of operation resides in that the pressure of the expansion gases can be raised to a greater extent. It thus becomes possi-

ble to adapt the method of this invention to the respective process conditions.

The process of this invention can be utilized in a wide variety of systems which need gas compression based on a very diversified requirements.

One utilization of gas compression can be to lower the partial pressure of acid gases in the cold-regenerated scrubbing medium of a sour gas scrubbing operation without the use of stripping gas or vacuum compressor means.

In water scrubbing operations for the removal of CO₂ from gases, it is possible, for example, to reduce the load in loaded scrubbing water by lowering its pressure and thereby lower the load of the regenerated water. Since the regenerated scrubbing medium recycled to the scrubbing operation has a lower load, the result is a lower residual CO₂ content in the scrubbed product due to greater absorption. This is accomplished without the use of stripping gas or vacuum, i.e., without a compressor, while at the same time improving the yield of removed CO₂.

A further utilization of gas compression can be to increase the CO₂ yield in physical scrubbing operations before a subsequent stripping operation. Gas compression can also be suitable for obtaining additional peak cold in physical scrubbing processes.

Finally, it is possible by gas compression to reduce the CO₂ load in chemical scrubbing operations prior to thermal regeneration, or to reduce the CO₂ load of the scrubbing medium that is regenerated solely by expansion. Thereby, the quantity of scrubbing medium circulated can be reduced on account of greater CO₂ absorption. Also, the energy requirement of a subsequent fine scrubbing step can be reduced since, with the more highly regenerated scrubbing medium, the CO₂ can be scrubbed out more advantageously from the crude gas in the rough or initial scrubbing stage.

BRIEF DESCRIPTION OF THE DRAWINGS

Various other objects, features and attendant advantages of the present invention will be more fully appreciated as the same becomes better understood when considered in conjunction with the schematically illustrated flowsheets in the accompanying drawings, in which like reference characters designate the same or similar parts, and wherein:

FIG. 1 shows removal of CO₂ from a gas stream by a water scrubbing operation,

FIG. 2 shows how the process of the invention can be used to increase CO₂ yield in a physical scrubbing operation,

FIG. 3 shows the generation of peak cold in a cold scrubbing operation, and

FIG. 4 illustrates step-wise expansion of a loaded scrubbing medium using several liquid-jet compressors in series.

DETAILED DESCRIPTION OF THE DRAWINGS

According to FIG. 1, a scrubbing column 2 is charged via conduit 1 with 10,000 Nm³/h of crude gas under a pressure of 40 bar and at a temperature of 30° C. The crude gas has the following composition:

H ₂	65 vol-%
N ₂ + CO	5 vol-%

-continued

CO ₂	30 vol-%
-----------------	----------

Via conduit 3, the scrubbing column 2 is charged with 500 t/h of scrubbing water which absorbs CO₂ from countercurrently rising crude gas. Via conduit 4, pure product gas at a temperature of about 30.5°-31° C. can thus be withdrawn containing at this point merely maximally about 1 vol-% of CO₂.

Via conduit 5, the CO₂-loaded water is discharged at a temperature which is almost identical with the temperature of the pure product gas, because of the large amount of water the increase of temperature is very small. In order to obtain a residual CO₂ content of below about 1 vol-% in the product gas, the loaded water is first expanded to 10 bar in an expansion turbine 6. After separating the thus-released gaseous phase via conduit 8, namely 926 Nm³/h, consisting of:

CO ₂	76 vol-%
H ₂ + CO	24 vol-%

in a separator 7, the water is expanded to a pressure of 1.2 bar by way of a liquid-jet compressor 9. The thereby released gaseous phase (2,063 Nm³/h, containing 95.4 vol-% of CO₂, remainder being H₂ + CO) is withdrawn from a further phase separator 10 via conduit 11 under a pressure of 1.2 bar. By way of conduit 12, the water is withdrawn from phase separator 10 and further expanded to 0.3 bar in a conventional throttle valve 13. In a downstream phase separator 14, the thus-obtained expansion gas (321 Nm³/h of CO₂ + H₂) is discharged via conduit 15 and raised to a pressure of 1.2 bar in the liquid-jet compressor 9. The regenerated water is withdrawn from phase separator 14 via conduit 3 with pump 16 and reintroduced into the scrubbing column 2.

In the conventional processes, the amount of gas withdrawn from separator 14 is compressed in conduit 15 by means of a vacuum compressor. The energy expenditure required therefor amounts to about 22 kW. The primary advantages of the process of this invention as compared with the process customary heretofore resides not only in a saving of energy but more importantly it permits the elimination of the vacuum compressor, i.e., a piston compressor with movable parts, representing a wear-prone part of the installation requiring frequent servicing. The liquid-jet compressor, in contrast thereto, represents a substantially cheaper and a less maintenance intensive solution. FIG. 4 illustrates an embodiment similar to that of FIG. 1 except that the expansion turbine 6 is replaced by a liquid-jet compressor 6a. Loaded scrubbing medium is thus expanded by two liquid-jet compressors, i.e. 6a and 9, connected in series.

FIG. 2 illustrates a section of the regenerating portion of an installation for the separation and obtaining of gaseous components, especially CO₂, COS and H₂S, from a gaseous mixture by means of a physical scrubbing operation.

By way of conduit 17, 55 t/h of scrubbing methanol, almost saturated with CO₂, from a scrubbing column (not shown) is conducted at a temperature of -40° and under a pressure of 25 bar. By way of conduit 18, 45 t/h of scrubbing methanol is introduced which has been loaded in the scrubbing column (not shown) with CO₂, H₂S and COS and which is obtained under a pressure of

likewise 25 bar and at a temperature of -7°C . The methanol stream in conduit 18, loaded with CO_2 and also sulfur compounds, represents the first scrubbing medium stream; the methanol stream in conduit 17, loaded only with CO_2 , represents the second scrubbing medium stream.

The first scrubbing medium stream is expanded in a liquid-jet compressor 20 to a pressure of 3 bar. The second scrubbing medium stream is expanded in a throttle valve 19 to a pressure of 3 bar. Both scrubbing medium streams are then introduced into an enrichment column 21. The first scrubbing medium stream is introduced into the middle portion and the second stream is introduced into the head portion which is equipped with separator. During expansion, gaseous fractions are liberated in each case. In the case of the second scrubbing medium stream, a fraction of predominantly CO_2 is liberated and discharged from the column head via conduit 27, and in the case of the first scrubbing medium stream, a fraction containing CO_2 as well as H_2S and COS is liberated which then rises upwardly in the interior of the column. The proportions of H_2S and COS are scrubbed from this gaseous, rising fraction by a portion of the second scrubbing medium stream that is returned from the separator at the column head via a conduit 22 into the upper section of the enrichment column 21. This returned portion of the second scrubbing medium stream amounts to 35 t/h.

The scrubbing medium streams collected above a flue plate in the lower part of the enrichment column 21 are discharged via a conduit 24, brought to an intermediate pressure of 0.7 bar in a throttle valve 36, and then introduced into a separator 37. The gaseous fraction released after intermediate expansion in valve 36, containing CO_2 and small proportions of H_2S and COS , is passed on via conduit 38 to a conduit 33 which will be described below. The bottoms liquid from separator 37 is conveyed with the aid of a liquid pump 25 through heat exchangers 26 and 29 and subsequently returned, by way of a conduit 23, into the lowermost section of the enrichment column 21 which functions as a separator. During this step, the scrubbing medium is heated from -55°C to -43°C in heat exchanger 26 by heat exchange with methanol, fed via a conduit 30 (the methanol in conduit 30 being completely regenerated in a subsequent part of the installation, not shown); and then heated to -33°C in heat exchanger 29 by heat exchange with a methanol stream, previously loaded with CO_2 , which yields heat of absorption. In a separator 39, a gaseous fraction is separated which contains CO_2 , H_2S and COS and is introduced via a conduit 40 into the enrichment column 21. This gaseous fraction introduced by conduit 40 containing mostly CO_2 with minor proportions of H_2S and COS , flows through the enrichment column 21 in the upward direction and during this step is likewise freed of the sulfur-containing components by means of the second scrubbing medium stream. Gases separated in the lowermost section of the enrichment column flow through the flue plate and upwardly through the enrichment column whereby they are also scrubbed by the second scrubbing medium stream.

Via a conduit 27, 23,316 Nm^3/h of a first product gas is withdrawn from the head of the enrichment column 21. This product gas consists of 23,000 Nm^3/h of CO_2 , 247 Nm^3/h of H_2 , 49 Nm^3/h of CO , Ar , CH_4 , and 20 Nm^3/h of N_2 . The purity of the gaseous CO_2 product is 98.7 vol-%. The CO_2 product is obtained at a temperature of -49°C .

From the bottom of the enrichment column 21, 80 t/h of methanol, now enriched in H_2S and COS , is withdrawn at a temperature of -33°C . This methanol entrains, in dissolved form, 7,232 Nm^3/h of CO_2 as well as 1,025 Nm^3/h of H_2S and COS . This enriched methanol is introduced via a conduit 28 into a throttle valve 31, expanded to a pressure of 0.7 bar, and then introduced into a separator 32. The gaseous fraction liberated during this expansion (4,896 Nm^3/h of CO_2 and 295 Nm^3/h of H_2S and COS) is removed from separation 32 via conduit 35 and is delivered via conduit 33 to the liquid-jet compressor 20 and, after recompression to the pressure of the enrichment column 21, fed into the latter.

In accordance with conventional operation prior to this invention, this released gaseous fraction in conduit 33 was taken in by means of a compressor and then recompressed. During the recompression step, the gaseous fraction was heated up. Since the H_2S - and COS -containing fraction, however, must be desulfurized in the enrichment column 21 at low temperatures with cold methanol, the heating-up occurring during compression represents a cold value loss which must be compensated for again with external refrigeration in an exchanger. However, this raises the energy requirement for increasing the CO_2 yield. With the use of the process according to this invention, the compressor and re-cooler can be replaced by a liquid-jet compressor which also simultaneously takes over the task of a heretofore employed throttle valve for expanding the methanol introduced via conduit 18. In this throttle valve of prior processes, the entire energy which was liberated during expansion of the first scrubbing medium stream loaded with H_2S , COS and CO_2 was dissipated. This energy is now usefully exploited for the compression of produced expansion gases. Besides saving compression energy and external cold, this also makes it possible to replace an expensive part of the installation which frequently required servicing, i.e., the compressor.

From the bottom for the separator 32, 80 t/h of scrubbing methanol is withdrawn via conduit 34 and fed into a stripping column (not shown) for complete regeneration. The portion of the second scrubbing medium stream (20 t/h) that has not been recycled is withdrawn via conduit 41 and utilized for rescrubbing sulfur-containing components in the stripping column (not shown).

FIG. 3 likewise shows only a section of an entire installation for obtaining hydrogen and CO_2 .

By way of a conduit 42, 55 t/h of methanol is conveyed from a scrubbing column (not shown). This methanol, representing a first scrubbing medium stream, is obtained under a pressure of 27 bar and at a temperature of -50°C . It is almost saturated with CO_2 . By way of conduit 43, 45 t/h of methanol loaded with CO_2 and H_2S from the same scrubbing column is fed under a pressure of 27 bar and at a temperature of -6°C . This methanol stream represents a second scrubbing medium stream.

The first scrubbing medium stream is expanded in a throttle valve 44 to 3.5 bar and introduced into the upper section, serving as a separator, of a CO_2 column 45. The second scrubbing medium stream is expanded to 3.5 bar in a liquid-jet compressor 46 and fed into the middle section of the CO_2 column. Concentration and recovery of CO_2 take place in the CO_2 column. For this purpose, the gas rising in the column is scrubbed with 35 t/h of H_2S -free scrubbing medium introduced via conduit 47. The gases rising in the CO_2 column 45 are

composed of the gases formed by expansion in the liquid-jet compressor 46 and of the gases passing through conduit 48 into the lowermost section of the column 45. For producing these last-mentioned gases, liquid is withdrawn via conduit 49 from a fine plate arranged in the lower section of column 45, expanded in a throttle valve 50 to 1.0 bar and conveyed, after separation of the thus-liberated expansion gases in separator 51, through a heat exchanger 53 by means of a pump 52. In heat exchanger 53 the liquid stream is heated from -63°C . to -35°C . From heat exchanger 53 the heated stream is reintroduced into the column via conduit 48. In heat exchanger 53, the liquid is heated, for example, against a methanol stream (the methanol stream being completely regenerated in a subsequent stripping column, not shown), so that a portion of the components dissolved in the liquid is released, and a gas-liquid mixture passes, via conduit 48, into the column 45.

By expansion in valve 50, a portion of the concomitantly dissolved gases is liberated in gaseous phase with simultaneous cooling of the scrubbing medium. These expansion gases, separated in separator 51, are taken in by the liquid-jet compressor 46 and compressed to the pressure of column 45. By use of the liquid-jet compressor, cold can be obtained at the lowest temperature level for the scrubbing operation without additional energy, and thus the entire scrubbing process can be made more economical. The cold, obtained at a low temperature level, has the effect that the regenerated scrubbing medium can be subcooled to a greater extent, and accordingly the amount of scrubbing medium or the number of plates in the scrubbing column can be reduced.

The methanol introduced into the column 45 via conduit 42 can be fed at the very top into the head of the column, since the gas released during expansion in 44 is pure CO_2 which need no longer be scrubbed. In the head of the CO_2 column 45, the temperature is -45°C . The gas being withdrawn through conduit 56 (25,405 Nm^3/h) has the following composition:

H_2	1.34 mol-%
$\text{N}_2 + \text{Ar} + \text{CH}_4$	0.14 mol-%
CO	0.12 mol-%
CO_2	98.04 mol-%
H_2S	2 ppm

The cold inherent in the gas is normally furthermore transferred to the crude gas by heat exchange (not illustrated) so that the CO_2 product is then available at a temperature of 27°C . and under a pressure of 2 bar.

From the bottoms of the column, via conduit 54, 80 t/h of methanol containing 1,841 Nm^3/h of H_2S and 7,232 Nm^3/h of CO_2 is discharged and fed to a stripping column (not shown) for complete regeneration. Via conduit 55, 20 t/h of methanol is introduced to the stripping column for rescrubbing H_2S .

The ejectors that can be used in the preceding embodiments are generally described in "Ullmanns Encyclopadie de technischen Chemie", 4th edition, Vol. 3, p. 107-110 and p. 172.

From the foregoing description, one skilled in the art can easily ascertain the essential characteristics of this invention, and without departing from the spirit and scope thereof, can make various changes and modifications of the invention to adapt it to various usages and conditions.

What is claimed is:

1. In a process for regenerating a scrubbing medium which has been utilized in a scrubbing operation, said scrubbing operation comprising contacting a gaseous mixture with a scrubbing medium to obtain a scrubbed gas fraction and said scrubbing medium loaded with scrubbed-out components from said gaseous mixture, and said process for regenerating comprising expanding the loaded scrubbing medium to release scrubbed-out components therefrom, the improvement comprising:

expanding said loaded scrubbing medium in a first stage and at least one downstream expansion stage to obtain resultant regenerated scrubbing medium, wherein said loaded scrubbing medium is expanded in said first stage through a liquid-jet compressor to form resultant expanded loaded scrubbing medium and first stage expansion gases, and

further expanding the resultant expanded loaded scrubbing medium in one of said at least one downstream expansion stage to form further expanded scrubbing medium and downstream expansion gases, separating said downstream expansion gases from said further expanded scrubbing medium, and passing said downstream expansion gases into said liquid-jet compressor.

2. A process according to claim 1, wherein said loaded scrubbing medium is expanded stepwise in a series of expansion stages by means of series-connected liquid-jet compressors.

3. A process according to claim 1, wherein said scrubbing operation comprises a process for scrubbing acid gas from said gaseous mixture.

4. A process according to claim 1, wherein said scrubbing operation comprises a water scrubbing process for the removal of CO_2 from said gaseous mixture.

5. A process according to claim 1, wherein said scrubbing operation operates at a substantially higher pressure than the pressure of the expanded loaded scrubbing medium.

6. A process according to claim 1, wherein said loaded scrubbing medium is expanded in a preliminary expansion stage upstream of said first stage to form a liberated gaseous phase, said liberated gaseous phase being returned to said scrubbing operation, and scrubbing medium discharged from said preliminary stage is introduced into said liquid-jet compressor of said first stage for further expansion.

7. A process according to claim 6, wherein said preliminary expansion stage comprises a gas turbine.

8. A process according to claim 6, wherein said resultant expanded scrubbing medium and first stage expansion gases are discharged from said first stage and introduced into a separator, and resultant expanded scrubbing medium is removed from said separator and delivered to said at least one downstream expansion stage.

9. A process according to claim 8, wherein said regenerated scrubbing medium is returned to said pressure scrubbing operation.

10. A process according to claim 1, wherein, after said loaded scrubbing medium is expanded in said first stage, resultant expanded scrubbing medium is delivered to a separator wherein a gaseous phase and a partially expanded scrubbing medium phase are separated, said partially expanded scrubbing medium phase is further expanded in a subsequent expansion stage and delivered to a second separator wherein expansion gases are separated from regenerated scrubbing medium, said

expansion gases being delivered to said first stage and pressurized by said liquid-jet compressor.

11. A process according to claim 10, wherein said regenerated scrubbing medium is returned to said scrubbing operation.

12. A process according to claim 1, wherein said scrubbing operation employs a physical scrubbing medium.

13. A process according to claim 12, wherein said gaseous mixture contains CO₂ and H₂S and said scrubbing operation comprises removing CO₂ and H₂S from said gaseous mixture by scrubbing with said physical scrubbing medium.

14. A process according to claim 13, wherein said physical scrubbing medium comprises methanol.

15. A process according to claim 13, wherein, after expansion of the loaded physical scrubbing medium stream within said liquid-jet compressor of said first stage, resultant expanded physical medium stream and liberated gases are delivered to an enrichment column having a top portion, a middle portion and a bottom portion, said resultant expanded physical medium stream and liberated gases being introduced into said middle portion, a CO₂ product stream being discharged

from said top portion, and a physical medium stream enriched in H₂S and COS being discharged from said bottom portion.

16. A process according to claim 15, wherein said liberated gases rise upwardly within said enrichment column in countercurrent contact with a second scrubbing stream delivered to the top portion of said enrichment column.

17. A process according to claim 15, wherein the enriched physical medium stream is removed from said bottom portion of said enrichment column and delivered to a stripping column for further regeneration.

18. A process according to claim 15, wherein said resultant expanded physical medium stream is removed from said column and further expanded in said downstream expansion stage to form said downstream expansion gases and further expanded physical medium stream, and said downstream expansion gases are delivered to said liquid-jet compressor of said first stage and pressurized.

19. A process according to claim 18 wherein said further expanded physical medium stream is delivered to said bottom portion of said column.

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[54] ACID GAS REMOVAL SYSTEM
EMPLOYING REGENERATOR WITH
INTERNAL FLASH SECTION

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[21] Appl. No.: 167,237

[22] Filed: Dec. 15, 1993

[51] Int. Cl.⁴ B01D 50/00; F28D 21/00;
C01B 17/16

[52] U.S. Cl. 422/171; 422/178;
422/205; 422/228; 422/236; 55/228; 423/223

[58] Field of Search 422/171, 173, 175, 178,
422/205, 206, 228, 236; 55/228, 229; 95/187,
191, 192, 203, 264; 96/194, 197, 198; 423/223,
228, 229, 232, 242, 243

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Primary Examiner—Robert J. Warden

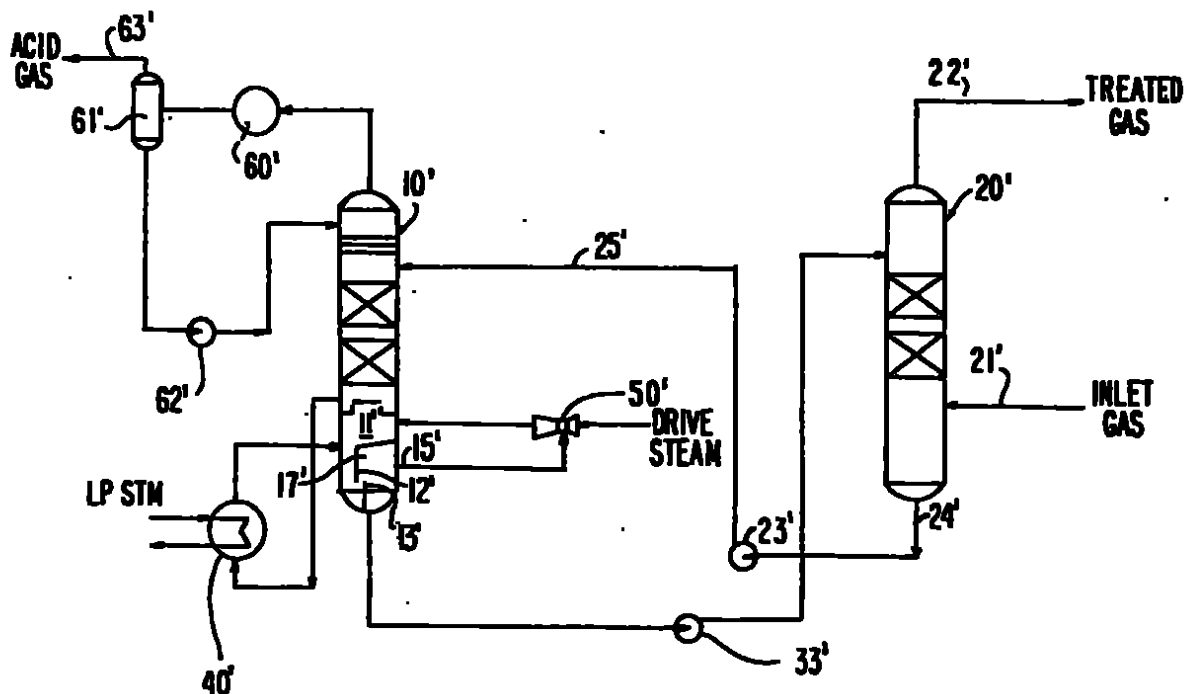
Assistant Examiner—Christopher Y. Kim

Attorney, Agent, or Firm—Bryan Cave

[57] ABSTRACT

An acid gas removal process employing a regenerator with an internal flash section is described. The regenerator uses gravitational flow to feed the flash section, eliminating the need for an external flash tank, and charge pump even without vacuum. The internal flash section has a first pressure side, a second pressure side having a lower pressure than the first pressure side during gas flow and a pressure reduction zone disposed therebetween. The pressure drop in the pressure reduction zone is due to liquid flow pressure losses occurring as liquid flows from said first pressure side to said second pressure side. A liquid seal between the first pressure and second pressure sides eliminates the need for a pressure let down valve. Vacuum control is provided by the first pressure side liquid level.

20 Claims, 4 Drawing Sheets



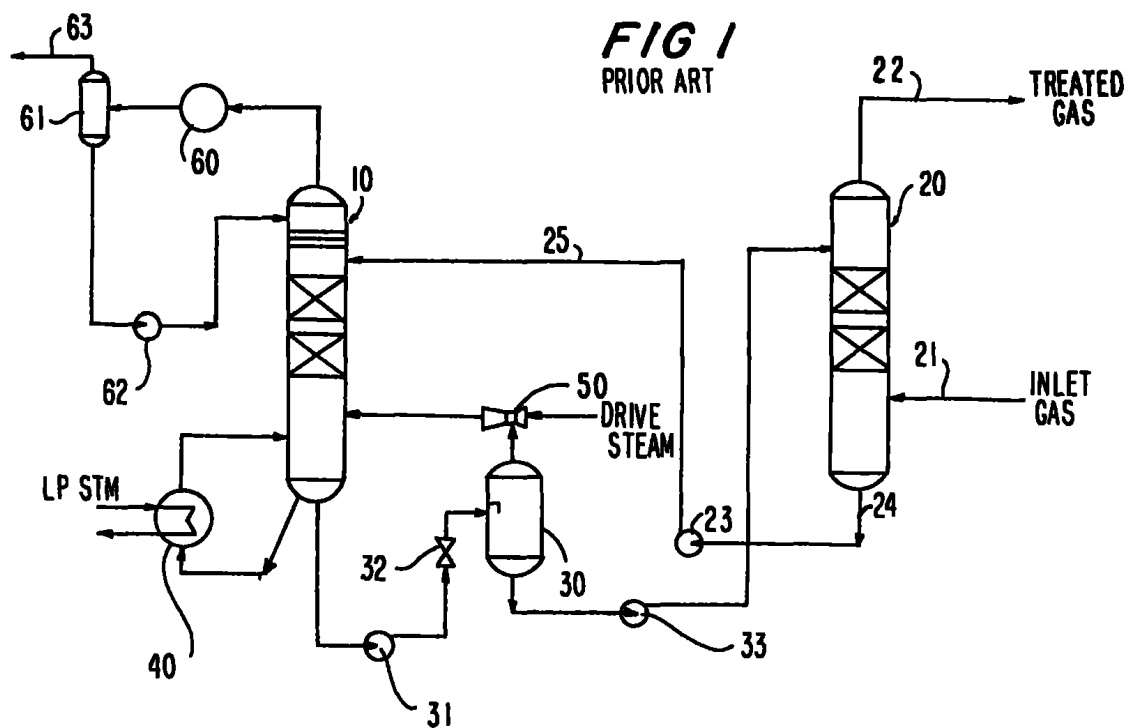
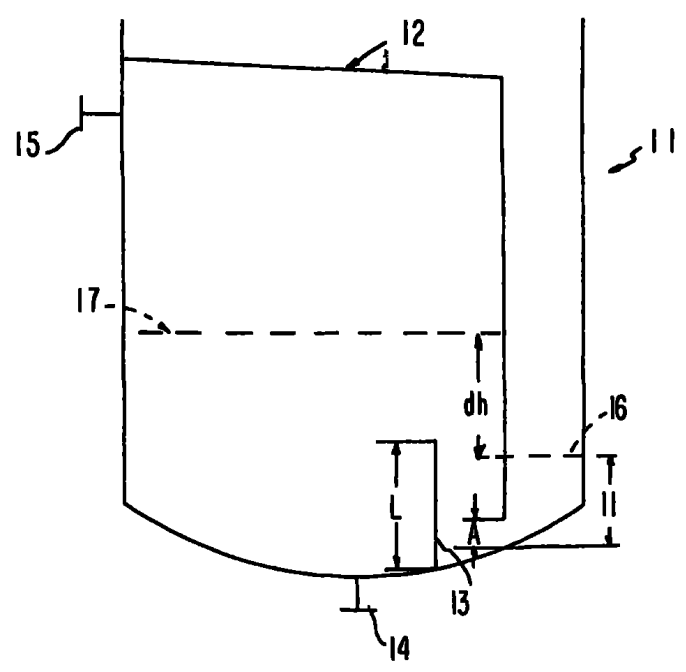


FIG 2



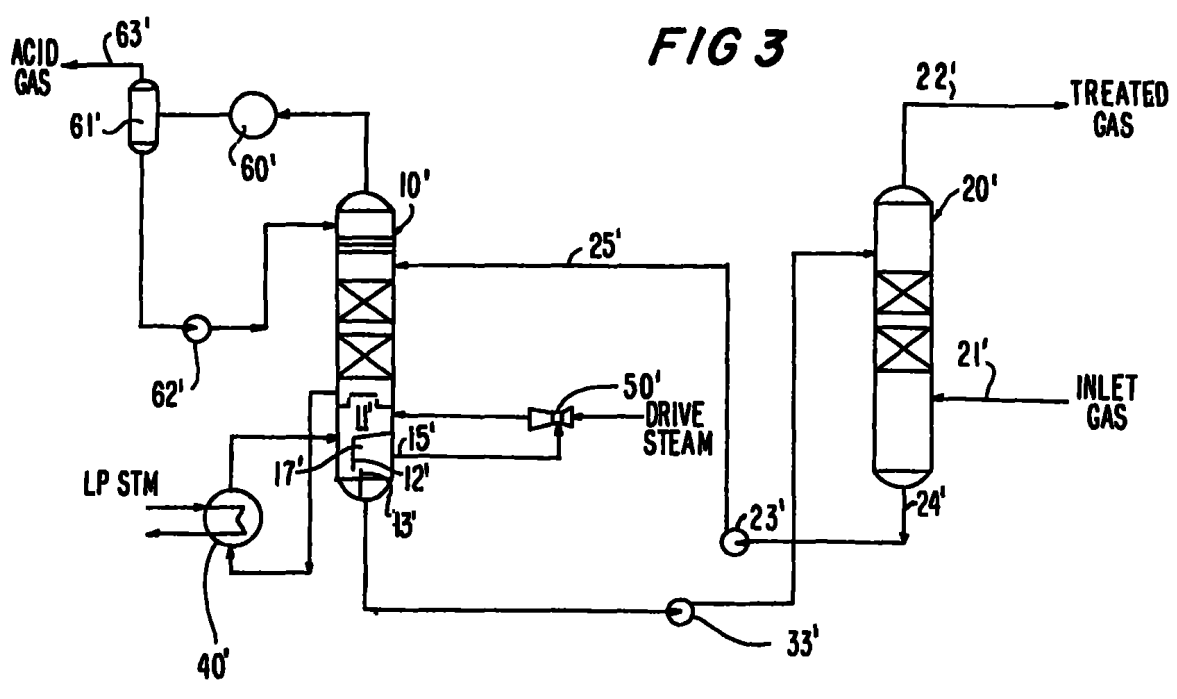
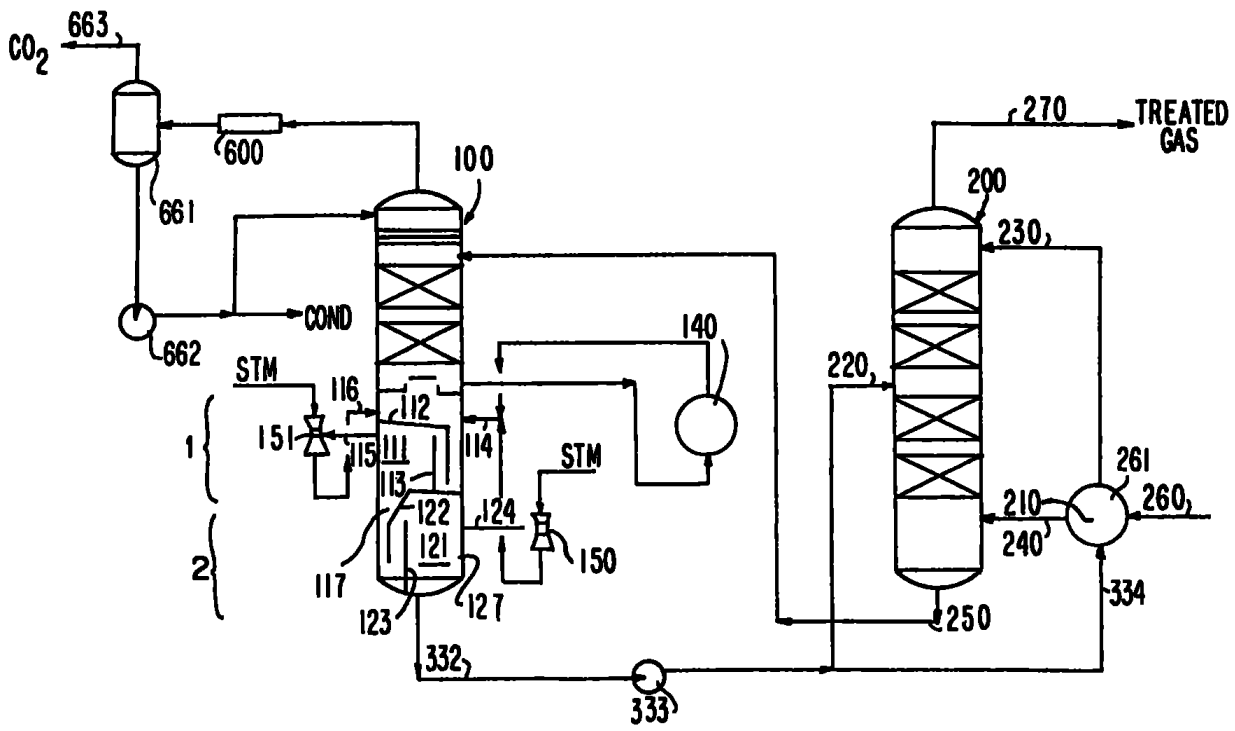


FIG 4



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July 25, 1995

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ACID GAS REMOVAL SYSTEM EMPLOYING REGENERATOR WITH INTERNAL FLASH SECTION

FIELD OF THE INVENTION

The invention relates to an improved process and apparatus for the removal of acid gases from gas mixtures.

BACKGROUND

Processes employing a modified potassium salt solution containing various types of catalysts and corrosion inhibitors are known. See U.S. Pat. Nos. 5,145,658, 4,430,312, 4,271,132, 3,932,582, 3,896,212 and 3,851,041. The catalysts greatly increase the rate of absorption and desorption of acid gases, thereby reducing equipment sizes, capital costs, utility requirements and residual CO₂ and H₂S contents.

The most common use is in ammonia, hydrogen, LNG and ethylene oxide plants. Users can purify sour gas streams by removing CO₂, H₂S or COS. Purified gas under adequate pressure can contain less than 4 ppm of H₂S and between 0.005% and 2% of CO₂. In the absence of sulfur compounds, the recovered high purity CO₂ is suitable for making urea, liquid CO₂ or dry ice for beverage use and enhanced oil recovery.

DESCRIPTION OF THE DRAWINGS

FIG. 1 is a process flow diagram of a prior art acid gas removal process using a conventional flash tank configuration for heat recovery from the hot lean solution;

FIG. 2 is a cross-sectional view of a single-stage internal flash section of a regenerator in accordance with the invention;

FIG. 3 is a process flow diagram of an acid gas removal process employing a regenerator with a single-stage internal flash section in accordance with the invention; and

FIG. 4 is a process flow diagram of an acid gas removal process employing a regenerator with a multiple-stage internal flash section in accordance with the invention.

DETAILED DESCRIPTION OF THE PREFERRED EMBODIMENTS

A typical single stage acid gas removal installation is shown in FIG. 1. As shown in FIG. 1 lean solution from the regenerator 10 is fed via a charge pump 31 to a flash tank 30. Vapors leaving the flash tank are recompressed in a steam jet 50 and returned to the regenerator to reduce the heat duty of the reboiler 40. Lean solution from the flash tank is fed via a lean pump 33 to the absorber 20. The treated gas 22 is returned from the absorber 20.

In single-stage designs, the inlet gas 21 is fed into the absorber 20 and is scrubbed countercurrently by the lean solution in the absorber 20. The rich solution from the absorber 20 is fed by line 24 to the rich pump 23 and then fed by line 25 to the regenerator 10, where the rich solution is regenerated countercurrent to steam from the reboiler 40 in the regenerator 10. Overhead vapors are fed from the regenerator 10 to the OVHD (overhead) Cooler 60 which compresses the vapors to form a solution and gas which is then fed to a storage tank, K.O. Drum 61. The acid gas is removed by a line 63 from the K.O. Drum 61 and the solution is fed by a line

from the K.O. Drum 61 to a Cond (condensate) pump 62, which in turn feeds the condensate back to the regenerator 10.

Both the absorber 20 and the regenerator 10 can be either packed towers or trayed towers. This design can be used when a very high purity is not required in the treated gas. (CO₂ partial pressure of about 2 PSI or more.)

The lean and/or semi-lean solution from the regenerator 10, which is at its boiling point, is flashed to a lower pressure in the flash tank 30. The flashed vapors are mostly steam with some acid gas. The vapors are compressed to the regenerator and used to reduce the reboiler heat duty. They can be compressed using either a steam jet 50 (thermo-compressor) or a mechanical compressor.

The use of such technology reduces the overall heat requirements and, therefore, the heat transfer surface in the reboiler and the overhead acid gas cooler. Additionally, it can lower the circulation rate and therefore the pump cost and the diameter of the absorber.

The additional equipment in FIG. 1, which is necessary to operate the prior art heat recovery system includes the (vertical or horizontal) flash tank 30, the charge pump 31, a pressure let down valve 32 in the line from the charge pump to the flash tank 30, an additional level control system (not shown) in the flash tank and the appurtenant external piping.

The disadvantages of this type of system are the higher investment in equipment and operating expense, additional mechanical maintenance, the need for a larger solution inventory, larger space requirements and limitations on turndown. In this latter regard, turndown is limited by the need to observe a minimum flow rate of the charge pump, the need to avoid control valve cavitation, vertical riser stability and over vacuum on the flash tank.

Accordingly, the present invention improves [a need exists for improving] upon existing flash tank configurations for heat recovery in the foregoing process of FIG. 1 as disclosed in more detail herein below. An improvement upon existing flash tank configurations is illustrated in FIG. 2 by an internal flash section according to the present invention, i.e. the bottom of a modified regenerator. The present invention as illustrated by FIGS. 3 and 4 improve on the foregoing process by the use of a modified regenerator incorporating an internal flash system.

As shown in FIG. 2, the regenerator in accordance with the invention includes an internal flash section 11 at the bottom of the regenerator. The flash section 11 includes side wall portions and a bottom partition 12 appended to a side wall portion, which bottom partition 12 is defined by a horizontal member and a vertical member depending from the horizontal member. FIG. 2 further illustrates an optional baffle 13. The internal flash section 11 also comprises a first pressure side, a second pressure side having a lower pressure than the first pressure side during gas flow, and a pressure reducing means for generating a pressure drop between a first pressure side and a second pressure side during gas flow, wherein the pressure drop is due to the second pressure side liquid head and liquid flow pressure losses occurring as liquid flows from the first pressure side to the second pressure side through a spacing between the vertical member of the bottom partition and the bottom of the internal flash section. Vapor is withdrawn via

vapor outlet 15 from the space defined by a side wall of the flash section 11 comprising vapor outlet 15, portions of the horizontal and vertical members of a bottom partition 12 and a second pressure side 17 liquid level. There is a differential liquid head dh between a first pressure side 16 and a second pressure side 17.

In FIG. 2, the spacing A (defined by a bottom portion of the vertical member of the bottom partition 12 and the bottom of the flash section 11) provides a pressure drop at design flow rates which contributes to the pressure differential dh . As the flow rate increases, the pressure head H on the first pressure side 16 will tend to increase. Conversely, as flow rate drops H will drop as well. The lower effective limit of the first pressure head H is the spacing A between the vertical member of bottom partition 12 and the bottom of the flash section 11. Thus, at extremely low flow rates as H drops to height A , the liquid seal between the first pressure side 16 and the second pressure side 17 will be broken thereby preventing overvacuum.

Spacing A can be designed to achieve any desired pressure differential dh within design flow rates. The pressure drop due to spacing A can be readily determined using known engineering calculations that will be appreciated by persons skilled in the art.

Alternative pressure drop generating structures such as orifices and the like can be substituted for, or augment, the spacing A . For example, the vertical member of bottom partition 12 can incorporate orifices, e.g. orifice plates, disposed therein (not shown). If desired, the orifices can be removably mounted in the vertical member of bottom partition 12 for substitution of varying sized orifices depending on the desired flow rates and pressure drops. An access panel can be provided in the regenerator 10 for changing orifice plates after installation. The vertical member of bottom partition 12 can also incorporate a slidable gate (not shown) which can be used to vary the spacing A .

The baffle 13 is not required in single stage flash sections as shown in FIG. 2. However, the baffle 13 is useful in multiple stage flash sections, described below. In this latter regard, the baffle 13 cooperates with the vertical member of bottom partition 12 to create a riser therebetween. In operation the riser will be filled with lean solution and its overflow at the top will proceed to accumulate and flow back into the space between the vertical member of bottom partition 12 and the side wall of flash section 11 of the first pressure side of the previous flash stage to form the first pressure side 16 liquid level, or during steady state operation will simply flow out to the liquid outlet 14 as shown in FIG. 2.

In multi-stage internal flash section designs, the length L of the baffle 13 in at least the upper flash stages should be chosen to provide a riser having a minimum length (measured from the bottom of the vertical member of bottom partition 12 to the top of the baffle 13) at least equal to, and preferably greater than, the [low] second pressure side liquid level 17 ($H + dh$) at the lowest expected flow rate.

A single stage system including the modified regenerator in accordance with the invention is shown in FIG. 3. Vapor is withdrawn from the second pressure side 17' of the flash section 11' via vapor outlet 15' (wherein structures 11', 15' and 17' of FIG. 3 correspond to FIG. 2 structures 11, 15 and 17, respectively; as described above for FIG. 2) by a steam jet 50'. This vapor is recompressed and returned to the regenerator 10' thereby reducing the heat duty on the reboiler 40'.

In FIG. 3 (as described above for FIG. 1), the inlet gas 21' is fed into the absorber 20' and is scrubbed countercurrently by the lean solution in the absorber 20'. The rich solution from the absorber 20' is fed by line 24' to the rich pump 23' and then fed by line 25' to the regenerator 10', where the rich solution is regenerated countercurrent to steam from the reboiler 40' in the regenerator 10'. Overhead vapors are fed from the regenerator 10' to the OVHD (overhead) Cooler 60' which compresses the vapors to form a solution and gas which is then fed to a storage tank, K.O. Drum 61'. The acid gas is removed by a line 63' from the K.O. Drum 61' and the solution is fed by a line from the K.O. Drum 61' to a Cond (condensate) pump 62', which in turn feeds the condensate back to the regenerator 10'.

The regenerator in accordance with the invention uses gravitational flow to feed the internal flash section at the bottom of the regenerator, eliminating the need for an external flash tank, and charge pump even without vacuum. The seal between the first pressure and second pressure sides, which is provided by the liquid head H , also eliminates the need for a pressure let down valve. Vacuum control is provided by the first pressure side liquid level. As described above, this also avoids overvacuum at reduced capacities.

Additionally, since the flash section at least in part occupies what would otherwise be regenerator solution surge space, this results in greater equipment economy. Elimination of the external flash system also reduces the total volume of regenerator solution required to operate the system.

Other process flow schemes can be used wherein the regenerator 10' is modified to include an internal flash section 11' to enhance its operation. Examples of such alternative process flow schemes include split-cooled lean designs and two stage designs.

In split-cooled lean designs, a small portion of the lean solution is cooled and sent to the top of the absorber permitting a greater purity of the treated gas. The bulk of the lean solution is sent hot to the middle of the absorber, thereby improving the heat efficiency of the system. CO_2 partial pressures of about 0.3 PSI are achievable in the treated gas.

Two-stage designs are generally the most heat efficient configurations when a high purity of treated gas is required, such as 4 PPM H_2S and/or 50-1500 PPM CO_2 . Two stages of absorption and regeneration may be used. The bulk of the solution is circulated as a semi-lean solution. A minor portion (about 10-30%) of the total solution is then thoroughly regenerated and cooled to be circulated as a lean solution.

A particularly preferred process flow scheme is shown in FIG. 4. The regenerator 100 includes two flash sections 111 and 121, each flash section comprising a first pressure side and a second pressure side having a lower pressure than the corresponding first pressure side during gas flow. The flash sections 111 and 121 comprise, respectively, bottom partitions 112, 122 (wherein each bottom partition is defined by a horizontal member and a vertical member depending from the horizontal member) and baffles 113, 123. Vapor from the respective second pressure sides 117, 127 is recompressed and returned to the regenerator 100. Vapor is withdrawn from second pressure side 117 of flash section 111 via vapor outlet 115, which is located on the second pressure side wall of flash section 111, by a steam jet 151 and returned to the regenerator 100 via regenerator inlet 116. Likewise, vapor is withdrawn

from second pressure side 127 of flash section 121 via vapor outlet 124, which is located on the second pressure side wall of flash section 121, by a steam jet 150 and returned to the regenerator 100 via regenerator inlet 114.

Lean solution is fed from the regenerator 100 via line 332 to lean pump 333 and is then fed via line 230 to an intermediate point in absorber 200. The lean solution is also fed from lean pump 333 via line 334 through a heat exchanger 261 and on through line 230 to the top of absorber 200 where the treated gas 270 then exits absorber 220. The lean solution which is fed by line 334 into the heat exchanger 261 is cooled in the lean cooler 210 of heat exchanger 261 by incoming gas from in line 260. The thus heated gas passes through line 240 into the bottom of absorber 200 and passes countercurrent to the lean solution fed via lines 230, 230. Rich solution exits the bottom of the absorber 200 and is fed via line 250 to the top of the regenerator 100.

In FIG. 4 overhead vapors from the regenerator 100 are treated in a manner analogous to that described in FIG. 1 for corresponding overhead vapors and structures. Overhead vapors are fed from the regenerator 100 to the OVHD (overhead) Cooler 600 which compresses the vapors to form a solution and gas which is then fed to a storage tank, K.O. Drum 661. The CO₂ gas is removed by a line 663 from the K.O. Drum 661 and the solution is fed from the K.O. Drum 661 to a Cond (condensate) pump 662, which in turn feeds the condensate back to the regenerator 100. Also, as shown in FIG. 4, steam is removed from the regenerator 100 and fed to the reboiler 140 in a manner analogous to that illustrated in FIG. 1 for the regenerator 10 and the reboiler 40. In turn, the reboiler 140 returns steam to the regenerator inlet 114 on the side wall of the regenerator 100.

Any number of multiple flash sections can be accommodated in the regenerator. In addition, if desired, one or more external flash systems can be incorporated as an adjunct to the internal flash section or sections in the regenerator in accordance with the invention. The external flash systems can be used as first, intermediate or final flash stages in systems wherein multiple flash stages are desired. For example, in an existing plant which is already equipped with one or more external flash systems, the regenerator can be modified by incorporating one or more flash sections in accordance with the invention. In the latter case, the external flash system could be bypassed completely when reduced capacity is warranted, e.g. during plant turndown when the entire regeneration system is operating at lower capacity.

As will be apparent to persons skilled in the art, various modifications to these specific embodiments of the invention can be envisioned which nevertheless employ the inventive concepts disclosed herein. Accordingly, the scope of the invention is not intended to be limited to these specific disclosed embodiments, but rather is intended to be defined by the claims set forth below.

What is claimed is:

1. An absorbent solution regenerator for removing acid gases from gas mixtures, said regenerator comprising at least one rich solution inlet, at least one overhead vapor outlet, at least one lean solution inlet, at least one steam inlet, at least one steam outlet, and at least one internal flash section at the bottom portion of the regenerator;

said at least one internal flash section comprising side wall portions, at least one vapor inlet, at least one

vapor outlet, at least one liquid outlet, at least one bottom partition appended to a side partition and said bottom partition is defined by a horizontal member and a vertical member depending from the horizontal member, and having a vapor space formed therein defined by a portion of at least one side wall of said side wall portions and portions of the horizontal and vertical members of said at least one bottom partition,

said internal flash section including a first pressure side, a second pressure side having a pressure lower than the first pressure side during gas flow and a pressure reducing means comprising a spacing between the vertical member of the bottom partition and bottom of said internal flash section for generating a pressure drop between the first pressure and second pressure side during gas flow; and said internal flash section including a means for preventing overvacuum at low flow rates of said gas mixtures.

2. The absorbent solution regenerator of claim 1, wherein the internal flash section includes a bottom partition defined by a horizontal member and a vertical member depending from the horizontal member separating the first pressure side from the second pressure side of the flash section, the vertical member being displaced from the bottom of the flash section by a spacing, said spacing and a vapor removal means connected to the vapor outlet of the flash section providing the pressure reducing means, which limits the pressure differential between the first pressure side and the second pressure side during gas flow.

3. The absorbent solution regenerator of claim 2, further including a baffle mounted on the bottom of the flash section and so spaced from the bottom partition as to define a riser therebetween.

4. The absorbent solution regenerator of claim 3, wherein the length of the baffle is sufficient to define a riser between the baffle and the bottom partition having a length measured from the bottom of the bottom partition to the top of the baffle at least equal to the low pressure side liquid level at the lowest expected flow rate.

5. The absorbent solution regenerator of claim 4, comprising a plurality of flash sections arranged in series, each flash section comprising a bottom partition dividing the high pressure side from the low pressure side of the flash section and a baffle mounted on the bottom of the flash section and spaced from the bottom partition, and the low pressure side of the first and any succeeding flash section defining the high pressure side of the second and any succeeding flash section.

6. An absorbent solution regenerator according to claim 3, wherein the baffle is removable.

7. An absorbent solution regenerator according to claim 3, wherein the baffle has a means for adjusting the length of the baffle.

8. The absorbent solution regenerator of claim 1, wherein the at least one bottom partition of the at least one internal flash section comprises a means for varying the spacing between the bottom of the internal flash section and the vertical member of said at least one bottom partition.

9. The absorbent solution regenerator of claim 8, wherein said means for varying the spacing between the internal flash section and the vertical member of said at least one bottom partition comprises a removable slidable gate.

10. The absorbent solution regenerator of claim 8, wherein said means for varying the spacing between the internal flash section and the vertical member of said at least one bottom partition comprises at least one orifice.

11. The absorbent solution regenerator of claim 8, wherein said means for varying the spacing between the internal flash section and the vertical member of said at least one bottom partition comprises at least one orifice plate having a means for varying the number or size of at least one orifice.

12. An absorbent solution regenerator for removing acid gases from gas mixtures, said regenerator comprising: at least one rich solution inlet, at least one overhead vapor outlet, at least one lean solution inlet, at least one steam inlet, at least one steam outlet, and a single stage internal flash section comprising:

- (1) a bottom partition defined by a horizontal member and a vertical member depending from the horizontal member, said bottom partition dividing the flash section into a first pressure side and a second pressure side with a pressure lower than the first pressure side during gas flow;
- (2) vapor outlet means in the low pressure side of the flash section disposed above the liquid level therein;
- (3) liquid outlet means in the low pressure side of the flash section disposed below the liquid level therein; and
- (4) pressure reducing means during gas flow defined by a spacing between the vertical member of the bottom partition and the bottom of the flash section, a vapor removal means connected to the vapor outlet means of the flash section, and a liquid removal means connected to the liquid outlet means of the flash section, said pressure reducing means comprising a means for generating a pressure drop between the first pressure side and second pressure side defined by a spacing between the vertical member of the bottom partition and the bottom of the flash section, said spacing being a means for limiting the pressure differential between the first pressure side and the second pressure side during gas flow.

13. The absorbent solution regenerator of claim 12, further including a baffle mounted on the bottom of the flash section and so spaced from the vertical member of the bottom partition as to define a riser therebetween.

14. An absorbent solution regenerator according to claim 13, wherein the baffle is removable.

15. An absorbent solution regenerator according to claim 13, wherein the baffle has a means for adjusting the length of the baffle.

16. An absorbent solution regenerator for removing acid gases from gas mixtures, said regenerator comprising: at least one rich solution inlet, at least one overhead vapor outlet, at least one lean solution inlet, at least one steam inlet, at least one steam outlet, and a two-stage internal flash section including:

- (a) a first internal flash section comprising:
 - (i) a bottom partition defined by a horizontal member and a vertical member depending from the horizontal member, said bottom partition dividing the flash section into a first pressure side and a second pressure side with a pressure lower than the first pressure side during gas flow;
 - (ii) vapor outlet means in the second pressure side of the flash section disposed above the liquid level therein;

(iii) liquid outlet means in the second pressure side of the flash section disposed below the liquid level therein and communicating with the first pressure side of the second stage internal flash section;

(iv) pressure reducing means defined by a spacing between the vertical member of the bottom partition and the bottom of the flash section, a vapor removal means connected to the vapor outlet means of the flash section, and the liquid outlet communicating with the second stage internal flash section, said pressure reducing means comprising a means for generating a pressure drop between the first pressure side and second pressure side defined by a spacing between the vertical member of the bottom partition and the bottom of the flash section, the spacing being a means for limiting the pressure differential between the high pressure side and the low pressure side during gas flow; and

(v) a baffle mounted on the bottom of the flash section and so spaced from the vertical member of the bottom partition as to define a riser therebetween, the length of the baffle being such that the length of the riser measured from the bottom of the bottom partition to the top of the baffle is at least equal to the second pressure side liquid level at the lowest expected flow rate;

(b) a second internal flash section comprising

(i) a second bottom partition defined by a horizontal member and a vertical member depending from the horizontal member, said second bottom partition dividing the flash section into a second first pressure side communicating with the second pressure side of the first internal flash section and into a second second pressure side with a pressure lower than the second first pressure side during gas flow;

(ii) second vapor outlet means in the second pressure side of the second flash section disposed above the liquid level therein;

(iii) second liquid outlet means in the second pressure side of second flash section disposed below the liquid level therein;

(iv) a second pressure reducing means comprising a spacing between the vertical member of the second bottom partition and the bottom of the second flash section, a second vapor removal means connected to the second vapor outlet means of the second flash section, and a second liquid removal means connected to the second liquid outlet means of the second flash section, said pressure reducing means comprising a means for generating a pressure drop between the first pressure side and second pressure side of the second flash section defined by a spacing between the vertical member of the second bottom partition and the bottom of the second flash section, the spacing being a means for limiting the pressure differential between the first pressure side and the second pressure side of the second flash section during gas flow; and

(v) a second baffle mounted on the bottom of the second flash section and so spaced from the vertical member of the second bottom partition as to define a second riser therebetween, the length of the second baffle being such that the length of the second riser measured from the bottom of the

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- vertical member of the second bottom partition to the top of the second baffle is at least equal to the second second pressure side liquid level at the lowest expected flow rate.

17. The absorbent solution regenerator of claim 16, wherein the baffle mounted on the bottom of the first flash section is removable.
18. The absorbent solution regenerator of claim 16, wherein the second baffle mounted on the bottom of the second flash section is removable.

19. The absorbent solution regenerator of claim 16, wherein the baffle mounted on the bottom of the first flash section has a means for adjusting the length of the baffle.

20. The absorbent solution regenerator of claim 16, wherein the second baffle mounted on the bottom of the second flash section has a means for adjusting the length of the baffle.

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[54] REMOVAL OF ACID GASES FROM HOT GAS MIXTURES

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[73] Assignee: Bestfield Corporation, Pittsburgh, Pa.

[21]. Appl. No.: 094,200

[22] Filed: Mar. 7, 1978

[51] Int. Cl.² B01D 53/34

[52] U.S. CL 423/220; 423/223;
423/232; 423/228

[58] Field of Search 423/222, 223, 224, 226-229,
423/232, 233, 230

[56] References Cited

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3,823,222	7/1974	Benson	423/223
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[57] **ABSTRACT**

A process for removing acid gases such as CO_2 and H_2S from gas streams using an aqueous alkaline scrubbing solution which is circulated between an absorption and regeneration stage wherein the stripping steam for regeneration is derived at least partly from the hot feed gas to be purified. The thermal efficiency of the process is substantially improved by first heat-exchanging the hot feed gas with the scrubbing solution to raise stripping steam and then heat-exchanging the hot feed gas with water (which is preferably condensate water produced in the scrubbing process) in a second heat exchanger operating under reduced pressure, thus recovering additional low-level heat from the feed gas and producing low-pressure steam which is compressed and then utilized as additional stripping steam. In a preferred embodiment, additional low-pressure steam is generated by also flashing off low-pressure steam from the regenerated scrubbing solution and compressing this steam together with the steam produced in the second heat exchanger for use as stripping steam.

12 Claims, 2 Drawing Figures

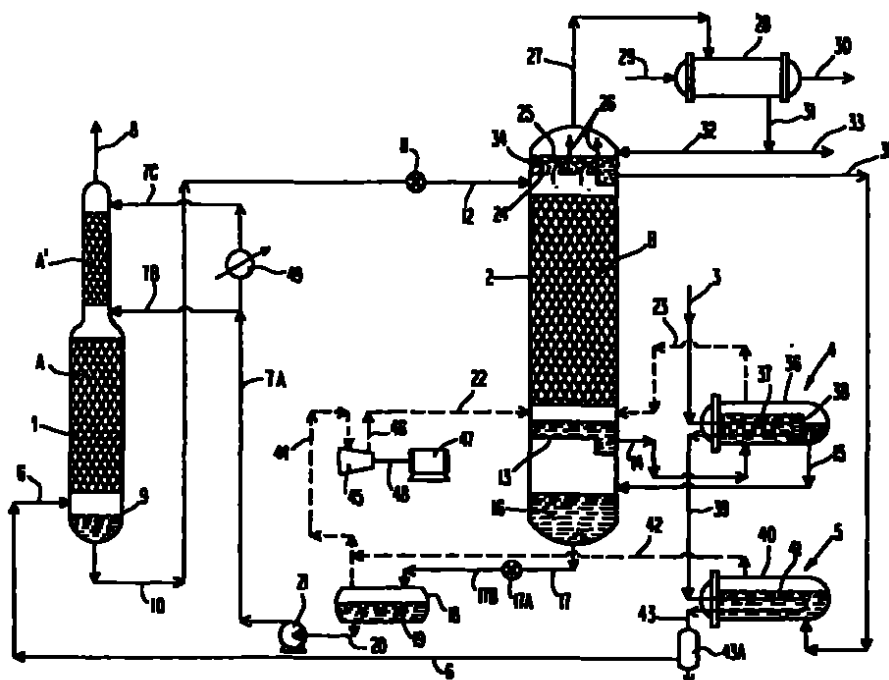


Fig 1

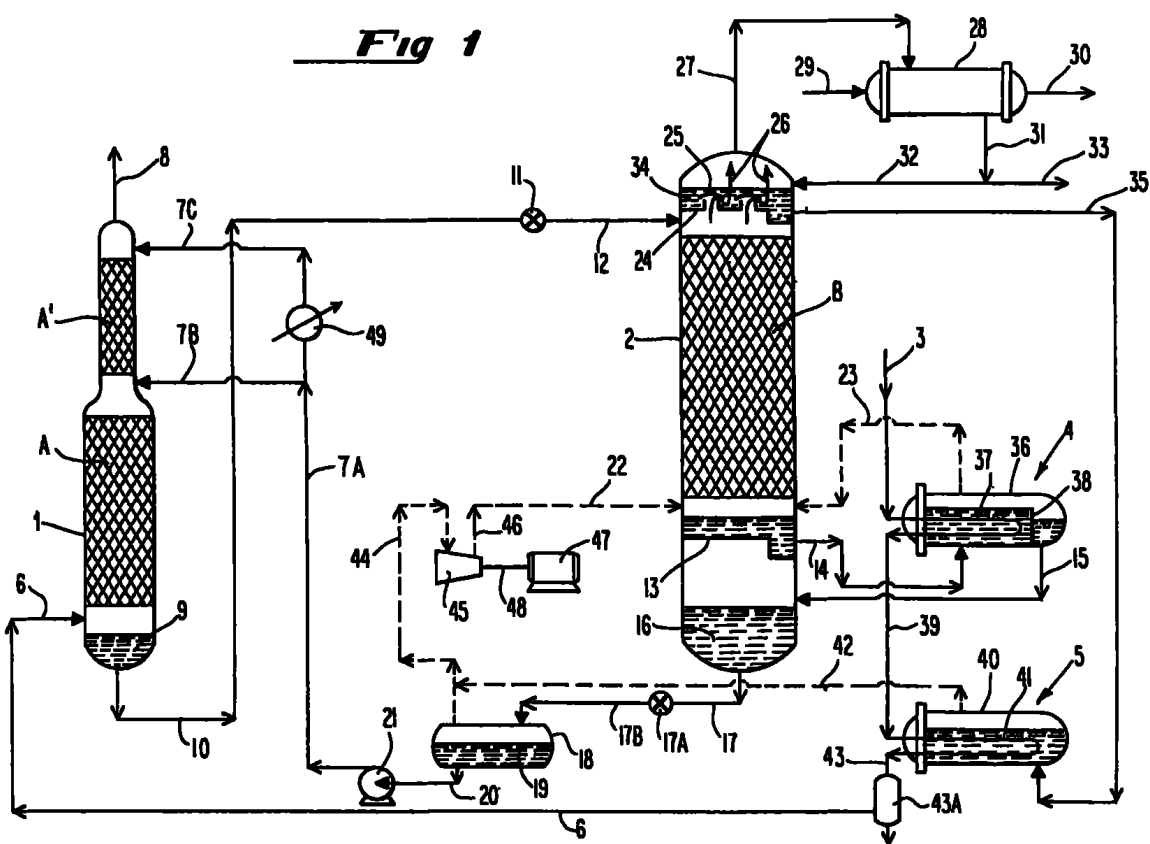
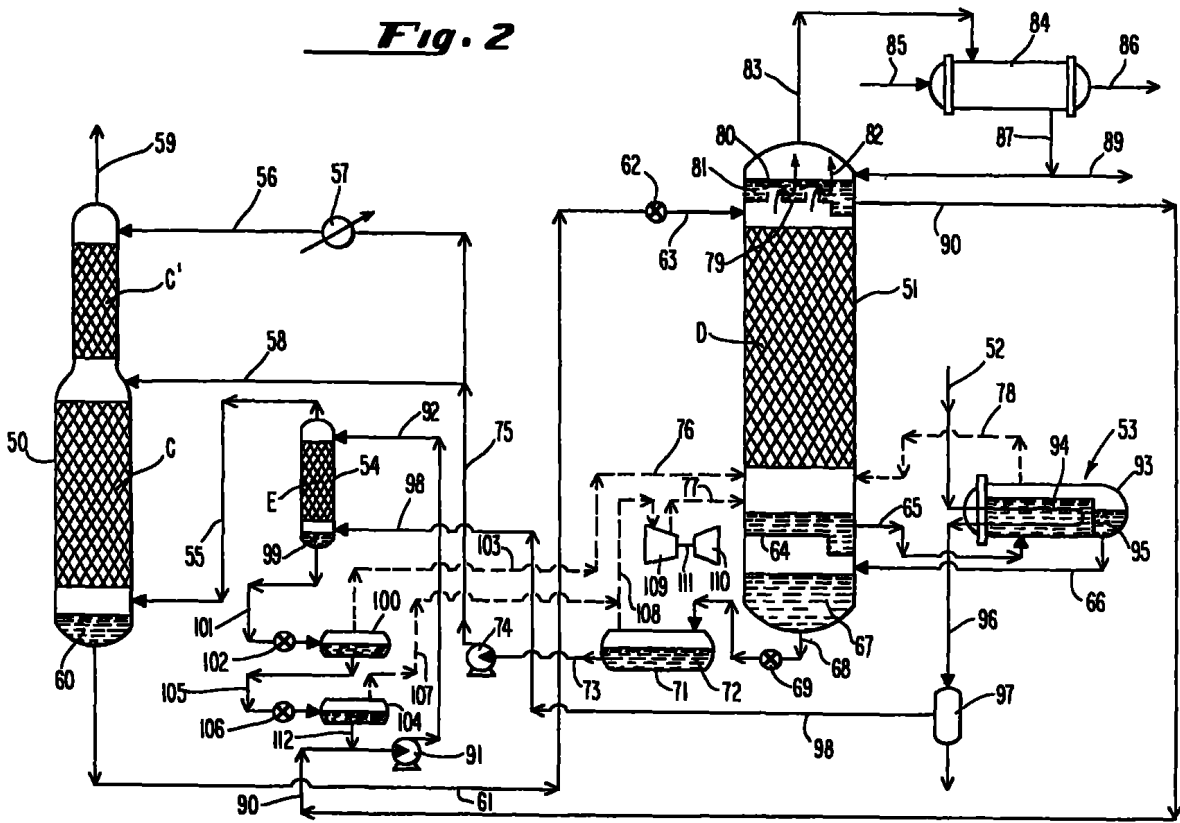


Fig. 2



REMOVAL OF ACID GASES FROM HOT GAS MIXTURES

BACKGROUND OF THE INVENTION AND PRIOR ART

This invention relates to an improved process for the removal of acid gases, such as CO_2 and H_2S , from hot, steam-containing gas mixtures.

The industrial importance of gas scrubbing processes involving the bulk removal of acid gases, particularly CO_2 and H_2S , from various raw gas mixtures is steadily increasing. As the demand for synthetic fuels and synthesis gases derived from fuel sources such as natural gas, oil and coal increases, there is an ever-increasing need for efficient processes for the removal of CO_2 and H_2S from the raw gas mixtures that are generated. For example, in the production of synthetic methane from naphtha, fuel oil or coal, the starting material is subjected to reforming or partial oxidation, producing a raw gas containing e.g. from 20% to as much as 50% of CO_2 together with smaller amounts of H_2S where a sulfur-containing starting material is employed. Likewise, in the reforming of natural gas to produce hydrogen for ammonia synthesis or hydrogenation reactions, a raw gas is produced containing usually from 16% to 20% CO_2 , all of which must be removed prior to the ammonia synthesis step.

The acid gas-containing raw gas mixtures produced by such processes as steam-reforming and partial oxidation are at elevated temperatures (and usually at elevated pressures) and contain large amounts of steam. Good thermal efficiency demands the efficient recovery and utilization of the heat content of such raw gases. In this connection, the extent to which such heat content can be efficiently utilized to provide the energy required to remove the large quantities of acid gases they contain is a highly significant factor in determining overall energy efficiency.

In modern practice, the most widely used process for the bulk removal of CO_2 and H_2S from such gas mixtures involves scrubbing of the gas with aqueous alkaline scrubbing solution which is continuously recirculated between an absorption stage where the acid gases are absorbed and a regeneration stage in which the acid gases are desorbed from the solution by means of steam-stripping. For most applications, the most efficient type of such cyclic process utilizes a substantially isothermal absorption and regeneration cycle, i.e. the absorption and regeneration stages are operated at or close to the same temperature, viz. a temperature in the vicinity of the atmospheric boiling temperature of the scrubbing solution. By eliminating the heating and cooling that is required by non-isothermal processes, heat losses are greatly reduced.

In any such process, whether isothermal or non-isothermal, the major energy requirement in the process is the stripping steam for regenerating the solution, and it is accordingly highly desirable to reduce the regeneration heat requirements and/or derive such regeneration heat from heat sources that may have little or no utility for other purposes.

It is particularly desirable that maximum utilization be made of the heat content of the raw feed gas to provide the heat energy required to produce the necessary stripping steam and that such heat extracted from the raw process gas be at the lowest possible energy level so that the higher energy-level heat in the process gas can

be utilized for other purposes. An efficient recovery of the low-level heat contained in the raw process gas will result in a lower-temperature feed gas to the absorption stage which, in turn, will improve the efficiency of the absorption stage.

There have been prior proposals for increasing the efficiency of heat recovery from hot, steam-containing gas mixtures which are to be treated for the removal of acid gases. See, for example, U.S. Pat. No. 3,823,222 to Homer E. Benson, issued July 9, 1974. In the process shown in that patent, the hot, steam-containing feed gas is passed in series through two heat exchangers, the first of which boils water to produce medium-pressure steam to operate a steam ejector, and the second of which raises relatively low-pressure steam in a heat exchanger where scrubbing solution is heated. The steam ejector is employed to raise additional low-pressure steam by subjecting regenerated scrubbing solution to a reduced pressure, and then to compress such low-pressure steam and inject it into the regenerator as additional stripping steam. The overall result is an enhancement in the net thermal efficiency of the process and a reduction in the amount of heat abstracted from the hot, steam-containing feed gas to produce the necessary stripping steam.

GENERAL DESCRIPTION OF INVENTION AND PREFERRED EMBODIMENTS

In accordance with this invention, a method has been discovered which is capable of still further enhancing the amount of low-level heat than can be abstracted from the hot, steam-containing feed gas and usefully converting such heat into regeneration stripping steam.

Described in its broader aspects, the invention employs a cyclic scrubbing process utilizing an aqueous alkaline scrubbing solution which is continuously recycled between an absorption stage, where the acid gases contained in the hot, steam-containing feed gas are absorbed, and a regeneration stage in which the absorbed acid gases are desorbed by steam-stripping. The stripping steam required is at least partially derived from the feed gas through a series of heat recovery steps which maximize the amount of low-level heat in the feed gas (i.e. heat available at a relatively low temperature) that can be recovered and converted to useful stripping steam. The first heat recovery step involves an indirect heat exchange between the hot feed gas and the scrubbing solution, thereby heating the scrubbing solution to its boiling point and producing steam that is utilized in the regeneration stage as stripping steam. The partially cooled hot gas is then conducted to a second heat exchanger where the gas is passed in heat-exchange relationship with water (which is preferably water condensate produced in the scrubbing process). This second heat-exchange step may be carried out by direct or indirect contact of the gas with the water. In the course of this second heat-exchange step, the water is subjected to a reduced pressure, thus lowering its boiling temperature; and the water of reduced boiling temperature is brought into heat-exchange relationship with the hot feed gas. The increased temperature differential between the water and the gas resulting from this pressure reduction and the consequent lowering of the boiling point causes further cooling of the feed gas and the transfer of greater amounts of heat from the feed gas to the water. The steam produced by heating of the water under reduced pressure is generated at a lower pressure than the pressure in the regeneration stage, and this low-pressure steam is compressed to a level at least

equal to that in the regeneration stage and injected into the regenerator as stripping steam. Suitable means for producing the reduced pressure over the water in the second heat-exchange stage and for compressing the low-pressure steam produced include mechanical compressors and thermal compressors, such as steam ejectors. The feed gas leaving the second heat exchanger after heat exchange with the water of reduced boiling temperature departs at a substantially lower temperature than in prior practice, and is passed to the absorption stage for removal of its acid gas content. The net effect of this heat-exchange sequence is the recovery of substantial additional quantities of low-level heat in the feed gas and effective conversion of this low-level heat into useful stripping steam. In typical commercial applications, savings in higher-level heat from the hot gas otherwise required to produce the necessary stripping steam may range from 15% to 20%. The effective utilization of the low-level heat in the feed gas for the production of stripping steam improves the overall efficiency of heat recovery from the feed gas and preserves higher-level heat (i.e. heat available at higher temperatures) for other uses such as the raising of medium-pressure steam.

The invention has particularly valuable application to the type of scrubbing process in which the absorption and regeneration stages are carried out at or close to the same temperatures and in which the absorption occurs under substantial superatmospheric pressures of at least 100 pounds per square inch gage (psig) and typically 200 to 1000 psig, and in which the regeneration occurs at close to atmospheric pressure. Preferred scrubbing solutions are aqueous solutions of alkali metal carbonates, particularly relatively concentrated solutions of potassium carbonate. Scrubbing processes of these preferred types are described, for example, in U.S. Pat. No. 2,886,405 of H. E. Benson et al. issued May 12, 1959.

In a particularly preferred embodiment, the system of the invention is used in combination with means for abstracting additional low-pressure steam from hot regenerated solution by reducing the pressure on the hot solution causing the production of further quantities of low-pressure steam which is compressed and injected into the regenerator as stripping steam. In such preferred embodiment, the same equipment employed for reducing the pressure on the water employed in the second heat-exchange step and for compressing the resulting low-pressure steam can also be used in producing the reduced pressure over the hot regenerated solution and compressing the low-pressure steam thus evolved.

Reference is now made to the accompanying drawings wherein FIG. 1 is a diagrammatic flowsheet which illustrates a particularly preferred embodiment of the invention wherein low-pressure steam is generated both from the water which is heat-exchanged with the hot feed gas and from the hot regenerated solution, both sources of steam being compressed and fed to the regenerator as stripping steam.

FIG. 2 is a diagrammatic flowsheet illustrating an embodiment of the invention in which the heat exchange between the hot, steam-containing feed gas and the water takes place by direct contact between the two.

Referring now to FIG. 1, the scrubbing system illustrated comprises an absorption column 1 and a regeneration column 2 between which the scrubbing solution is continuously circulated. Hot, steam-containing feed gas

enters the system by line 3, passes through a first heat-exchange stage generally designated by the reference numeral 4; then through a second heat-exchange stage designated generally by the reference numeral 5 and then is conducted to the bottom of the absorption column 1 by line 6.

The absorption column 1 is suitably equipped in the cross-hatched portions designated by the letters A and A' with means for producing intimate gas-liquid contact. For such purposes, packing materials are used such as Raschig rings, Berl saddles, Intalox saddles, Pall rings or other types of packing bodies exposing a large surface of liquid to the gas flowing through the packing. In place of packing, other means such as plates e.g. sieve trays, may be employed for insuring intimate gas-liquid contact. In the regenerator 2, the cross-hatched section labelled B similarly contains packing materials or plates to insure intimate contact between the scrubbing solution and stripping steam passing through this column.

In the preferred embodiment of the invention, the absorber column is maintained at substantial superatmospheric pressures of at least 100 pounds per square inch gage (psig) and preferably at least 200 psig. Absorption pressures in typical applications of the invention will generally range from 250 to 1500 pounds per square inch.

In the absorption column, a portion of the regenerated solution, after passing through line 7A and cooler 49, is supplied to the top of the column by line 7C and flows downwardly through the gas-liquid contact section A' countercurrently to the feed gas entering the bottom of the column by line 6. Another portion of the regenerated scrubbing solution flowing in line 7A is supplied by line 7B to an intermediate level of the absorption column and mixes with solution entering the top of the column and then flows downwardly through gas-liquid contact section A countercurrently to the feed gas. Acid gases such as CO₂ and/or H₂S contained in the feed gas are absorbed by the solution and the purified feed gas leaves the top of the tower by line 8.

The scrubbing solution containing absorbed acid gas collects at the bottom of the absorber in sump 9 and is conducted by line 10 to a pressure letdown valve 11 where the pressure in the solution is reduced to that prevailing at the top of the regenerator tower 2, and the let-down solution then flows into the top of regenerator tower 2 by line 12.

As indicated previously, the invention is applicable generally to systems using any regenerable aqueous alkaline scrubbing solution including e.g. aqueous solutions of alkali metal carbonates, particularly potassium carbonate, aqueous solutions of ethanolamines, or alkali metal phosphates. Particularly preferred are relatively concentrated potassium carbonate solutions having potassium carbonate concentrations of 15 to 45% and preferably from 20 to 35% by weight (these concentrations being calculated on the assumption that all the potassium present is present as potassium carbonate). Potassium carbonate solutions are preferably activated by the addition of additives such as amines, particularly ethanolamines, alkali metal borates such as potassium borate or sodium borate, As₂O₃, amino acids such as glycine or other additives which tend to increase the rates of absorption and desorption of acid gas in the potassium carbonate solution.

Particularly preferred additives for potassium carbonate solutions are the alkanolamines added in amounts ranging from 1 to 10% and particularly from 1

to 6% by weight. Diethanolamine is particularly preferred from the standpoint of cost, relatively low volatility and effectiveness.

In the regeneration column 2, the pressurized solution containing absorbed acid gases entering by line 12 is immediately freed of a portion of its content of acid gases through the reduction in pressure occurring at the top of the regeneration column. Further desorption of the acid gases is carried out in section B of the regeneration column 2 by countercurrent contact of the scrubbing solution with stripping steam introduced into the bottom of the column and rising up through the gas-liquid contact section B countercurrent to the descending solution.

The absorption and desorption reactions that occur in columns 1 and 2, respectively, are well known in the art, being described, for example, in U.S. Pat. No. 2,886,403.

The regenerated solution, stripped of most of the absorbed acid gases, is collected at the bottom of column 2 on a tray 13 and is fed to the heat exchanger system designated generally by reference numeral 4 by line 14. Scrubbing solution leaves heat exchanger 4 by line 15 and is fed to the bottom of the regenerator where it collects in sump 16. The regenerated solution leaves the bottom of the regenerator column 2 by line 17 and, after passing through letdown valve 17A, is conducted by line 17B to a flashing tank 18 operating under a reduced pressure as will be described in detail below. The scrubbing solution collects at the bottom of tank 18 in sump 19, and is conducted by line 20, recycle pump 21, and line 7A to the absorber column 1. The major portion of the solution is introduced without cooling by line 7B into an intermediate level in the absorber, while a minor portion is first cooled in cooler 49 and then introduced by line 7C into the top of the absorber.

Stripping steam for regeneration of the solution is introduced into the regeneration column through lines 22 and 23, such steam being generated in a manner to be described in detail below. The stripping steam from these sources rises in column 2 countercurrently to descending scrubbing solution and the mixture of steam and desorbed gases at the top of column 2 passes through a contact tray 24 equipped with bubble caps 25 as indicated by the arrows 26. The mixture of steam and desorbed gas leaves the top of column 2 by line 27 and is conducted to a condenser 28, cooled by a cooling medium supplied by line 29 and leaving by line 30. In the condenser 28, most of the steam content of the mixture is condensed and the water condensate is withdrawn from the condenser by line 31. A portion or all of the condensate is returned to the top of column 2 by line 32. Depending on the water balance in the system, a portion of the condensate may be removed by line 33. The condensate introduced into column 2 collects on contact tray 24 as shown by reference numeral 34. The mixture of steam and desorbed gas bubbling up through condensate 34 on tray 24 heats it to a temperature of e.g. 190° F. to 210° F. The preheated aqueous condensate is fed to heat-exchanger system 5 by line 35.

Referring specifically now to heat-exchanger system 4 which provides for indirect heat exchange between the hot feed gas stream and the scrubbing solution, this includes shell 36. Regenerated scrubbing solution collecting on plate 13 is withdrawn by line 14 and introduced into the bottom of shell 36. The scrubbing solution flows over a bundle of tubes, indicated by reference numeral 37, through which hot feed gas is introduced

by line 3. Heat from the feed gas is transferred through the tube bundle to the scrubbing solution, heating it to its boiling point. The steam generated passes through line 23 and is introduced into the bottom of the stripping column as stripping steam. The scrubbing solution overflows weir 38, then is withdrawn from shell 36 by line 15 and introduced into the bottom of the regenerator column 2 where it collects in sump 16. In the heat-exchanger system 4, the hot feed gas is partially cooled, yielding up a portion of its heat by indirect heat exchange with the scrubbing solution in heat exchanger 4, leaves heat exchanger 4 by line 39 and is introduced into heat-exchanger system 5. In the embodiment shown, this consists of a shell 40 supplied with aqueous condensate by line 35. The feed gas from line 39 passes through a tube bundle, shown diagrammatically and indicated by reference numeral 41, which is immersed in the aqueous condensate. In heat-exchanger system 5, the heat is transferred to the water in shell 40 through tube bundle 41 causing boiling to take place, and the steam thus generated leaves shell 40 by line 42. The hot feed gas, in passing through tube bundle 41, is further cooled and leaves heat-exchanger system 5 by line 43. After passing through a knockout pot 43A or similar device for removing condensed water, the cooled feed gas is introduced into the bottom of the absorber through line 6.

In order to enhance heat recovery from the hot, steam-containing feed gas and produce additional useful stripping steam, the boiling temperature of the water in shell 40 is reduced by connecting shell 40 through line 43 and line 44 to the suction side of a mechanical compressor 45 which reduces the pressure over the water in shell 40 to a pressure lower than that prevailing in the bottom portion of regeneration column 2, thus correspondingly lowering the boiling temperature of the water in shell 40. The effect of this pressure reduction and consequent lowering of the boiling point increases the temperature differential between the hot feed gas flowing in tube bundle 41 and the water, thereby making it possible to transfer more heat from the hot gas (which is generally at this point saturated with steam). The increased heat recovery from the hot feed gas generates additional amounts of low-pressure steam. This low-pressure steam is converted into useful stripping steam by compressing the steam in compressor 45 to a pressure at least equal to that prevailing in the bottom of the regenerator. The compressed steam leaves compressor 45 by line 46 and is introduced into the bottom of regenerator column 2 by line 22. Compressor 45 in the embodiment shown is driven by an electric motor 47 connected to the compressor by shaft 48. Alternate drives for the compressor 45 may, of course, be used, such as a steam turbine drive.

In the embodiment shown in FIG. 1, the compressor 45 is utilized not only to generate low-pressure steam by reduction of the pressure over the boiling water in heat-exchanger system 5, but is also employed to produce additional low-pressure steam by reducing the pressure over hot regenerated solution which is fed into flash tank 18 through line 17. The solution leaving the bottom of regenerator column 2 may typically be at a temperature of 248° F. or more, and upon reduction of the pressure over the solution, substantial quantities of low-pressure steam may be obtained. Flash tank 18 is connected to the suction of compressor 45 by line 44, and the mixture of flashed steam from heat exchanger 5 and the steam flashed from flash tank 18 flows by line 44 into the compressor 45 and, after compression, flows by

lines 46 and 22 into the bottom of the regenerator 2 to provide stripping steam. The flashing of steam in tank 18 results in cooling of the solution and the solution leaving tank 18 by line 20 for recycle to the absorber is typically at a temperature of 10° F. to 30° F. lower than the temperature of the solution leaving the bottom of the regeneration column.

The conservation of heat by operation in accordance with the invention is particularly advantageous in that not only is a substantial amount of heat recovered from the hot feed gas stream, but the heat recovered is at a low energy level (i.e. at a low temperature) which makes more higher energy-level heat in the gas stream available for other uses. In many cases, the low-level heat recovered from the feed gas stream as useful stripping steam in accordance with the invention is at too low an energy level to be recovered economically and is wasted. The conservation of the higher-level heat in the feed gas stream is reflected by the substantial reduction in the temperature of the process gas required for providing net regeneration heat requirements for the scrubbing process.

Other important advantages of the invention include a significant reduction in the size and thus the cost of certain equipment normally employed such as a reduction in the size of the overhead condenser 28 and the solution cooler 49 in the embodiment shown in FIG. 1. These advantages result from the additional cooling of the process gas stream by heat-exchanger system 5 and the cooling of the regenerated solution by flashing in flash tank 18. A further advantage of cooling the process gas stream and the regenerated solution is a somewhat reduced solution temperature at the bottom of the absorber which in turn allows the solution to absorb higher quantities of acid gases (CO₂ and/or H₂S) per unit volume of circulating solution. This in turn reduces solution circulation rates, and thus reduces pumping costs, and also increases overall thermal efficiency of the process.

The above advantages of the invention are illustrated in the following detailed examples. Example 1 illustrates the invention using the system shown in FIG. 1. Example 2 illustrates the invention using the system shown in FIG. 1 except that the regenerated solution leaving the bottom of the regenerator by line 17 is fed directly to recycle pump 21 rather than passing through flash tank 18. In Example 2, accordingly, only the flashed steam generated in heat-exchange system 5 is fed to compressor 45 for compression and introduction into regenerator 2 as stripping steam. In order to illustrate the advantages of the invention compared to conventional operation, Example A is included in which all regeneration steam is provided by heat-exchanging the hot feed gas against the scrubbing solution in the heat-exchange system 4 of FIG. 1 with no stripping steam being provided by heat-exchanger system 5 or flash tank 18.

So that Example A and Examples 1 and 2 will be directly comparable, the same inlet and outlet feed gas compositions were employed and the same solution composition used. The feed gas in all cases was an ammonia synthesis gas of the following composition (dry basis):

Component	Flow Rate in Pounds Mols Per Hour	Volume %
CO ₂	2,700	18.0
H ₂	9,000	60.0
N ₂	3,175.5	21.17
CO	52.5	0.35
CH ₄	37.5	0.25
Argon	34.5	0.23
TOTALS	15,000	100.00

In each case the feed gas was saturated with steam at the temperature entering heat-exchanger system 4 and enters the system at a pressure of 410 psia (pounds per square inch absolute).

In each case the CO₂ content of the feed gas was reduced to 0.1% by volume leaving the absorber. The solution composition in each case was an aqueous solution of potassium carbonate containing 30% by weight of K₂CO₃ (assuming all carbonate present is present as K₂CO₃) and 3% by weight of diethanolamine. The absorption and regeneration columns were of the same size and contained the same volume and type of tower packing. In each case the absorber pressure was 400 psia and the pressure at the bottom of the regenerator was 24.7 psia. The portion of the scrubbing solution entering the top of the absorber in each case was 25% of the total circulating solution and was cooled in each case to 158° F. in cooler 49 before introduction into the top of the absorber. The main portion of the circulating solution (75%) entered by line 7B at an intermediate level of the absorber. The solution temperature leaving the bottom of the regenerator was in each case 248° F.

In Example 1, employing both heat-exchanger system 5 and flash tank 18, the pressure above the boiling water in shell 40 and the pressure above the regenerated solution in flash tank 18 was reduced to 15 psia in contrast to the pressure of 24.7 psia prevailing at the bottom of the regenerator column 2. At the 15 psia pressure prevailing in shell 40, the boiling temperature of the water was reduced to 213° F., and the temperature of the process gas leaving heat-exchange system 5 by line 43 was 231° F. At the 15 psia pressure prevailing in flash tank 18, the temperature of the solution was reduced from 248° F. entering the tank to 220° F. leaving the tank due to the endothermic steam flashing. Under these conditions steam was evolved at the rate of 23,260 pounds per hour from heat-exchange system 5 while 54,105 pounds per hour of steam was evolved from the solution in flash tank 18. The combined low-pressure steam from these two sources entered the suction side of compressor 45 and, after compression in the pressure prevailing at the bottom of regenerator column 2, was fed into the column by line 22. In Example 1 an additional 68,315 pounds per hour of steam was generated in heat-exchanger system 4 and introduced into the bottom of the regenerator as stripping steam. In Example 2, using heat-exchanger systems 4 and 5 but not flash tank 18, the pressure in shell 40 was reduced to 15 psia and 23,260 pounds per hour of steam was evolved which was compressed in compressor 45 and fed into the bottom of the regenerator, while an additional 124,737 pounds per hour of stripping steam was required from heat-exchanger system 4. In Example A, where all stripping steam was generated in heat-exchange system 4, 151,050 pounds per hour of stripping steam was required from that source.

Table I below summarizes the results of Examples 1 and 2 and comparative Example A. In all cases the regeneration heat required is expressed in millions of BTU per hour (MM BTU/hr.) assuming a pound of steam is equivalent to 950 BTU.

TABLE I

	Example A	Example 1	Example 2
Regeneration heat generated by heat-exchange system 4, MM BTU/hr.	143.5	64.9	118.5
Regeneration heat generated by heat-exchange system 5, MM BTU/hr.	—	22.1	22.1
Regeneration heat generated in flash tank 18, MM BTU/hr.	—	51.4	—
Total regeneration heat required, MM BTU/hr.	143.5	138.4	140.6
Temperature of feed gas at inlet to heat-exchange system 4, °F.	362.1	325.7	352.5
Heat saved in hot feed gas for other uses, MM BTU/hr.	—	78.6	25.1
Heat saved as a percentage of total regeneration heat	—	54.8	17.5
Cooling duty of cooler 49, MM BTU/hr.	45.2	29.0	43.9
Cooling duty of overhead condenser 28, MM BTU/hr.	130.1	71.3	110.0
Solution circulation rate, gallons per minute	4310	3960	4185

The advantages of the invention are apparent from Table I. The principal advantage is the amount of heat saved in the hot feed gas and made available for other uses. In Example 1 where heat-exchange system 5 and flash tank 18 are used in combination, the heat saved in the hot feed gas compared to Example A is 78.6 MM BTU/hr., equivalent to 82,737 pounds per hour of steam. The heat saved in the hot feed gas is reflected by the lower temperature at which the hot feed gas enters the heat-exchange system to provide the necessary regeneration heat requirements. In Example A, using the conventional heat-exchange system 4, the temperature of the hot, steam-saturated feed gas entering the system is 362.1° F., while in Example 1 the inlet temperature of the steam-saturated feed gas to the heat-recovery system is reduced to 325.7° F. This means that, in Example 1, the sensible and latent heat content of the feed gas stream between the temperatures of 325.7° F. and 362.1° F. (equivalent to 78.6 MM BTU/hr.) is available for other uses. The 78.6 MM BTU/hr. of higher-level heat thus saved is equivalent to 54.8% of the total regeneration heat requirements of the conventional system of Example A. In Example 2, using heat-exchange systems 4 and 5 but without flashing of the regenerated solution, the amount of heat saved in the hot feed gas is smaller but still highly significant, amounting to 25.1 MM BTU/hr., a heat savings equivalent to 17.5% of the total regeneration heat requirements of the conventional system of Example A.

A further advantage of the invention is that the total regeneration heat required (and the corresponding amount of stripping steam necessary) is significantly reduced. While the total regeneration heat requirements in Example A is 143.5 million BTU per hour, in Examples 1 and 2 the total requirement is reduced to 138.4 and 140.6 million BTU/hour, respectively. Still further advantages are substantial reductions in the size of the

solution cooler (cooler 49 in FIG. 1) and in the size of the overhead condenser (condenser 28 in FIG. 1) as well as the solution circulation rate (thus reducing pump size and pumping energy), all as shown in Table I. Thus, not only does the invention substantially enhance the thermal efficiency of the process by increasing the recovery of heat from the hot feed gas stream; but, at the same time, it provides substantial savings in capital cost for the gas scrubbing plant as a whole.

Reference is now made to FIG. 2 which illustrates an embodiment of the invention similar to that shown in FIG. 1 except that the heat-exchange system 5 (where the heat from the process gas is transferred indirectly to the water) is replaced by a heat-exchange system in which the heat exchange between the hot steam-containing feed gas and the water takes place by direct contact between the two. In FIG. 2, the scrubbing solution is continuously circulated between an absorption column 50 and a regeneration column 51. Hot, steam-containing feed gas enters the scrubbing system by line 52 and passes first through a heat-exchange system generally designated by the numeral 53. It then passes through a direct-contact column 54 where it is brought into direct contact with water circulating through column 54 after which it is conducted to the bottom of the absorption column by line 55.

The absorption column 50 is suitably equipped in the cross-hatched portions designated by the letters C and C' with means for producing intimate gas-liquid contact as in column 1 described in FIG. 1. In the regenerator column 51, the cross-hatched section D similarly contains means for producing intimate contact between the scrubbing solution and the stripping steam passing through this column.

The regenerated solution from regenerator 51 is introduced into the absorber 50 in two streams. A minor portion (usually 15 to 35% of total circulating solution) is introduced into the top of the column through line 56 after passing through cooler 57. The major portion of the regenerated solution is introduced into the absorber at an intermediate level through line 58. The feed gas, containing the acid gases such as CO₂ and/or H₂S to be removed, is introduced into the bottom of the absorber column by line 59 and rises countercurrently to the descending scrubbing solution and the purified gas leaves the top of absorption column 50 by line 59.

The scrubbing solution, containing absorbed acid gases, collects in a sump 60 at the bottom of the absorber and is conducted by line 61 to a pressure letdown valve 62 where the pressure on the solution is reduced to that prevailing at the top of regenerator column 51 after which the letdown solution is introduced into the top of the regenerator column by line 63.

At the top of regenerator column 51 a portion of the absorbed acid gases together with steam flashes off and the partially regenerated solution then descends through section D countercurrently to rising stripping steam introduced into the bottom of the column as will be described below.

The regenerated solution, now containing a relatively small amount of absorbed acid gases, collects at the bottom of column 51 on a trapout tray 64 and is then fed to the heat-exchanger system generally designated by the numeral 53 by line 65. Scrubbing solution leaves heat exchanger 53 by line 66 and is fed to the bottom of the regenerator where it collects in a sump 67. The regenerated solution leaves the bottom of the regenerator

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tor column by line 68 and, after passing through pressure letdown valve 69, is conducted by line 70 to a flashing tank 71 operating under a reduced pressure. The flashed solution 72 in tank 71 is then conducted by line 73, recycle pump 74 and line 75 to the absorber column 58. As previously described, the regenerated solution is split into two streams, a minor stream which is first cooled in cooler 57 and introduced into the top of the absorber by line 56, and a major stream which is introduced into the absorber at an intermediate level by line 58.

Stripping steam for the desorption of acid gases from the solution is introduced into the bottom of the stripping column 51 by lines 76, 77 and 78. The manner in which these three sources of stripping steam are generated will be described in detail below. The stripping steam introduced from these three sources rises in column 51 countercurrently to descending scrubbing solution. The mixture of steam and desorbed acid gases at the top of column 51 passes through a contact tray 79 equipped with bubble caps 80. Contact tray 79 is supplied with aqueous condensate 81 as will be described below. The mixture of steam and desorbed gases passes upwardly through the layer of aqueous condensate 81 as indicated by arrows 82, and then is removed from the top of the regeneration column by line 83, and is conducted to a condenser 84, cooled by a cooling medium supplied by line 85 and leaving by line 86. In the condenser 84 most of the steam content of the mixture is condensed and the water condensate withdrawn by line 87. A portion or all of the condensate is returned to the tray 79 at the top of the regenerator column by line 88. Depending on the water balance in the system, a portion of the condensate may be removed by line 89. The condensate, which leaves the condenser 84 at temperatures, e.g. 130° to 150° F., is heated on tray 79 by contact with the mixture of steam and desorbed gas bubbling up through the condensate to temperatures of e.g. 190° to 210° F. The preheated aqueous condensate is withdrawn from tray 79 by line 90 and is then circulated by means of pump 91 and line 92 to the top of direct-contact column 54 where the preheated condensate is brought into direct contact with hot, steam-containing feed gas as will be described in more detail below.

Reference is now made to heat-exchanger system 53 which provides for indirect heat exchange between the hot feed gas stream and the scrubbing solution to generate stripping steam. This includes shell 93. Shell 93 is supplied with scrubbing solution collecting on tray 64 by line 63. The solution flows over a bundle of tubes designated by reference numeral 94 through which hot feed gas is introduced by line 52. Heat from the feed gas is transferred through the tube bundle to the scrubbing solution, heating it to its boiling point. The steam generated passes through line 78 and is introduced into the bottom of column 51 as stripping steam. The scrubbing solution overflows weir 95 and is then withdrawn from the shell by line 66 and introduced into the bottom of regenerator column 51 where it collects in sump 67. In heat-exchanger system 53, the hot feed gas is partially cooled, yielding up a portion of its heat by indirect heat exchange with the scrubbing solution and leaves the heat-exchanger system 53 at a reduced temperature by line 96. After passing through a knockout pot 97 for removal of condensed water vapor, it is conducted by line 98 to the bottom of direct contactor 54. The feed gas stream passes upwardly through direct contactor 54

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supplied with packing, trays or other means for producing intimate gas-liquid contact in the cross-hatched portion of the column designated by the reference letter E. The upwardly rising process gas flows countercurrently to a descending stream of water introduced into the top of the direct contactor 54 by line 92. As a result of the direct contact between the hot, steam-containing feed gas and the water introduced by line 92, the water is heated, e.g. to temperatures of 245° to 260° F., with corresponding cooling of the process gas and condensation of some of its steam content. The thus cooled process gas leaving the top of direct contactor 54 is introduced by line 85 into the bottom of absorber 58.

The water heated by direct contact with the hot process gas in direct contactor 54 to temperatures e.g. of 245° to 260° F. collects at the bottom of the contactor column in sump 99. This heated water is then subjected to flashing in two stages at two different pressure levels. The first stage of flashing occurs in flash tank 100 which is operated at a pressure slightly above the pressure prevailing at the bottom of regenerator column 51. The hot water from sump 99 is introduced into flash tank 100 by line 101 after passing through pressure letdown valve 102 which reduces the pressure to that prevailing in tank 100. The steam evolved by the flashing of the solution in tank 100 is removed from the tank by line 103 and introduced into the bottom of regeneration column 51 by line 76.

The second flashing of the water heated in direct contactor 54 occurs in a second flash tank 104. The somewhat cooled water from flash tank 100 is conducted by line 105 and pressure letdown valve 106 to flash tank 104. Flash tank 104 is connected by line 107 and line 108 to the suction side of a compressor 109 which reduces the pressure above the water in flash tank 104 to a pressure lower than that prevailing at the bottom of regeneration column 51. As a result of the reduction of the pressure over the water in tank 104, further flashing of steam occurs and the evolved steam is conducted by lines 107 and 108 to the compressor 109 where it is compressed to a pressure at least slightly above that prevailing at the bottom of regenerator column 51 and then introduced into the bottom of the column by line 77 as stripping steam. Compressor 109 is driven by a steam turbine (or other suitable drive means) 110 connected to compressor 109 by shaft 111.

The water collecting in tank 104, further cooled by the endothermic steam flashing occurring in tank 104, is removed by line 112 and introduced into line 90 and then recirculated by circulation pump 91 and line 92 to the top of direct contactor column 54.

The direct contactor system described above thus includes the direct contactor column 54, flash tank 100, flash tank 104, and recirculation pump 91. Water is continuously circulated by recycle pump 91 through the direct contactor column 54 where the cooled water is brought into contact with hot process gas entering the bottom of the contactor column by line 98 whereby the process gas yields up a portion of its heat to the circulating water. The hot water thus produced is then flashed in two stages in flash tank 100 and 104, resulting in evolution of steam which is employed as stripping steam in column 51, and the cooled water being then recirculated by circulation pump 91 for renewed contact with the hot process gas stream.

In the embodiment shown in FIG. 2, the water circulating in the direct contactor system described above is generated from two sources. Some of the water is gen-

erated by condensation of the water vapor in the hot feed gas stream as it contacts the circulating water in direct contactor 54. Another portion of the water is supplied by the condensate collecting at the top of the regeneration column 51 which, after preheating on tray 79 as described above, is conducted by line 90 to recirculating pump 91. The relative proportions of the water supplied to the direct contactor system described above by condensate from the hot feed gas and condensate from the top of the regeneration column 51 will depend on the steam content in the hot feed gas and other factors affecting the overall water balance in the system. The water balance can be suitably adjusted by the amount of condensate withdrawn from the system by line 89.

The suction side of compressor 109 is also connected by line 108 to flash tank 71 fed by hot regenerated solution from the bottom of the regenerator column 1. The pressure over the regenerated solution in tank 71 is reduced to a pressure lower than that prevailing at the bottom of the regenerator, causing flashing of low-pressure steam from the solution. The flashed, low-pressure steam flows by line 108 to the suction side of compressor 109 where it is compressed to a pressure at least slightly above that prevailing at the bottom of the regenerator and is introduced by line 77 into the bottom of the regenerator as stripping steam.

The system shown in FIG. 2, wherein the indirect heat-exchanger system 5 of FIG. 1 is replaced by the direct contactor system described above, makes possible the recovery of still further amounts of heat from the hot feed gas stream since it makes possible a closer temperature approach between the feed gas stream and the water which is to be converted into stripping steam. As a result, more heat is transferred to the water which can be converted into useful stripping steam and further cooling of the process feed gas occurs which still further increases the overall efficiency of the scrubbing process.

The advantages of the system shown in FIG. 2 are illustrated by the following Example 3. The same scrubbing solution is employed as in the previous Examples as well as the same feed gas composition. As in the previous examples, the CO₂ content of the feed gas is reduced to 0.1% by volume leaving the absorber. The absorption and regeneration columns were of the same size and contained the same size and type of tower packing. Other conditions such as absorber and regenerator pressures, the temperature of the solution entering the top of the absorber were maintained the same such that the results of Example 3 can be directly compared to those of the previous examples.

In Example 3, the feed gas enters the system of FIG. 2 through line 52 at a temperature of 322.6° F. and a pressure of 410 psia saturated with steam. In heat-exchanger system 53, indirect heat exchange between the feed gas and the scrubbing solution produces 63,260 pounds per hour of steam which is fed into the regeneration column as stripping steam with resulting cooling of the process gas to a temperature of 263.6° F. The partially cooled process gas then enters direct contactor column 54 and is brought into direct contact with water entering the top of the column by line 92 at a temperature of 214° F. As a result of the direct contact between the feed gas and the water, the feed gas is cooled to a temperature of 223° F. leaving the top of column 54 and the water is heated to 256.6° F. leaving the bottom of

column 54. The feed gas enters absorber column 58 at a temperature of 223° F. and a pressure of 404 psia.

The heated water from direct contactor 54 is introduced into flash tank 100 maintained at a pressure of 25.0 psia, slightly above the pressure of 24.7 psia prevailing at the bottom of regenerator column 51. Under these conditions, 10,804 pounds per hour of steam are generated in flash tank 100 which flows by way of lines 103 and 76 into the bottom of the regeneration column. As a result of the endothermic steam flashing in tank 100, the water leaving the bottom of tank 100 by line 105 is cooled to 240.1° F. and is conducted to flash tank 104 maintained at a pressure of 15.3 psia. As a result of the reduction in pressure, further steam flashing occurs at the rate of 17,090 pounds per hour of steam which is conducted by lines 107 and 108 to the suction side of compressor 109 where it is compressed to the pressure prevailing at the bottom of regenerator column 51, and introduced into the bottom of the column by line 77. Thus, the total steam generated in the two flash tanks 100 and 104 amounts to 27,895 pounds per hour.

The aqueous condensate from the top of column 51, preheated to a temperature of 210° F., is supplied by line 90 to the direct contactor system at the rate of 11,925 pounds per hour, whereas condensate produced by the condensation of water in the hot feed gas produces 15,970 pounds per hour of the total water requirements of 27,895 pounds per hour.

The remaining stripping steam requirements are supplied by the steam generated in flash tank 71. The hot, regenerated solution collecting at the bottom of regeneration column 51 at a temperature of 248° F. is conducted to flash tank 71 maintained at a pressure of 15.0 psia. Under these conditions, steam flashes from the solution at the rate of 53,790 pounds per hour and the solution cools to a temperature of 220.2° F. The flashed steam is conducted by line 108 to the suction side of compressor 109 where it is compressed to a pressure just above the pressure prevailing at the bottom of the regeneration column 51, and introduced by line 77 as stripping steam.

The solution leaving tank 71 is conducted by recycle pump 74 to the absorber where it is split into two streams, one of which (75% of the total) enters the intermediate portion of the absorber by line 58 at a temperature of 220.2° F. and the second portion (25% of the total) is cooled in cooler 57 to a temperature of 158° F. and enters the top of the absorber by line 56.

Table II below summarizes the results of Example 3 and also shows the results of comparative Example A which, as in conventional processes, produces all of the stripping steam requirements by indirect heat exchange between the hot feed gas and the scrubbing solution in a heat-exchanger system such as that of system 4 in FIG. 1 or system 53 in FIG. 2. In all cases the regeneration heat required is expressed in millions of BTU per hour (MM BTU/hour) assuming a pound of steam is equivalent to 950 BTU.

TABLE II

	Example A	Example 3
Regeneration heat generated by heat exchange system 53 MM BTU/hr.	143.5	60.1
Regeneration heat generated by direct contactor heat exchange system MM BTU/hr.	—	26.5
Regeneration heat generated in flash tank 71		

TABLE II-continued

	Example A	Example 3
MM BTU/hr.	—	51.1
Total regeneration heat required, MM BTU/hr.	143.5	137.7
Temperature of feed gas at inlet to heat-exchange system 53, °F.	362.1	322.6
Heat saved in hot feed gas for other uses, MM BTU/hr.	—	83.4
Heat saved in hot feed gas as a percentage of total regeneration heat requirements	—	58.1
Cooling duty of cooler 57 MM BTU/hr.	45.2	28.8
Cooling duty of overhead condenser 84, MM BTU/hr.	130.1	67.7
Solution circulation rate, gallons per minute	4310	3940

As may be seen from Table II, the heat saved from the hot feed gas is 83.4 MM BTU/hour equivalent to an increased recovery of 87,790 pounds per hour of steam. This amounts to 58.1% of the total regeneration heat requirements of the conventional system of Example A. Note that the temperature of the hot feed gas entering the scrubbing system is 322.6° F. in Example 3 while a temperature of 362.1° F. is required in Example A. The heat content of the steam-saturated feed gas stream between the temperature levels of 362.1° F. and 322.6° F. is equivalent to the heat savings of 83.4 MM BTU/hour which is conserved for other uses by operation in accordance with the invention. As in Examples 1 and 2, these valuable heat savings are also accompanied by important savings in the duty of the solution cooler (cooler 57 in FIG. 2), the overhead condenser (condenser 84 in FIG. 2) and reduction in the solution circulation rates.

The amount of steam that can be recovered by flashing of the water in indirect heat-exchange system 5 in FIG. 1 or the direct contactor system of FIG. 2 depends on the extent to which the pressure is reduced over the water. The practical limit of heat recovery in this manner is related to the cost of producing the reduction in pressure over the water and subsequently compressing the steam thus generated to the pressure prevailing at the bottom of the regenerator column. The greater the reduction in pressure, the greater is the quantity of steam that will be evolved by flashing; but the power requirements to produce the reduced pressure and to recompress the steam become disproportionately high as the pressure is reduced below levels corresponding to a water boiling point of 180° F. Generally, the practical limits for producing flash steam from the water will correspond to reduced boiling temperatures of from 194° F. to 250° F., and more usually from 200° F. to 220° F. Thus, the indirect heat exchanger 5 in FIG. 1 will generally be operated at pressures of not less than 10 psia (corresponding to a water boiling temperature of 194° F.) and preferably not less than 11.5 psia (corresponding to a water boiling temperature of 200° F.). Similarly, the flash tank 104 of the direct contactor system of FIG. 2 will be subject to the same practical limitations of pressure reduction.

In carrying out the flashing operation on the regenerated solution in flash tank 18 of FIG. 1 and flash tank 71 of FIG. 2, the amount of flash steam evolved will also be dependent on the final pressure over the regenerated solution in the flash tank. In general, the same practical limits for the lower limit of pressure reduction will

apply in the case of flashing the regenerated solution as in the case of flashing the water as discussed above.

The amount of heat recovered by flashing of the regenerated solution will also, of course, depend on the temperature of the regenerated solution as it leaves the regeneration column. The higher this temperature in general, the more heat that may be recovered by flashing of the regenerated solution. In most applications, the regeneration column will operate at slightly above atmospheric pressures such that the temperature of the regenerated solution will typically be of the order of 230° F. to 270° F. Under these conditions, and subject to the limitations in the extent to which the pressure can be reduced over the solution, the amount of flash steam that will be produced will generally correspond to a drop in the solution scrubbing temperature (as a result of the endothermic steam flashing) up from about 7° F. to 35° F., and in most cases from 10° F. to 30° F.

In the embodiment shown in FIGS. 1 and 2, the reduction of the pressure over the water and over the regenerated solution is produced by a mechanical compressor which produces the pressure reduction and then compresses the low-pressure steam produced to the pressure prevailing at the bottom of the regeneration column. Other means for producing the pressure reduction and compressing the evolved steam may also be employed such as a steam-ejector system such as that shown in U.S. Pat. No. 3,823,222 of Homer E. Benson. When using a steam ejector system, a supply of high- or medium-pressure steam is supplied to the ejector as motive steam which produces the reduced pressure and compresses the mixture of motive and flash steam to the pressure prevailing in the regenerator column. When using the steam ejector system, some of the steam-stripping requirements will be supplied by the motive steam. The choice between a mechanical compressor system as shown in FIGS. 1 and 2 and alternate means such as a steam ejector will depend on a variety of considerations such as the relative capital cost, the availability of suitable motive steam, water-balance considerations, and others.

As pointed out previously, it is advantageous to supply the water to be employed for direct or indirect heat exchange against the hot feed gas from water which is produced in the course of the scrubbing process. When employing an indirect heat exchange between the water and the hot feed gas such as heat-exchanger system 5 in FIG. 1, it is highly preferred to employ preheated aqueous condensate which is produced by condensation of the overhead vapors from the regenerator. The preheating of the condensate at the top of the regenerator increases heat efficiency. Another advantage is that the necessity for providing an outside source of treated boiler feed water is eliminated. The aqueous condensate from the overhead condenser may be contaminated with components of the scrubbing solution, but is nevertheless suitable for reuse in the scrubbing system, since the contaminants are thus kept within the scrubbing system.

When using a direct contactor system such as shown in FIG. 2 for exchanging heat between the hot process gas and the water, a portion of the aqueous condensate is automatically produced from the steam condensing from the hot feed gas in direct contactor 54. This source of condensate may be supplemented by condensate from the overhead condenser of the regenerator as illustrated in FIG. 2. Here again, the generation of the

required water from aqueous condensate produced within the scrubbing system, itself, eliminates the need for providing pretreated boiler feed water.

In the system of FIG. 1, still further quantities of low-level heat may be recovered from the hot condensate condensing from the process gas which collects in the knockout pot 43A. The condensate leaving knockout pot 43A may be introduced into a separate flash tank (not shown) with a suitable pressure letdown valve between the knockout pot and the flash tank. The flash tank receiving the hot condensate is connected to the suction side of the compressor 45 which results in reduction of the pressure over the condensate to a pressure lower than that at the bottom of the regenerator 51, whereby low-pressure steam flashes off from the hot condensate and, after compression in compressor 45, is fed into the bottom of the regenerator through line 23 together with compressed steam raised in heat-exchange system 5 and flash tank 18.

In the system of FIG. 2, additional quantities of condensate from the hot process gas can be fed to direct-contact tower 54 by eliminating knockout pot 97, thus carrying the condensate otherwise removed in knockout pot 97 to the direct-contact tower 54 and recovering additional heat from this condensate. In that event, the condensate from the hot process will produce a greater proportion of the total water needed to supply the requirements of flash tanks 100 and 104, correspondingly reducing the amount of condensate supplied by line 90 from the top of the regenerator 51.

The invention, of course, is not limited to the specific embodiments shown. It may be applied to other types of scrubbing systems than those specifically illustrated. For example, all of the scrubbing solution may enter the absorber at the top of the column with or without cooling rather than in two separate streams as shown in FIGS. 1 and 2. Likewise, a so-called two-stage regenerator may be employed in which a portion of the scrubbing solution is subjected to an intermediate degree of regeneration (i.e. leaving a higher proportion of absorbed gases in the solution) while a smaller fraction of the solution is subjected to additional steam-stripping to produce a more thoroughly regenerated scrubbing solution such as the system shown in FIG. 7 of U.S. Pat. No. 2,886,403.

We claim:

1. In a cyclic process for the removal of acid gases from a feed gas stream consisting of a hot, steam-containing gas mixture by means of an aqueous alkaline scrubbing solution which is continuously recycled between an absorption stage in which said acid gases are absorbed by said scrubbing solution and a regeneration stage in which said acid gases are desorbed by steam-stripping, and in which said stripping steam is at least partially derived from said hot, steam-containing feed gas, an improved method for maximizing heat recovery from said feed gas and thereby increasing the thermal efficiency of said regeneration stage comprising the steps of

- (a) passing said feed gas in indirect heat exchange with said scrubbing solution, thereby heating said solution and producing steam;
- (b) utilizing the steam produced in step (a) as stripping steam for regenerating said solution;
- (c) utilizing the partially cooled feed gas from step (a) to heat water in a heat exchange system separate from that employed in step (a);

(d) reducing the pressure over the water heated in step (c), thus reducing the boiling temperature of said water and producing steam at a pressure lower than the pressure in said regeneration stage and bringing the cooled water resulting from said pressure reduction into heat-exchange relationship with said feed gas, thus further cooling said feed gas;

(e) compressing the low-pressure steam produced in step (d) to at least the pressure in said regeneration stage and utilizing such compressed steam as stripping steam for regenerating said solution;

(f) passing the cooled feed gas from step (d) to said absorption stage for removal of acid gases therefrom.

2. A process in accordance with claim 1 wherein the water employed in steps (c) and (d) is provided at least in part by aqueous condensate produced in the course of said scrubbing process.

3. A process in accordance with claim 1 wherein the heat exchanger in step (c) is an indirect heat exchange system in which the hot feed gas is brought into indirect heat exchange with said water.

4. A process in accordance with claim 1 in which the heat exchanger in step (c) is a direct-contact heat exchange system in which the hot feed gas is brought into direct contact with cooled water resulting from said pressure reduction over said water.

5. A method in accordance with claim 1 in which the scrubbing solution comprises an aqueous solution of potassium carbonate.

6. A method in accordance with claim 1 in which the absorption stage is conducted at a substantial superatmospheric pressure and in which the temperatures in the absorption stage are close to those in the regeneration stage.

7. In a cyclic process for the removal of acid gases from a feed gas stream consisting of a hot, steam-containing gas mixture by means of an aqueous alkaline scrubbing solution which is continuously recycled between an absorption stage in which said acid gases are absorbed by said scrubbing solution and a regeneration stage in which said acid gases are desorbed by steam-stripping, and in which said stripping steam is at least partially derived from said hot, steam-containing feed gas, an improved method for maximizing heat recovery from said feed gas and thereby increasing the thermal efficiency of said regeneration stage comprising the steps of

- (a) passing said feed gas in indirect heat exchange with said scrubbing solution, thereby heating said solution and producing steam;
- (b) utilizing the steam produced in step (a) as stripping steam for regenerating said solution;
- (c) utilizing the partially cooled feed gas from step (a) to heat water in a heat exchange system separate from that employed in step (a);
- (d) reducing the pressure over the water heated in step (c), thus reducing the boiling temperature of said water and producing steam at a pressure lower than the pressure in said regeneration stage and bringing the cooled water resulting from said pressure reduction into heat-exchange relationship with said feed gas, thus further cooling said feed gas;
- (e) reducing the pressure over the hot, regenerated solution in a flashing zone, thereby producing steam at a pressure lower than the pressure in said

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regeneration stage with concomitant cooling of said solution, and passing the cooled solution to said absorption zone;

(f) compressing the low-pressure steam produced in steps (d) and (e) to at least the pressure in said regeneration stage and utilizing such compressed steam as stripping steam for regenerating said solution;

(g) passing the cooled feed gas from step (d) to said absorption stage for removal of acid gases therefrom.

8. A process in accordance with claim 7 wherein the water employed in steps (c) and (d) is provided at least in part by aqueous condensate produced in the course of said scrubbing process.

9. A process in accordance with claim 7 wherein the heat exchange system in step (c) is an indirect heat

exchange system in which the hot feed gas is brought into indirect heat exchange with said water.

10. A process in accordance with claim 7 in which the heat exchange system in step (c) is a direct contact heat exchange system in which the hot feed gas is brought into direct contact with cooled water resulting from said pressure reduction over said water.

11. A method in accordance with claim 7 in which the scrubbing solution comprises an aqueous solution of potassium carbonate.

12. A method in accordance with claim 7 in which the absorption stage is conducted at a substantial super-atmospheric pressure and in which the temperatures in the absorption stage are close to those in the regeneration stage.

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PATENT SPECIFICATION

(11) 1 484 050

1 484 050

(21) Application No. 55851/74 (22) Filed 24 Dec. 1974

(23) Complete Specification filed 23 Dec. 1975

(44) Complete Specification published 24 Aug. 1977

(51) INT CL² B01D 53/34 C10K 1/12

(52) Index at acceptance

C1A S171 S181 S18Y S410 S412 S413 S41Y S420 S46Y S492
S642 S681 S712 SB

C5E D14



(54) IMPROVEMENTS TO THE PROCESSES FOR REMOVING ACID GASES FROM GASEOUS MIXTURES

(71) I, FRANCIS VAN HECKE, a Belgian citizen of Avenue de Tervueren 183-B-1040 Brussels, Belgium, do hereby declare the invention for which I pray that a patent may be granted to me, and the method by which it is to be performed, to be particularly described in and by the following statement:—

This invention relates to improvement to the regeneration of regenerable aqueous scrubbing solutions, which are used for the bulk removal of acidic gases, such as CO₂, H₂S, HCN, COS, SO₂ and mercaptans, by absorption from gas mixtures containing these acidic gases, in a cyclic process in which said aqueous scrubbing solutions are regenerated by steam stripping before being recycled to the absorption stage.

The invention relates particularly to a process for the regeneration of regenerable aqueous scrubbing solutions which are used for the bulk removal of acidic gases, by absorption from gas mixtures containing these acidic gases, in a cyclic process in which said aqueous scrubbing solutions are regenerated in a regeneration system by steam stripping before being recycled to the absorption stage, a part of the scrubbing solutions being regenerated in a main regeneration section of the regeneration system, whereas the other part of the scrubbing solutions is regenerated in an auxiliary regeneration section of the regeneration system.

The gas mixture to be purified may be, for example, either a natural gas stream or an industrial process gas.

The most commonly used aqueous scrubbing solutions used in cyclic processes are solutions of alkaline carbonates, particularly potassium carbonate which may contain any of the well known activation additives such as arsenites, borates, aminoacids, alkanolamines or other additives which increase the rates of absorption and desorption of the acidic gas in the scrubbing solution. In addition to alkaline carbonate solutions, other regenerable solutions such as aqueous

solutions of ethanolamines and other alkanolamines, of alkali metal salts of aminoacids, of alkaline phosphates, alkaline phenates, alkaline borates and other similar components may be used either separately or in admixture. For the absorption of SO₂ it is sometimes preferred to use a solution of alkali metal sulfites and bisulfites. It is common practice to incorporate in these solutions a corrosion inhibitor.

The regeneration of the spent scrubbing solution by steam stripping and boiling takes place at a regeneration pressure which is usually close to atmospheric pressure, corresponding to boiling temperatures which are commonly in the range of 70°C to 150°C.

The stripping steam may be produced by boiling the aqueous scrubbing solution in an indirectly heated reboiler, or be obtained from an outside steam source supplying live steam. The spent scrubbing solution which leaves the absorber where the absorption of the acidic gas has normally been carried out under a superatmospheric pressure, is first submitted to a pressure reduction and flashing before being introduced into the steam stripping zone of the regeneration section.

The regeneration equipment which is commonly used comprises a regeneration column equipped with packing materials, bubble cap plates, sieve trays or other suitable means for ensuring intimate contact between the solution and the stripping steam. Similar equipment is used for the absorption stage.

According to the temperature at which the regenerated scrubbing solution is recycled to the absorber and at which the spent scrubbing solution leaves the said absorber, the cyclic process may be defined as "isothermal", "optimal" or "classic". In the classic cycle the absorber operates at a moderate temperature which may be close to ambient temperature. As the regeneration by steam stripping takes place at a higher temperature, several heat exchangers and/or solution heaters and coolers will be required in

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addition to the usual absorption and regeneration equipment. Such a classic cycle is for instance preferred for the removal of CO₂, when the scrubbing solution is an aqueous monoethanolamine solution.

When the gaseous mixture which is to be purified is available at such a pressure that the partial pressure of the acidic gas constituent which must be removed is relatively high, it becomes possible to increase the temperature of the absorption stage. The preferred cyclic process will then be either of the isothermal or the optimal type.

The difference between the so-called isothermal cycle and the optimal cycle can most easily be shown in the case of a two stage absorption and regeneration process. In the usual two stage absorption and regeneration cycle, the total flow of spent scrubbing solution is sent to the top of the regenerator, from where a major part of partially regenerated solution is withdrawn at an intermediate level, whereas the remaining minor part of solution is more fully regenerated. Only this minor part of "lean" solution is sent to the top of the absorber, while the major stream of "semi-lean" solution is recycled to the absorber at an intermediate level.

In the so-called "isothermal" cycle only the minor stream of lean solution is cooled before being introduced at the top of the absorber, whereas the major stream of semi-lean solution is recycled to the absorber without any intermediate cooling whatsoever. In the "optimal" cycle, the minor stream of lean solution is cooled before recycling to the top of the absorber, in order to adjust the vapour pressure of the solution to a value corresponding to the required degree of purification, whereas the temperature of the semi-lean solution which is recycled to an intermediate level is adjusted by cooling to such a value that the optimal temperature profile for an efficient absorption process can be achieved.

This optimal thermal profile is sometimes important in order to control or limit some unwanted side reactions.

When the optimal cycle is to be applied to a single stage split-flow regeneration system, both the minor flow of regenerated solution which is recycled to the top of the absorber and the major flow of solution which is recycled to an intermediate level of the absorber are cooled although to different degrees. By comparison, in the so-called isothermal single stage split-flow system only the minor flow of regenerated solution is cooled, whereas the major flow of regenerated solution is recycled to an intermediate level of the absorber without cooling.

It is already known that the overall heat requirement of these cyclic processes can be reduced if a dual-pressure regeneration system is used, in which the major part of the spent

scrubbing solution is regenerated in a main regeneration column operating at a higher pressure level, while the minor part of the solution is regenerated in a second auxiliary regeneration column operating at a lower pressure level, by a flow of stripping steam which has been produced exclusively by flashing, following pressure let-down, of the major part of regenerated solution from the higher pressure level to the lower pressure level.

According to this system essentially all the heat required in order to produce the stripping steam used for regeneration both in the higher pressure and in the lower pressure regeneration columns is introduced in the system at the higher pressure level, where the temperature increase of the solution serves as a heat storage from which upon pressure let-down to the lower pressure level, the stripping steam for the lower pressure regeneration column will be obtained.

The thermal efficiency of this dual-pressure regeneration system increases when the pressure difference between the two pressure levels increases. As in general the acidic gases which are desorbed during the regeneration process have to be collected for further treatment, it is usually preferable to operate the lower pressure regeneration column at a pressure which is slightly superatmospheric, so as to decrease the energy requirement for moving the desorbed gas to the next processing stage.

This means that in practice the relative heat saving will be dependent on the increase of pressure, and therefore of temperature, at the column operating at the higher pressure level.

This increase in temperature sometimes constitutes an important drawback in view of some well known undesirable side-reactions such as the thermal degradation of various scrubbing agents, notably the alkanolamines or aminoacids, or the formation of formates, as well as a more important risk of corrosion.

According to a modification of the above mentioned dual-pressure regeneration process, part of the stripping steam required in the column operating at the lower regeneration pressure can be produced in a solution reboiler operating at said lower regeneration pressure level. If therefore milder operating conditions are desired, it is no longer required that the major part of the solution should be regenerated at the higher pressure level nor is it necessary to increase the pressure in the higher pressure level to the same extent.

However such decreases of the severity of operation can only be obtained at the expense of an increase of the overall heat requirement. This increased heat requirement is introduced in the system through two solution reboilers, one at the higher pressure level and one at the lower pressure level. In this two-reboiler version of the dual-pressure re-

generation system, the specific problems associated with solution reboilers, i.e. risk of solution overheating and reboiler corrosion are increased in relation to the increased reboiler duty.

According to the present invention, it was found that, contrary to all expectations, the severity of operation and the overall heat requirement can be decreased simultaneously together with a reduction of the solution reboiler duty.

Thus, according to one feature of the present invention, there is provided a process for the regeneration of regenerable aqueous scrubbing solutions which are used for the bulk removal of acidic gases, by absorption from gas mixtures containing these acidic gases, in a cyclic process in which said aqueous scrubbing solutions are regenerated in a regeneration system by steam stripping before being recycled to the absorption stage, a part of the scrubbing solutions being regenerated in a main regeneration section of the regeneration system, whereas the other part of the scrubbing solutions is regenerated in an auxiliary regeneration section of the regeneration system, said process comprising the steps of selecting the pressure in the auxiliary regeneration section independently from the pressure in the main regeneration section and obtaining at least a part of the stripping steam required in the auxiliary regeneration section through flashing of the scrubbing solution in a reduced pressure zone connected to the suction side of a steam-jet thermocompressor, the mixture of thermocompressor motive steam and recompressed flashed vapour being directly discharged into said auxiliary regeneration section.

In a particular embodiment of the process according to this invention, the pressure in the auxiliary regeneration section is lower than the pressure in the main regeneration section. Preferably a part of the stripping steam required in said auxiliary regeneration section is obtained by direct flashing of the scrubbing solution upon pressure reduction from the higher pressure of the main regeneration section to the lower pressure of the auxiliary regeneration section.

Contrary to the known practice, it was surprisingly found that the pressure in the auxiliary regeneration section may also, according to this invention, be equal to or higher than the pressure in the main regeneration section.

A particular embodiment of the process according to this invention enables a substantial improvement of the thermal efficiency of the regeneration procedure, due to the fact that the stripping steam required at the lower regeneration pressure level is only partially obtained through flashing resulting from a pressure let-down of the scrubbing solution from the higher pressure regeneration section

to the lower pressure regeneration section, whereas the remainder of the stripping steam required in the said lower pressure regeneration section is obtained through additional flashing of the scrubbing solution to a still lower pressure level and recompression of the flashed vapour mixture by way of a steam-jet thermocompressor. This technique avoids all risk of solution overheating in an additional solution reboiler, and recovers in the form of useful stripping vapour a certain amount of heat stored in the solution which would otherwise be lost in the coolers which are used to trim the temperature of the solution before recycle to the absorber.

According to an additional feature of the invention, the flashing to the reduced pressure induced by the operation of the steam-jet ejector is performed on a partially or totally regenerated scrubbing solution obtained from either the main regeneration section or the auxiliary regeneration section.

According to another feature of the invention, a part of the mixture of thermocompressor motive steam and recompressed flashed vapour is discharged in the main regeneration section.

Other details and features of the invention will appear from the following detailed description, wherein reference is made to Figures 1 to 4 of the attached drawings which represent flow diagrams illustrating, by way of non-limitative examples, four embodiments of the process according to this invention.

In the following description, reference is made, by way of example, to a specific operation, i.e. the bulk removal of CO₂ from gas mixtures, using, as scrubbing solution, an aqueous concentrated solution of potassium carbonate containing known appropriate additives.

The flow diagram shown in Figure 1 refers to an optimal cycle of regeneration of the spent scrubbing solution, which due to the possibility of selective temperature adjustment of both the lean and the semi-lean scrubbing solution streams allows enough flexibility to achieve the desired temperature profile in the absorber.

It is thus possible to keep constant the spent solution outlet conditions in the absorption stage even when modifications in the regeneration section tend to modify the temperature of the semi-lean and lean scrubbing solution streams which are to be recycled.

The process gas which has to be purified is introduced by line 30 at the base of the absorber 1 and leaves the top of said absorber by line 31 after acidic gas removal.

The hot spent scrubbing solution leaving the bottom of absorber 1 through line 21 is sent partly to a main regeneration column 2 through a pressure let-down valve 52 and line 22 and partly to an auxiliary regeneration

column 3 through a pressure let-down valve 53 and a line 23.

In the flash zone 7 of the main regeneration column 2 which operates at the higher pressure level, a certain amount of steam and acidic gas is given off, while the remaining solution, which has been cooled by flashing to its boiling temperature under the prevailing pressure, flows downward towards the stripping zone 15.

In this stripping zone the solution flows counter-currently to the stripping steam and is progressively heated by said stripping steam, thereby maintaining the solution at its boiling temperature as it progressively regenerates.

Part of the solution is withdrawn at an intermediate level from draw-off pan 40 after having been partially regenerated, whilst the remainder of the solution flows down through the stripping section 16 to the base of the main column 2 where it is heated in the solution reboiler 90.

The fully regenerated "lean" solution leaving the base of the regeneration column 2 by line 25 flows through the pressure reducing valve 57 to the expansion vessel 4 which is maintained at the same pressure as the base of the auxiliary regeneration column 3. The flashed gas mixture liberated in 4 flows to the lower part of column 3 where it is used as stripping steam, whereas the lean solution, which has been cooled by flashing, leaves vessel 4 by line 26 and flows to exchanger 99 where it preheats the regeneration condensate before being pumped by pump 81 to the top of the absorber 1 after temperature adjustment in cooler 91.

The partially regenerated "semi-lean" solution which is withdrawn from draw-off pan 40 flows through the pressure reducing valve 56 and line 24 to the flash zone 9 at the base of the auxiliary column 3.

The gas mixture which is generated by flashing, mixes with the vapour coming from expansion vessel 4 and flows upwards in column 3 where it is used as stripping steam.

At the top of said auxiliary column 3, the solution which has been fed through a pressure let-down valve 53 flows through a flash zone 8 towards a regeneration section 17 where it is regenerated by the flow of stripping steam which rises from the base of the column.

After regeneration, the solution which collects in a draw-off pan 41 is withdrawn through a line 271 and mixes with solution from the flash zone 9 which flows through a line 27. These combined streams flow through a pressure reducing valve 58 to an expansion vessel 5 which is connected to the suction side of a steam-jet thermocompressor or ejector 80. The steam-jet ejector 80 is operated by the motive steam supplied by a line 34, in order to maintain in the expansion

vessel 5 a suction pressure which is lower than the pressure prevailing in the auxiliary regeneration column 3, while the discharge pressure of the ejector 80 is adjusted so that the mixture of flashed vapour exhausted from expansion vessel 5 is recompressed to the pressure prevailing in the auxiliary regeneration column 3.

The recompressed mixture of flashed vapour and motive steam is delivered through line 35 to the auxiliary regeneration column 3 where it is used as stripping steam.

The total amount of stripping steam available in the auxiliary stripping column 3 will thus be made up in part of the steam liberated by flashing of the solution supplied from the main stripping column 2, through lines 24 and 33 and in part by the mixture of flashed vapour and motive steam provided by the steam-jet thermocompressor 80 through line 35.

Depending on the relative pressure differences which exist between the columns 2 and 3 on the one hand, and the column 3 and expansion vessel 5 on the other hand, the relative amounts of these steam supplies will vary in importance.

After flashing in expansion vessel 5, the semi-lean solution flows through a line 28 and is pumped by a pump 82 to an intermediate level of the absorber 1 through a line 281 and a cooler 92 where its temperature can be adjusted to the desired value. The mixture of desorbed acidic gas and residual stripping steam at the top of the auxiliary column 3 flows through a line 32 where it mixes with the desorbed acidic gas and steam mixture released by the main column 2 through a pressure reducing valve 54.

The combined flows are cooled in a cooler 93 where the major part of the steam condenses. The regeneration condensate which is collected in a reflux drum 6, from where the acidic gas leaves the unit, through a line 37, flows through a line 29 to a pump 83. After preheating in an exchanger 99, it is either delivered through a valve 51 to a boiling kettle 95 where it is vaporized to produce motive steam to be used in the steam ejector, or returned through a valve 62 to the main regeneration column 2 as process reflux, or if desired evacuated out of the system through a valve 63. Purging of the steam kettle 95 occurs through the blow down valve 46. The preheating of the condensate which in this case occurs in an exchanger 99, can be realized in various ways.

Sometimes it is preferred to recover for this purpose the heat contained in the acidic gas-steam mixture of line 32 before it is cooled in exchanger 93, or it may be chosen to use the hot semi-lean solution of line 28 as the heating medium. These various modifications have in common that the preheating of the condensate is achieved with heat which

is available within the regeneration system, in order to reduce the overall external heat requirement. If a separate cooler and condensate drum are provided for the overhead vapour from column 2, it is possible to avoid any pressure reduction of said acidic gas stream, which can thus be delivered at the same higher pressure level at which it has been desorbed.

When the process feed gas which has to be purified in absorber 1 is available at a sufficiently high temperature as is the case for some industrial process gases, it can conveniently be used as the heating medium in kettle 95 where the motive steam can be produced. As the motive steam is usually required at a pressure of 5 to 6 kg/cm², the temperature of the process gas leaving the kettle 95 through line 36 will still be high enough to be used as heating medium in the solution reboiler 90.

After having passed through the heating coils 36 of the reboiler 90, the process gas has to be cooled further and this can for instance occur in a boiler feed water heater.

After such a series of cooling operations, the process gas can then be introduced in the absorption column 1 through line 30.

When the process gas which must be purified is available at a lower temperature, as is the case with some natural gas streams, it is usually more convenient to use medium or low pressure steam as a heating medium in the solution reboiler 90 and the steam kettle 95.

The flow-sheet of figure 1 shows clearly the difference between the process according to this invention and the previously known dual-pressure regeneration methods, according to which the solution from the auxiliary column 3, which operates at the lower pressure level, flows directly through the lines 27 and 271 to the line 229, through valve 59 and pump 82 from where it is returned to the absorber 1 after temperature adjustment in cooler 92.

One obvious difference between the two methods of operation is that in the new method of this invention part of the cooling of the solution is realized in expansion vessel 5 and that consequently an important part of the heat which must otherwise be wasted in cooler 92 is recovered in a useful manner by way of the steam-jet thermocompressor 80.

An advantage of the process according to this invention is that this heat saving can be used either to decrease the overall heat requirement of the system or to decrease the severity of the operating conditions without however increasing the heat consumption.

Further research has shown that the same overall thermal efficiency can be achieved even if the pressure difference between the main regeneration column 2 and the auxiliary

regeneration column 3 is reduced, provided that a sufficient pressure difference is applied by the steam-jet thermocompressor between the expansion vessel in which the solution flashes and the regeneration column in which the recompressed flashed mixture is discharged.

It is thus possible, as shown in figure 2, to adjust the flow-sheet of the regeneration process in such a way that the direct flashing of the solution from the pressure level of the main regeneration column 2 to the pressure level of the auxiliary regeneration column 3 is avoided, so that all the flashing occurs upon pressure let-down between the pressure levels of either one of the two regeneration columns and the pressure level of an expansion vessel connected to the suction side of a steam-jet ejector.

As the transfer of heat and of stripping steam from the main regeneration column 2, which receives the reboiler heat, to the auxiliary column 3 which has no reboiler is now controlled by the operation of steam-jet thermocompressors 80 and 88, it becomes possible to vary the relative pressure levels of the main and of the auxiliary columns in such a way that, if desired, the main regeneration column 2 with solution reboiler 90 may operate at a lower pressure level than the auxiliary regeneration column 3.

It has also been found that it is possible to adjust the operating conditions so that the reboiler heat duty of the solution reboiler 90 of the main regeneration column 2 is decreased to such an extent that less than half of the total heat requirement for the production of stripping steam in the regeneration system is supplied through the solution reboiler. Although it is always possible to eliminate completely the solution reboiler by using live steam, such a practice requires that an equivalent quantity of process condensate should be removed from the system. The reduction of reboiler duty to which reference is made corresponds to a water balance of the regeneration system which is maintained in equilibrium, so that there is no need to evacuate process condensate out of the system.

As shown in Figure 1 and 2, the spent scrubbing solution from the absorber 1 is split in two fractions which are regenerated separately, in two regeneration columns 2 and 3. As the pressure may be adjusted independently on each column, it also follows that the two fractions of desorbed acidic gas may be obtained at two different pressure levels.

This implies that, for the fraction which is obtained at the higher pressure level, the cost of recompression for further processing is decreased.

As shown in Figure 2, the two streams of desorbed acidic gas from columns 2 and 3 are cooled separately in coolers 93 and 96 and evacuated through separate lines 37 and 39.

The process condensates which have been collected separately in two different reflux drums 6 and 10 may however conveniently be mixed in line 29 for further reuse.

5 If it is desired to obtain the totality of the desorbed acidic gas at the same pressure, it may be more advantageous to deliver all the spent scrubbing solution coming from the absorber 1 to a single flashing zone after
10 which one fraction of the flashed solution will be regenerated in the main regeneration column 2 and a second fraction will be regenerated in the auxiliary column 3, in which, according to the present invention, the stripping
15 steam is supplied either exclusively or partially by flashing of the solution which has been induced by a pressure reduction produced by a steam-jet ejector which also recompresses the flashed vapour mixture to the
20 required pressure level, the other part of the steam supply being obtained by flashing of the solution upon pressure let-down from the higher regeneration pressure level to the lower regeneration pressure level.

25 Further research has shown that the process of this invention also applies if, after the initial flashing of the single stream of spent solution coming from the absorber 1, some regeneration has been performed in the main
30 regeneration column and the separation of the solution in two fractions occurs in the form of the withdrawal of a side-stream of already partially regenerated solution which is then further regenerated separately in an
35 auxiliary column with stripping steam which is supplied according to the method of this invention.

40 For a better understanding of this embodiment of the invention, reference is made to figure 3 of the drawings.

The spent scrubbing solution leaving the base of absorber 1 by line 21 flows through the pressure reducing valve 52 to the main
45 regeneration column 2. After flashing in the flash zone 7, the solution flows downwards to the regeneration zone 18. A first part of the partially regenerated solution is withdrawn from draw-off plate 42, whilst the remainder
50 of the solution flows to the regeneration zone 15. A second fraction of the solution may be withdrawn from the column at the draw-off plate 40, and the remainder of the solution flows to the regeneration section 16, and from
55 there to the solution reboiler 90, and the bottom of the regeneration column 2.

60 According to the embodiment of the invention, as represented in Figure 3, the auxiliary regeneration column 3 is situated at a lower elevation level than the draw-off pan 42. The first fraction of solution, withdrawn from draw-off pan 42, flows through line 23
65 to the valve 53 and auxiliary column 3. The difference in level between draw-off pan 42 and the auxiliary regeneration column 3 can easily be adjusted so that the hydrostatic head

created in line 23 is sufficient to overcome the higher operating pressure which is maintained in the auxiliary column 3.

70 As the solution coming from draw-off plate 42 enters in column 3, there is no flashing in flash zone 8, because of the higher operating pressure. In the auxiliary column 3, the solution flows to the regeneration zone 17, where
75 it is heated to its boiling temperature under the prevailing higher pressure, and regenerated by the counter-current stream of stripping steam which rises from the base of the column 3.

80 The stripping steam and the desorbed acidic gas flow through line 321 and valve 55 to the main regeneration column 2. The regenerated solution which collects at the base of column 3 flows through line 27 and pressure reduction valve 58 to the expansion
85 vessel 5 which is connected to the suction side of the steam-jet ejector 80, which maintains in said vessel the desired lower pressure.

90 The vapour mixture which is given off by flashing is recompressed in the steam-jet thermocompressor 80 and discharged together with the motive steam through line 35 at the base of the auxiliary column 3.

95 The flashed solution collecting at the base of expansion vessel 5 flows through line 28 and is pumped by pump 82 to an intermediate level of absorber 1 after temperature adjustment in cooler 92. A second fraction of partially regenerated solution which may be
100 withdrawn from the main regenerator 2 at draw-off pan 40 flows through pressure reducing valve 56 and line 24 to the expansion vessel 5 where the flashed vapour mixture is evacuated and recompressed by the
105 steam-jet ejector 80 which discharges through line 35 in column 3. The lean solution which collects at the base of the main regeneration tower 2 flows through line 25 and pressure reducing valve 57 to the expansion vessel 4
110 which is connected to the suction side of steam-jet ejector 88. The flashed vapour mixture is evacuated and recompressed by the steam-jet ejector 88 to the pressure prevailing in column 3 and the mixture of motive
115 steam and recompressed flashed vapour is discharged through line 33 and valve 60 to the auxiliary column 3.

120 The supply of stripping steam in the auxiliary regeneration column 3 is thus made up of the discharge flow of the steam-jet thermocompressors 80 and 88.

125 As shown in Figure 3, it is possible, if the total stripping steam supply exceeds the requirements of column 3, to discharge part or all of the output of the steam-jet thermocompressor 88 to column 2 through valve 61.

130 The lean flashed solution in expansion vessel 4 flows through line 26 to pump 81 and is pumped in split-flow partly through line 261 to the top of the absorber 1 after

cooling in cooler 91, and partly through line 262 to an intermediate level of the absorber 1 after temperature adjustment in cooler 94. At the top of the main regeneration column 2, mixture of desorbed acidic gas and overhead steam is cooled in the overhead cooler-condenser 93. The acidic gas is evacuated from the condensate reflux drum 6 through line 37. The regeneration condensate flows through line 29 and pump 83 to the steam kettle 95.

The application of the invention is not limited to the regeneration of one single scrubbing solution. It is also suited for improving the regeneration of an acidic gas scrubbing system where two different scrubbing solutions are used, in order to achieve a more complete removal of the acidic gas.

Reference is made to Figure 4 of the drawings which shows the flow-sheet of a CO₂ removal unit, in which a first scrubbing solution is used for the removal of the major fraction of the acidic gas, and where a second aqueous solution is used to remove more completely the remaining fraction of acidic gas.

The first scrubbing solution is for instance a potassium carbonate solution containing any of the usual activating additives, such as diethanolamine, while the second solution may be another aqueous solution capable of reducing more effectively the residual partial pressure of the acidic gas in the process gas which must be purified. It is most convenient to use as second scrubbing solution an aqueous solution of diethanolamine or an aqueous solution of the potassium salt of an amino acid such as dimethylglycine.

The process gas stream containing the acidic gas is introduced through line 30 at the base of the absorber 1, where it flows upwards successively through a first absorption zone 14 which is usually maintained at a temperature close to the atmospheric boiling temperature of the solution and a second absorption zone 12 which is maintained at a significantly lower temperature, and leaves the top of the absorber 1, after purification through line 31.

The spent hot scrubbing solution from the first absorption zone 14 collects at the base of the absorber 1 and flows through line 21, in part to the main regeneration column 2 and in part to the auxiliary column 3. The part of the solution which flows to column 2 through the pressure reducing valve 52 and line 22 flashes in flash zone 7 and flows towards the regeneration zone 15 where it is regenerated by the stripping steam rising from the lower part of the column, and collects in draw-off pan 40. This regenerated solution leaves column 2 through line 24 and flows through the pressure reducing valve 56 to the expansion vessel 5 which is main-

tained at the lower pressure by the suction side of the steam-jet ejector 80. 65

The second part of the scrubbing solution from line 21 flows through the pressure reducing valve 53 and line 23 to the auxiliary column 3, where after passing through the flash zone 8 it flows downwards to the regeneration zone 17, counter-currently to the rising flow of stripping steam. The regenerated solution collects at the base of column 3 and flows through line 27 and the pressure reducing valve 58 to the expansion vessel 5, where the steam-jet ejector 80 maintains a lower pressure. 70

The total flow of the first scrubbing solution is thus collected in expansion vessel 5 and flows through line 28 to pump 82 from where it is recycled to the first absorption zone 14 of absorber 1, after temperature adjustment in cooler 92. 75

The second scrubbing solution from the colder second absorption zone 12 leaves the absorber 1 at draw-off pan 43 by line 20, and flows to the heat exchanger 100 where it is reheated by the recycled regenerated solution, and from there through the pressure reducing valve 64 to the lower regeneration section of the main regeneration column 2. After flashing in flash zone 9, it flows to the regeneration zone 16 and to solution reboiler 90 where the stripping steam is produced. The regenerated solution leaves the base of column 2 by line 25 and flows through pressure reducing valve 57 to expansion vessel 4 where a lower pressure is maintained by the suction side of the steam-jet ejector 88. 80

The flashed vapour is evacuated and recompressed by the steam-jet ejector 88 which discharges the mixture of motive steam and recompressed flashed vapour through line 33. 85

Depending on the relative steam requirements of the two regeneration columns 2 and 3, it is possible to supply the stripping steam mixture produced by thermocompressor 88 to either column, or to apportion part of it to each. 90

After flashing, the regenerated solution leaves expansion vessel 4 through line 26 and is pumped by pump 81 to heat exchanger 100 where it preheats the spent solution of line 20 which flows towards the regeneration section. After final temperature adjustment in cooler 91 it is returned to the second absorption zone 12 of the absorber 1. 95

The acidic gas which has been desorbed in the auxiliary column 3 is cooled in cooler 96 and is evacuated from the condensate reflux drum 10 by line 39, whereas the acidic gas desorbed in the main column 2 flows through line 32 to heat exchanger 99 where it preheats the process condensate and then to cooler-condenser 93. It leaves the condensate drum 6 through line 37. The regeneration condensate from the two con- 100

densate drums 10 and 6 flows to pump 83 and after preheating in heat exchanger 99, is sent to vapourizing kettle 95 or may also be returned to the process as reflux through valve 62. The steam produced in kettle 95 is delivered as motive steam in steam collector 34 for the steam-jet ejectors 80 and 88.

It will be noted that in column 2 the stripping steam required in the second regeneration zone 16 is obtained mainly by reboiling the second stripping solution in reboiler 90.

After passing through this lower regeneration zone 16, the mixture of stripping steam and acidic gas which has been desorbed in said zone flows upwards to the first regeneration zone 15 where it is used to regenerate the first scrubbing solution. As in the absorber 1 only a minor fraction of the acidic gas has been absorbed in the second scrubbing solution, the quantity of acidic gas in the stripping steam emerging from the second regeneration zone 16 remains small, and said stripping steam mixture can be used efficiently in the first regeneration zone 15.

The stripping steam required in the auxiliary column 3 is supplied by the discharge of the thermocompressors 80 and 88.

As it was indicated previously, the invention makes it possible to select for each regeneration column the most appropriate pressure level, as the determining factor is the pressure difference which is applied between each of the regeneration columns and the expansion vessels.

As already pointed out, it has been chosen to describe the various single solution flow-sheet modifications with reference to an absorber operating always under the same terminal conditions (temperature of top and bottom of the absorber), according to an "optimal" cycle which incorporates the possibility of temperature adjustment of all the recycled solution streams. This particular feature was chosen only for the purposes of a meaningful comparison of the various modifications.

It is however clear to anyone skilled in the art that the application of the invention is not limited to a so-called "optimal" absorption cycle. The benefits of the invention, which refers specifically to the regeneration of the scrubbing solution, are equally applicable to any other operating conditions of the absorber, whether a temperature adjustment or cooling of the recycled solution streams is applied or not. The invention refers in fact to the method according to which the spent solution stream or streams is or are regenerated before recycling.

The following examples which illustrate the invention refer to the removal of CO₂ from an industrial process gas resulting from a catalytic steam reforming of natural gas, and which after CO₂ removal will be used for the synthesis of ammonia.

It is well known that CO₂ is an acidic gas for which the kinetics of absorption and desorption are relatively slow, as a result of which the regeneration of the spent scrubbing solution is somewhat more difficult than in the case of other acidic gases such as H₂S. It is thus appropriate to illustrate the invention with reference to a CO₂ removal process.

EXAMPLE 1

Reference is again made to Figure 1 of the drawings which shows a cyclic acidic gas removal process, in which an aqueous potassium carbonate scrubbing solution is used. The process gas stream from which CO₂ must be removed is available at a temperature of 250°C, a pressure of 27.5 kg/cm², and a flow rate of 45,700 Nm³/hr (dry basis). It contains 0.44 Nm³ of water vapour per Nm³ of dry gas.

As an inlet temperature to the absorber of 80°C is desired, it is customary to recover the heat contained in the process feed gas by use of one or more heat exchangers, which may if desired be used to supply part or all of the heat requirements of the acidic gas removal system. For instance the hot process gas may be used first to produce low pressure steam at a pressure of 6 kg/cm² in the steam kettle 95 in which it enters through line 47. The process gas leaving kettle 95 through line 36 may then be sent to the solution reboiler, for the production of stripping steam in the main regeneration column 2, after which it may be further cooled to 80°C in a boiler-feed water heater (not shown).

The relative quantities of steam produced in kettle 95 and solution reboiler 90 will depend on the intermediate temperature of the process gas in line 36. For a temperature of 167°C, the steam production in kettle 95 will amount to 5900 kg/hr.

Before entering the absorber 1 which operates at a pressure of 27 kg/cm², the condensate which has been formed upon cooling of the feed gas is removed in a separator drum (not shown).

The process gas which enters the absorber 1 through line 30 contains 17.6% by volume of CO₂. The CO₂ which is absorbed by the scrubbing solution reacts with the potassium carbonate to form potassium bicarbonate according to the reaction:



The scrubbing solution thus contains a mixture of unconverted potassium carbonate and of potassium bicarbonate. The relative proportions of K₂CO₃ and KHCO₃ are indicated by the percent conversion.

The "potassium bicarbonate fraction" or "percent conversion" as used herein means the proportion of original K₂CO₃ expressed in percent which has been converted to potas-

sium bicarbonate by reaction with CO_2 . For example, a solution having a percent conversion of 20% is obtained by the conversion of 20 mols percent of the potassium carbonate content of the original solution to potassium bicarbonate, such that the ratio of the potassium ions present as K_2CO_3 to the potassium ions present as KHCO_3 is 80/20. Since two mols of KHCO_3 are produced for each mol of K_2CO_3 , the mol ratio of $\text{K}_2\text{CO}_3/\text{KHCO}_3$ at 20% conversion is 80/40. As used herein the concentration of the solution, indicated as weight percent potassium carbonate, is referred to the original solution free from bicarbonate, i.e. with zero percent conversion.

The spent scrubbing solution which collects at the base of absorber 1 has a concentration of 29% by weight of K_2CO_3 and contains in addition to the usual corrosion inhibitors 3% by weight of diethanolamine. As the process feed gas flows upwards in the absorber 1 through the absorption zones 14 and 12, it is scrubbed by the hot semi-lean solution which is introduced at an intermediate level and by the cooled lean solution which is introduced at the top of the absorber at a temperature of 80°C. The residual CO_2 content of the cleaned process gas which leaves the absorber at a temperature of 80°C through line 31, is not more than 0.1% by volume.

The lean solution which has a conversion of 22% is introduced in the absorber at a flow rate equivalent to 18% of total flow rate of the spent scrubbing solution. The semi-lean solution which is introduced in the absorber at a temperature of 101°C and a flow rate corresponding to 82% of the total flow rate of the spent scrubbing solution has a conversion of 42.8%.

The spent scrubbing solution which leaves the base of the absorber 1 through line 21 at a flow rate of 378,000 kg/hr has a conversion of 84% and a temperature of 108°C. The load factor of the solution corresponds to 0.45 mole of CO_2 per mole of K_2CO_3 which is equivalent to an overall percent conversion change of 45%.

The spent scrubbing solution flow is divided in two parts. The major fraction of 60% flows through pressure reducing valve 52 and line 22 to the main regeneration column 2 and flash zone 7, where the operating pressure is set at 1.9 kg/cm². The minor fraction of 40% flows through pressure reducing valve 53 and line 23 to the auxiliary regeneration column 3 and flash zone 8 where the pressure is set at 1.3 kg/cm². In the main regeneration column 2, the solution flows downwards, after flashing, towards the regeneration zone 15 where it is scrubbed by the countercurrent flow of stripping steam which desorbs the acidic gas.

At the draw-off pan 40, a fraction of 70%

of the partially regenerated solution flow is withdrawn at a temperature of 120°C and conversion of 46.3%, while the remaining fraction of 30% flows downwards to regeneration zone 16 and the solution reboiler 90. The heat which is supplied in the reboiler 90 by the hot feed gas supplied by line 36, upon cooling from 167°C to 128°C corresponds to a steam production of 10,300 kg/hr which satisfies the stripping steam requirement in the main regeneration column 2. The CO_2 desorption rate in column 2 corresponds to 60% of the total acidic gas removal rate of 8,000 Nm³/hr of CO_2 . The stripping steam requirement in column 2 thus corresponds to approximately 2.14 kg of steam per Nm³ of CO_2 desorbed in the column. Part of the steam which is supplied at the base of the column is condensed in the regeneration zone in order to satisfy the endothermic desorption heat requirement and also in order to maintain the solution at its boiling temperature. The remainder emerges at the top of the column as residual stripping steam.

In view of the fact that the heat content of the steam supply may vary according to the temperature and pressure of the system, it is preferable to refer to the regeneration heat requirement in terms of kilocalories per Nm³ of desorbed CO_2 , which in this case corresponds to 1,165 kcal/Nm³ CO_2 .

The lean solution which collects at the base of column 2 at a temperature of 125°C has a conversion of 22%.

The minor fraction of 40% of the spent scrubbing solution in the auxiliary column 3 must be regenerated to a conversion of 39% in order to achieve the specified load factor or overall percent conversion change of 45%.

The heat requirement for this regeneration duty corresponds to 1,090 kcal/Nm³ of CO_2 desorbed in the auxiliary column 3, or approximately 6,400 kg of steam for the desorption duty of 3,200 Nm³ CO_2 , i.e. 40% of the total acidic gas removal duty, which is to be performed in said auxiliary column.

The semi-lean solution from draw-off pan 40 flows at the rate of 156,000 kg/hr, i.e. 42% of the total solution flow, through the pressure reducing valve 56 and line 24 to the flash zone 9 at the base of the auxiliary column 3 where the operating pressure of 1.4 kg/cm² is slightly higher than at the top of the column. By flashing the solution produces 2,000 kg/hr of steam thereby cooling to 110°C. In addition to the steam, a small amount of CO_2 is released. Due to the slow kinetics of the desorption process the quantity of CO_2 which is given off by the flashing process is sufficiently small so that it does not interfere in a significant way with the regeneration process of the auxiliary column, and there is therefore no need, for the understanding of this example, to take it in account.

The lean solution which leaves the base of column 2 through line 25 and pressure reducing valve 57, flows to expansion vessel 4 where essentially the same pressure of 1.4 kg/cm² prevails as in flash zone 9 at the base of column 3. Steam is generated by flashing at the rate of 1,000 kg/hr thereby cooling the solution to 113°C.

The ratio of CO₂ to steam in the vapour mixture generated by flashing is in this case even smaller than in the case of the flashing of the semi-lean solution, and the small amount of CO₂ thus liberated can be neglected for the purpose of the present example.

The total amount of flashed steam produced both in flash zone 9 and in expansion vessel 4 corresponds to 3,000 kg/hr or approximately 0.94 kg of steam per Nm³ of CO₂ desorbed in column 3. As the heat thus supplied only amounts to 506 kcal/Nm³ CO₂, an additional heat input of 581 kcal/Nm³ CO₂ will be needed in order to satisfy the regeneration heat requirement.

According to the present invention, this additional heat supply is obtained through use of thermocompressor 80 which is operated with motive steam from line 34 produced at a pressure of 6 kg/cm² in kettle 95 by vapourization of process condensate.

The minor fraction of 40% of scrubbing solution which is regenerated in regeneration zone 17 of the auxiliary column 3 is collected on draw-off pan 41 at a temperature of 111°C and a percent conversion of 39%.

It flows through line 271 and pressure reducing valve 58 to the expansion vessel 5, together with the semi-lean solution leaving the bottom of column 3 through line 27. This combined flow of 305,000 kg/hr produces upon flashing to the reduced pressure of 1.2 kg/cm² which is maintained by the suction side of the steam-jet ejector 80 in vessel 5, 1,400 kg/hr of steam thereby cooling the combined solution flow which has now a conversion of 42.8% to 107°C.

The motive steam at a pressure of 6 kg/cm² used by the steam-jet ejector 80, in order to recompress the flashed steam to the pressure of 1.4 kg/cm² which exists at the base of column 3, amounts to 2,000 kg/hr.

The combined flow of 3,400 kg/hr of recompressed flashed steam and motive steam is discharged through line 35 at the base of column 3. As explained earlier, the small amount of CO₂ liberated by flashing may also be neglected for the purpose of the present explanation; it may be noted that its relative importance is further decreased by the diluting effect of the motive steam.

By use of this steam-jet thermocompressor 80, which uses 1 kg of motive steam at a pressure of 6 kg/cm² to recompress 0.7 kg of flashed steam, thus producing 1.7 kg of stripping steam, the external heat supply

which is required in order to satisfy the regeneration heat demand of column 3 amounts to 59% only of the additional heat demand of 581 kcal/Nm³ CO₂.

The external heat input thus amounts to 341 kcal/Nm³ CO₂ of motive steam, the balance being obtained from the solution itself by flashing and recompression.

The overall regeneration heat requirement of the cyclic process is defined by combining the external heat supply to the main regeneration tower 2 through the solution reboiler 90, and the external heat supply to the auxiliary column 3 under form of motive steam.

To the external regeneration heat supply in the solution reboiler of 1,165 kcal/Nm³ CO₂ for the desorption of 4,800 Nm³/hr of CO₂ in the main column 2, must be added the supply of 341 kcal/Nm³ CO₂ for the 3,200 Nm³/hr of CO₂ desorbed in the auxiliary column 3. This corresponds to an overall heat requirement of 836 kcal/Nm³ of CO₂ calculated on the basis of the total acidic gas removal duty of 8,000 Nm³/hr of CO₂.

The semi-lean solution which leaves expansion vessel 5 through line 28 is pumped by pump 82 through line 281 to the intermediate level of absorber 1, after having been cooled to 101°C in cooler 92.

The lean solution from expansion vessel 4 flows to heat exchanger 99, where it preheats the process condensate to 100°C. It is pumped by pump 81 through line 261 to the cooler 91 where it is cooled to 80°C and from there to the top of absorber 1.

The desorbed CO₂ and the residual stripping steam from the auxiliary column 3 flow through line 32 and cooler condenser 93 to the condensate separator drum 6 which are all maintained at essentially the same pressure level of 1.3 kg/cm². The desorbed CO₂ from the main column 2, which represents 60% of the total CO₂ or 4,800 Nm³/hr flows together with the residual stripping steam through the pressure reducing valve 54 into line 32 where it mixes with the flow coming from the auxiliary regeneration column 3. The combined stream of 8,000 Nm³/hr of desorbed CO₂ is evacuated at a temperature of 60°C from the unit through line 37, while the overhead process condensate leaves vessel 6 through line 29 at a rate of 7,400 kg/hr. The process condensate is pumped by pump 83 to the exchanger 99, where it is heated from 60°C to 100°C, after which part of it is sent to the kettle 95 to produce the required amount of motive steam for the steam-jet ejector 80 and the remainder is returned as process reflux to the main regeneration column, through valve 62.

It is quite obvious that, if desired, the major flow of 4,800 Nm³/hr of CO₂ which has been desorbed at a pressure of 1.9 kg/cm² can easily be discharged from the acid gas

removal unit at this higher pressure by way of a suitable piping set up such as is illustrated in figure 2.

This example can easily be compared with the prior art practice of regeneration.

It was indicated that the stripping steam demand of the auxiliary column 3 amounted to approximately 2 kg of steam per Nm³ of CO₂ desorbed in the column, or a total of 6400 kg/hr, whereas the steam produced upon flashing in flash zone 9 and expansion vessel 4 amounted only to 3000 kg/hr. The additional requirement of 3,400 kg/hr could of course be satisfied by way of a solution reboiler.

Instead of sending the semi-lean solution streams from column 3 to the expansion vessel 5, they could be circulated through a solution reboiler, where an additional heat input of 581 kcal/Nm³ CO₂ would produce the required amount of additional stripping steam.

As compared with the method of the invention, the use of the reboiler requires a greater heat input, together with the increased solution temperature which prevails in a reboiler, as against a solution cooling in expansion vessel 5.

According to another method of the prior art, it is possible to operate the cyclic process of Figure 1 without an additional reboiler or without use of the present invention, if the combined semi-lean solution flow from column 3 is sent directly through line 229 to pump 82. The operating conditions must then be modified in order to balance the steam demand and supply in column 3. If the same regeneration pressure levels of 1.9 kg/cm² and 1.3 kg/cm² are maintained, the major fraction of solution regenerated in column 2 must be increased to 76.2% and the solution

flow to the auxiliary column decreased to 23.8%.

The overall heat input to the system, which is now limited to the single solution reboiler 90, has increased in the ratio of solution flow to the main column and corresponds to 888 kcal/Nm³ CO₂. It is however possible to increase the thermal efficiency of this method of operation by increasing the pressure on the main regeneration column in order to decrease as much as possible the heat loss resulting from the flashing in flash zone 7. If the pressure is increased to a value which is equivalent to the vapour pressure of the spent scrubbing solution, essentially no flashing would occur in flash zone 7. In the case of this example, a pressure of 3 kg/cm² would be required. There is a corresponding increase of the heat requirement in column 2, which now amounts to 1,435 kcal/Nm³ of desorbed CO₂, but the overall heat requirement is decreased to 865 kcal/Nm³ as only 60.3% of the spent scrubbing solution must be regenerated in column 2. The temperature of the solution at the base of column 2 has however increased to 139°C.

The data concerning Example 1 and the comparison with prior art practice are summarized in Table 1. Experiment No. 1 refers to the prior art practice as shown in Figure 1 of the drawings, where the combined semi-lean solution streams leaving column 3 through lines 27 and 271 flow through line 229 directly to pump 82 and are recycled to the absorber 1 without further treatment. Experiment No. 2 refers to the prior art practice where an additional reboiler is used for the auxiliary column 3. Experiment No. 3 refers to Example 1 of the present invention.

TABLE 1

Experiment No.	1	2	3
1 Pressure at top of column 2/3 kg/cm ²	3/1.3	1.9/1.3	1.9/1.3
2 Temperature at base of column 2/3 °C	139/111	125/111	125/111
3 Total external heat input kcal/Nm ³ CO ₂	865	931	836

EXAMPLE 2

Reference is made to Figure 2 of the drawings. The same process feed gas and scrubbing solution are used as in Example 1. The same amount of CO₂ is removed from the feed gas in absorber 1 where the operating conditions are the same as in Example 1, except for the following modification. The lean solution flow rate has been decreased to 15% of the equivalent total flow rate of spent scrubbing solution. Its temperature and conversion are maintained at 80°C and 22%. The semi-lean solution flow rate is now equivalent to 85% of the total flow rate of spent scrubbing solution. Its conversion is now 42% and the temperature 100°C. The spent scrubbing solution collects at the base of the absorber at the same temperature and con-

version of 108°C and 84% and its flow rate is unchanged at 378,000 kg/hr. The load factor of 0.45 mole CO₂ per mole K₂CO₃ is unchanged.

The spent scrubbing solution flowing through line 21 is split in two equal fractions of 50%, which are regenerated separately in the main regeneration column 2 and in the auxiliary column 3. The operating pressure of the main regeneration column is set at 1.3 kg/cm² while the operating pressure of the auxiliary column is adjusted to the slightly higher pressure of 1.5 kg/cm². The solution flowing through valve 52 and line 22 to the main regeneration column 2 flashes in flash zone 7 and flows to the regeneration zone 15. A fraction of 70% of the partially regenerated semi-lean solution with percent conversion

of 46.3% is withdrawn from draw-off plate 40, at a temperature of 109°C and the remaining fraction of 30% equivalent to 15% of the total solution flow flows to regeneration zone 16 of the main regeneration column 2 and the solution reboiler 90.

After reboiling, the lean solution with a conversion of 22% and a temperature of 115°C collects at the bottom of the column 2.

The second fraction of 50% of the spent scrubbing solution which flows through pressure reducing valve 53 and line 23 to the auxiliary regeneration column 3 flashes in flash zone 8 and flows towards the regeneration zone 17.

The semi-lean solution from column 3 must be regenerated to a conversion of 39% in order to achieve the specified overall percent conversion change of 45%. This requires a regeneration heat supply of 1,115 kJ/Nm³ of CO₂ desorbed in the column 3, which is equivalent to approximately 8,200 kg/hr of stripping steam. This stripping steam will be supplied by the steam-jet thermocompressor 80 which discharges through line 35 into column 3.

The suction side of the steam-jet ejector 80 is connected to the expansion vessel 5 where a pressure of 1.1 kg/cm² is maintained. The semi-lean solution flow of 130,000 kg/hr which is withdrawn from draw-off pan 40 flows through line 24 and pressure reducing valve 56 to expansion vessel 5 where 900 kg/hr of steam are generated by flashing, thereby cooling the solution to 103°C. The semi-lean solution which collects at base of auxiliary column 3 at a temperature of 115°C and conversion of 39% flows at the rate of 185,000 kg/hr through line 27 and pressure reducing valve 58 to expansion vessel 5 where 2,200 kg/hr of steam are produced by flashing thereby cooling the solution to 105°C. The motive steam of 6 kg/cm² pressure produced in kettle 95 is supplied to the steam-jet ejector 80 by line 34 at a flow rate of 5,100 kg/hr and the recompressed mixture of flashed vapour and motive steam is discharged by the thermocompressor 80 through line 35 at the pressure of 1.64 kg/cm² which prevails at the base of column 3, at a flow rate of 8,200 kg/hr.

The regeneration heat demand of column 3 has thus been satisfied to an extent of 62% by an external supply of motive steam and to an extent of 38% by heat extracted from the solution itself by the thermocompression action of a steam-jet ejector.

The external heat supply to auxiliary column 3 therefore amounts to 62% of the regeneration heat requirement of 1,115 kJ/Nm³ CO₂, i.e. 692 kJ per Nm³ of CO₂ desorbed in column 3.

As the heat demand of auxiliary column 3 is fully satisfied by the steam supplied by

the steam-jet thermocompressor 80, it is possible to use the additional steam supply discharged by thermocompressor 88 as stripping steam in the main regeneration column 2.

The lean solution at the base of column 2 flows through line 25 and pressure reducing valve 57 to the expansion vessel 4, which is connected to the suction side of the steam-jet ejector 88, which maintains in said vessel a reduced pressure of 1 kg/cm². Steam is generated by flashing at a rate of 600 kg/hr thereby cooling the solution to 106°C.

Motive steam produced at a pressure of 6 kg/cm² in kettle 95 is supplied through line 34 at the rate of 1000 kg/hr to the steam-jet ejector 88 and the recompressed mixture of flashed vapour and motive steam is discharged through valve 61 at the pressure of 1.5 kg/cm² which prevails at the base of column 2.

When it is thus chosen to return the total output of 1,600 kg/hr of steam from thermocompressor 88 to column 2, valve 60 of line 33 is closed. The regeneration heat requirement in main column 2 amounts to 1,044 kJ per Nm³ of CO₂ desorbed in the column, which is equivalent to a stripping steam supply of 7,700 kg/hr. As the steam-jet thermocompressor 88 already supplies approximately 20% of the total demand of column 2, the heat supply by solution reboiler 90 will be reduced in the same proportion and will amount to 838 kJ per Nm³ of CO₂ desorbed in column 2. The total external regeneration heat supply to column 2 will thus be made up of 838 kJ/Nm³ CO₂ from reboiler 90 and 62 kJ/Nm³ CO₂ under form of motive steam, or a total of 900 kJ/Nm³ CO₂ desorbed in column 2. As the total regeneration duty is divided in equal parts between the main column 2 and the auxiliary column 3, where the respective external heat requirements are 900 kJ/Nm³ CO₂ and 692 kJ/Nm³ CO₂, the combined overall external heat supply will amount to 796 kJ per Nm³ of CO₂ removed from the spent scrubbing solution.

After flashing in vessel 4, the lean solution flows through exchanger 99 where the regeneration condensate is preheated to 100°C and is pumped by pump 81 through line 261 to the cooler 91 where it is cooled to 80°C before entering the top of absorber 1. The combined semi-lean solution flow from expansion vessel 5 with a percent conversion of 42% flows through line 28 to pump 82 from where it is recycled to absorber 1 after cooling in cooler 92 to a temperature of 100°C. The overhead mixture of steam and CO₂ from the main column 2 flows through line 32 to cooler 93 and reflux drum 6 where the condensate is separated while the cooled CO₂ is discharged at a pressure of approximately 1.3 kg/cm² and a temperature of 60°C through line 37, at a flow rate of 4,000

Nm³/hr. The other fraction of acidic gas which is desorbed in auxiliary column 3 is evacuated at the higher pressure of 1.5 kg/cm² through line 39 after having been cooled to 60°C in cooler 96. The condensate from reflux drum 10 flows through pressure reducing valve 50 into line 29 through which it flows jointly with the condensate from reflux drum 6 at a combined flow rate of 8,700 kg/hr and temperature of 60°C to pump 83. From there it is pumped to heat exchanger 99 where it is reheated to 100°C after which it flows at a rate of 6,100 kg/hr to vaporizing kettle 95 where the corresponding amount of motive steam is produced at a pressure of 6 kg/cm² by the hot feed gas which is supplied by line 47 and leaves the kettle 95 by line 36 at a temperature of 164°C. The remainder of the regeneration condensate is returned as reflux to the main column 2 through valve 62.

It will be noted that, according to this mode of operation, the solution reboiler is operated at the lower regeneration pressure, and that the reboiler heat duty has been reduced to 419 kcal/Nm³ CO₂.

EXAMPLE 3

Reference is made to Figure 3 of the drawings. The scrubbing solution composition and load factor is the same as in Example 2, as well as the percent conversion, temperature and flow rate of the streams of lean and semi-lean solutions entering the absorber and of spent scrubbing solution leaving the absorber. The recycle solution line 262 and cooler 94 are not used. The same feed gas is used as in Example 2.

The spent scrubbing solution, with a conversion of 84%, and temperature of 108°C, flows through line 21 and pressure reducing valve 52 to the flash zone 7 of the main regeneration column 2 where the pressure is set at 1.5 kg/cm².

At draw-off pan 42, below regeneration zone 18, where the pressure is approximately 1.6 kg/cm², a fraction of 50% of the partially regenerated solution is withdrawn at a temperature of 110°C and conversion of 55%. The remaining 50% fraction flows down to regeneration zone 15 and draw-off pan 40 where a fraction of 70% of the remaining semi-lean solution, i.e. 35% of the total flow entering the top of the column is withdrawn at a temperature of 113°C and conversion of 46.3%. The remaining fraction of 15% flows to regeneration zone 16, and after passing through solution reboiler 90, collects at the bottom of the column at a temperature of 118°C and conversion of 22%.

The partially regenerated solution withdrawn from draw-off pan 42 flows through line 23 to the auxiliary column 3, which is situated at a lower elevation, which is such that the hydrostatic pressure of the solution

in line 23 is sufficient to overcome the difference in pressure between columns 2 and 3. In this example, the difference in height between draw-off pan 42 and pressure reducing valve 53 is 2 m., which corresponds to a hydrostatic head of approximately 0.25 kg/cm², and the pressure in auxiliary column 3 is 1.7 kg/cm², i.e. 0.1 kg/cm² more than at draw-off pan 42.

Because of the increased pressure in flash zone 8, there is no flashing of the solution, which after regeneration in regeneration zone 17 collects at the base of the column.

In order to satisfy the constant load factor of the solution of 45% overall percent conversion change, the solution stream in column 3 must be regenerated to a conversion of 39%. For this regeneration duty, an increased amount of stripping steam is however required, so that the partial pressure of acidic gas in the vapour mixture which emerges at the top of regeneration zone 17 and is subsequently returned to the main column 2, by line 321, is maintained at a value which is sufficiently low to ensure that said vapour mixture will still have an adequate stripping efficiency when transferred in the main column 2.

As a conversion change of only 16% is performed on a fraction of 50% of the total solution, this will correspond to a desorption of 1,420 Nm³/hr of CO₂. The stripping steam which must be supplied to column 3 corresponds to 5.7 kg steam per Nm³ of CO₂ or approximately 8,100 kg/hr.

According to the invention, this stripping steam demand is satisfied by the discharge flow of the steam-jet ejector 80. The semi-lean solution regenerated in column 3 to a conversion of 39% and a temperature of 116°C flows through line 27 and pressure reducing valve 58 to the expansion vessel 5, in which the suction side of the steam-jet ejector 80 maintains a reduced pressure of 1.2 kg/cm². The flashing induced by this decrease in pressure generates 2,100 kg/hr of steam thereby cooling the solution to 105°C. The semi-lean solution which is withdrawn from the main column 2 at draw-off pan 40, flows through line 24 and pressure reducing valve 56 to expansion vessel 5, where the flashing generates 1,200 kg/hr of steam; thereby cooling the solution to 105°C.

The motive steam, which is supplied from kettle 95 through line 34, at a pressure of 6 kg/cm² and a flow rate of 4,800 kg/hr recompresses the combined flashed vapours to a pressure of 1.8 kg/cm² which corresponds to the pressure at the base of column 3, where it discharges the mixture of motive steam and recompressed flashed vapour through line 35. The regeneration heat requirement of the auxiliary column 3 has thus been satisfied by an external heat supply under form of live

motive steam, equivalent to 1,851 kcal per Nm³ of CO₂ desorbed in said column.

The regeneration heat demand in the main column 2 is now satisfied in part by the stripping steam supplied by the auxiliary column 3 through line 321, and the additional requirement corresponds to a stripping steam supply of 7,800 kg/hr at the base of the column, of which part will be supplied by the thermocompressor 88 through line 33 and valve 61, and part through the solution reboiler 90.

The lean solution at the bottom of the main column 2 flows through line 25 and pressure reducing valve 57 to the expansion vessel 4 where the suction side of the steam-jet ejector 88 maintains a reduced pressure of 1.15 kg/cm². The induced flashing generates 700 kg/hr of steam thereby cooling the solution to 108°C. The motive steam which is supplied at a pressure of 6 kg/cm² by kettle 95 through line 34, at a flow rate of 1,000 kg/hr recompresses the flashed vapour to the pressure of 1.7 kg/cm² which prevails at the base of column 2, where it discharges the mixture of motive steam and recompressed flashed vapour through line 33 and valve 61 at a flow rate of 1,700 kg/hr. During this operation valve 60 is closed so that no discharge steam from thermocompressor 88 is allowed to enter the auxiliary column 3.

The balance of 6,100 kg/hr of stripping steam supply is obtained from reboiler 90 which is heated by the hot synthesis feed gas.

If the heat supply to the reboiler is related to the overall acidic gas removal duty of 8,000 Nm³/hr of CO₂, it corresponds to 417 kcal/Nm³ CO₂ desorbed (overall). The motive steam used in the steam-jet thermocompressors 80 and 88 at a combined flow rate of 5,800 kg/hr corresponds to a heat supply, which is introduced in the regeneration system through vapourizing kettle 95, of 397 kcal/Nm³ CO₂ desorbed overall.

The combined regeneration heat demand of this regeneration method thus amounts to 814 kcal/Nm³ CO₂ removed. All the CO₂ which has been desorbed in the regeneration system is collected at the top of the main regeneration column 2 at a pressure of 1.5 kg/cm², from where it flows through line 32 to cooler 93 and discharge line 37 where it leaves the unit at a temperature of 60°C.

The overhead regeneration condensate which collects in reflux drum 6 at a rate of 8,000 kg/hr flows through line 29 and is pumped by pump 83 to heat exchanger 99 where it is preheated to 100°C, after which part of it is supplied to kettle 95 for the production of motive steam and the remainder is returned to the main column 2 as process reflux.

According to Figure 3, it is shown that this reflux is returned through valve 62 at the top of column 2. It is however well known that it is equally satisfactory to return said

reflux either to an intermediate level, or at the base of the regeneration column, or even directly into the side stream flowing into the solution reboiler 90.

According to another well known practice, it is also possible to return the cold process condensate into the system, without preheating in exchanger 99, directly into line 26.

These various modifications are equally applicable within the scope of the present invention.

After passing through heat exchanger 99, the lean solution is recycled by pump 81 and line 261 to the top of the absorber 1 after cooling to 80°C in cooler 91.

The semi-lean solution from vessel 5 with conversion of 42%, is recycled to the absorber 1 by pump 82 through line 281 and cooler 92 where its temperature is adjusted to 100°C.

EXAMPLE 4

Reference is made to Figure 4 of the drawings, which shows a dual solution scrubbing system in which an aqueous potassium carbonate solution is used for the bulk removal of CO₂ in the first absorption zone 14 of absorber 1, and a 20% by weight aqueous diethanolamine (DEA) solution is used in the second absorption zone 12 of absorber 1 for final cleaning of the gas. The same process feed gas is used as in Example 1. After passing through the kettle 95, where upon cooling to 163°C it vaporizes 7,200 kg/hr of motive steam at a pressure of 6 kg/cm², solution reboiler 90 and another exchanger (not shown), the process feed gas enters absorber 1 which operates at a pressure of 27 kg/cm² through line 30 at a flow rate of 45,700 Nm³/hr (dry basis). After passing through the first absorption zone 14, where the CO₂ content is reduced to 1.6% by volume, the gas is further cleaned in the second absorption zone after which it leaves the absorber through line 31 at a temperature of 45°C and with a residual CO₂ content of not more than 0.01% by volume. The regenerated potassium carbonate solution is recycled to the absorber 1 through line 281 at a temperature of 101°C and a conversion of 41.5%. The spent potassium carbonate solution which collects at the base of absorber 1 has a concentration of 29% by weight of potassium carbonate, and contains, in addition to a small amount of corrosion inhibitor, 3% by weight of diethanolamine. It leaves through line 21, at a temperature of 110°C and a conversion of 84% at a flow rate of 371,000 kg/hr. The lean DEA (diethanolamine) solution is recycled to the top of the absorber 1 through line 261 at a temperature of 45°C and an acidic gas load of 0.08 mole CO₂ per mole of DEA. The spent DEA solution is withdrawn from absorber 1 at draw-off pan 43, at a temperature of 77°C and an acidic gas load of 0.38 mole per mole of DEA.

It flows through line 20, at a flow rate of 47,000 kg/hr to heat exchanger 100, where it is preheated to 95°C, and after pressure reduction in valve 64, enters the lower regeneration section of the main regeneration column 2. After flashing in flash zone 9, where the operating pressure is 1.45 kg/cm², the solution flows to regeneration zone 16 and to the solution reboiler 90. The heat supplied by the hot process feed gas from line 36, generates 4,800 kg/hr of steam. Part of this stripping steam is used in regeneration zone 16 to supply the heat for the endothermic desorption process and to heat the solution to its boiling temperature, while the remainder, together with the desorbed CO₂, flows upwards to the upper regeneration section where the potassium carbonate solution is regenerated.

The spent potassium carbonate solution leaving absorber 1 is divided in two parts. A fraction of 30% is sent through pressure reducing valve 52 and line 22 to the main regeneration column and flash zone 7 where the pressure is set at 1.3 kg/cm².

After flashing the solution flows to regeneration zone 15 and to draw-off pan 40 from where it is withdrawn at a temperature of 109°C and a conversion of 46%. The regeneration steam demand in zone 15 amounts to 4,600 kg/hr of which only 2,550 kg/hr are rising from the lower regeneration section. In order to satisfy the additional requirement, the lean DEA solution which leaves the bottom of the main column 2 through line 25 at a temperature of 112°C, is led to expansion vessel 4, where the suction side of the steam ejector 88 maintains a reduced pressure of 1.05 kg/cm². The flashing of this solution upon pressure reduction through valve 57, generates 800 kg/hr of steam, which are recompressed by the thermocompressor 88, which is operated with 1,250 kg/hr of motive steam at a pressure of 6 kg/cm², supplied by kettle 95 through the steam collector 34. The thermocompressor 88 discharges the mixture of recompressed flashed vapour and motive steam through line 33 and valve 611 to the flash zone 9 at a flow rate of 2,050 kg/hr. During this operation, the valves 61 and 60 are maintained closed.

The second fraction of 70% of the spent potassium carbonate solution flows through the pressure reducing valve 53 and line 23 to the auxiliary regeneration column 3 and flash zone 8 where the pressure is set at 1.4 kg/cm².

In order to satisfy the specified overall conversion change of 42.5%, the regeneration in the auxiliary column must be carried through to a conversion of 39%. The corresponding regeneration steam demand of 10,100 kg/hr is supplied by the steam-jet ejector 80.

The regenerated solution from draw-off pan 40 flows through line 24 and pressure

reducing valve 56 to the expansion vessel 5 where the suction side of the steam-jet ejector 80 maintains a reducing pressure of 1 kg/cm². Upon flashing steam is generated at the rate of 1,000 kg/hr thereby cooling the solution to 101°C. The solution which has been regenerated in the auxiliary column 3, flows by line 27 at a temperature of 112°C through the pressure reducing valve 58, to the expansion vessel 5, where a steam flow of 3,100 kg/hr is generated, thereby cooling the solution to 101°C.

A supply of 6,000 kg/hr of motive steam is supplied by line 34 to the steam-jet thermocompressor 80, which recompresses the flashed vapour to the pressure of 1.54 kg/cm² which prevails at base of column 3, and discharges the mixture of motive steam and recompressed vapour at a flow rate of 10,100 kg/hr through line 35.

The combined flow of regenerated potassium carbonate solution, with a conversion of 41.5%, flows at a temperature of 101°C through line 28 to pump 82 and is recycled to absorber 1 through line 281.

No cooling is required in cooler 92. The lean DEA solution which leaves expansion vessel 4 at a temperature of 102°C flows through line 26 to pump 81, and is pumped to heat exchanger 100 where it preheats the spent scrubbing solution. From there it flows to cooler 91 where its temperature is adjusted to 45°C. The flow of desorbed CO₂ from the auxiliary column 3 is cooled in cooler 96 and leaves the regeneration unit at a pressure of 1.5 kg/cm² by line 39. The CO₂, which has been desorbed in the main column flows through line 32 to the heat exchanger 99 where it preheats the process condensate and then to cooler 93 before leaving the unit by line 37 at a pressure of 1.3 kg/cm².

The regeneration condensate which collects in the condensate drums 10 and 6 flows through line 29 at a combined flow rate of 9,600 kg/hr to pump 83, from where it is pumped to heat exchanger 99. This heat exchanger could equally well be heated by the overhead CO₂ and vapour stream from column 3. After preheating to 94°C, the condensate flows to kettle 95 at a rate of 7,250 kg/hr and the remainder is returned through valve 62 to the lower section of the main regeneration column as process reflux.

The total external heat requirement of the regeneration system is satisfied by the solution reboiler steam production of 4,800 kg/hr and the motive steam supply of 7,250 kg/hr to the thermocompressors 80 and 88. A total quantity of 8,040 Nm³/hr of CO₂ has been removed from the process gas, and the specific heat consumption amounts to 825 kJ/Nm³ of CO₂ removed.

In this example, the two separate absorption zones 12 and 14 were combined in the same absorption column. It is obvious that

the regeneration method according to the present invention applies equally well if the separate absorption zones are arranged in different columns.

5 It is to be noted that the flow diagrams of Figures 1 to 4 do not include all the necessary devices needed for the operation of an industrial plant. However, the additional devices needed for that purpose will be obvious for those skilled in the art.

10 It is also to be noted that the invention is not limited to the four above described embodiments and that the flow diagrams may be modified within the scope of this invention.

WHAT WE CLAIM IS:—

1. A process for the regeneration of regenerable aqueous scrubbing solutions which are used for the bulk removal of acidic gases, by absorption from gas mixtures containing these acidic gases, in a cyclic process in which said aqueous scrubbing solutions are regenerated in a regeneration system by steam stripping before being recycled to the absorption stage, a part of the scrubbing solutions being regenerated in a main regeneration section of the regeneration system, whereas the other part of the scrubbing solutions is regenerated in an auxiliary regeneration section of the regeneration system, said process comprising the steps of selecting the pressure in the auxiliary regeneration section independently from the pressure in the main regeneration section and obtaining at least a part of the stripping steam required in the auxiliary regeneration section through flashing of the scrubbing solution in a reduced pressure zone connected to the suction side of a steam-jet thermocompressor, the mixture of thermocompressor motive steam and recompressed flashed vapour being directly discharged into said auxiliary regeneration section.

2. A process according to claim 1, in which all the stripping steam required in the auxiliary regeneration section is obtained by flashing of the scrubbing solution in at least one reduced pressure zone connected to the suction side of a steam-jet thermocompressor, the mixture of thermocompressor motive steam and recompressed flashed vapour being directly introduced into said auxiliary regeneration section.

3. A process according to anyone of claims 1 and 2, in which the pressure in the auxiliary regeneration section is lower than the pressure in the main regeneration section.

4. A process as claimed in claim 3 wherein a part of the stripping steam required in said auxiliary regeneration section is obtained by direct flashing of the scrubbing solution upon pressure reduction from the higher pressure of the main regeneration section to the lower pressure of the auxiliary regeneration section.

5. A process according to either of claims 1 and 2, in which the pressure in the auxiliary regeneration section is equal to or higher than the pressure in the main regeneration section.

6. A process according to any one of the preceding claims, in which the flashing to a reduced pressure, induced by the operation of the steam-jet ejector, is performed on a partially or totally regenerated scrubbing solution obtained from either the main regeneration section or the auxiliary regeneration section.

7. A process according to any one of the preceding claims, in which a part of the mixture of thermocompressor motive steam and recompressed flashed vapour is discharged in the main regeneration section.

8. A process according to any one of the preceding claims, in which the spent scrubbing solution is separated into two fractions which are introduced directly into the main regeneration section and into the auxiliary regeneration section respectively.

9. A process according to any one of claims 1 to 7, in which all the spent scrubbing solution is introduced into the main regeneration section and a side stream of partially regenerated scrubbing solution is withdrawn from the main regeneration section and introduced into the auxiliary regeneration section.

10. A process according to any one of the preceding claims, in which two different scrubbing solutions which have been used in independent gas absorption sections are regenerated in a single regeneration system comprising one main regeneration section and one auxiliary regeneration section, the stripping steam which has been used for the regeneration of one of the spent scrubbing solutions being reused as stripping steam for the regeneration of at least part of the other spent scrubbing solution.

11. A process according to claim 10, in which the stripping steam used in the auxiliary regeneration section is obtained by flashing of one or both of the different scrubbing solutions in one or more corresponding reduced pressure zones connected to the suction side of one or more steam-jet thermocompressors, each of the mixtures of thermocompressor motive steam and recompressed flashed vapour being directly discharged into said auxiliary regeneration section.

12. A process according to claim 10, in which the stripping steam used in the auxiliary regeneration section is obtained by flashing of one or both of the different scrubbing solutions in one or more corresponding reduced pressure zones connected to the suction side of one or more steam-jet thermocompressors, any or each of the mixtures of thermocompressor motive steam and recompressed flashed vapour being discharged partly or wholly into said auxiliary regenera-

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tion section or partly or wholly into the main regeneration section.

5 13. A process as claimed in claim 1 substantially as herein described in any one of Examples 1 to 4.

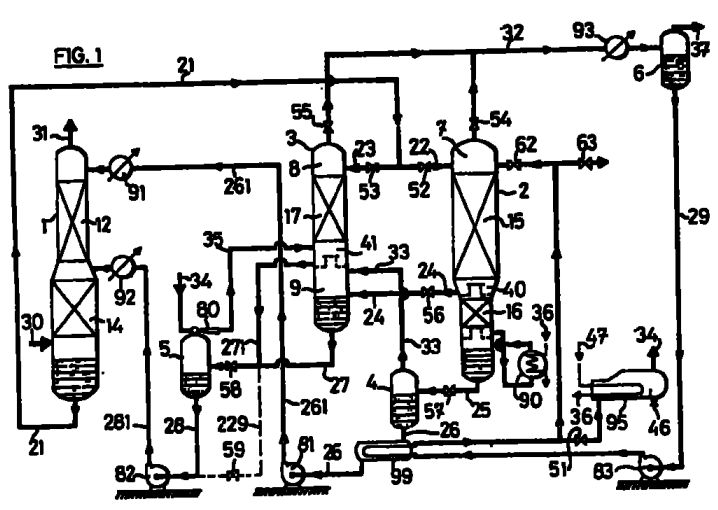
14. A process as claimed in claim 1 substantially as herein described with reference to Figures 1 to 4 of the attached drawings.

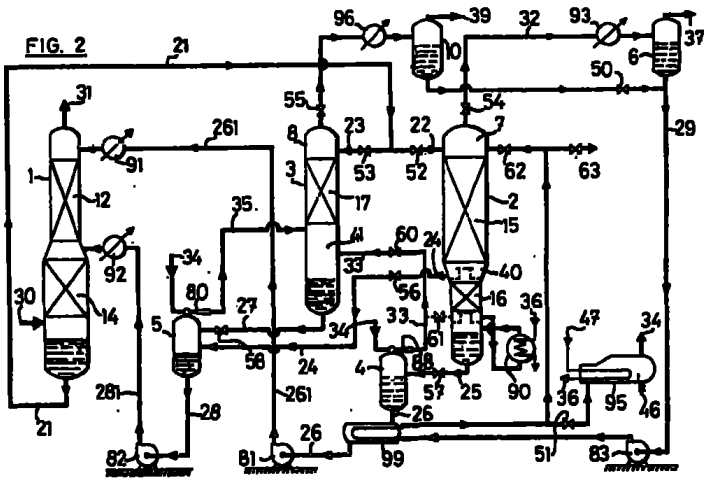
15. A process as claimed in claim 1 substantially as herein described.

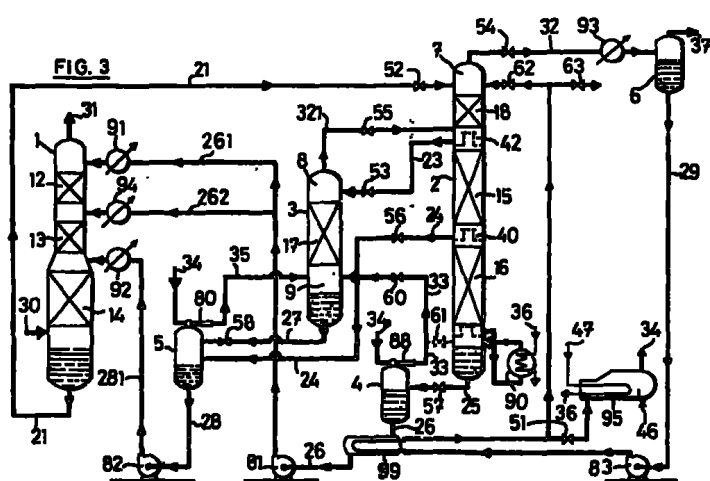
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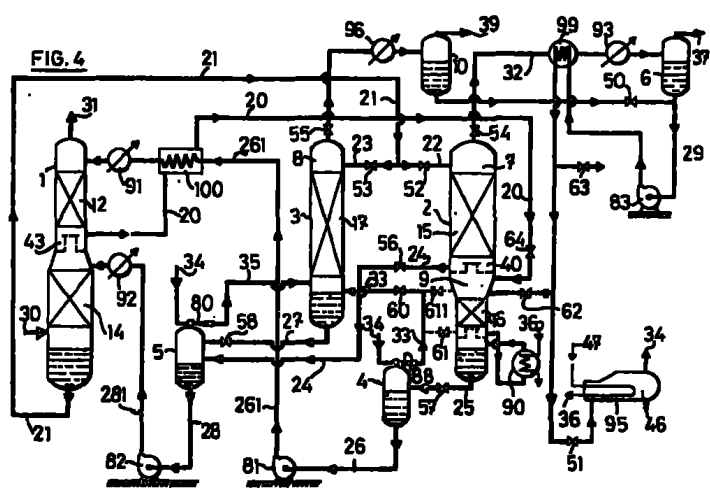
For the Applicants,
FRANK B. DEHN & CO.,
Imperial House,
15-19 Kingsway,
London WC2B 6UZ.

Printed for Her Majesty's Stationery Office, by the Courier Press, Leamington Spa, 1977
Published by The Patent Office, 25 Southampton Buildings, London, WC2A 1AY, from which copies may be obtained.









[54] **PROCESS OF REGENERATING ALKALI CARBONATE SOLUTIONS FORMED BY THE DESULFURIZATION OF HOT GASES**

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[21] **Appl. No.:** 922,396

[22] **Filed:** Jul. 6, 1978

[30] **Foreign Application Priority Data**

Sep. 29, 1977 [DE] Fed. Rep. of Germany 2743753

[51] **Int. Cl.³** B01D 53/34

[52] **U.S. Cl.** 423/223; 423/232; 423/233

[58] **Field of Search** 423/210, 220, 223, 232, 423/233, 234

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[57] **ABSTRACT**

A process for the selective removal of hydrogen sulfide from a gas containing hydrogen sulfide and carbon dioxide wherein said gas is contacted at superatmospheric pressure with an alkali metal carbonate solution, the resultant solution which contains hydrosulfide and bicarbonate is subjected to pressure reduction, the total alkali metal bicarbonate content of the resultant solution is adjusted so that it is at least 55% of the total alkali content of the solution and the said solution is then stripped and recycled to the hydrogen sulfide absorption step.

12 Claims, 3 Drawing Figures

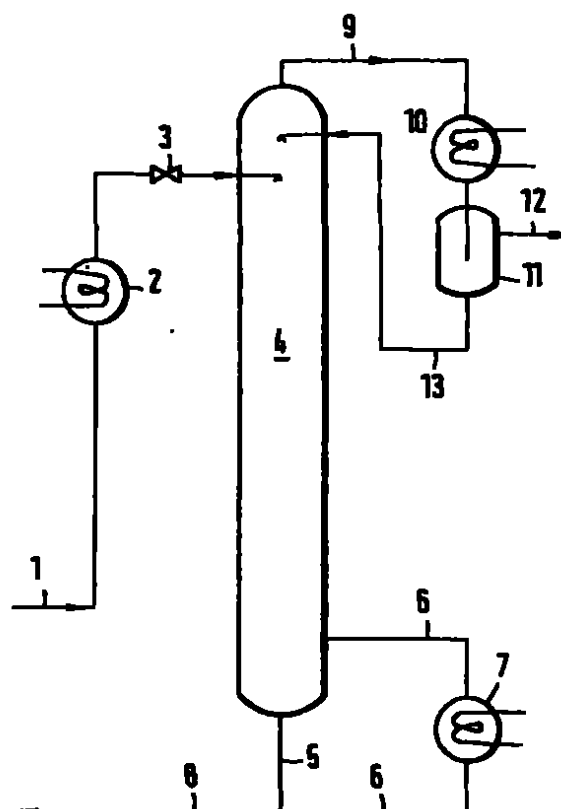
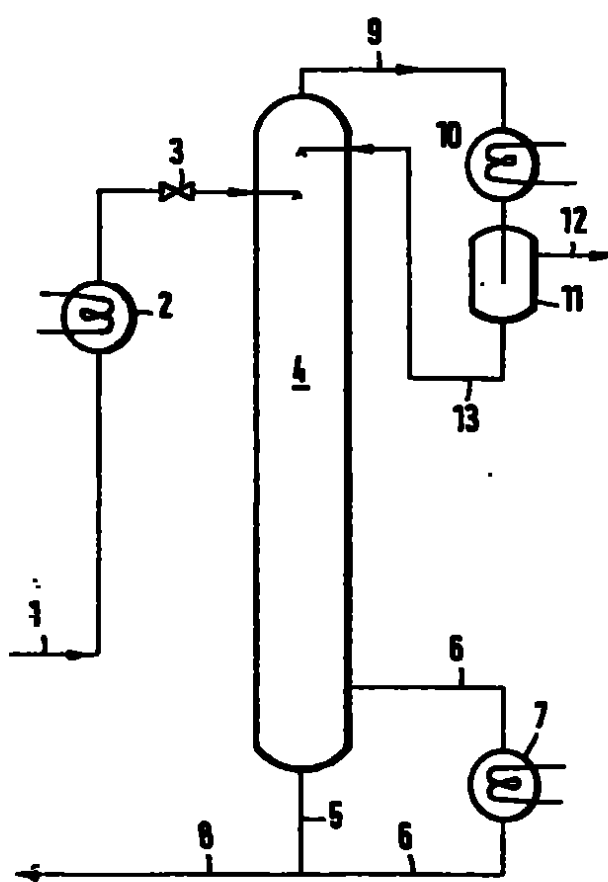
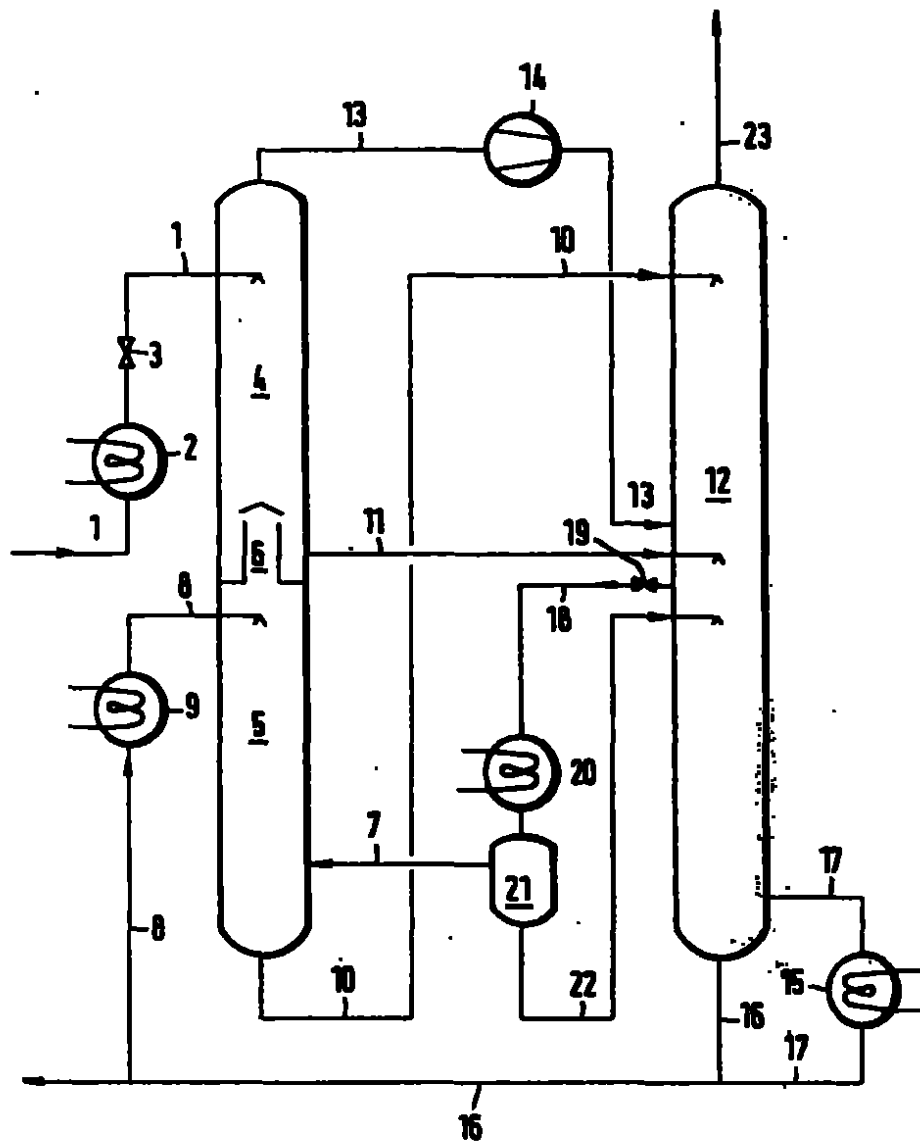


Fig. 1



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Fig. 2



PROCESS OF REGENERATING ALKALI CARBONATE SOLUTIONS FORMED BY THE DESULFURIZATION OF HOT GASES

This invention relates to a process of regenerating alkali carbonate solutions which contain hydrosulfides and bicarbonates and have been formed by the desulfurization of CO₂-containing hot gases by absorption under superatmospheric pressure and comprises pressure-relieving and stripping the solutions.

It is known that gases formed, e.g., by the reaction of liquid or solid carbonaceous fuels with gases which contain free oxygen and with water vapor can be desulfurized by being scrubbed under superatmospheric pressure with a concentrated aqueous solution of one or more alkali salts of weak inorganic acids at a temperature which is near the atmospheric boiling point of the solution and the laden solution can be regenerated by being pressure-relieved, heated, and stripped with water vapor (Printed German Application No. 2,127,768).

In the known process steam is consumed at a high rate to regenerate the solution if only a small sulfur content is permissible in the regenerated solution in order to enable a substantial desulfurization of the gas. The gas cannot be desulfurized satisfactorily if steam is available only at a restricted rate for the regeneration. Besides, the H₂S content of the exhaust gas formed by the regeneration will necessarily depend on the rate at which CO₂ is scrubbed off too as the gas is desulfurized. Particularly when the gas to be desulfurized has a low H₂S content, the processing of the resulting exhaust gas to form elementary sulfur with a high conversion, e.g., in Claus process plants, will involve a high expenditure if a substantial emission of polluting sulfur compounds is to be avoided.

It is an object of the invention to avoid these and other disadvantages of the prior art, to reduce the rate at which steam is consumed in regenerating the laden absorbent to improve the desulfurization of the gas, and to enable the recovery of an exhaust gas having a higher H₂S concentration.

This object is accomplished according to the invention in that the alkali bicarbonate content of the pressure-relieved solution which is to be stripped is adjusted to at least 55% of the total alkali content of the solution, the solution is then stripped at its boiling temperature to remove a predominant part of its H₂S content and part of its CO₂ content, and the stripped solution is re-used to desulfurize gas without being cooled before.

According to a preferred further feature of the invention, a decrease of the alkali bicarbonate content of the laden solution before the stripping is avoided in that the solution before being pressure-relieved is cooled below its boiling temperature at the pressure relief and pressure, preferably to 80°-95° C.

According to a further preferred feature of the invention the cooled solution is treated under normal or superatmospheric pressure at 50° to 100° C. with a gas which consists predominantly of CO₂ and the CO₂-saturated solution is then stripped at its boiling temperature with a gas which consists predominantly of water vapor.

In another embodiment of the invention, the stripped solution is cooled once more and is stripped at 50° to 100° C. with a gas which consists predominantly of CO₂ and is subsequently stripped with a gas which

consists predominantly of water vapor. Within the scope of the invention that multi-stage treatment can be repeated several times.

The stripping gas consisting predominantly of water vapor is desirably obtained by an indirect heating of the stripped solution so that the solution is stripped with its own vapor, suitably in a countercurrent.

Within the scope of the invention, a suitable high-CO₂ gas for saturating the cooled solution can be obtained in that the exhaust gas, which is formed by the stripping step and consists mainly of water vapor and CO₂ and contains also H₂S is cooled, the aqueous condensate is separated and the cooled gas is scrubbed with a partial stream of the alkali carbonate solution at a temperature below 60° C., preferably 25° to 35° C.

In this way most of the hydrogen sulfide contained in the cooled gas is selectively scrubbed off so that all or part of the scrubbed low-H₂S gas which consists predominantly of CO₂ can be used to saturate the cooled solution that is to be regenerated.

Within the scope of the invention, the saturating step, in which part of the CO₂ is transferred from the gas into the solution, may be succeeded by the use of the non-absorbed gas for stripping the partial stream of the alkali carbonate solution which has been used to scrub the cooled exhaust gas formed by the stripping step and the non-absorbed gas may then be discharged whereas the stripped partial stream is returned to the main cycle of the alkali carbonate solution.

As a result, virtually all hydrogen sulfide which has been stripped from the solution to be regenerated is contained in the exhaust gas which is discharged. The concentration of hydrogen sulfide in that exhaust gas will be increased if only part of the scrubbed gas, which consists predominantly of CO₂, is used first to saturate the cooled solution to be regenerated, and subsequently to strip the high-H₂S partial stream of the solution, and the remaining portion is discharged as a low-H₂S exhaust gas.

In accordance with the invention the high-H₂S partial stream of the alkali carbonate solution which has been used to scrub the cooled exhaust gas formed by the stripping step can be regenerated by stripping said partial stream at its boiling temperature with a gas consisting predominantly of water vapor so as to remove substantially all hydrogen sulfide, whereafter the resulting gas, which contains hydrogen sulfide, is discharged and the stripped partial stream is returned to the main cycle of the alkali carbonate solution.

According to a preferred feature of the invention the gas consisting predominantly of CO₂ and used to saturate the cooled solution is formed by the regeneration of a second cycle of alkali carbonate solution used to scrub the desulfurized, CO₂-containing hot gas under superatmospheric pressure. In this embodiment, the first desulfurizing cycle is succeeded in the gas flow path by a second cycle of alkali carbonate solution, which supplies the CO₂ required to regenerate the laden solution in the first cycle. Besides, the second scrubbing of the gas under superatmospheric pressure with absorbent flowing in a second cycle causes additional hydrogen sulfide to be removed from the gas so that its desulfurization can also be improved. The alkali carbonate solutions in the two cycles may differ in composition. For instance, a concentrated solution may be used in the first cycle and a more dilute solution in the second or vice versa.

Within the scope of the invention it is preferable to use an alkali carbonate solution to which sodium tetraborate has been added, preferably in an amount of 1 to 20 mole percent of the total alkali content of the solution, i.e., 1 to 20% of the total alkali which can be titrated with acid should be added as sodium tetraborate, $\text{Na}_2\text{B}_4\text{O}_7$.

The advantages afforded by the invention reside in that an alkali carbonate solution which contains hydrosulfide and bicarbonate and has been obtained by the desulfurization of CO_2 -containing hot gases under superatmospheric pressure can be selectively regenerated by the process described in that almost all of the combined hydrogen sulfide contained in the solution as hydrosulfide but only part of the combined carbon dioxide present as bicarbonate are released and stripped off. This will be particularly significant if a selective removal of hydrogen sulfide from such gas is desired, i.e., when it is desired to scrub as much hydrogen sulfide as possible and as little CO_2 as possible from the gas, as is required in various applications, e.g., in the desulfurization of fuel gas for a gas turbine process in which the mass of the humid gas should be preserved as far as possible before it is burnt. Besides, the process renders the driven-off hydrogen sulfide available in a higher concentration. Moreover, the further processing is facilitated and the investment and supplies required for the further processing are reduced. Finally, the quantity of polluting gases emitted into the environment may be decreased.

The reactions taking place in the alkali carbonate solution are known and may be summarized as follows:

Absorption:



The reactions take place in the opposite direction during the regeneration:



From equation (4) it is apparent that bicarbonate is required for the decomposition of the hydrosulfide which has been absorbed. Such alkali carbonate solutions formed by the desulfurization of CO_2 -containing gases contain in more cases more bicarbonate than hydrosulfide. But when the solutions are regenerated in the conventional manner by being pressure-relieved, heated and stripped with steam, equal percentages of bicarbonate and hydrosulfide are always decomposed and driven off as CO_2 and H_2S . In that case, when 90% of the hydrosulfide are to be removed by the regeneration, about 90% of the bicarbonate must also be decomposed and driven off. The regenerated solution will then contain only 10% of the hydrosulfide content but will also contain only 10% of the bicarbonate content of the unregenerated solution. Such regenerated solution contains almost only alkali carbonate and for this reason can form much bicarbonate when it is used to desulfurize the CO_2 -containing gas under superatmospheric pressure. As a result, CO_2 is scrubbed off at a high rate so that the desulfurization of the gas is less selective, with all disadvantages listed herein before.

FIG. 1 is a flow diagram illustrating one embodiment of the invention.

FIG. 2 is a flow diagram illustrating a preferred embodiment of the invention.

FIG. 3 is a flow diagram illustrating a further preferred embodiment of the invention.

Whereas much bicarbonate is required for a substantial regeneration of alkali carbonate solutions which contain hydrosulfide, part of the CO_2 required for this purpose may be recirculated and re-used in the regenerating stage in the process according to the invention if this is required. For this reason the regenerated solution may have a relatively high bicarbonate content and yet be freed almost completely from hydrosulfide, i.e., subjected to a selective desulfurization.

A preferred feature of the invention is based on the recognition that the purified gas which has been obtained by the desulfurization can be used to special advantage in a combined power plant process. For that purpose the hot desulfurized hot gas which is under superatmospheric pressure is used for an ecologically satisfactory production of electric power in a combined power plant process, in which energy is recovered from the gas by a pressure relief and/or combustion and is utilized to drive gas turbines and/or to produce steam.

That practice affords the advantage that the CO_2 left in the desulfurized gas greatly increases the efficiency of the combined power plant process because the energy contained in the CO_2 under superatmospheric pressure can be utilized.

Another advantage of the invention resides in that the H_2S -containing exhaust gas formed by the regeneration of the absorbent has a higher H_2S content than the exhaust gas formed in a regenerating process known in the art. As a result, the plants for processing such exhaust gas may be smaller and less expensive and can operate more economically.

The invention will be described more fully hereinafter. The invention is shown diagrammatically and by way of example in FIG. 1 to 3. The drawings show only the most important elements of the process, whereas pumps and valves, e.g., have been omitted for the sake of clearness.

EXAMPLE 1

The process according to the invention is shown in its simplest form in FIG. 1. That embodiment may be used if the alkali carbonate solution which has been formed by the desulfurization of the gas and is to be regenerated contains at least 55% of its alkali as bicarbonate. This solution is fed through conduit 1 at a temperature of 115° C. under a pressure of 20 bars and consists of an aqueous solution of 251 grams KHCO_3 /liter, 5.0 grams KHS /liter 62.5 grams K_2CO_3 /liter and 60 grams $\text{Na}_2\text{B}_4\text{O}_7 \cdot 10 \text{H}_2\text{O}$ /liter. The KHCO_3 content thus amounts to 66% of the total content of which rises in the regenerator 4. Gases and vapors from the top of regenerator 4 flow through conduit 9, are cooled and partially condensed in the indirect cooler 10. Gases and condensate are separated in the separator 11. Remaining gases are removed through the conduit 12. The condensate flows back to the top of the regenerator 4 through conduit 13. The solution is cooled to 90° C. in the cooler 2 and is pressure-relieved in the pressure-relief valve 3 and then enters the regenerating column 4, which is under a pressure of 1.5 bars. Almost no gas and water vapor are released by the pressure relief, which does not change the bicarbonate and hydrosulfide contents.

In the regenerator 4, the solution is heated and decomposed by the gases which rise from the lower portion and consist predominantly of water vapor. As a result, the solution releases CO_2 and H_2S . In the lower portion, the solution is circulated in conduits 5 and 6 and the indirectly heated heater 7, whereby mainly water vapor is produced, which rises in the regenerator 4. The regenerator solution runs off through conduit 8 and now contains only 72.3 grams KHCO_3 /liter and 1.1 grams KHS /liter. The borax content has not been changed and the K_2CO_3 content has increased to 189.9 grams per liter. The bicarbonate content has thus been decreased to 28.8% of its initial value and the hydrosulfide content to 22.0% of its initial value. It is apparent that the hydrosulfide content has been decreased more than the bicarbonate content, which means that the regeneration is selective with respect to hydrosulfide or H_2S . The solution may now be re-used for a selective desulfurization of gas.

CONTROL EXAMPLE

When the cooler 2 is not operated, in accordance with the state of the art, the incoming solution at 115°C . is pressure-relieved in the valve 3 to a KHCO_3 content of 49% of the total alkali content, and the heater 7 is supplied with steam at the same rate as in the preceding Example, the resulting regenerated solution contains 77.8 grams KHCO_3 /liter, 1.5 grams KHS /liter, 60 grams borax/liter and 185.7 grams K_2CO_3 /liter. In this operation, which is in accordance with the prior art, the KHCO_3 content has been decreased by 69% and the KHS content by 70%, which means that the portions driven off are about the same. The regeneration is not selective. When the regenerated solution is used to desulfurize gas, this unselective regeneration results in a higher residual H_2S content of the scrubbed gas, in the removal of more CO_2 by scrubbing, and in a lower concentration of H_2S in the H_2S -containing exhaust gas.

EXAMPLE 2

This Example is illustrated in FIG. 2, which represents a preferred embodiment of the invention. The solution to be regenerated is supplied through conduit 1 under a pressure of 20 bars and a temperature of 140°C . and per liter contains 205.4 grams potassium hydrogen carbonate, 5.7 grams potassium hydrogen sulfide, 60.0 grams borax and 93.6 grams potassium carbonate. The solution is cooled to 92°C . in the indirect cooler 2 and is pressure-relieved in the valve 3. Virtually no CO_2 and no H_2S are released by the pressure relief. After the pressure relief the solution enters the CO_2 saturator 4, which is under a pressure of 1.2 bars and in which the solution at 85°C . absorbs CO_2 from the gas which rises through the gas transfer pipe 6, and consists predominantly of CO_2 . As a result, the KHCO_3 content increases to 235.8 grams/liter and the K_2CO_3 content is decreased. 62% of the total alkali are now present as KHCO_3 . The remaining gases are conducted through a conduit 13 and a compressor 14 into the upper portion of a regenerator 12 which is under a pressure of 1.5 bars. The solution to be regenerated, which is saturated with CO_2 , flows through conduit 11 into the lower portion of the regenerator 12 and is stripped there at its boiling temperature by the gas which rises from the bottom and consists mainly of water vapor. That gas is produced in that the regenerated solution which runs off through conduit 16 is circulated in the conduit 17 and heated by an indirectly heated reheater 15.

Under the inlet 11 for the solution, part of the gas which rises in the lower portion of the regenerator and consists predominantly of water vapor is withdrawn through conduit 18, which incorporates a control valve 19, and the withdrawn gas is cooled in the indirect cooler 20. The resulting aqueous condensate is separated in a separator 21 and is recycled to the regenerator 12 through a conduit 22. The gas leaving the separator 21 consists mainly of CO_2 and also contains H_2S and is conducted through conduit 7 to the H_2S scrubber 5. A partial stream of the regenerated solution is fed through conduit 8 and the indirect cooler 9 to the scrubber 5, in which a major part of the H_2S content of the gas is scrubbed off at 30°C . and under a pressure of 1.3 bars so that the gas entering the CO_2 saturator 4 through the gas transfer pipe 6 contains only 200 ppm H_2S . In the solution running off through the conduit 10 the H_2S which has been removed is contained as KHS . That solution is fed to the top of the regenerator 12, in which the H_2S is driven off and exits together with CO_2 through conduit 23.

In addition to borax, the regenerated solution running off through conduit 16 contains only 0.58 grams KHS /liter and 68.5 grams KHCO_3 /liter but 193.0 grams K_2CO_3 /liter. It is apparent that relative to the composition of the solution in conduit 1, 90% of the KHS content have been decomposed and driven off as H_2S but only 67% of the KHCO_3 content. The solution which has been selectively regenerated according to the invention can be used at 140°C . in an absorber under a pressure of 20 bars to desulfurize a gas containing 0.3% H_2S by volume to a residual content of 200 ppm, which means that more than 93% of the H_2S content have been scrubbed off, whereas only 20 to 30% of the CO_2 also contained in the gas are also scrubbed off.

EXAMPLE 3

Another preferred embodiment of the invention is diagrammatically shown in FIG. 3. A solution which contains 235.8 grams KHCO_3 /liter (=62% of the total alkali content), 21.6 grams KHS /liter, 75 grams borax/liter and 51.9 grams K_2CO_3 /liter is supplied at a temperature of 120°C . and under a pressure of 21 bars through conduit 1. The solution is cooled to 91°C . in the indirect cooler 2, and pressure-relieved in valve 3 to the pressure of 1.6 bars which prevails in the stripper 4, to which the pressure-relieved solution is supplied. A gas consisting predominantly of water vapor is conducted from the stripper 6 and enters the stripper 4 through the gas transfer pipe 5. The solution is stripped at its boiling temperature and after being stripped contains 83.7 grams KHCO_3 /liter, 4.8 grams KHS /liter, 75.0 grams borax/liter, and 173.1 grams K_2CO_3 /liter. The solution then flows through conduit 7 into the indirect cooler 8 and is cooled therein to 85°C . and subsequently enters the CO_2 saturator 9, which is under a pressure of 1.4 bars. A gas consisting predominantly of CO_2 enters the CO_2 saturator through a gas transfer pipe 10 and is used to treat the solution at 80° to 90°C . The treated solution contains 247.2 grams KHCO_3 /liter, 4.4 grams KHS /liter, 75.0 grams borax/liter and 60.6 grams K_2CO_3 /liter and is fed in conduit 11 to the second stripper 6. In the latter the solution is regenerated at its boiling temperature under a pressure of 1.6 to 1.7 bars by being stripped with a gas which consists mainly of water vapor and is formed in that the solution circulated in conduits 13 and 14 is indirectly heated in the reheater 12. The regenerated solution running off through con-

duit 13 contains only 0.7 grams KHS/liter as well as 87.5 grams KHCO_3 /liter, 74.8 grams borax/liter and 174.4 grams K_2CO_3 /liter. It is apparent that a total of 96.8% of the hydrosulfide KHS and 62.9% of the bicarbonate KHCO_3 fed through conduit 1 have been decomposed and driven off and that the regeneration has been highly selective for KHS. The regenerated solution can be used to desulfurize gases and need not be cooled before.

A partial stream of the gases rising in the stripper 4 is withdrawn through conduit 16, which incorporates a control valve 15, and is cooled to 25° C. in the cooler 17. The resulting aqueous condensate is withdrawn from the separator 18 and is recycled through conduit 19 to the stripper 4. The cooled gas is fed through the conduit 20 to the H_2S scrubber 21 and is scrubbed there under a pressure of 1.4 to 1.5 bars with a partial stream of the alkali carbonate solution. That partial stream is provided in that part of the solution in conduit 7 is branched off through conduit 22 and cooled in the indirect cooler to 23° to 26° C. In the scrubber 21, the H_2S is removed from the gas at 26° to 30° C. to a residual content of 280 ppm. The scrubbed gas, which consists predominantly of CO_2 , serves to treat the solution in the CO_2 saturator 9. The non-absorbed gas is withdrawn through conduit 24 and is fed by a compressor 25 into the stripper 4.

The high- H_2S solution running off from the scrubber 21 is fed through conduit 26 to the top of the stripper 4 for driving off the H_2S . The exhaust gas contains water vapor and is withdrawn through conduit 27. It contains all hydrogen sulfide that has been driven off in the regenerating stage and also the CO_2 which has been driven off.

In the manner which has been described, the solution can be regenerated so that a regenerated solution can be obtained which can be used to desulfurize to a residual content of 200 to 300 ppm H_2S a gas which contains 1.0% H_2S by volume and 10.2% CO_2 by volume at 120° C. under a pressure of 21 bars, which means that 97% of the H_2S are scrubbed off. Only 25% of the CO_2 are scrubbed off at the same time.

We claim:

1. In a process for selective removal of H_2S from hot gases containing H_2S and CO_2 , wherein the gases are contacted at superatmospheric pressure with an alkali metal carbonate solution and the resultant alkali metal carbonate solution which contains hydro-sulfides and bicarbonates is regenerated by pressure relieving the same, the so pressure relieved solution is thereafter stripped and re-used for selective removal of H_2S from hot gases containing the same and carbon dioxide, the improvement which comprises adjusting the alkali metal bicarbonate content of the pressure-relieved solution to be stripped so that it is at least 55% of the total alkali content of the solution and reusing said solution, following stripping, without cooling the same.

2. A process according to claim 1, characterized in that a decrease of the alkali bicarbonate content of the laden solution before the stripping thereof is avoided in that the solution before being pressure-relieved is cooled below its boiling temperature at the pressure relief and pressure preferably to 80°-95° C.

3. A process according to claim 2, characterized in that the cooled solution is treated under normal or su-

peratmospheric pressure at 50° to 100° C. with a gas which consists predominantly of CO_2 and the CO_2 -saturated solution is then stripped at its boiling temperature with a gas which consists predominantly of water vapor.

4. A process according to claim 3, characterized in that the stripped solution is cooled once more and is stripped at 50° to 100° C. with a gas which consists predominantly of CO_2 and the solution is subsequently stripped with a gas which consists predominantly of water vapor.

5. A process according to claim 4, characterized in that the multi-stage treatment of the solution is repeated several times.

6. A process according to claim 1, characterized in that the solution to be regenerated is stripped at its boiling temperature with a gas which consists predominantly of water vapor and has been obtained by an indirect heating of the stripped solution.

7. A process according to claim 1, characterized in that the exhaust gas formed by the stripping is cooled, the condensate is separated, the cooled gas is scrubbed with a partial stream of the alkali carbonate solution at a temperature below 60° C., preferably 25° to 35° C., and all or part of the scrubbed gas consisting predominantly of CO_2 is used to saturate the cooled solution which is to be regenerated.

8. A process according to claim 2, characterized in that all or part of the scrubbed gas which consists predominantly of CO_2 is first used to saturate the cooled solution to be regenerated and is subsequently used to strip the heated partial stream of the alkali carbonate solution formed by the scrubbing of the cooled exhaust gas formed by the stripping step and is then discharged and the stripped partial stream is returned to the main cycle of the alkali carbonate solution.

9. A process according to claim 4, characterized in that a partial stream of the alkali carbonate solution is withdrawn from the scrubber for scrubbing the cooled exhaust gas formed by the stripping step and said partial stream is stripped at its boiling temperature with a gas consisting predominantly of water vapor to remove a predominant portion of hydrogen sulfide from said partial stream, the resulting gas, which contains hydrogen sulfide, is removed and the stripped partial stream is returned to the main cycle of the alkali carbonate solution.

10. A process according to claim 1, characterized in that the gas consisting predominantly of CO_2 is formed by the regeneration of a second cycle of alkali carbonate solution used to scrub the desulfurized, CO_2 -containing hot gas under superatmospheric pressure.

11. A process according to claim 2, characterized in that the alkali carbonate solution contains sodium tetraborate in an amount of 1 to 20 mole percent of its total alkali content.

12. A process according to claim 1 characterized in that the desulfurized hot gas under superatmospheric pressure is used for an ecologically satisfactory production of electric power in a combined power plant process in which energy is recovered from the gas by a pressure relief and/or combustion of the gas and is utilized to drive gas turbines and/or to produce steam.

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**UNITED STATES PATENT AND TRADEMARK OFFICE
CERTIFICATE OF CORRECTION**

PATENT NO. :4,258,019

DATED :Mar. 24, 1981

INVENTOR(S) :Heinz Hiller et al

It is certified that error appears in the above-identified patent and that said Letters Patent are hereby corrected as shown below:

Col. 8, line 21 Delete "1" and insert --3--.

Col. 8, line 29 Delete "2" and insert --3--.

Col. 8, line 38 Delete "4" and insert --7--.

Col. 8, line 49 Delete "1" and insert --3--.

Signed and Sealed this

Sixth Day of October 1981

[SEAL]

Attest:

GERALD J. MOSSINGHOFF

Attesting Officer

Commissioner of Patents and Trademarks

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AUSTIN INTL

Date of mailing (day/month/year)
23 September 2004 (23.09.2004)

Applicant's or agent's file reference
UTFB:718WO

IMPORTANT NOTICE

International application No.
PCT/US2004/006580

International filing date (day/month/year)
04 March 2004 (04.03.2004)

Priority date (day/month/year)
10 March 2003 (10.03.2003)

Applicant

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(19) World Intellectual Property
Organization
International Bureau



(43) International Publication Date
23 September 2004 (23.09.2004)

PCT

(10) International Publication Number
WO 2004/080573 A1

(51) International Patent Classification⁷: B01D 53/14
(21) International Application Number: PCT/US2004/006580
(22) International Filing Date: 4 March 2004 (04.03.2004)
(25) Filing Language: English
(26) Publication Language: English
(30) Priority Data: 60/453,279 10 March 2003 (10.03.2003) US

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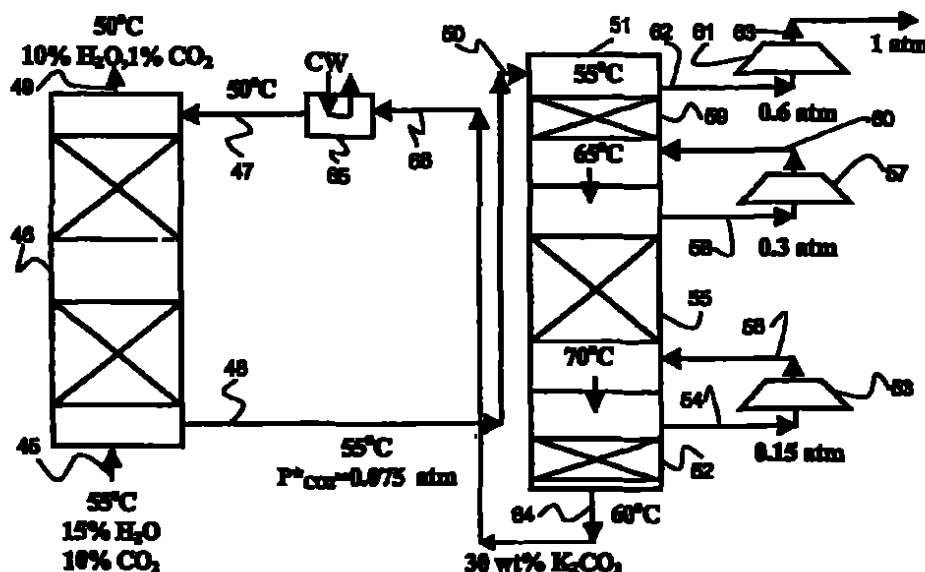
(81) Designated States (unless otherwise indicated, for every kind of national protection available): AE, AG, AL, AM, AT, AU, AZ, BA, BB, BG, BR, BW, BY, BZ, CA, CH, CN, CO, CR, CU, CZ, DE, DK, DM, DZ, EC, EE, EG, ES, FI, GB, GD, GE, GH, GM, HR, HU, ID, IL, IN, IS, JP, KE, KG, KP, KR, KZ, LC, LK, LR,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

(84) Designated States (unless otherwise indicated, for every kind of regional protection available): ARIPO (BW, GH, GM, KE, LS, MW, MZ, 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, FR, GB, GR, HU, IE, IT, LU, MC, 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).

Published:
— with international search report

[Continued on next page]

(54) Title: REGENERATION OF AN AQUEOUS SOLUTION FROM AN ACID GAS ABSORPTION PROCESS BY MULTISTAGE FLASHING AND STRIPPING



(57) Abstract: An acid gas such as carbon dioxide, hydrogen sulfide, or a mixture thereof is removed from gaseous streams using aqueous absorption and stripping processes. By replacing the conventional stripper used to regenerate the aqueous solvent and capture the acid gas with a multipressure stripper (51) that combines acid gas compression with stripping, less energy is consumed. The multipressure stripper is a multistage flash (52, 55, 59) in which the total vapor flow from each stage is compressed and fed to the bottom of the previous flash stage at a higher pressure. In this process, the heat in the water content of the vapor exiting each stage is utilized at a higher pressure in the previous stage. The described stripping process generates the acid gas at a higher pressure without operating the stripper at a higher temperature, thereby reducing the energy consumption of the system.



— *before the expiration of the time limits for amending the claims and to be republished in the event of receipt of amendments*

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REGENERATION OF AN AQUEOUS SOLUTION FROM AN ACID GAS ABSORPTION PROCESS BY
MULTISTAGE FLASHING AND STRIPPING

5 This application claims priority to, and incorporates by reference, U.S. Provisional Patent Application Serial No. 60/453,279 filed March 10, 2003. The present invention relates generally to the removal and capture of acid gases such as carbon dioxide, hydrogen sulfide and mixtures thereof from gases containing same through aqueous absorption and stripping processes. More particularly, it provides methods for reducing the energy consumption of such absorption and
10 stripping processes.

 A common viewpoint held by a significant segment of the environmental community is that carbon dioxide released into the air plays a major role in global climate change. Thus, global climate change initiatives such as the Kyoto Protocol have identified the curtailment of carbon dioxide releases from fossil fuel combustion and other point sources as a primary means
15 of reducing global climate change. Extensive programs already in place are beginning to demonstrate the economic and technical feasibility of sequestering carbon dioxide by approaches such as injection in underground reservoirs (see, Bergman, P.D. et al, "Disposal of Power Plant CO₂ in Depleted Oil and Gas Reservoirs in Texas," presented at the Third International Conference on Carbon Dioxide Removal, Cambridge, MA, Sept 9-11, 1996) and disposal in the
20 deep ocean (Fujioka, Y. et al., "Cost Comparison of Various CO₂ Ocean Disposal Options," presented at the Third International Conference on Carbon Dioxide Removal, Cambridge, MA, Sept 9-11, 1996).

 One method of curtailing carbon dioxide releases in the industrial arena involves removing carbon dioxide from combustion gases and other gases. Carbon dioxide is emitted in
25 large quantities from fuel combustion by mobile and stationary sources. Carbon dioxide capture/sequestration will be most effective if applied to large stationary sources. The largest single sources of carbon dioxide are conventional coal-fired power plants. These sources represent 30 to 40% of the carbon dioxide emissions in the United States. Technology developed for such sources should also be applicable to CO₂ capture from gas and oil fired
30 boilers, combined cycle power plants, coal gasification, and hydrogen plants. Absorption/stripping is primarily a tail-end technology and is therefore suitable for both existing and new boilers. Specifically, it can be used with existing coal-fired boilers, especially if they already have scrubbers for SO₂-abatement.

The use of absorption and stripping processes with aqueous solvents such as alkanolamines and promoted potassium carbonate is a known, effective technology for the removal and capture of carbon dioxide from fine gas, natural gas, hydrogen, synthesis gas, and other gases. U.S. Patents 4,477,419 and 4,152,217, each of which is incorporated herein by reference, describe aspects of this technology. Alkanolamine absorption/stripping is one proven and effective technology for carbon dioxide capture from gas. The first generation of this technology uses aqueous solutions of monoethanolamine (MEA). Advances in this technology have provided other alkanolamine solvents for carbon dioxide treating in various industries. Monoethanolamine (MEA), diethanolamine (DEA), and the hindered amine AMP are used alone in an aqueous solution. Typical solvent blends include a methyldiethanolamine (MDEA) solution promoted by piperazine or other secondary amines. Also, potassium carbonate solvents are commonly promoted by DEA or other reactive amines.

Gas absorption is a process in which soluble components of a gas mixture are dissolved in a liquid. Stripping is essentially the inverse of absorption, as it involves the transfer of volatile components from a liquid mixture into a gas. In a typical carbon dioxide removal process, absorption is used to remove carbon dioxide from a combustion gas, and stripping is subsequently used to regenerate the solvent and capture the carbon dioxide contained in the solvent. Once carbon dioxide is removed from combustion gases and other gases, it can be captured and compressed for use in a number of applications, including sequestration, production of methanol, and tertiary oil recovery.

The conventional method of using absorption/stripping processes to remove carbon dioxide from gaseous streams is described in U.S. Patent 4,384,875, which is incorporated herein by reference. In the absorption stage, the gas to be treated, containing the carbon dioxide to be removed, is placed in contact, in an absorption column, with the chosen absorbent under conditions of pressure and temperature such that the absorbent solution removes virtually all the carbon dioxide. The purified gas emerges at the top of the absorption column and, if necessary, it is then directed towards a scrubber employing sodium hydroxide, in which the last traces of carbon dioxide are removed. At the bottom of the absorption column, the absorbent solution containing carbon dioxide (also called "rich solvent") is drawn off and subjected to a stripping process to free it of the carbon dioxide and regenerate its absorbent properties.

To effect the regeneration of the absorbent solution, the rich solvent drawn off from the bottom of the absorption column is introduced into the upper half of a stripping column, and the rich solvent is maintained at its boiling point under pressure in this column. The heat necessary for maintaining the boiling point is furnished by reboiling the absorbent solution contained in the stripping column. The reboiling process is effectuated by indirect heat exchange between part of

the solution to be regenerated located in the lower half of the stripping column and a hot fluid at appropriate temperature, generally saturated water vapor. In the course of regeneration, the carbon dioxide contained in the rich solvent to be regenerated maintained at its boiling point is released and stripped by the vapors of the absorbent solution. Vapor containing the stripped carbon dioxide emerges at the top of the stripping column and is passed through a condenser system which returns to the stripping column the liquid phase resulting from the condensation of the vapors of the absorbent solution which pass out of the stripping column with the gaseous carbon dioxide. At the bottom of the stripping column, the hot regenerated absorbent solution (also called "lean solvent") is drawn off and recycled to the absorption column after having used part of the heat content of the solution to heat, by indirect heat exchange, the rich solvent to be regenerated, before its introduction into the stripping column.

In simple absorption/stripping as it is typically practiced in the field, aqueous rich solvent is regenerated at 100-120°C in a simple, countercurrent, reboiled stripper operated at a single pressure, which is usually 1-2 atm. The rich solvent feed is preheated by cross-exchange with hot lean solvent product to within 5-30°C of the stripper bottoms. The overhead vapor is cooled to condense water, which is returned as reflux to the countercurrent stripper. When used for carbon dioxide sequestration and other applications, the product carbon dioxide is compressed to 100-150 atm.

A major problem with the existing absorption/stripping process described above is that it is very energy intensive, and this is largely because the heat required for the heat reboiler is significant. In application on a coal-fired power plant, the required heat of such a process can reduce net power production by as much as 15 to 30%. (Herzog, H., E. Drake, & E. Adams, "CO₂ Capture, Reuse, and Storage Technologies for Mitigating Global Climate Change," final rept, DOE Order No. DE-AF22-96PC01257, 1997). Therefore, it is important to maximize energy efficiency in the design and operation of these systems. The primary method for enhancing energy efficiency is the recovery of useful heat from the overhead condenser, as the overhead vapor can contain one to ten moles of water vapor for every mole of carbon dioxide. Common forms of heat recovery currently practiced include vapor recompression and multieffect strippers. In vapor recompression, the overhead vapor is compressed by a factor of two to ten and then exchanged with the bottoms liquid to provide heat for the reboiler. With multieffect strippers, two or more strippers are operated in parallel, but each stripper is operated at a significantly different pressure. The vapor from a higher pressure stripper is used to heat the reboiler of a lower pressure stripper in a cascade arrangement. Unfortunately, both of these configurations result in a loss of energy in the required heat exchanger.

As noted in U.S. Patent 4,152,217, several attempts have been made to reduce the overall cost associated with the regeneration of absorbent liquid streams. By devising a system in which the spent absorbent/lean absorbent heat exchanger, the overhead cooler-condenser, the reflux drum, and the reflux pump, ordinarily constructed and used with a conventional stripper or regenerator, could be eliminated, the patentee in U.S. Patent 3,690,861 sought to reduce capital investment costs. While capital investment was considerably reduced in the disclosed process, no consideration was given to how the elimination of heat exchangers would affect overall heat requirements of the system. As described in U.S. Patent 4,152,217, it has been found that elimination of heat exchangers increases the overall heat requirements of the system. Thus, while initial capital expenditures are considerably lessened, long term operating expenses, especially in view of rising energy costs, would be higher in an absorption/stripping process that eliminates heat exchangers.

Other patents have disclosed various methods for improving the cost effectiveness of carbon dioxide removal systems by reducing the energy requirements of such systems. One such patent, U.S. Patent 4,553,984, which is incorporated herein by reference, discloses a method in which the rich solvent laden with carbon dioxide is regenerated without the use of a stripping column simply by flashing in one or more flash stages. The disclosed method is said to substantially reduce both capital costs and energy costs. However effective at reducing costs such a technology may be, it may be ineffective for applications which require the captured carbon dioxide to be produced at a higher pressure for use in sequestration, production of methanol, tertiary oil recovery, or other applications. At the very least, such a technology may require significant capital expenditure for the addition of a compressor to enable the captured carbon dioxide to be compressed for use in applications that require higher pressure carbon dioxide.

In light of the above, it would be advantageous to provide for technology in which carbon dioxide can be removed from combustion gases and other gases by an absorption/stripping process that is significantly more energy efficient than the processes currently practiced. The ideal system would generate carbon dioxide at a higher pressure without operating the stripper at a greater temperature. Such a system would, in turn, significantly reduce the energy cost associated with systems in which the carbon dioxide must be compressed for sequestration, production of methanol, tertiary oil recovery, or other applications.

Any problems or shortcomings enumerated in the foregoing are not intended to be exhaustive but rather are among many that tend to impair the effectiveness of previously known techniques. Other noteworthy problems may also exist; however, those presented above should

be sufficient to demonstrate that apparatus and methods appearing in the art have not been altogether satisfactory and that a need exists for the techniques disclosed herein.

The present invention overcomes deficiencies in the prior art by providing, among other things, a method for using aqueous absorption and stripping processes to remove an acid gas from gaseous streams in a manner that generates said acid gas at a higher pressure while consuming less energy than the existing technology. The method involves replacing the conventional single-pressure stripper used to regenerate the aqueous solvent and capture the carbon dioxide with a multipressure stripper that combines acid gas compression with stripping. By generating the acid gas at a higher pressure without operating the stripper at a greater temperature, the method reduces the energy consumption of systems in which the carbon dioxide must be compressed for sequestration, production of methanol, tertiary oil recovery, or other applications.

In one embodiment, the present invention concerns an improved method for removing an acid gas from a gaseous stream containing same using aqueous absorption and stripping equipment, the improvement comprising:

- (a) passing the acid gas-rich solvent stream exiting the absorbing equipment through a multipressure stripper in which the acid gas-rich solvent stream passes through multiple flash stages such that the vapor obtained by flashing the acid gas-rich solvent stream in each flash stage strips the acid gas from the acid gas-rich solvent stream, resulting in an acid gas-rich gaseous stream exiting the multipressure stripper and an acid gas-lean solvent stream exiting the multipressure stripper, wherein the multipressure stripper is operated at multiple pressure levels such that the vapor obtained in each flash stage is compressed in a compressor and fed to the previous flash stage at a higher pressure; and
- (b) recycling the acid gas-lean solvent stream exiting the multipressure stripper in step (a) back to the absorption equipment at least once.

In another embodiment of the present invention, each stage of the multipressure stripper from the stripping process is a countercurrent contactor.

In another embodiment of the present invention, one or more of the stages of the multipressure stripper from the stripping process is a cocurrent contactor.

In another embodiment of the present invention, one or more of the stages of the multipressure stripper from the stripping process is a mixed contactor.

In another embodiment of the present invention, additional heat is supplied at one or more of the stages of the stripping column.

5 In another embodiment of the present invention, the invention involves a stripping equipment for stripping an acid gas from an acid gas-rich solvent stream. This equipment includes a stripping column containing multiple flash stages. The equipment also includes a compressor connected between each of the flash stages such that vapor produced in each flash stage is compressed and fed to the previous flash stage at a higher pressure.

In another embodiment of the present invention, each stage of the stripping column from the stripping equipment is a countercurrent contactor.

10 In another embodiment of the present invention, one or more of the stages of the stripping column from the stripping equipment is a cocurrent contactor.

In another embodiment of the present invention, one or more of the stages of the stripping column from the stripping equipment is a mixed contactor.

15 In another embodiment of the present invention, additional heat is supplied at one or more of the stages of the stripping column.

In the present invention, in an absorption process stage, the gaseous stream is contacted with an aqueous solvent (such as an aqueous amine, an aqueous alkanolamine or mixtures thereof, or an amine promoted aqueous potassium carbonate) in an absorption equipment such that the acid gas in the gaseous stream is transferred from the gaseous stream to the solvent, resulting in a purified gaseous stream exiting the absorption equipment and an acid gas-rich solvent stream exiting the absorption equipment. In a stripping stage, the acid gas-rich solvent stream exiting the absorption equipment is passed through a multipressure stripper in which the acid-gas-rich solvent stream passes through multiple flash stages such that the vapor obtained by flashing the acid gas-rich solvent stream in each flash stage strips the acid gas from the acid gas-rich solvent stream, resulting in an acid gas-rich gaseous stream exiting the multipressure stripper and an acid gas-lean solvent stream exiting the multipressure stripper. In the stripping stage, the multipressure stripper is operated at multiple pressure levels such that the vapor obtained in each flash stage is compressed in a compressor and fed to the previous flash stage at a higher pressure. In the final step of the method of the present invention, the acid gas-lean solvent stream exiting the multipressure stripper is recycled back to the absorption equipment.

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Conveniently, the gaseous stream to be treated by the present method is comprised of flue gas, natural gas, hydrogen gas, or synthesis gas.

As used in the specification, "a" or "an" may mean one or more. As used in the claim(s), when used in conjunction with the word "comprising", the words "a" or "an" mean one or more than one unless explicitly stated otherwise. As used herein "another" may mean at least a second or more.

5 Other objects, features, and advantages of the present invention will become apparent from the following detailed description. It should be understood, however, that the detailed description and the specific examples, while indicating illustrative embodiments of the invention, are given by way of illustration only, since various changes and modifications within the spirit and scope of the invention will become apparent to those skilled in the art from this detailed
10 description.

The following drawings form part of the present specification and are included to further demonstrate certain aspects of the present invention. The invention may be better understood by reference to one or more of these drawings in combination with the detailed description of specific embodiments presented herein.

15 FIG. 1 is a schematic drawing of CO₂ capture by potassium carbonate absorption/stripping utilizing one possible vapor recompression for energy integration.

FIG. 2 is a schematic drawing of CO₂ capture by potassium carbonate absorption/stripping utilizing conventional turbine exhaust steam for energy integration.

20 FIG. 3 is a schematic drawing of CO₂ capture by potassium carbonate absorption/stripping utilizing multipressure stripping for energy integration according to embodiments of the present invention.

FIG. 4 is a schematic diagram of the multipressure stripping process according to embodiments of the present invention.

25 According to the present invention it is possible to minimize the heat energy requirements associated with the removal and capture of carbon dioxide from a gaseous stream by aqueous absorbent and stripping processes. This is attained by combining aqueous absorption with multipressure stripping. Because the present invention results in the captured acid gas being in a compressed state, the invention may be most beneficial in systems requiring compressed acid gas for applications such as sequestration, production of methanol, and tertiary
30 oil recovery.

The present invention utilizes aspects of the process of aqueous absorption, as described in U.S. Patent 6,139,605, which is incorporated herein by reference. Through this process, an

acid gas such as carbon dioxide, hydrogen sulfide, or a mixture thereof is removed from gaseous streams by contacting the gaseous stream with a liquid absorbent that absorbs the acid gas. As industrial applications expected to benefit from the present invention require the concentration of the acid gas to be reduced to a very low level, the absorbent is generally selected to be one that reacts with the acid gas. Examples of absorbent liquids suitable for use in the present invention for the absorption of the acid gas include, but are not limited to, amine promoted aqueous potassium carbonate and aqueous solutions of amines and alkanolamines. Non-limiting examples of alkanolamines suitable for use in the present invention are monoethanolamine (MEA), diethanolamine (DEA), and methyl diethanolamine (MDEA).

The absorption step may be carried out by contacting the gaseous stream at a relatively low temperature but at an elevated pressure in an absorption column with a stream of the absorbent liquid, referred to at this stage as "lean solvent," flowing counter-current to the gaseous stream. As described in U.S. Patent 4,384,875, the purified gaseous stream emerges from the top of the absorber, while the absorbent liquid containing the acid gas, referred to at this stage as "acid gas-rich solvent," emerges from the bottom of the absorber. To capture the acid gas and regenerate the absorbent liquid so it can be recirculated back to the top of the absorber column as lean solvent, the rich solvent is treated with a stripping process.

In conventional absorption/stripping, the aqueous solvent is regenerated in a simple, countercurrent, reboiled stripper operated at a single pressure. To carry out the stripping process in the present invention, a multipressure stripper is employed. The multipressure stripper integrates the acid gas compression with stripping. The stripper itself is a multistage flash in which the total vapor flow from each stage is compressed and fed to the previous flash stage at a higher pressure. In this process, the heat in the water content of the vapor exiting each stage is utilized at a higher pressure in the previous stage. This is significant because the overhead vapor can contain one to ten moles of water vapor for every mole of the acid gas. The described stripping process generates the acid gas at a higher pressure without operating the stripper at a higher temperature, thereby reducing the energy consumption of the system.

The multipressure stripping process of the present invention may be undesirable in certain distillation and stripping applications because compressors are currently expensive to the point of being cost prohibitive. However, in applications where a compressor is already required to achieve a desired end result, the multipressure stripping process can be implemented with little additional cost.

The present invention finds application in a vast array of fields. Because it integrates compression with stripping to produce a compressed acid gas product, the invention may be particularly attractive in systems that require compression of the product acid gas for

applications such as sequestration, production of methanol, and tertiary oil recovery. Such systems will already have a compressor incorporated into their current configuration, and thus it will not be necessary to expend the significant capital necessary to purchase a compressor for use in the present invention.

5 In addition, as environmental concerns magnify, industry may be forced to further explore and adopt various technologies that effectively reduce acid gas (such as, for example, carbon dioxide) emissions in an energy-efficient manner. In such a scenario, the present invention will be highly desirable to an even greater spectrum of industry due to its energy-efficient method of operation.

10 The following non-limiting examples are included to demonstrate specific embodiments of the invention. It should be appreciated by those of skill in the art that the techniques disclosed in the examples which follow represent techniques discovered by the inventor to function well in the practice of the invention, and thus can be considered to constitute specific modes for its practice. However, those of skill in the art should, in light of the present disclosure, appreciate
15 that many changes can be made in the specific embodiments which are disclosed and still obtain a like or similar result without departing from the spirit and scope of the invention.

FIG. 1, FIG. 2, and FIG. 3 show the same carbon dioxide capture process utilizing three different modes of energy integration. The solvent used in the example process is an amine promoted aqueous potassium carbonate, although other solvents known in the art may be used as
20 well. FIG. 1 and FIG. 2 show processes utilizing known modes of energy integration, whereas FIG. 3 shows a process utilizing multipressure stripping for energy integration according to embodiments of the present invention.

Comparative Example 1

FIG. 1 shows a process utilizing one possible vapor recompression for energy integration.
25 Referring to FIG. 1, a gas which contains a high concentration of CO₂, for example a natural gas or a flue gas, is passed, via line 1, into the bottom of absorption column 2 and flows up through column 2 counter-current to absorbent flowing down the column. At the same time, lean absorption solvent, in this example a promoted potassium carbonate, is passed via line 3 to the upper end of the absorption column and semi-lean absorption solvent is fed via line 4 to a
30 location intermediate the upper and lower ends of column 2. The absorption solvent, which is fed counter-current to the gas, becomes laden with CO₂, and the rich absorption solvent is taken off at the bottom of the absorption column via line 5. The washed gas is taken off at the top of the absorption column via line 6.

Thus, the gas to be treated is scrubbed against semi-lean absorbent near the middle and lower end of absorption column 2 and against lean absorbent near the top of absorption column 2 to absorb most of the specified gas while passing most of the other gases in the gas to be treated out of the absorption step. The bulk of the absorption takes place near the bottom of column 2 by using semi-lean absorbent. Only the last traces of the CO₂ are removed from the gas in the top section of absorption column 2.

The CO₂-rich absorbent solvent is then fed, via line 7, to the upper end of a vacuum stripping column 8, where the rich absorbent flows down column 8. Heat is provided to the stripper by compressing the stripper overhead exiting column 8 via line 9 in compressors 10 and 11 and condensing its contained water vapor against a recirculating stream from the stripper bottoms in condensers 12 and 13. The recirculating stream from the stripper bottoms exits column 8 via line 15, and a portion of it is fed through condensers 12 and 13 via line 16. The recirculating stream is fed back to column 8 via line 14. Compressed gas containing the CO₂ stripped from the rich absorbent liquid exits the system via line 21. The heat provided to the stripper is used to generate the vapor necessary for stripping the CO₂ from the rich absorbent. The vapor generated from heating the rich absorbent rises up through the stripping column 8 countercurrent to the rich absorbent liquid flowing down column 8. As the vapor rises to the top of the stripping column, it strips the CO₂ from the rich absorbent. The vapor rising from the stripper bottom is mainly water vapor since aqueous solvents generally have a low volatility. At the top of column 8, the vapor carrying the CO₂ stripped from the rich absorbent is passed via line 9 through compressor 10 as described previously.

Semi-lean absorbent liquid is withdrawn from a location intermediate the lower and upper ends of stripping column 8 via line 17 and is recycled to a location intermediate the upper and lower ends of absorption column 2. Thus, the semi-lean absorbent withdrawn via line 17 is fed to heat exchanger 18, where it is cooled with cooling water to a temperature and is then fed to absorption column 2 via line 4. A portion of the lean absorbent liquid taken from the bottom of stripping column 8 via line 15 is recycled to the upper end of absorption column 2. Thus, some of the lean absorbent withdrawn via line 15 is fed to heat exchanger 20 via line 19, where it is cooled with cooling water to a temperature and is then fed to absorption column 2 via line 3.

Comparative Example 2

FIG. 2 shows a process utilizing conventional turbine exhaust steam for energy integration. Referring to FIG. 2, a gas which contains a high concentration of CO₂, for example a natural gas or a flue gas, is passed, via line 22, into the bottom of absorption column 23 and flows up through column 23 counter-current to absorbent flowing down the column. At the

same time, lean absorption solvent, in this example a promoted potassium carbonate, is passed via line 24 to the upper end of the absorption column and semi-lean absorption solvent is fed via line 25 to a location intermediate the upper and lower ends of column 23. The absorption solvent, which is fed counter-current to the gas, becomes laden with CO₂, and the rich absorption solvent is taken off at the bottom of the absorption column via line 26. The washed gas is taken off at the top of the absorption column via line 27.

Thus, the gas to be treated is scrubbed against semi-lean absorbent near the middle and lower end of absorption column 23 and against lean absorbent near the top of absorption column 23 to absorb most of the specified gas while passing most of the other gases in the gas to be treated out of the absorption step. The bulk of the absorption takes place near the bottom of column 23 by using semi-lean absorbent. Only the last traces of the CO₂ are removed from the gas in the top section of absorption column 23.

The CO₂-rich absorbent solvent is then fed, via line 28, to the upper end of a vacuum stripping column 29, where the rich absorbent flows down column 29. Heat is provided to the stripper by using exhaust steam from steam turbine 30. High pressure steam is fed to turbine 30 via line 36, and the turbine produces some form of work, such as electricity. The exhaust steam from turbine 30 is fed to condenser 31 via line 32, where it is condensed against a recirculating stream from the stripper bottoms. The recirculating stream from the stripper bottoms exits column 29 via line 33, and a portion of it is fed through condenser 31 via line 34. The recirculating stream is fed back to column 29 via line 35. The heat provided to the stripper is used to generate the vapor necessary for stripping the CO₂ from the rich absorbent. The vapor generated from heating the rich absorbent rises up through the stripping column 29 countercurrent to the rich absorbent liquid flowing down column 29. As the vapor rises to the top of the stripping column, it strips the CO₂ from the rich absorbent. The vapor rising from the stripper bottom is mainly water vapor since aqueous solvents generally have a low volatility. At the top of column 29, the vapor carrying the CO₂ stripped from the rich absorbent is passed via line 36 through condenser 37, where heat exchange with cooling water causes most of the water in the vapor to condense out. The remaining vapor, which is high in CO₂ concentration, is fed to compressor 38 via line 39. The compressed CO₂-rich gas exits the system via line 44.

Semi-lean absorbent liquid is withdrawn from a location intermediate the lower and upper ends of stripping column 29 via line 40 and is recycled to a location intermediate the upper and lower ends of absorption column 23. Thus, the semi-lean absorbent withdrawn via line 40 is fed to heat exchanger 41, where it is cooled with cooling water to a temperature and is then fed to absorption column 23 via line 25. A portion of the lean absorbent liquid taken from the bottom of stripping column 29 via line 33 is recycled to the upper end of absorption column 23.

Thus, some of the lean absorbent withdrawn via line 33 is fed to heat exchanger 43 via line 42, where it is cooled with cooling water to a temperature and is then fed to absorption column 23 via line 24.

Example 1

FIG. 3 shows a process utilizing multipressure stripping for energy integration according to embodiments of the present invention. Referring to FIG. 3, a gas which contains a high concentration of CO₂, for example a natural gas or a flue gas, is passed, via line 45, into the bottom of absorption column 46 and flows up through column 46 counter-current to absorbent flowing down the column. At the same time, lean absorption solvent, in this example a promoted potassium carbonate, is passed via line 47 to the upper end of the absorption column. The absorption solvent, which is fed counter-current to the gas, becomes laden with CO₂, and the rich absorption solvent is taken off at the bottom of the absorption column via line 48. The washed gas is taken off at the top of the absorption column via line 49.

The CO₂-rich absorbent solvent is then fed, via line 50, to the upper end of a multipressure stripping column 51, where the rich absorbent flows down column 51. The multipressure stripping column is a multistage flash, whereby the lowest flash stage 52 provides stripping steam by flashing the absorption solvent. The stripping steam then contacts the rich absorbent liquid in each flash stage and strips the CO₂ from the rich absorbent liquid. The total vapor flow produced in lowest stage 52 is fed through compressor 53 via line 54. The compressed vapor exiting compressor 53 is fed to the bottom of intermediate flash stage 55 via line 56. The total vapor flow produced in intermediate flash stage 55 is fed through compressor 57 via line 58. The compressed vapor exiting compressor 57 is fed to the bottom of top flash stage 59 via line 60. The total vapor flow produced in top flash stage 59 is fed through compressor 61 via line 62. The CO₂-rich compressed vapor exiting compressor 62 exits the system via line 63.

The lean absorbent liquid taken from the bottom of multipressure stripping column 51 via line 64 is recycled to the upper end of absorption column 46. Thus, some of the lean absorbent withdrawn via line 64 is fed to heat exchanger 65 via line 66, where it is cooled with cooling water to a temperature and is then fed to absorption column 46 via line 47.

Using known vapor recompression as shown in FIG. 1, heat will be provided to the stripper by compressing the stripper overhead and condensing its contained water vapor against a recirculating stream from the stripper bottoms. The compression work will also be recovered as heat in the stripper bottoms. This feature replaces steam with equivalent power. Vapor recompression may require an axial flow compressor with high volume capacity for water vapor.

The capital cost of this compressor may be significant. It will be comparable in size to the air compressor of a large gas turbine for a power plant of the equivalent size, but the shaft power requirement will be an order of magnitude smaller because of the low density of the vacuum stream.

5 Using conventional turbine exhaust steam as shown in FIG. 2, a steam turbine could produce electricity and could provide the additional work required to compress the CO₂ from 0.15 atm to 1 atm. The capital cost of compressors would be significantly reduced relative to that for vapor recompression. However, the capital cost of a steam turbine may be significant.

10 Compared to the conventional modes of energy integration shown in FIG. 1 and FIG. 2, the multipressure stripping mode of energy integration of the present invention shown in FIG. 3 offers the minimum energy configuration with the minimum amount of heat exchange. The bottom stage provides stripping steam by flashing the solvent. The top stage recovers heat and drives the overall stripper to a greater temperature and pressure. All of the energy is provided as work, either from off-peak electricity or from steam turbines.

15 Example 2

A schematic diagram of a multipressure stripping process according to embodiments of the present disclosure is shown in FIG. 4. Referring to FIG. 4, CO₂-rich absorbent solvent is fed, via line 87, through cross exchanger 88. The solvent is then fed, via line 67, through cross
20 exchanger 68. The rich absorbent is then fed, via line 69, to the top stage 70 of a multipressure stripping column, where the rich absorbent flows down the column via lines 71 and 73. The multipressure stripping column is a multistage flash, whereby the lowest flash stage 74 provides stripping steam by flashing the absorption solvent. The stripping steam then contacts the rich absorbent liquid in each flash stage and strips the CO₂ from the rich absorbent liquid. The total
25 vapor flow produced in lowest stage 74 is fed through compressor 76 via line 75. The compressed vapor exiting compressor 76 is fed to the bottom of intermediate flash stage 72 via line 77. The total vapor flow produced in intermediate flash stage 72 is fed through compressor 79 via line 78. The compressed vapor exiting compressor 79 is fed to the bottom of top flash stage 70 via line 80. The CO₂-rich compressed vapor exiting top flash stage 70 exits the system
30 via line 81.

A portion of the lean absorbent liquid exiting lowest flash stage 74 via line 82 is recycled to lowest flash stage 74. Thus, some of the lean absorbent withdrawn via line 82 is fed to cross exchanger 84 via line 83, where it is then fed to lowest flash stage 74 via line 85. The remainder

of the lean absorbent liquid exiting lowest flash stage 74 via line 82 that is not recycled to lowest flash stage 74 is fed to cross exchanger 88 via line 86. The lean absorbent liquid exiting cross exchanger 88 is then recycled back to the absorption stage of the process via line 89.

5 The stripper configuration depicted in FIG. 4 was simulated assuming equilibrium contacting in a spreadsheet simulation with a hypothetical solvent. The nonlinear equilibrium of the system was represented by the simple equation:

$$\ln P_{\text{CO}_2} = a + B \cdot \ln g - \Delta H/RT$$

10 The constant B was selected to give the specified capacity ($m = \text{gmol/kg water}$) for CO_2 absorption at the nominal absorber temperature, T. The heat of absorption, ΔH , was also varied to simulate different solvents; a value of 22 kcal/gmol is typical of monoethanolamine (MEA); a value of 15 kcal/gmol would represent an alternative solvent. The heat rate, Q, was provided at a maximum temperature of 120°C with a 10°C driving force. The equivalent work of the heat was calculated from the Carnot equation:

$$\text{Equivalent Work of Steam} = Q (T_{\text{steam}} - 313)/T_{\text{steam}}$$

15 The approach to equilibrium at the rich and lean end of the stripper was taken to be 33% (typical of MEA systems) and 75% (possible with reactive alternative solvent). Table 1 shows the results of the aforementioned spreadsheet simulation.

TABLE 1. Simulated Energy Requirements for Multipressure Stripper

ΔH (kcal/gmol)	Approach (%)	Q (kcal/gmol)	Compressor Work (kcal/gmol)	Equivalent Work of Steam (kcal/gmol)
15	33	32	2.2	8.9
15	75	28.6	1.4	7.3
22	33	34	1.5	7.5
22	75	34	1.0	6.8
MEA simple	33 rich, 5 lean	50	2.0	12

Parameters: 10% CO₂, 40° Absorber, 90% removal, compression to 8 atm CO₂
0.5 m capacity, 5°C cross-x approach, 4 contact stages (3 compressor stages)

As shown in Table 1, the multipressure stripper requires 6.8 to 8.9 kcal equivalent work/mole CO₂ removed. This is 25 to 40% less than the conventional MEA stripper (12 kcal/gmol). The actual compressor work to get a CO₂ product at 8 atm is 1.0 to 2.2 kcal/gmol. Even though the multipressure stripper requires some compression for water vapor, the compressor work required is not significantly different from the simple MEA stripper because much of the CO₂ is produced at higher pressure.

The net heat requirement for the multipressure stripper (29-34 kcal/gmol CO₂) is approximately the heat of absorption of the CO₂ plus 5°C sensible heat to drive the cross exchanger 88. It is significantly less than the net heat required (40-50 kcal/gmol) for the state-of-the-art MEA technology. Therefore, the compressor capital cost and power requirement may not be significantly different from the conventional MEA stripper, but the heat requirement may be significantly reduced. A closer approach to equilibrium in the absorber reduces the equivalent total work by 10 to 20%. The solvent with the lower heat of absorption appears to be 10 to 15 % better.

While the present invention may be adaptable to various modifications and alternative forms, specific embodiments have been shown by way of example and described herein. However, it should be understood that the present invention is not intended to be limited to the particular forms disclosed. Rather, it is to cover all modifications, equivalents, and alternatives falling within the spirit and scope of the invention as defined by the appended claims. Moreover, the different aspects of the disclosed equipment and methods may be utilized in various combinations and/or independently. Thus the invention is not limited to only those combinations shown herein, but rather may include other combinations, as well.

CLAIMS

1. An improved method for removing an acid gas from a gaseous stream containing same using aqueous absorption and stripping equipment, the improvement comprising:
 - (a) passing the acid gas-rich solvent stream exiting the absorbing equipment through a multipressure stripper in which the acid gas-rich solvent stream passes through multiple flash stages such that the vapor obtained by flashing the acid gas-rich solvent stream in each flash stage strips the acid gas from the acid gas-rich solvent stream, resulting in an acid gas-rich gaseous stream exiting the multipressure stripper and an acid gas-lean solvent stream exiting the multipressure stripper, wherein the multipressure stripper is operated at multiple pressure levels such that the vapor obtained in each flash stage is compressed in a compressor and fed to the previous flash stage at a higher pressure; and
 - (b) recycling the acid gas-lean solvent stream exiting the multipressure stripper in step (a) back to the absorption equipment at least once.
2. The method of claim 1, wherein each stage of the multipressure stripper is a countercurrent contactor.
3. The method of claim 1, wherein one or more of the stages of the multipressure stripper is a cocurrent contactor.
4. The method of claim 1, wherein one or more of the stages of the multipressure stripper is a mixed contactor.
5. The method of any one of claims 1 to 4, wherein the gaseous stream comprises flue gas, natural gas, hydrogen gas, or synthesis gas.
6. The method of any one of claims 1 to 4, wherein the acid gas comprises carbon dioxide.
7. The method of any one of the claims 1 to 4, wherein the acid gas comprises hydrogen sulfide.
8. The method of any one of claims 1 to 4, wherein the solvent is an aqueous amine, an aqueous alkanolamine, or mixtures thereof.
9. The method of any one of claims 1 to 4, wherein the solvent is an amine promoted aqueous potassium carbonate.
10. The method of any one of claims 1 to 4, wherein additional heat is supplied at one or more of the stages of the multipressure stripper.
11. An equipment for stripping an acid gas from an acid gas-rich solvent stream, comprising:
 - (a) a stripping column comprising multiple flash stages; and
 - (b) at least one compressor connected between each of the flash stages such that vapor produced in each flash stage is compressed and fed to the previous flash stage at a higher pressure.

12. The equipment of claim 11, wherein each stage of the stripping column is a countercurrent contactor.
13. The equipment of claim 11, wherein one or more of the stages of the stripping column is a cocurrent contactor.
- 5 14. The equipment of claim 11, wherein one or more of the stages of the stripping column is a mixed contactor.
15. The equipment of claim 11, wherein additional heat is supplied at one or more of the stages of the multipressure stripper column.

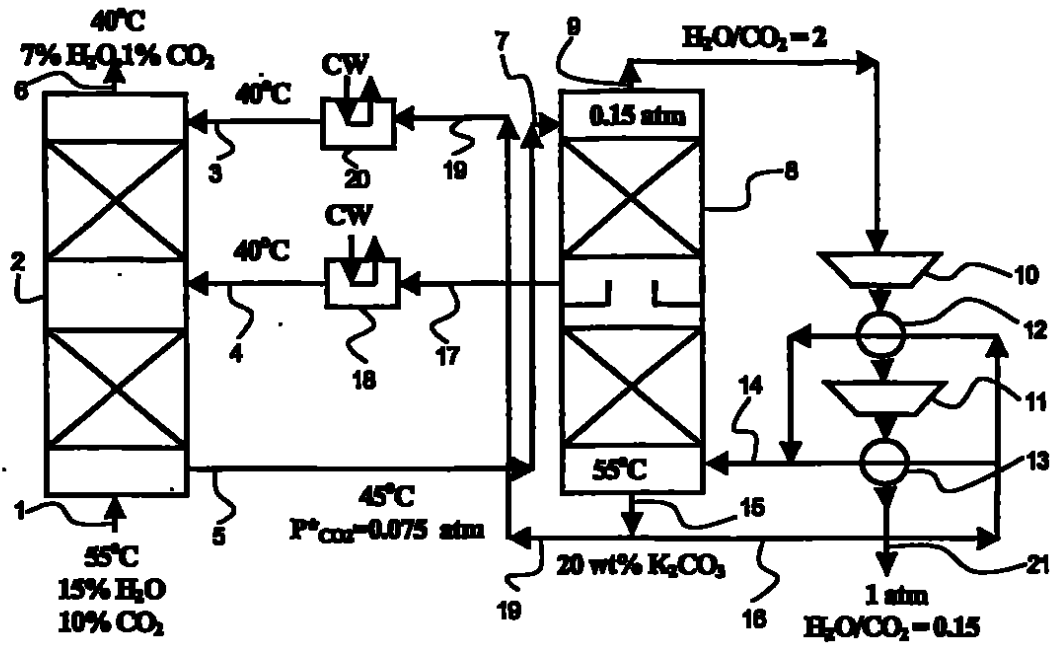


FIG. 1

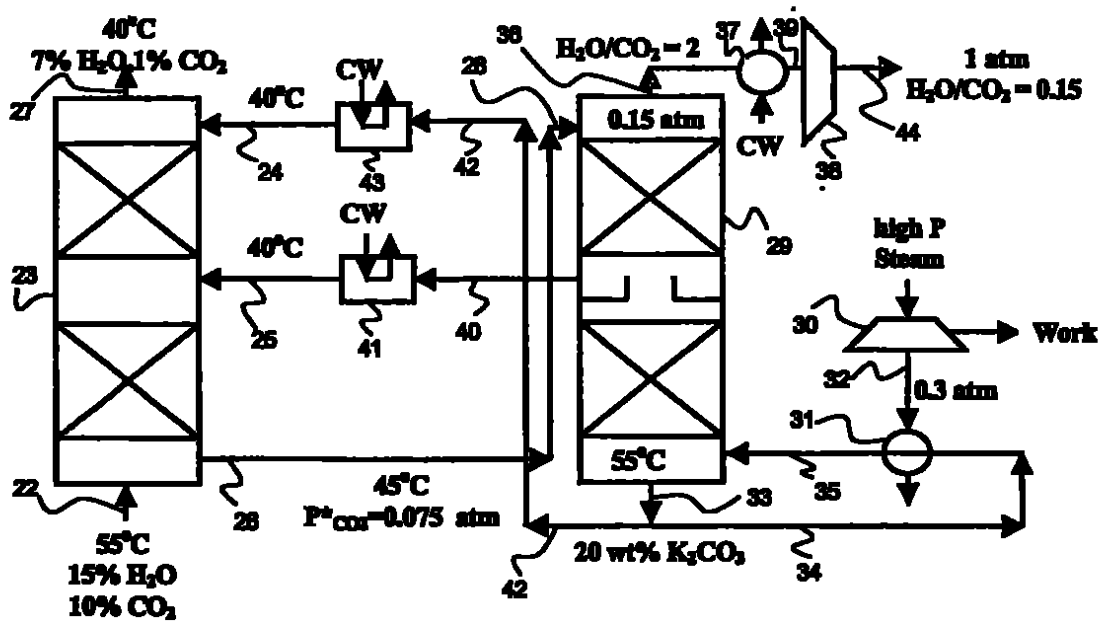


FIG. 2

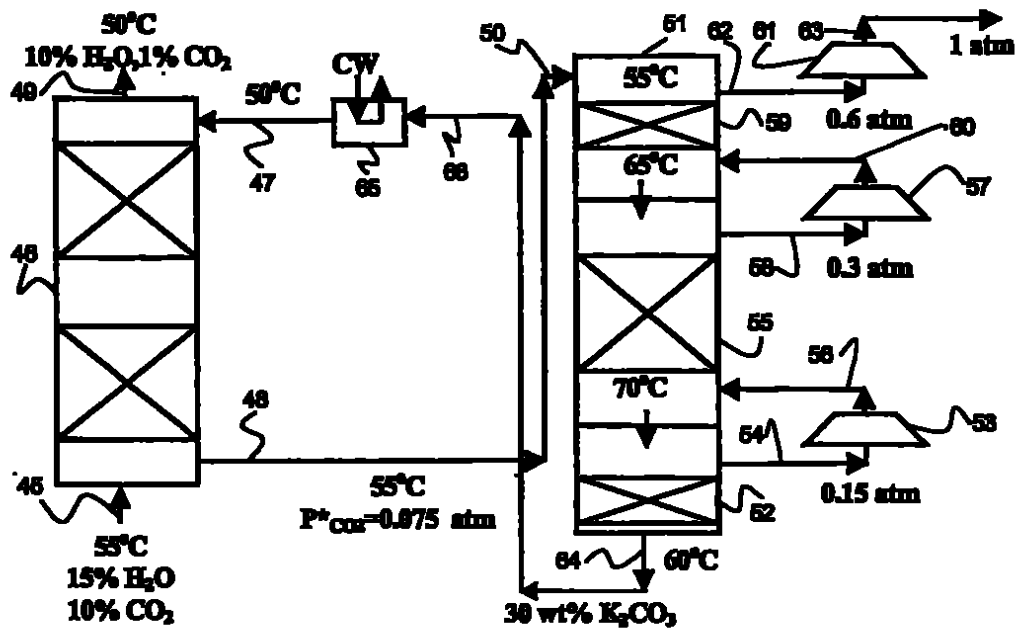


FIG. 3

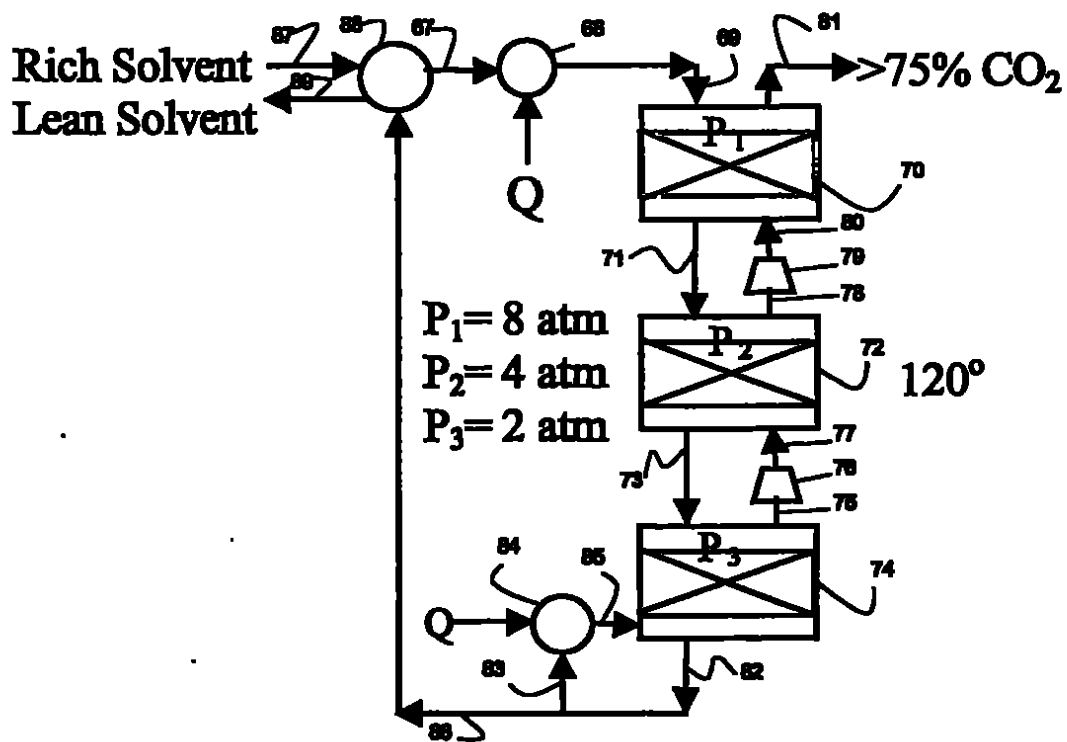


FIG. 4

INTERNATIONAL SEARCH REPORT

International Application No
PCT/US2004/006580

A. CLASSIFICATION OF SUBJECT MATTER
IPC 7 B01D53/14

According to International Patent Classification (IPC) or to both national classification and IPC

B. FIELDS SEARCHED

Minimum documentation searched (classification system followed by classification symbols)
IPC 7 B01D

Documentation searched other than minimum documentation to the extent that such documents are included in the fields searched

Electronic data base consulted during the international search (name of data base and, where practical, search terms used)
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C. DOCUMENTS CONSIDERED TO BE RELEVANT		
Category *	Citation of document, with indication, where appropriate, of the relevant passages	Relevant to claim No.
X	US 4 780 115 A (RANKE GERHARD) 25 October 1988 (1988-10-25)	1-7
Y	figure 4	8-15
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A	page 1, line 9 - line 52; figures 1-4	1,2,5,6, 10-15
	— — — — — -/-	

☒ Further documents are listed in the continuation of box C. ☒ Patent family members are listed in annex.

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Date of the actual completion of the international search 14 July 2004	Date of mailing of the international search report 26/07/2004
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PCT/US2004/006580

C.(Continuation) DOCUMENTS CONSIDERED TO BE RELEVANT

Category *	Citation of document, with indication, where appropriate, of the relevant passages	Relevant to claim No.
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Information on patent family members

International Application No

PCT/US2004/006580

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18 November 2004 (18.11.2004)Applicant's or agent's file reference
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IMPORTANT NOTICE

International application No.
PCT/US2004/006580International filing date (day/month/year)
04 March 2004 (04.03.2004)Priority date (day/month/year)
10 March 2003 (10.03.2003)

Applicant

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The International Bureau transmits herewith the following documents:

- ☐ copy of the international application as published by the International Bureau on under No. WO
- ☒ copy of international application as republished by the International Bureau on 18 November 2004 (18.11.2004) under No. WO 2004/080573
For an explanation as to the reason for this republication of the international application, reference is made to INID codes (15), (48) or (88) (as the case may be) on the front page of the attached document.

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(19) World Intellectual Property
Organization
International Bureau



(43) International Publication Date
23 September 2004 (23.09.2004)

PCT

(10) International Publication Number
WO 2004/080573 A1

- (51) International Patent Classification⁷: B01D 53/14
- (21) International Application Number:
PCT/US2004/006580
- (22) International Filing Date: 4 March 2004 (04.03.2004)
- (25) Filing Language: English
- (26) Publication Language: English
- (30) Priority Data:
60/453,279 10 March 2003 (10.03.2003) US
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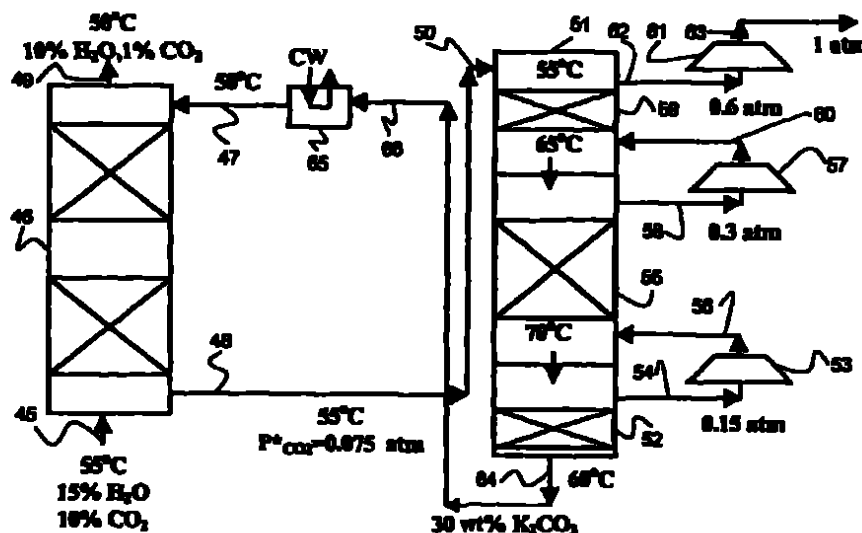
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(81) Designated States (unless otherwise indicated, for every
kind of national protection available): AE, AG, AL, AM,
AT, AU, AZ, BA, BB, BG, BR, BW, BY, BZ, CA, CH, CN,
CO, CR, CU, CZ, DE, DK, DM, DZ, EC, EE, EG, ES, FI,
GB, GD, GE, GH, GM, HR, HU, ID, IL, IN, IS, JP, KE,
KG, KP, KR, KZ, LC, LK, LR, LS, LT, LU, LV, MA, MD,
MG, MK, MN, MW, MX, MZ, NA, NI, NO, NZ, OM, PG,
PH, PL, PT, RO, RU, SC, SD, SE, SG, SK, SL, SY, TJ, TM,
TN, TR, TT, TZ, UA, UG, US, UZ, VC, VN, YU, ZA, ZM,
ZW.

(84) Designated States (unless otherwise indicated, for every
kind of regional protection available): ARIPO (BW, GH,
GM, KE, LS, MW, MZ, SD, SL, SZ, TZ, UG, ZM, ZW),
Eurasian (AM, AZ, BY, KG, KZ, MD, RU, TJ, TM), Euro-
pean (AT, BE, BG, CH, CY, CZ, DE, DK, EE, ES, FI, FR,
GB, GR, HU, IE, IT, LU, MC, NL, PL, PT, RO, SE, SI, SK,

[Continued on next page]

(54) Title: REGENERATION OF AN AQUEOUS SOLUTION FROM AN ACID GAS ABSORPTION PROCESS BY MULTI-STAGE FLASHING AND STRIPPING



(57) Abstract: An acid gas such as carbon dioxide, hydrogen sulfide, or a mixture thereof is removed from gaseous streams using aqueous absorption and stripping processes. By replacing the conventional stripper used to regenerate the aqueous solvent and capture the acid gas with a multipressure stripper (51) that combines acid gas compression with stripping, less energy is consumed. The multipressure stripper is a multistage flash (52, 55, 59) in which the total vapor flow from each stage is compressed and fed to the bottom of the previous flash stage at a higher pressure. In this process, the heat in the water content of the vapor exiting each stage is utilized at a higher pressure in the previous stage. The described stripping process generates the acid gas at a higher pressure without operating the stripper at a higher temperature, thereby reducing the energy consumption of the system.



TR), OAPI (BF, BJ, CF, CG, CI, CM, GA, GN, GQ, GW, ML, MR, NE, SN, TD, TG).

(15) Information about Correction:

see PCT Gazette No. 47/2004 of 18 November 2004, Section II

Published:

— with international search report

(48) Date of publication of this corrected version:

18 November 2004

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.



Expediente No		OS103104	No de Folio
Item		No de unidades	
1	CD		
2	DVD		
3	REVISTA		
4	LIBRO		
5	PERIODICO		
6	DISQUETES		
7	SOBRE DE MANILA	1	
8	SOBRE BLANCO TAMAÑO CARTA		
9	SOBRE BLANCO TAMAÑO OFICIO		
10	OTROS:		

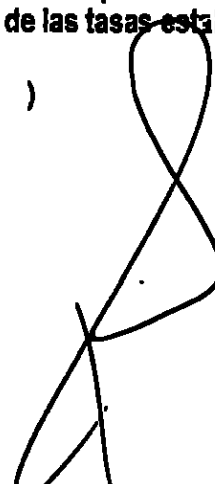
Observaciones: A la Final

REQUISITOS NECESARIOS PARA PRESENTACION Y AGILIZACION DE: TRAMITE

	USUARIO	RADICADOR
PATENTE DE INVENCION <i>RCT</i>	()	()
- Indicación de que se solicita la concesión de una patente.	()	(X)
- Datos de identificación del solicitante o de la persona que se presenta la solicitud.	()	(X)
- Descripción de la invención.	()	(X)
- Dibujos de ser estos pertinentes.	()	(X)
- Comprobante de Pago de las tasas establecidas.	()	()
COMPLETA (X)	INCOMPLETA ()	()
 DISEÑO INDUSTRIAL ()		
- Indicación de que se solicita el registro de Diseño Industrial	()	()
- Datos de identificación del solicitante o de la persona que presenta la solicitud.	()	()
- Representación gráfica o fotográfica del Diseño Industrial o muestra del Material que incorpora el diseño.	()	()
- Comprobante de pago de las tasas establecidas	()	()
COMPLETA ()	INCOMPLETA ()	()
 ESQUEMA DE TRAZADO ()		
- Indicación de que solicita el registro de un Esquema de trazado.	()	()
- Datos de identificación del solicitante o de la persona que presenta La solicitud.	()	()
- Representación gráfica del esquema de trazado	()	()
- Comprobante de pago de las tasas establecidas.	()	()
COMPLETA ()	INCOMPLETA ()	()

Firma Responsable:

Fecha:



700

REPUBLICA DE COLOMBIA
SUPERINTENDENCIA DE INDUSTRIA Y COMERCIO

Bogotá,
2020

Doctor(a)
RODRIGUEZ D'ALEMAN DILIA MARIA

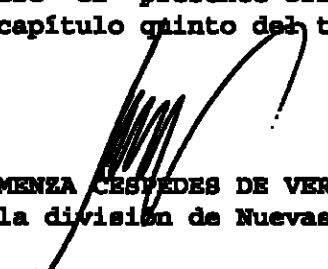
REFERENCIA	Radicación No.	5 103104
	Solicitud internacional	US2004/006580
	Fecha inicio fase nacional	10/10/2005
	Trámite	11
	Evento	379
	Actuación	430
	Oficio No.	10424

Como resultado del examen de forma, realizado con fundamento en el artículo 38 de la Decisión 486 de la Comisión de la Comunidad Andina, artículo 27 del Tratado de Cooperación en Materia de Patentes (PCT), regla 51 bis del Reglamento y Circular Única de la Superintendencia de Industria y Comercio, se le comunica que:

1. Si fuere el caso, el solicitante debe adjuntar la traducción de los anexos del reporte de examen preliminar internacional (Art. 36.2.b) del tratado PCT.
2. La solicitud no contiene la copia del documento en que consta la cesión del derecho a la patente del inventor al solicitante o a su causante. (Art. 26-k) Decisión 486.
3. La solicitud no contiene el poder debidamente otorgado, con el lleno de los requisitos exigidos por los artículos 63 y subsiguientes del Código de Procedimiento Civil. Adicionalmente, deberá presentar el documento que acredita la existencia y representación legal de la sociedad solicitante, cuando ésta fuere persona jurídica.

En cumplimiento de lo dispuesto por el artículo 39 de la Decisión 486, se le requiere para que dentro del término de dos meses siguientes a la fecha de notificación del presente oficio, complete los requisitos anteriormente enunciados. En caso de no dar cumplimiento al presente requerimiento, la solicitud se considerará abandonada y perderá su prelación.

Notifíquese el presente oficio conforme a lo dispuesto en el numeral 5.2, literal e), del capítulo quinto del título primero de la Circular Única No.10 de 2001.


ALIX CARMONA CESPEDES DE VERGEL
Jefe de la división de Nuevas Creaciones

El anterior requerimiento fue notificado por fijación en lista de fecha 04 Nov. 2020
adpall