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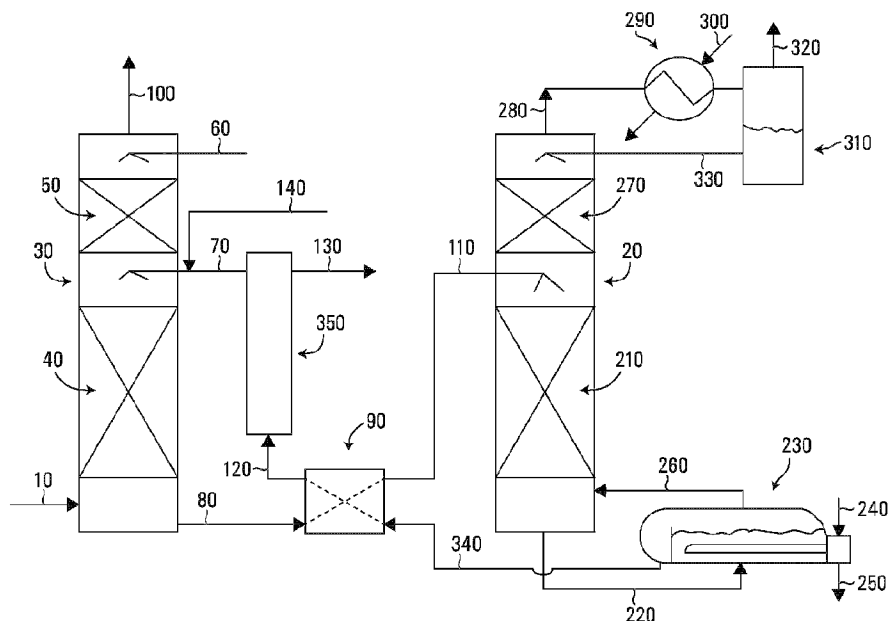
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(54) Title: PROCESS FOR CAPTURING SULFUR DIOXIDE FROM A GAS STREAM



(57) **Abrégé/Abstract:**

A process for removing sulfur dioxide from a feed gas stream, which comprises (i) contacting the feed gas stream with an aqueous lean absorbing medium comprising a chemical solvent comprising a regenerable absorbent, a physical solvent, and one or more heat stable salts. The regenerable absorbent is an amine. The ratio of the wt% of the physical solvent over that of the regenerable absorbent is 0.5 to 2.5. The ratio of the wt% of heat stable salts over that of the regenerable absorbent is 0.29 to 0.37. The pH of the lean absorbing medium is 6 or less. With the process SO₂ can be selectively removed. When the absorbing medium is stripped, a reduced amount of energy is required as compared to known processes.

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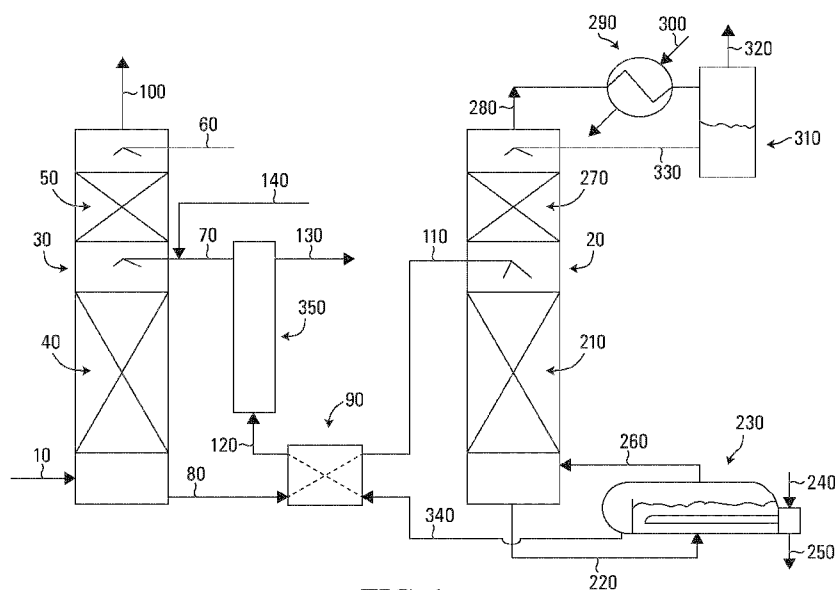


FIG. 1

(57) Abstract: A process for removing sulfur dioxide from a feed gas stream, which comprises (i) contacting the feed gas stream with an aqueous lean absorbing medium comprising a chemical solvent comprising a regenerable absorbent, a physical solvent, and one or more heat stable salts. The regenerable absorbent is an amine. The ratio of the wt% of the physical solvent over that of the regenerable absorbent is 0.5 to 2.5. The ratio of the wt% of heat stable salts over that of the regenerable absorbent is 0.29 to 0.37. The pH of the lean absorbing medium is 6 or less. With the process SO₂ can be selectively removed. When the absorbing medium is stripped, a reduced amount of energy is required as compared to known processes.

- 1 -

PROCESS FOR CAPTURING SULFUR DIOXIDE FROM A GAS STREAM

FIELD

The present invention relates to a process for capturing sulfur dioxide (SO_2) from a feed gas stream. The present invention especially relates to a process
5 suitable to selectively capture sulfur dioxide (SO_2) from a feed gas stream, more especially to remove SO_2 from a gas stream while not at the same time removing CO_2 from the gas stream.

BACKGROUND

10 It is known that SO_2 is more soluble in water than many other components of feed gas streams. For example, measured at 1.013 bar 0°C , the solubility of SO_2 in water is 228 g/L whereas the solubility of carbon dioxide and hydrogen sulfide in water is 3.369 g/L and 7.100 g/L,
15 respectively.

The solubility of SO_2 in many other pure solvents has also been widely studied. See, for example, Fogg and Gerrard, 1991 (Solubility of Gases in Liquids, John Wiley and Sons, Chichester, U.K.) for a summary of the
20 literature solubility data of SO_2 .

Regenerable absorbents can be used to remove SO_2 from feed gas streams. Typically, a lean aqueous medium comprising the absorbent is exposed to a SO_2 containing feed gas stream, and then SO_2 is absorbed by the medium
25 producing a SO_2 lean gas stream and a spent absorbing medium. Removal (recovery) of the absorbed SO_2 from the spent absorbing medium to regenerate the aqueous medium and to provide gaseous SO_2 is typically effected by gaseous stripping using steam generated in situ.

- 2 -

Amine-based absorbents can be used for SO₂ removal. See, for example, US5019361 which discloses the use of an aqueous absorbing medium containing a water-soluble half salt of a diamine. US7214358 discloses the use of an aqueous absorbing medium containing a water-soluble half salt of a diamine and an elevated level of heat stable salts (HSS). Physical solvents can also be used as SO₂ absorbents.

Commercially available steam-regenerable SO₂ capture technologies include those that rely on chemical solvents or physical solvents, such as Cansolv DSTM (amine-based absorbent-containing chemical solvent), LabsorbTM (inorganic absorbent-containing chemical solvent), ClausMasterTM (non-aqueous physical solvent), and Sea water process (chemical solvent).

Use of a combination of solvents has also been disclosed.

Indian Patent Application No. 2381/DEL/2006 describes a process for the removal of SO₂ using a solvent blend comprising chemical and physical solvents.

US20130039829 describes a process for the capture of sulfur dioxide from a gaseous stream utilizing a regenerable diamine absorbent comprising a diamine and a weak organic acid, such as formic acid.

The energy required for regenerating absorbing medium in a SO₂ removal process, in particular the energy required for stripping absorbed SO₂ from absorbing medium, accounts for a significant portion of the operating cost of SO₂ removal from a feed gas. For example, the net present value of the existing SO₂ capture technologies is strongly dependent on steam cost. Therefore, there remains a need to reduce regeneration

energy consumption in processes for removing SO₂ from a gas stream.

DESCRIPTION

The invention relates to a process for removing sulfur dioxide from a feed gas stream, which process comprises:
5 (i) contacting the feed gas stream with an aqueous lean absorbing medium to absorb sulfur dioxide and to form a sulfur dioxide lean treated gas stream and a spent absorbing medium;

10 wherein the aqueous lean absorbing medium comprises:

(a) a chemical solvent comprising a regenerable absorbent,

(b) a physical solvent, and

(c) one or more heat stable salts;

15 wherein the regenerable absorbent is an amine, preferably a mono amine, a diamine, a polyamine, or a mixture thereof, most preferably a diamine;

wherein the ratio of the weight percentage of the physical solvent in the lean absorbing medium over that of the
20 regenerable absorbent is in the range of from 0.5 to 2.5, preferably from 1.1 to 2.2;

wherein the ratio of the weight percentage of heat stable salts in the lean absorbing medium over that of the
regenerable absorbent is in the range of from 0.29 to 0.37,
25 preferably from 0.31 to 0.34; and

wherein the pH of the lean absorbing medium is 6 or less, preferably 5.6 or less, more preferably in the range of from 4.5 to 5.6, even more preferably in the range of from 5.2 to 5.6.

In accordance with one aspect there is provided a process for removing sulfur dioxide from a feed gas stream, which process comprises:

(i) contacting the feed gas stream with an aqueous lean absorbing medium to absorb sulfur dioxide and to form a sulfur dioxide lean treated gas stream and a spent absorbing medium;

wherein the aqueous lean absorbing medium comprises:

(a) a chemical solvent comprising a regenerable absorbent, wherein the regenerable absorbent is a diamine or polyamine which in half salt form has a pK_a value for the free nitrogen atom of 3.0 to 5.5, at a temperature of 20°C in an aqueous medium;

(b) a physical solvent, wherein the physical solvent has a vapour pressure less than 0.1 mmHg at 20°C with a boiling point equal to or higher than 240°C; and

(c) one or more heat stable salts;

wherein the ratio of the weight percentage of the physical solvent in the lean absorbing medium over that of the regenerable absorbent is in the range of from 0.5 to 2.5; wherein the ratio of the weight percentage of heat stable salts in the lean absorbing medium over that of the regenerable absorbent is in the range of from 0.29 to 0.37; and

wherein the pH of the lean absorbing medium is in the range of from 4.5 to 5.6;

(ii) stripping absorbed sulfur dioxide from the spent absorbing medium to produce a regenerated aqueous absorbing medium and a gaseous sulfur dioxide;

and optionally

(iii) recycling the regenerated aqueous absorbing medium from step (ii) to step (i).

With the process of the current invention SO₂ can be removed selectively, that is, SO₂ is removed from gas. CO₂ and other components are not or hardly removed from

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- 4 -

the gas. Furthermore, during step (i) the absorbing medium is present in a single liquid phase. No liquid-liquid phase separation takes place.

5 The process preferably comprises the following additional steps:

(ii) stripping, preferably steam stripping, absorbed sulfur dioxide from the spent absorbing medium to produce a regenerated aqueous absorbing medium and a gaseous sulfur dioxide;

10 and optionally

(iii) recycling the regenerated aqueous absorbing medium from step (ii) to step (i).

During step (ii) the absorbing medium is present in a single liquid phase. No liquid-liquid phase separation
15 takes place.

In some embodiments, the absorbing medium comprises at least 14wt% of the physical solvent. The absorbing medium may comprise up to 35wt% of the physical solvent.

In some embodiments, the sulfur dioxide amine
20 absorbent is a mixture of 4-[hydroxyethyl]piperazine (Hep) and 1,4-bis[hydroxyethyl]piperazine (DiHep). In some embodiments, the lean absorbing medium may be an aqueous medium comprising 18wt% of a mixture of Hep and DiHep, 1.2 eq./mole HSS and 14wt% PEGDME. In some
25 embodiments, the lean absorbing medium may be an aqueous medium comprising 13wt% of a mixture of Hep and DiHep, 1.2 eq./mole HSS and 17wt% PEGDME.

In some embodiments, the sulfur dioxide amine absorbent is 2-piperazinone 1,4-bis[2-hydroxyethyl]
30 (Amide-DiHep). In some embodiments, the lean absorbing medium may be an aqueous medium comprising 25wt% Amide-DiHep, 0.35 eq./mole HSS and 20wt% PEGDME.

- 5 -

In some embodiments, the sulfur dioxide amine absorbent is 3-aminopyrazole. In some embodiments, the lean absorbing medium may be an aqueous medium comprising 22wt% 3-aminopyrazole, 0.1 eq./mole HSS and 32wt% PEGDME.

5 In some embodiments, the step of stripping absorbed sulfur dioxide may use steam. With the current invention it proved to be possible to use less steam during stripping than in a corresponding step in a process that does not use a physical solvent. This results in a
10 significant energy reduction.

In some embodiments, the processes as described herein may further comprise a step of removing heat stable salts from the regenerated aqueous absorbing medium before recycling the regenerated aqueous absorbing
15 medium. The step of removing heat stable salts may comprise using a weak base anion resin, ion pairing, crystallisation or precipitation.

In some embodiments, the processes as described herein further comprise a step of recovering the gaseous
20 sulfur dioxide.

With the process of the invention a pure SO₂ stream can be obtained that can be used for sulfuric acid make, or for use in a sulfur reduction unit in a Claus application. The pure SO₂ stream is not or hardly
25 contaminated with CO₂ or mercaptans which would contaminate sulfuric acid, or which would contaminate a Claus unit.

Drawings

Figure 1 shows a flow diagram of a regenerative SO₂ removal process according to one invention embodiment.
30

Figure 2 shows regeneration energy consumptions of hybrid solvents and an amine-based solvent according to an embodiment of the invention.

- 6 -

Figure 3 shows SO₂ emission levels of the different tested solutions in Example 5 (Test 2) (tested at 1.5% pp SO₂ and liquid/gas ratio (L/G) of 0.98 L/Nm³).

Figure 4 shows SO₂ delta loading in solution vs. liquid/gas ratio in absorber of the various tested solutions in Example 5 (Test 2).

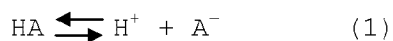
Figure 5 shows regeneration energy consumptions of Solutions A, B and C in Example 5 (Test 2) (tested at 1.5% pp SO₂).

Figure 6 shows regeneration energy consumptions of Solutions B and B.1 in Example 6 (Test 3).

SO₂ removal

In general, a suitable indicator for an appropriate choice of absorbent (e.g., a chemical solvent) for use in the capture of a given gaseous acid gas contaminant (such as SO₂) in a feed gas is the difference in the pK_a values between the acid gas in water and the absorbent.

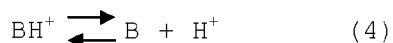
The pK_a of an acid is defined as the negative logarithm to the base 10 of the equilibrium constant K_a for the ionization of the acid HA (e.g., H₂SO₃), where H is hydrogen and A is a radical capable of being an anion:



$$K_a = [\text{H}^+][\text{A}^-]/[\text{HA}] \quad (2)$$

$$\text{pK}_a = -\log_{10} K_a \quad (3)$$

For a basic absorbent B, the pK_a is for the ionization reaction of the conjugate protonated acid of B, the species BH⁺:



- 7 -

The reaction involved in the absorption of the acid gas contaminant HA by the basic absorbent B can be shown as follows:

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Reaction (5) is reversible:

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When SO₂ is dissolved in water, following reaction (1), bisulphite ions (HSO₃⁻) and protons are formed. The proton may be ionically associated with the absorbent (for example, when an amine-based absorbent is used, the proton may be ionically associated with the sorbing nitrogen of the absorbent). The absorbed SO₂/desorbed SO₂ equilibrium is illustrated in the above reaction (6). Absorbed SO₂ can be "stripped" from the spent absorbing medium as gaseous SO₂, for example and without limitation, by the application of steam. In this stripping process, desorbed SO₂ is released from the spent absorbing medium. "Stripping" is used herein to broadly encompass removal of absorbed SO₂ from the spent absorbing medium, and should be understood as also, more specifically, encompassing releasing desorbed SO₂ from the spent absorbing medium.

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It has been found that contacting a feed gas stream with a lean absorbing medium comprising a chemical solvent and a physical solvent may reduce the energy consumption for stripping absorbed SO₂ from spent absorbing medium, or may reduce the energy consumption for releasing desorbed SO₂ from spent absorbing medium,

- 8 -

thereby reducing regeneration energy consumption in a process for removing SO₂ from the gas stream. The reduction of regeneration energy consumption achieved by the methods of the invention is understood to be relative to a method that does not use a physical solvent.

As used herein, regeneration energy relates to the amount of energy required to regenerate an absorbing medium used to absorb SO₂ in a process for removing SO₂ from a feed gas stream. The absorbing medium, according to the invention, comprises a chemical solvent and a physical solvent.

Stripping

Preferably step (ii) is performed in a reboiler. More preferably step (ii) is performed in a kettle reboiler, forced circulation reboiler, fired reboiler, falling film reboiler, direct steam reboiler, or thermosyphon, most preferably in a thermosyphon.

The reboiler may be heated by hot oil, electricity or steam, preferably steam. Alternatively, direct steam addition can be utilized.

Preferably at least 97 vol%, more preferably at least 99 vol%, even more preferably at least 99.9 vol% of the spent absorbing medium formed in step (i) is stripped, preferably steam stripped, in step (ii).

Chemical Solvent

Chemical solvents for use in the invention comprise a regenerable absorbent that selectively absorbs SO₂. In some embodiments, the chemical solvent comprises an aqueous medium and the absorbent.

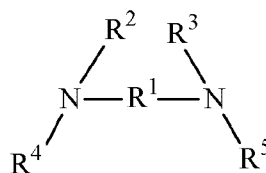
In general, a suitable chemical solvent may have one or more of the following properties: high capacity for the absorption of SO₂; ready and substantially complete release of absorbed SO₂; little tendency to cause

- 9 -

oxidation of SO₂; low heat of absorption; high boiling point; low specific heat; and high stability at temperatures required for the release of SO₂.

The chemical solvent is or comprises an amine. The amine may be a mono amine, a diamine, a polyamine, or a mixture thereof. Suitable amines include, but are not limited to, 1,4-bis[hydroxyethyl]piperazine, 4-[hydroxyethyl]piperazine, 1,4-diazabicyclo[2.2.2]octane (DABCO), 2-[2-aminoethyl]pyridine, 2-aminomethylpyridine, 3-amino 5-methylpyrazole, 3-aminopyrazole, 3-methylpyrazole, N,N,N',N'-tetraethyldiethylenetriamine, N,N,N',N'-tetramethyldiethylenetriamine, 2-piperazinone 1,4-bis[2-hydroxyethyl], or a combination thereof.

The amine-based absorbent may be a diamine represented by the structural formula:

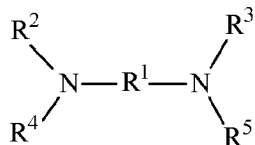


wherein R¹ is an alkylene of two or three carbon atoms as a straight chain or as a branched chain, R², R³, R⁴, and R⁵ may be the same or different and can be hydrogen, alkyl (e.g., lower alkyl of 1 to 8 carbon atoms including cycloalkyls), hydroxyalkyl (e.g., lower hydroxy alkyl of 2 to 8 carbon atoms), aralkyl (e.g., 7 to 20 carbon atoms), aryl (may be, for example, monocyclic or bicyclic), or alkaryl (e.g., 7 to 20 carbon atoms), and any of R², R³, R⁴, and R⁵ may form cyclic structures.

The diamines may also be tertiary diamines. For instance, the tertiary diamine may be of the formula:

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- 10 -



wherein R¹ is as defined above, and R², R³, R⁴, and R⁵ are as defined above with the exception that none are hydrogen. In an exemplary embodiment, each of R², R³, R⁴, and R⁵ is the same or different and is an alkyl group (e.g., methyl or ethyl) or a hydroxy-alkyl group (e.g., 2-hydroxyethyl).

Other diamines in which one or both of the nitrogen atoms is primary or secondary and which otherwise meet the parameters discussed herein may also be suitable, provided mild oxidative or thermal conditions exist to minimize side reactions of the solvent, including oxidation.

Suitable diamines, according to some invention embodiments, have one amine with a lower pK_a and the other amine with a higher pK_a wherein the higher pK_a is above 6.5 and, in some instances, above 7.5 and the lower pK_a is less than 5.0 and, in some instances, less than 4.0. The stronger amine (the one with the higher pK_a) may react to form heat stable salts (HSS). For instance, the stronger amine may react with a strong acid (e.g., sulfuric acid) to obtain a HSS. In some embodiments, the lean amine-based absorbent, which is exposed to the gas stream, is therefore in its half-salt form. Accordingly, only the weaker, more moderate amine is available for reacting with the feed gas stream and releasably absorbing SO₂.

In some embodiments, the diamine in half salt form has a pK_a value for the free nitrogen atom of 3.0 to 5.5 and, in some instances, 3.5 to 4.7 at a temperature of

- 11 -

20 °C in an aqueous medium. The free amine form of the amine salt absorbent may have a molecular weight less than 300 g/mol and, in some instances, less than 250 g/mol.

5 In some embodiments, the amine salt absorbents have a hydroxyalkyl group as a substituent on an amine group. Without being limited by theory, it is believed that a hydroxy substituent may increase the solubility of the amine salt absorbents in water. Without being limited by
10 theory, it is further believed that a hydroxy substituent may retard the oxidation of sulphite or bisulphite to sulfate, which can result in the formation of HSS. As discussed below, it may be desirable to minimize the formation of HSS.

15 Suitable diamine compounds may include, but are not limited to, N,N',N'-(trimethyl)-N-(2-hydroxyethyl)-ethylenediamine ($pK_a=5.7$); N,N,N', N'-tetrakis(2-hydroxyethyl)ethylenediamine ($pK_a=4.9$); N,N'-dimethylpiperazine ($pK_a=4.8$); N,N,N',N'-tetrakis(2-
20 hydroxyethyl)-1,3-diaminopropane; and N',N'-dimethyl-N,N-bis(2-hydroxyethyl)ethylenediamine. Useful diamines may also comprise, in some embodiments, heterocyclic compounds, such as piperazine ($pK_a=5.8$), N-(2-hydroxyethyl)piperazine, N,N'-di(2-
25 hydroxyethyl)piperazine and 1,4-diazabicyclo[2.2.2]octane ($pK_a=4.9$). The pK_a values identified in the brackets are for the weaker, sorbing nitrogen.

According to some embodiments of the invention, the diamine may be selected from the group comprising
30 hydroxyethyl piperazine, bis-hydroxyethyl piperazine, piperazine, hydroxyethylethylenediamine, bis-hydroxyethylethylenediamine and mixtures thereof. For example, the diamine may comprise 1,4-

- 12 -

bis[hydroxyethyl]piperazine, 4-[hydroxyethyl]piperazine, or a combination thereof.

Without being limited by theory, it is believed that use of amine-based absorbents that generate HSS at a
5 controllable low level may permit an increase in the concentration of the physical solvent in absorbing medium of the present invention, and maintain a one-phase solution of the absorbing medium. One example of such an amine is 2-piperazinone 1,4-bis(2-hydroxyethyl) (Amide-
10 DiHep).

It will be appreciated that, in some embodiments, one or more amines may be used as the absorbent and one or more amines may be used with other heat regenerable sulfur dioxide absorbents.

15 The amine-based absorbent may be in an amount sufficient to provide a spent absorbing medium containing at least 180 grams of SO₂ per kilogram of absorbing medium. The amount of amine-based absorbent, however, may not be so great as to either (a) unduly increase the
20 viscosity of the absorbing medium such that undesirable pressure drops are incurred in the feed gas stream passing through an absorber vessel or (b) render the absorbing medium difficult to atomize in, for example, a Waterloo scrubber.

25 In some other embodiments, the chemical solvent may comprise an organic acid. The organic acid may have a pKa such that, at the pH of the lean aqueous medium, the organic acid is substantially in its basic form and, at the pH of the spent absorbent medium, the organic acid is
30 substantially in its acidic form. For example, if the organic acid is formic acid, then at the pH of lean absorbent stream, the formic acid is present as formate and, at the pH of the spent absorbing medium (SO₂ rich

- 13 -

absorbent stream), the organic acid is substantially in the form of formic acid. By substantially, it is meant that at least 30% or, in some instances, at least 50%, of the organic acid is in the particular form at the specified pH.

The organic acid may have a pK_a of 1.2-6 and, in some instances, 3.5-5.5.

The organic acid may comprise one or more of formic acid, acetic acid, glycolic acid, malonic acid, propanoic acid, succinic acid, phthalic acid, citric acid, adipic acid, tartaric acid, malic acid, and oxalic acid. In some embodiments, the organic acid comprises one or more of formic acid, acetic acid, malonic acid, malic acid, tartaric acid, citric acid, and adipic acid.

The chemical solvent may comprise a mixture of amine based absorbent and organic acid as described above.

Physical Solvent

Physical solvents for use in the invention may have one or more of the following characteristics: low volatility; water solubility; and low heat capacity.

The physical solvent may have a vapour pressure less than 0.1 mmHg at 20°C with a boiling point equal to or higher than 240°C.

Suitable physical solvents include, but are not limited to, a polyol, a polycarbonate, an N-formyl morpholine, or a combination thereof. The polyol may be a polyethylene glycol or an ether thereof, for instance, of the formula $R^6-O-(C_2H_4O)_n-R^7$, wherein n is 3 to 12, R^6 is hydrogen or lower alkyl (e.g., C_{1-8} alkyl), R^7 is hydrogen or lower alkyl (e.g., C_{1-8} alkyl), or R^6 is C_{6-10} aryl (e.g., phenyl) and R^7 is hydrogen or lower alkyl (e.g., C_{1-8} alkyl). For example, the physical solvent may be polyethyleneglycol dimethylether (PEGDME),

- 14 -

tetraethyleneglycol dimethylether (TetraEGDME),
triethyleneglycol monomethylether (TriEGMME),
tetraethylene glycol (TetraEG), or a combination thereof.

5 The miscibility of some physical solvents such as
PEGDME in the absorbing medium may be affected by the
concentration of the regenerable absorbent and/or the
amount of HSS. For example, the miscibility of PEGDME in
an aqueous diamine solution decreases as the
concentration of diamine increases and as the amount of
10 HSS increases. In some embodiments, it may be desirable
to lower the concentration of the regenerable absorbent
and/or the amount of HSS in order to increase the amount
of miscible physical solvents in the absorbing medium.

When the physical solvent is PEGDME, without being
15 limited by theory, it is believed that reduction of the
amount of HSS and increase of the concentration of PEGDME
may reduce hydrogen bonding and/or increase ether-sulfur
bridges in the absorbing medium, which renders the
absorbing medium more aprotic and potentially reduces the
energy consumption for stripping absorbed SO₂.
20

Without being limited by theory, it is believed that
physical solvents may reduce the energy consumption
required for releasing desorbed SO₂. Physical solvents
may compete with other components of spent absorbing
25 medium to attract SO₂. Physical solvents may further
reduce hydrogen bonding between SO₂ and spent absorbing
medium. Physical solvents may also reduce the polarity of
SO₂ in spent absorbent medium or make the medium more
aprotic. Physical solvents may even further change the
30 surface tension of spent absorbent medium.

Heat Stable Salts (HSS)

HSS may accumulate in the medium due to, for example,
sulfite/bisulfite oxidation or disproportionation, or due

- 15 -

to the absorption of acid mist from the feed gas. These salts are too stable to decompose under normal steam conditions for stripping SO₂ from spent absorbing medium. Examples of such heat stable salts are those salts that are formed from strong acids such as sulfuric acid, nitric acid, or hydrochloric acid. If allowed to accumulate, these heat stable salts would eventually completely neutralize the SO₂ absorption capacity of the absorbent. Therefore, management of HSS in the solution may be an important part of the SO₂ removal process to maintain performance over time

The amount of HSS formed may be affected by the absorbent used and/or the concentration of the absorbent. The amount of HSS for an absorbing medium may be controlled by using conventional means, such as an ion exchange resin, eletrodialysis unit or crystallization. Amine purification units (APU) that are currently used industrially utilize weak anionic resins capable of some selectivity between sulfate (a strong conjugated base) and weaker conjugated bases in the absorbing medium. The performance of such weak base resins varies depending on the concentration of sulfate in solution. These resins do not always perform well if there is a low concentration of HSS.

Ways to control the level of HSS for an organic acid/physical solvent mixture may also include ion exchange with cyclo[8]pyrrole as the functional groups or by crystallization of alkaline sulfate salts (e.g. Na₂SO₄), where the cation can be sodium or potassium, most often sodium. Another way of controlling the level of HSS in the organic acid/physical solvent mixture is precipitation of Ettringite (Ca₆Al₂(SO₄)₂(OH)₁₂•26H₂O).

- 16 -

In the alternative, HSS could also be removed by ion pairing. Without being limited by theory, it is believed that a low HSS amount in the absorbing medium, in accordance with some embodiments of the invention, may
5 reduce the efficiency of the exchange of HSS with a standard anionic weak base resin. In some embodiments, it may therefore be desirable to remove HSS by ion pairing, which may permit a higher rate of removal of HSS even when the amount of salts in solution is low. Ion pairing
10 may be achieved, for example, by using a dual function resin having different ionic functional groups (such as a combination of phenol and quaternary amine functional groups) or by liquid-liquid extraction.

Without being limited to theory, it is believed that
15 a strong base quaternary amine functional group insensitive to suppressed salt concentrations will attract opposite charged anions regardless of their type. During regeneration, the phenolic functional group which is the active exchange site in the above described dual
20 function resin, becomes negatively charged at a pH greater than 10.5, and repels the like charged anions.

The Absorbing Medium

The absorbing medium comprises the physical solvent and the chemical solvent. The absorbing medium is a one
25 phase solution during step (i) and during step (ii). It is aqueous.

The pH of the lean absorbing medium is 6 or less, preferably 5.6 or less, more preferably in the range of from 4.5 to 5.6. Even more preferably the pH of the lean
30 absorbing medium before contacting the feed gas is controlled in the range from 5.2 to 5.6.

The absorbing medium may contain at least one mole of water and usually more for each mole of SO₂ to be removed

- 17 -

from the gas stream. The water acts both as a solvent for the amine salt and as for a reactant to produce "sulfurous acid" H_2SO_3 from the SO_2 . The proportion of water present may be up to 95 weight percent of the absorbing medium and, in some instances, 60 to 90 weight percent of the absorbing medium.

The lean absorbing medium may comprise an organic acid and/or an anorganic acid, preferably an anorganic acid, more preferably one or more acids chosen from the group of nitric acid (HNO_3), hydrochloric acid (HCl), sulfuric acid (H_2SO_4) and sulfurous acid (H_2SO_3), even more preferably sulfuric acid (H_2SO_4) and/or sulfurous acid (H_2SO_3).

The viscosity of the absorbing medium may be below 1200 centipoise at 25°C , e.g., between 1 and 500 centipoise, and more specifically between 1 and 50 centipoise, at 25°C . Frequently, the solubility of the amine salt absorbent in water may be at least 0.01, often at least 0.1, mole per liter at 25°C . In some embodiments, the amine salt absorbent is miscible with water under the conditions in the process. However, the amine salt absorbent and water does not have to be miscible under the conditions of the process, nor does the amine salt absorbent have to be liquid under the conditions of the process.

In some embodiments, anti-foam agents known in the art may be used to reduce the foaming tendency of the mixture of the chemical solvent and the physical solvent. Anti-foam agents and amounts can be chosen and optimized in accordance with known practices. It may be desirable to choose an anti-foam agent compatible with the system chosen for HSS removal (e.g., compatible with an anionic resin used in a commercial installation).

- 18 -

Regeneration Energy Consumption

The methods of the present invention may reduce the regeneration energy consumption by 10% or more. In some embodiments, the regeneration energy consumption may be reduced by 15% or more, or even 20% or more.

It has been observed that the level of regeneration energy saving may vary over the concentration of the physical solvent or the ratio of the physical solvent over the regenerable absorbent. The ratio of the weight percentage of the physical solvent in the absorbing medium over that of the regenerable absorbent may be from 0.5 to 2.5. In some embodiments, the ratio may be from 1.1 to 2.2.

Without being limited by theory, it is believed that increasing the concentration of the physical solvent in the absorbing medium may decrease the level of regenerative energy required.

In some embodiments, it may be desirable that the absorbing medium comprises at least 14wt% of the physical solvent. The physical solvent may be present in the absorbing medium up to 35 wt%.

The amount of heat stable salts (HSS) may also affect the level of regeneration energy saving as it affects the solubility of the physical solvent in the absorbing medium. The presence of HSS may reduce the miscibility of the physical solvent in the absorbing medium. In some embodiments, it may be desirable that the absorbing medium comprises less than 0.4 equivalent/amine mole of HSS.

The ratio of the weight percentage of heat stable salts in the lean absorbing medium over that of the regenerable absorbent is in the range of from 0.29 to 0.37, preferably from 0.31 to 0.34.

- 19 -

SO₂ Removal Processes

The process of the invention is selective for SO₂ removal over other gases, such as CO₂, NO_x, etc. A process flow diagram for an exemplary embodiment of a process to remove or capture SO₂ is shown in FIG. 1.

FIG. 1 exemplifies a heat regenerable absorbent cycle. In general, a lean absorbent medium is exposed to a SO₂ containing feed gas stream whereby SO₂ is absorbed into the absorbent and removed from the feed gas stream. The SO₂ rich absorbent formed after SO₂ absorption is then regenerated by heat, such as in a steam-stripping column. The regenerated lean absorbent may then be cycled back to absorb more SO₂.

Referring to FIG. 1, a SO₂ containing feed gas stream 10 is treated to obtain a SO₂ rich absorbent stream 80 (the spent absorbent stream). The feed gas stream 10 may be any stream, which contains SO₂ at levels suitable for SO₂ removal before the gas is released to the atmosphere, such as, without limitation, flue gas from a fluid catalytic cracker unit, an acid plant tail gas, a coal fired power plant off-gas or the like.

SO₂ rich absorbent stream 80 is obtained by contacting feed gas stream 10 with any of the SO₂ absorbents taught herein and known in the art. The absorbent may be contacted with feed gas stream 10 using any means known in the art. As exemplified in FIG. 1, feed gas stream 10 flows into a gas-liquid contact apparatus 30, where intimate contact between feed gas stream 10 and lean absorbent stream 70 occurs. Apparatus 30 may be any gas-liquid contactor or absorption tower known in the art, such as a spray or packed tower. Illustrative contacting devices include, without limitation, countercurrent absorption columns including

- 20 -

packed columns and tray columns; countercurrent or co-current spray columns including Waterloo scrubbers and venturi scrubbers; thin film contactors; and semipermeable membranes. FIG. 1 illustrates a counter
5 current flow packed tower, wherein liquid gas contact is promoted by suitable random or structured packing 40 in the column. SO₂ is absorbed into the lean absorbent stream 70, producing rich SO₂ containing absorbent, which exits from the apparatus 30 as SO₂ rich absorbent stream
10 80.

The amount of absorbing medium employed per unit volume of gas and the contact time may be sufficient for effective removal of substantially all the SO₂ from the gas stream, or to leave a desired residual amount, e.g.,
15 less than 500 ppmv, or even less than 200 ppmv, or even less than 50 ppmv, SO₂. The process may be applicable to any SO₂ containing gas stream, e.g., up to 20 or 50 volume percent SO₂, including for application to flue gas streams from thermal generating plants, which contain 700
20 to 5000 ppmv SO₂, typically 1000 to 3000 ppmv SO₂.

In some embodiments, the feed gas stream 10 is saturated (e.g., 90 percent saturation or more) with water, which may prevent undue dehydration of the absorbing medium. In some embodiments, however, a
25 relatively water-unsaturated gas may be contacted with the absorbing medium in order to save capital investment or minimize the space required.

In some embodiments, the feed gas 10 may be relatively free of particulates such as fly ash, which
30 may minimize fouling of the gas-liquid contact equipment or provide materials that might catalyze the disproportionation reaction or the oxidation of sulphite or bisulphite.

- 21 -

The contact of the absorbing medium with the SO₂ containing gas stream may be effected within the temperature range from the freezing point of the absorbent up to 75°C, or from 10°C to 60°C, or from 10°C to 50°C.

In some embodiments, it may be desirable to optimize process conditions to provide a loading of SO₂ of at least 35 grams of sulfur dioxide per kilogram of absorbing medium, e.g., 50 grams to 150 grams or up to 300 grams.

The pH of the lean absorbent medium at the point of contact with feed gas stream 10 may be in the range of 4.5 to 6.5, e.g. 5 to 6.5 or 5 to 6. The pH of the absorbent at the end of the contacting stage (e.g., at the bottom of the absorption column) may be in the range of 3 to 5, e.g., 4 to 5.

Accordingly, the pH of the absorbing medium during the absorption process may vary from 6.5-3.0, e.g., 6.5-3.5 or 6.0-4.0. Usually the lean absorbing medium (lean absorbent stream 70) initially has a pH close to the upper end of this range, while the pH of the SO₂ rich absorbent (SO₂ rich absorbent stream 80) is on the lower end and may be determined by the absorption conditions, such as the partial pressure of SO₂ in the feed gas and the absorption temperature. Thus, as SO₂ is absorbed and the solution tends to become more acidic, the pH moves towards the lower end of the range.

In order to enhance the removal of sulfur dioxide and facilitate regeneration of the absorbent, a low temperature for the absorption, to enable significant absorption of SO₂, may be employed. For example, as the absorption temperature is increased, the amount of SO₂ absorbed per mole equivalent of sorbing nitrogen of an

- 22 -

amine-based absorbent is decreased. In some embodiments, the sorbing amines used are relatively weak bases (pK_a values of between 3.0 and 5.5), which may be regenerated with less energy consumption and at a lower temperature than, for example, stronger bases.

The time of contact between the gas and absorbing medium will depend upon the intimacy of contact between the phases and the rate of transfer of the SO₂ into the liquid phase. For spray-type scrubbers, the contact time may be less than 1 or 2 seconds. With absorption columns, the contact time may be 30 seconds or more. The pressure may vary widely, e.g., from sub-atmospheric to super-atmospheric pressures. Since higher pressures increase the partial pressure of a given concentration of SO₂, they may be favored from a thermodynamic standpoint. However, in many instances the gas to be treated is at a pressure slightly higher or lower than the ambient pressure and raising the pressure may be economically undesirable.

The feed gas stream 10, which is reduced in SO₂, may be optionally washed with, for example, water (stream 60), such as in another packed section 50, to remove absorbent that may have splashed or volatilized into the treated gas stream traveling upwardly through apparatus 30. The gas then leaves the apparatus 30 as treated feed gas stream 100 for, for example, release into the atmosphere or for further treatment or use.

The water balance in the overall process may be maintained by adding water, for example via stream 60, or withdrawing water from the process, such as by directing all or a part of stream 330 to waste.

In order to conserve energy, heated streams may be used to preheat cooler streams that are subsequently fed

- 23 -

to the process equipment. For example, as exemplified in FIG. 1, SO₂ rich absorbent stream 80 flows through an indirect cross flow heat exchanger 90, where it is indirectly heated by stream 340 (a heated lean stream from regeneration tower 20 which is recycled to apparatus 30). Stream 80 is then introduced into regeneration tower 20 as stream 110.

In FIG.1, heated SO₂ rich absorbent stream 110 is then treated at a temperature, which may be higher than the absorption temperature in apparatus 30, to regenerate the absorbent. The absorbent may be heated by any means known in the art. In some embodiments, the absorbent is reheated by means of steam. In such a case, regeneration tower 20 may be a steam-stripping tower. However, other sources of heat such as hot gas, heat transfer liquids and/or direct firing may be used. As exemplified in FIG. 1, SO₂ in downwardly moving heated SO₂ rich absorbent stream 110 is removed by upwardly moving stripping gas or steam to produce a SO₂ rich product stream 280 and a regenerated absorbent (heated lean absorbent stream 220). Inert gas stripping may also be practiced for stripping the SO₂ from heated SO₂ rich absorbent stream 110 in tower 20.

Regeneration tower 20 may be any conventional towers, for instance, having a packed or trayed design. A packed tower with a packing section 210 is shown in FIG. 1 below the SO₂ rich absorbent feed level (stream 110). The SO₂ rich absorbent is stripped of SO₂ as it flows downward in the tower and into optional reboiler 230. Reboiler 230 is heated by any means known in the art. In some embodiments, reboiler 230 is indirectly heated by stream 240 (which may be steam and may be obtained from any source) through, for example, a heat transfer tube

- 24 -

bundle, producing a steam condensate stream 250 which may be recycled to produce additional steam or used elsewhere in the plant. In some embodiments, reboiler 230 is a thermosyphon type reboiler. The boiling of an aqueous liquid (e.g., SO₂ lean absorbent) in reboiler 230 can produce a flow of steam 260 for introduction into the regeneration tower 20. The steam ascends through the tower, heating the downward flowing SO₂ rich absorbent and carrying upwards the SO₂ stripped from the SO₂ rich absorbent. The steam and gaseous mixture exits the tower as product stream 280.

The desorption (regeneration) process may be conducted under any temperature and pressure conditions known in the art. In some embodiments, it may be desirable to maintain a differential in temperature between the absorption and desorption steps of at least 30°C, and the desorption temperature may be less than 110°C, e.g., 50°C to 110°C, to provide a driving force for the desorption.

Desorption may be effected by gaseous stripping using steam generated in situ (e.g., steam 260) or by passing an inert gas or steam introduced into the system (not shown) through the spent absorbing medium, usually at near atmospheric pressure. Lower pressures somewhat favor desorption. The amount of stripping gas may vary from 0 to 100 liters per liter of absorbing medium.

The delta loading ratio of SO₂ at a SO₂ partial pressure $ppgSO_2=0.01$ bar may be 20 gSO₂/L absorbent to 90 gSO₂/L absorbent. The delta loading ratio is the amount of SO₂ gas which is releasably absorbed per unit of spent absorbent medium less the amount of SO₂ gas which is releasably absorbed per unit of regenerated absorbent medium.

- 25 -

During stripping, the pH of the solution usually rises as the acidic SO₂ is removed. The conditions maintained during the stripping operation may be selected to achieve the desired level of regeneration of the absorbent (e.g. the level of dissolved SO₂ left in the absorbent).

Product stream 280 may be treated to remove excess water vapor contained therein. The water vapor may be removed by condensation (e.g. by cooling with a cooling liquid). As shown in FIG. 1, a flow of cooling water 300 into overhead condenser 290 causes condensation of steam in product stream 280, producing a 2-phase mixture, which flows into the condensate accumulator 310. The gaseous phase, which is water saturated SO₂ leaves as product stream 320. Some or all of the condensed water may be returned to the regeneration tower 20 as stream 330, where it flows downward through optional packed section 270. The cool condensate of stream 330 serves to wash volatilized absorbent from the vapors before they leave the tower 20 as product stream 280. This may help to reduce loss of absorbent chemical with the gaseous SO₂ stream 320. It will be appreciated that additional treatment steps may be used to further limit the loss of absorbent from the process.

As noted above, hot lean absorbent stream 340 may be used to preheat SO₂ rich absorbent stream 80. However, it will be appreciated that stream 80 may be heated by other means, for example, by passing it through a reboiler (not shown) or heating stream 80 upon entry to tower 20 or any combination thereof. As shown in FIG. 1, SO₂ lean absorbent leaves regeneration tower 20 as stream 220 and enters the reboiler 230. The SO₂ lean absorbent may then leave the reboiler 230 by, e.g., overflowing a weir as

- 26 -

heated lean adsorbent stream 340, which passes through the cross flow heat exchanger 90 to preheat stream 80. The SO₂ lean absorbent leaves heat exchanger 90 as cooler lean absorbent stream 120, which may optionally be cooled further by a lean solvent trim cooler 350.

Optionally, the SO₂ absorbent may be treated to remove heat stable salts (HSS) that may build up therein. As exemplified in FIG. 1, a slipstream 130 may be drawn from lean solvent trim cooler 350 and sent to a HSS removal unit (not shown) and stream 140, which comprises SO₂ absorbent reduced in HSS from the HSS removal unit, joins the recycled cooled lean absorbent to form stream 70 (the SO₂ lean absorbent stream which is introduced into tower 30). HSS removal may be effected by any method known in the art, such as electrodialysis or ion exchange. In some embodiments, it may be desirable to employ ion pairing (or salt coupling). For example, a dual function resin having different ionic functional groups such as a combination of phenol and quaternary amine functional groups may be used or liquid-liquid extraction may be used. The stream 70 enters the absorption tower 30 for capturing SO₂ from the feed gas stream 10.

The process may be operated with any convenient pressure in the absorber 30. If the feed gas stream 10 is flue gas from a boiler, which usually is operated near atmospheric pressure, then tower 30 may be operated at atmospheric pressure or a bit below the pressure of feed gas stream 10 so as to favor the flow of feed gas stream 10 into tower 30. The regeneration tower 20 is often operated at a pressure slightly over atmospheric, generally not exceeding 3 bar absolute. An above-atmospheric pressure in the regenerator may help to strip

- 27 -

SO₂, due to the higher temperatures that can be achieved. Furthermore, the product SO₂ will be at a higher pressure, helping it to flow to a downstream unit without the aid of a fan or compressor.

5 EXAMPLES

 The invention will now be illustrated by the following examples.

1. Test apparatus and methodology

1.1 Description of the apparatus

10 A floor-scale pilot unit (hereinafter "pilot unit") was used for the testing of amine-based absorbents for removing SO₂ or carbon dioxide (CO₂) from a gas mixture. It is comprised of three columns: a pre-scrubber (to saturate and condition the gas at the desired
15 temperature), an absorber and a regenerator. The unit is fully instrumented in order to record temperatures, gas and liquid flows, temperature profiles in the absorber and the regenerator, the effect of inter-cooling in the absorber, CO₂, O₂ and SO₂ concentration profiles
20 throughout the absorber, as well as regeneration energy requirements.

 The main control points are: gas temperature, lean amine temperature, lean amine introduction point in the absorber, rich amine temperature and pressure, rich amine
25 introduction point in the stripper, intercooler temperature, and reboiler pressure and energy input.

1.2 Analytical methods

 The following methods and apparatus were used to characterize the solution's compositions: a Mettler-
30 Toledo™ DL38 Karl Fisher Titrator was used to analyze water content, amine concentration was analysed using a Mettler-Toledo DL25 Titrator, salt analysis was done by ion chromatography using Dionex™ ICS-2000 Ion

- 28 -

Chromatograph and accessories, Dionex IonPac AS15 4 mm x 250 mm column, Guard column IonPac AG15, Dionex Cation column and IonPac CS-17 precolumn IonPac AG-17, and Dionex automated sampler AS 40, absorber-side SO₂ concentration in the gas was measured using a Horiba™ VA3000 gas analyzer.

1.3 Optimization procedure

The following testing protocol was used for the optimization and comparison of energy consumption: fix inlet gas flow rate, inlet gas pressure, inlet gas temperature, lean amine temperature to absorber, desorption unit pressure, rich amine temperature to regenerator.

Begin the optimization with the determination of the minimum liquid circulation with over stripping conditions (very lean solvent). This point is identified when SO₂ begins to slip through the absorber or where SO₂ recovery lowers from 100% recovery to 99% recovery. Typically, the minimization of liquid circulation rate should maximize amine loading and reduce desorption energy requirements.

Reduce the reboiler heat duty until SO₂ recovery drops below 98.5% recovery.

Increase and decrease liquid circulation by 10 to 15% to determine if there is an enhancement in SO₂ recovery. If the recovery increases in either instance, it is then desired to reduce reboiler energy until 99% recovery is attained.

2. SO₂ solubility in PEGDME

SO₂ solubility in pure PEGDME is a function of its partial pressure in the gas. PEGDMEs that are commercially available from various manufacturers may be used in the invention. For gases having 100%vol SO₂ gas and 1.5%vol SO₂, SO₂ solubility in pure PEGDME fluctuated

- 29 -

from 28.2 wt% to 0.53 wt% whereas SO₂ solubility in a solution having 50% water/50% PEGDME under the same SO₂ partial pressure dropped to 8.5wt% and 0.22 wt%, respectively. See Table 1.

5

Table 1: SO₂ solubility versus PEGDME concentration

Test temperature 50°C			
Partial pressure SO ₂	%	1.5	100
Loading			
100% PEGDME	SO ₂ wt%	0.53	28.2
50% PEGDME in water	SO ₂ wt%	0.22	8.5
100% Water	SO ₂ wt%	0.15	3.9
Solution A	SO ₂ wt%	8.1	nd

Based on the results in Table 1, it is estimated that the possible contribution of PEGDME to the loading capacity of Solution A, an aqueous solution comprising 25wt% a mixture of Hep and DiHep and 1.2 equivalent/amine mole HSS, is about 2.6 wt%. It is noted that the SO₂ loading capacity of a pure physical solvent may not be maintained when the physical solvent is mixed with another solvent.

10

15

3. Test 1

Table 2 shows molecular properties for each pure solvent used in the test, including heat of reaction with SO₂ ("Heat R_x SO₂") and heat capacity ("Cp").

20

Table 2: Molecular properties

Pure solvent/absor	Mw	Cp	Heat (R _x) SO ₂	BP	Functional group
	g/mol	kJ/kg · K	kJ/mole	°C	
Water	18	4.2	18*	100	hydrogen bound

- 30 -

TriEGMME	164	2.19		246	hydroxyl/ ether
PEGDME	236	2.05	37-42 ^s	275	ether
TetraEG	194	2.18		315	hydroxyl /ether

* R.N. Goldberg and V.B. Parker, Thermodynamics of Solution of SO₂ (g) in Water and of Aqueous Sulfur Dioxide Solutions, National Bureau of Standards, Gaithersburg, MD 20899 Journal of Research of the National Bureau of Standards Volume 90, Number 5, September-October 1985

5 § Kurt A.G. Schmidt, Solubility of Sulphur Dioxide in Mixed Polyethylene Glycol Dimethyl Ethers, A thesis submitted to the faculty of graduate studies and research in partial fulfillment of the requirements of the degree of Master of Science in Chemical Engineering, University of Alberta, Fall 1997

10

Table 3 presents some of the solution properties. All the solutions formed one phase with the exception of the solution comprising 25wt% of the mixture of Hep and DiHep and 15wt% PEGDME.

15

Experimental heat capacity "(Cp)" of the solution comprising 25wt% of the mixture of Hep and DiHep is 3.51 MJ/kg K. All the other Cp values were calculated. The Aspen model, a calibrated computational model, was able to validate the experimental Cp value with 2.8% difference and 1% difference for the calculated value of the solution comprising 25wt% of the mixture of Hep and DiHep and 15wt% PEGDME.

20

Table 3: Solution properties

Mixture of Hep and DiHep	polyol	Viscosity (50°C)	Cp	Cp (Aspen)	Mw
wt%	wt%	Cst	kJ/kg K	kJ/kg K	g/mole
25	0	1.8	3.51	3.41	168
	PEGDME				
25	5	2.4	3.4		176.5
22	8	2.2	3.44		182
18	14	2.1	3.42		192

- 31 -

25	15 (2 phases)		3.19	3.16	188
	TetraEG				
25	15	3.6	3.2		176.9
25	25	9.3	3.00		180.1
	TriEGMME				
25	12	3.3	3.29		166.7
18	23	3.4	3.26		165.7

Tables 4 to 6 present several parameters from the tests done using the pilot unit at flow rates of 19, 25 and 30 g/min.

- 32 -

Table 4: Tests at 19g/min

Chemical Solvent	Physical Solvent	Lean	Lean	Rich	Delta	Cp	Regeneration	BP	Emission	Energy Saving
Mixture of Hep and DiHep	Polyol	Mole/mole	%	%	%	kJ/kg K	MJ/kg SO2	°C	ppm SO2	%
wt %	Selexol wt %									
25	0	0.15	1.66	5.89	4.23	3.5	9.9	112	141	0
25	5	0.09	1.26	5.53	4.27	3.4	9.9	114	100	
22	8	0.14	1.67	5.51	3.84	3.4	9.7	113	164	
18	14	0.06	0.75	4.64	3.89	3.4	9.7	113	68	
18	14	0.11	1.33	5.24	3.91	3.4	8.2	113	168	~17
	TetraEG wt %									
25	15	0.07	1.22	5.44	4.22	3	9.7	115	104	

Table 5: Tests at 25g/min

Physical solvent	Chemical solvent	Lean	Lean	Rich	Delta	Cp	Regeneration	BP	Emission	Energy saving
Polyol	Mixture of Hep and DiHep	mole/mole	%	%	%	kJ/kg K	MJ/kg SO2	°C	Ppm SO2	%
Selexol wt%	wt%									
0	25	0.13	1.49	4.11	2.6	3.5	14.3	112	103	0
14	18	0.07	0.92	4	3.1	3.4	14.2	113	83	
14	18	0.08	1.33	4.2	2.9	3.4	11.9	113	98	17

- 33 -

Table 6: Tests at 30g/min

Chemical solvent	Physical solvent	Lean	Lean	Rich	Delta	Cp	Regeneration	BP	Emission	Energy saving
Mixture of Hep and DiHep	Polyol	mole/mol e	%	%	%	kJ/kg K	MJ/kg SO ₂	°C	Ppm SO ₂	%
25	0	0.12	1.49	4.4	2.6	3.5	17.4	112	124	0
25	5	0.14	1.58	4.1	2.5	3.4	15.0	113	208	
22	8	0.12	1.55	4.3	2.7	3.4	14.6	113	179	
18	14	0.07	0.99	3.6	2.6	3.4	14.6	113	119	16
	TetraEG wt %									
25	15	0.11	1.29	4	2.7	3.2	14.6	115	137	16
25	25	0.08	0.93	3.6	2.6	3	17.2	117	99	0

For the test done at 30g/min (Table 5), solutions containing PEGDME, with similar Cp, boiling points, and similar regeneration energies, showed a decrease in SO₂ emission from 208 to 119 ppm as the amount of PEGDME increased from 5% to 14% in the solution.

Each solution tested comprising a mixture of chemical and physical solvents, except the solution comprising 25wt% of the mixture of Hep and DiHep and 5wt% PEGDME, provided a reduction in regeneration energy consumption compared to the bench mark solution. The results of the solution comprising 25wt% of the mixture of Hep and DiHep, the solution comprising 18wt% of the mixture of Hep and DiHep and 14wt% PEGDME, and the solution comprising 18wt% of the mixture of Hep and DiHep and 15wt% TetraEG shown in Tables 4 to 6 are also represented as Figure 2.

- 34 -

4. PEGDME solubility versus amine and salts levels

Table 7 shows the relationship between the level of HSS and PEGDME solubility when the amine concentration is the same: for the solution having 13wt% the mixture of Hep and DiHep, the level of miscible PEGDME was 22wt% at 1 equivalent/amine mole HSS versus 17wt% at 1.2 equivalent/amine mole HSS. Table 7 also shows that for the same level of HSS, the amine concentration also influences the miscibility of PEGDME to maintain a one-phase solution: for the solution containing 25wt% the mixture of Hep and DiHep to maintain a one-phase solution, the maximum concentration of PEGDME was 5wt%; whereas when the concentration of the mixture of Hep and DiHep was at 13wt%, the maximum concentration of PEGDME rose to 17wt%.

Table 7: PEGDME solubility in the mixture of Hep and DiHep

Mixture of Hep and DiHep	HSS	PEGDME
wt%	equi/mole	wt%
13	1	22
13	1.2	17
18	1.2	14
22	1.2	8
25	1.2	5

A polyol like TetraEG did not show solubility limitation versus amine concentration and TriEGMME showed intermediate solubility. Without being limited by theory,

- 35 -

it is believed that hydroxyl groups may confer better solubility than methyl groups.

5. Test 2

5.1 Tested Solutions

5 The following solutions were tested:

 Solution A, the benchmark solution, which is an aqueous solution comprising 25wt% the mixture of Hep and DiHep and 1.2 equivalent/amine mole HSS;

 Solution A-Org, which is an aqueous solution
10 comprising 25wt% the mixture of Hep and DiHep, 20wt% malic acid, and 6wt% NaOH;

 Solution DABCO, which is an aqueous solution comprising 25wt% DABCO and 1.0 equivalent/amine mole HSS;

 Solution B, which is an aqueous solution comprising
15 18wt% the mixture of Hep and DiHep, 14wt% PEGDME, and 1.2 equivalent/amine mole HSS;

 Solution C, which is an aqueous solution comprising 25wt% Amide-DiHep, 20wt% PEGDME, and 0.4 equivalent/amine mole HSS;

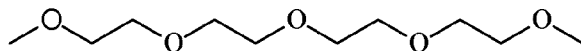
20 Solution D, which is an aqueous solution comprising 22wt% 3-aminopyrazole, 32wt% PEGDME, and 0.1 equivalent/amine mole HSS;

 Solution A-TriEGMME, which is an aqueous solution comprising 25wt% the mixture of Hep and DiHep, 12wt%
25 TriEGMME, and 1.2 equivalent/amine mole HSS; and

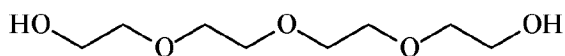
 Solution A-TetraEG, which is an aqueous solution comprising 25wt% the mixture of Hep and DiHep, 15wt% TetraEG, and 1.2 equivalent/amine mole HSS.

 A compilation of molecular structure diagrams of the
30 chemicals used in the various solutions described above is as follows.

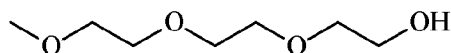
- 36 -



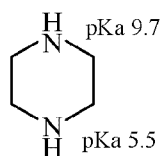
Polyethyleneglycol dimethylether (PEGDME)



5 Tetraethylene glycol (TetraEG)

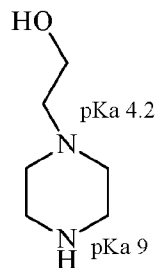


Triethyleneglycol monomethylether (TriEGMME)



10

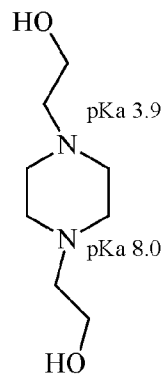
Piperazine



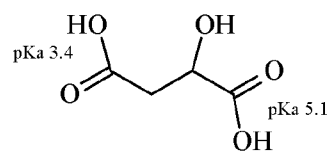
4-[hydroxyethyl]piperazine (Hep)

15

- 37 -

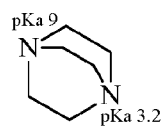


1,4-bis[hydroxyethyl]piperazine (DiHep)

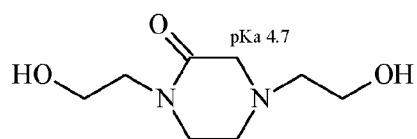


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Malic acid



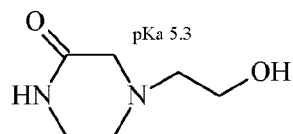
DABCO



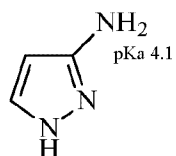
10

2-piperazinone 1,4-bis[2-hydroxyethyl] (Amide-DiHep)

- 38 -



2-piperazinone 4-(2-hydroxyethyl) (Amide-Hep)



3-aminopyrazole

5.2 Molecular properties of pure solvents

Pure SO₂ absorbents were initially ranked on the basis of their physical properties. Criteria included a low vapor pressure to minimize losses and a pK_a within the range of 3.5 to 4.7 for selective SO₂ removal. As shown in Table 9, with the exception of DABCO and Amide-HEP, all pure SO₂ absorbents have a vapor pressure of less than 0.01 mmHg and pKa in the range of 3.2 to 4.7. In commercial application, the DABCO amine is acidified to 1 equivalent resulting in a considerably suppressed vapor pressure.

- 39 -

Table 8: Molecular properties

Amine and solvent/ Property	Amide-Hep	Amide-DiHep	3-amino pyrazole	DABCO	PEGDME	TetraEG	TriEG MME
Molecular Weight	144	188	83	112	Mn~236	194	164
Density at 20°C, g/cc	1.152	0.9300	-	1.02	1.2650	1.125	1.026
Boiling Point, °C	374	409	290	174	>250	314	250
Vapor Pressure at 20°C, mm Hg	<0.01	<0.01	-	2.9 (50 °C)	<0.01	<0.01	<0.01
Freezing or Melting Point, °C	-	-	-	-	-	-	-
pKa at 25°C	5.3	4.7	4.1	3.2	-	-	-

5.3 Blending ratios

A relation between the salts level and the maximum concentration of miscible PEGDME in solution was shown in Table 10. When the concentration of PEGDME exceeds the maximum amount, phase separation may occur. Solutions 1 and 2 having increased HSS level from 1 to 1.2 eq. /mole showed corresponding decrease in the maximum concentration of miscible PEGDME from 22 to 17wt%. The salt effect is more pronounced with solutions 5, 6, and 7 where a decrease in the salt level from 1.2 to 0.1 eq. /mole corresponded with an increase of the maximum concentration of miscible PEGDME from 5 to 32wt%. Solutions 3 to 5 having increased amine concentrations and a constant salt level showed corresponding decrease in the PEGDME concentration from 14 to 8 wt%.

- 40 -

Table 9: Maximum miscible PEGDME concentration in solution as a function of amine concentration and HSS level

Solution	Amine Type	Amine	HSS	PEGDME
#		wt%	Eq/mole	wt%
1	Mixture of Hep and DiHep	13	1	22
2	Mixture of Hep and DiHep	13	1.2	17
3	Mixture of Hep and DiHep	18	1.2	14
4	Mixture of Hep and DiHep	22	1.2	8
5	Mixture of Hep and DiHep	25	1.2	5
6	Amide-DiHep	25	0.4	20
7	Aminopyrazole	22	0.1	32

5

5.4 Physical properties of the solutions

Solutions A-Org, B and C were shown to have densities that are comparable to that of Solution A. Solutions A-Org and B were shown to have viscosities that are comparable to that of Solution A.

10

5.5 Foam tendency and surface tension

Methods for performing foam measurements are described in the literature. See, for example, ASTM D1881, "Standard Test Method for Foaming Tendencies of Engine Coolants in Glassware" and ASTM D892, "Standard Test Method for Foaming Characteristics of Lubricating Oils", both of which are available from ASTM International, 100 Barr Harbor Drive, PC Box C700, West

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- 41 -

Conshohocken, PA 19428-2959 USA and are incorporated herein in their entirety.

For the present invention, samples were sparged with 1000 ml/min of nitrogen or air flowed through a diffuser stone (available from ASTM International, VWR International and Fisher Scientific). The height of the foam produced after one minute (± 1 second) was measured. After the gas flow was stopped, the time required for the foam to collapse was recorded. The results are shown in Table 10.

The target was a foam height of less than 150 ml and a break time of less than 15 seconds. Pure PEGDME did not foam. The mixture of PEGDME in water was measured to have a foam height of 250 ml and a break time of 18 seconds. Solutions B and C showed foaming tendency with more than 360 ml of foam and more than 58 seconds break time. Foaming was mitigated with the addition of an anti-foam agent such as DOW's Ucarsol™ GT-10 Antifoam (silicone based). Adding 50 to 150 ppm of an anti-foam agent in solution was shown to reduce the foaming height to below 100 ml with 8 seconds break time. Alternative anti-foam agents are available and could also be used, subject to compatibility with the process. For the surface tension, Solution A had a value of 45 Dyne/cm, pure PEGDME had a value of 36 Dyne/cm and Solution B had a value of 39 Dynes/cm.

- 42 -

Table 10: Foaming Tendency and Surface Tension (N2 Flow Rate 1 L/minute)

Solution		Foam Height (mL)	Foam Breakdown (sec)	Surface Tension Dyne/cm (at 22°C)
A		50	4	45
B		360	58	39
	With 50 ppm GT-10 antifoam	100	8	
C		370	80	nd
	With 150 ppm GT-10 antifoam	45	8	
PEGDME (100%)		0	0	36
PEGDME (25% in water)		250	18	

5 5.6 Operating conditions

The pilot unit was standardized for a SO₂ process with Solution A at 13, 19 and 25 g/min. Table 11 presents the process inputs and recorded performances of Solution A, the bench mark solution. The target shall be 100 ppmv (+/- 30ppmv) SO₂ emissions in all cases. The other tested solutions were run at 13, 19 and 25 g/min at the same operating conditions presented in Table 11.

- 43 -

Table 11: Conditions of Test 2

Feed gas SO2	wt%	1,5	1,5	1,5
Feed gasCO2	wt%	na	na	na
Flow N2	SL/min	17.33	17.33	17.33
Flow SO2	SL/min	0.26	0.26	0.26
Flow CO2	SL/min	na	na	na
Absorption packing section	inches	52	52	52
Stripper packing section	inches	26	26	26
MW	g/mole	168	168	168
FI-08 Flow	g/min	13	19	25
Pi-11 Pressure Regen outlet	psig	8	8	8
Te-03 Gas inlet	°C	32	31	33
Te-13 Rich line heating tape	°C	95	95	95
Energy input	%	25.5	29.5	34.5
Key Performance Indicators				
Average stripping factor	MJ/Kg	7.77	11.0	14.55
Lean loading	mole/mole	0.19	0.13	0.12
Delta loading	mole/mole	0.59	0.39	0.29
Lean pH	pH	5.1	5.3	53
Rich pH	pH	na	na	na
SO₂ gas emission (Horiba)	ppmv	130	108	112
Removal	%	99	99	99

5.7 Emission level

Among Solutions A, A-TriEGMME, A-TetraEG and B which have the same amine absorbent, Solution B demonstrated the lowest SO₂ emission under similar L/G (0.96 to 0.99 L/Nm³) and similar energy (9.7 to 9.8 MJ/kg SO₂). Solution C with an increased amount of PEGDME and a lower level of HSS demonstrated lower regeneration energy consumption compared to Solution A. See Figure 3.

- 44 -

5.8 Solvent capacity

The different solutions tested showed similar capacities for the range of L/G ratios tested. Solution A-Org had the highest capacity which may be due to more active sites for SO₂ scrubbing. All the others solutions with or without polyol showed very similar capacities (Figure 4). The presence of PEGDME from 14 to 22wt% in the solution seems to have low impact on the capacity of the solution.

5.9 Regeneration energy

Solution A, Solution B (14wt% PEGDME) and Solution C (20wt% PEGDME) were tested under similar test conditions presented in Table 11. Solutions B and C demonstrated 15 and 22% reduction of regeneration energy respectively over Solution A (Figure 5).

5.10 Amine purification unit (APU)

For Solution B, HSS removal tests were performed using a commercially available standard ion resin. As expected, due to its lower HSS amount, Solution B displayed a slower rate (by 21%) of sulfate removal than the bench mark solution. This is relevant to the APU bed volume, for instance, assuming the same generation rate of HSS, the APU bed volume will have to be larger (e.g., 21% larger).

Table 12 shows consumption results for a standard ion exchange against liquid-liquid (L/L) extraction on a per kilogram of sulfate removed. The water consumption per kilogram of sulfate was 30 times less for liquid-liquid extraction, and similarly 20 times less for the waste generated, compared to when the standard ion exchange was used. Amine loss per kilogram of sulfate was reduced by a factor of 10 for liquid-liquid extraction compared to the standard ion exchange.

- 45 -

Table 12: Comparison of anionic resin and ion pairing principle

100% base	Unit	IX	L/L extraction
		Anionic resin	Ion pairing
Amine	kg/kg sulfate	0.02	0.002
Water	kg/kg sulfate	1116	35
Waste generated	kg/kg sulfate	751	39

6. Test 3

5 Solution B.1 which is an aqueous solution comprising
13wt% the mixture of Hep and DiHep, 17wt% PEGDME, and 1.2
equivalent/amine mole HSS was tested. Figure 6 shows that
Solution B.1 was able to reduce regeneration energy
consumption more than Solution B did at two L/G (L/Nm³)
10 tested.

It should be understood that where a range is
disclosed herein, any range or specific value falling
within the broader range is intended to be encompassed as
if it was specifically disclosed.

15 Any patent, publication or other reference in
incorporated herein in its entirety.

Although the foregoing invention has been described
in some detail by way of illustrations and examples for
purposes of clarity of understanding, the scope of the
invention is not limited to the examples described herein
20 and should be given a broad interpretation consistent
with the specification as a whole, including the claims.

7. Test 4

25 A benchmark solution was prepared. It was an aqueous
solution comprising 25wt% of a mixture of Hep and DiHep,
1.2 equivalent/amine mole HSS.

A test solution D was prepared comprising 13wt% of a
mixture of Hep and DiHep, 1.1 equivalent/amine mole HSS,

- 46 -

20wt% PEGME, and its pH was adjusted in the range of between 5.2 and 5.6 using sulphuric acid.

Each solutions was used to absorb sulphur dioxide from a feed gas (step (i)). Then each solution was
5 subjected to stream stripping (step (ii)), and the regenerated aqueous absorbing medium was recycled.

With each solution an SO₂ stream was obtained. And the off-gas comprised a low amount of SO₂. Each process was set at an SO₂ emission level of 100 ppm, and then at
10 an SO₂ emission level of 60 ppm.

Neither the benchmark solution nor solution D showed liquid-liquid phase separation during step (i) or step (ii).

Solution D showed 28% decrease in energy consumption as compared to the benchmark solution. This was the case
15 as for the process with an SO₂ emission level of 100 ppm as well as for the process with an SO₂ emission level of 60 ppm.

20 8. Test 5

The test solution D was prepared comprising 13wt% of a mixture of Hep and DiHep, 1.1 equivalent/amine mole HSS, 20wt% PEGME, and its pH was adjusted in the range of between 5.2 and 5.6 using sulphuric acid.

25 A test solution E was prepared. It was an aqueous solution comprising 25wt% of a mixture of Hep and DiHep, 1.2 equivalent/amine mole HSS, and 25wt% PEGME.

Each solutions was used to absorb sulphur dioxide from a feed gas (step (i)). Then each solution was
30 subjected to stream stripping (step (ii)), and the regenerated aqueous absorbing medium was recycled.

Test solution E showed liquid-liquid phase separation during step (i) as well as during step (ii). Solution D

- 47 -

did not show liquid-liquid phase separation during step (i) or step (ii). For each test all absorbing medium was passed from step (i) to step (ii).

5 Solution D showed 28% decrease in energy consumption as compared to solution E.

CLAIMS:

1. A process for removing sulfur dioxide from a feed gas stream, which process comprises:

(i) contacting the feed gas stream with an aqueous lean absorbing medium to absorb sulfur dioxide and to form a sulfur dioxide lean treated gas stream and a spent absorbing medium;

wherein the aqueous lean absorbing medium comprises:

- (a) a chemical solvent comprising a regenerable absorbent, wherein the regenerable absorbent is a diamine or polyamine which in half salt form has a pK_a value for the free nitrogen atom of 3.0 to 5.5, at a temperature of 20°C in an aqueous medium;
- (b) a physical solvent, wherein the physical solvent has a vapour pressure less than 0.1 mmHg at 20°C with a boiling point equal to or higher than 240°C; and
- (c) one or more heat stable salts;

wherein the ratio of the weight percentage of the physical solvent in the lean absorbing medium over that of the regenerable absorbent is in the range of from 0.5 to 2.5;

wherein the ratio of the weight percentage of heat stable salts in the lean absorbing medium over that of the regenerable absorbent is in the range of from 0.29 to 0.37; and

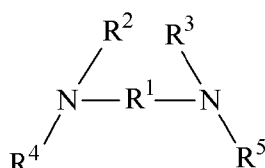
wherein the pH of the lean absorbing medium is in the range of from 4.5 to 5.6;

(ii) stripping absorbed sulfur dioxide from the spent absorbing medium to produce a regenerated aqueous absorbing medium and a gaseous sulfur dioxide; and optionally

(iii) recycling the regenerated aqueous absorbing medium from step (ii) to step (i).

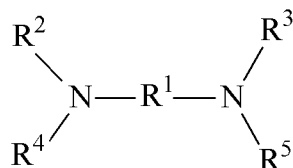
2. The process according to claim 1, wherein step (ii) is performed in a reboiler.

3. The process according to claim 1 or 2, wherein the regenerable absorbent is a diamine represented by the formula:



wherein R¹ is an alkylene of two or three carbon atoms as a straight chain or as a branched chain, R², R³, R⁴, and R⁵ may be the same or different and can be hydrogen, alkyl, hydroxyalkyl, aralkyl, aryl, or alkaryl, and any of R², R³, R⁴, and R⁵ may form cyclic structures.

4. The process according to any one of claims 1-3, wherein the regenerable absorbent is a tertiary amine represented by the formula:



wherein R¹ is an alkylene of two or three carbon atoms as a straight chain or as a branched chain, and R², R³, R⁴, and R⁵ can be alkyl, hydroxyalkyl, aralkyl, aryl, or alkaryl, and any of R², R³, R⁴, and R⁵ may form cyclic structures.

5. The process according to any one of claims 1-4, wherein the regenerable absorbent is piperazine, hydroxyethyl piperazine, bis-hydroxyethyl piperazine, hydroxyethylethylenediamine, bis-hydroxyethylethylenediamine, 1,4-diazabicyclo[2.2.2]octane (DABCO), 2-[2-aminoethyl]pyridine, 2-aminomethylpyridine, 3-amino 5-methylpyrazole, 3-aminopyrazole, 3-methylpyrazole, N,N,N',N'-tetraethyldiethylenetriamine, N,N,N',N'-tetramethyldiethylenetriamine, 2-piperazinone 1,4-bis[2-hydroxyethyl], or a combination thereof.

6. The process according to any one of claims 1-5, wherein the lean absorbing medium additionally comprises an anorganic acid.

7. The process according to any one of claims 1-6, wherein the physical solvent is a polyol, a polycarbonate, an N-formyl morpholine, or a combination thereof.

8. The process according to any one of claims 1-7, wherein the physical solvent is polyethyleneglycol dimethylether (PEGDME), tetraethyleneglycol dimethylether (TetraEGDME), tetraethylene glycol (TetraEG), triethyleneglycol monomethylether (TriEGMME), or a combination thereof.

9. The process according to any one of claims 1-8, wherein the physical solvent is polyethyleneglycol dimethylether (PEGDME), and

wherein the regenerable absorbent is
4-[hydroxyethyl]piperazine (Hep), or
1,4-bis[hydroxyethyl]piperazine (DiHep), or
3-aminopyrazole, or
a mixture of 4-[hydroxyethyl]piperazine (Hep) and
1,4-bis[hydroxyethyl]piperazine (DiHep).

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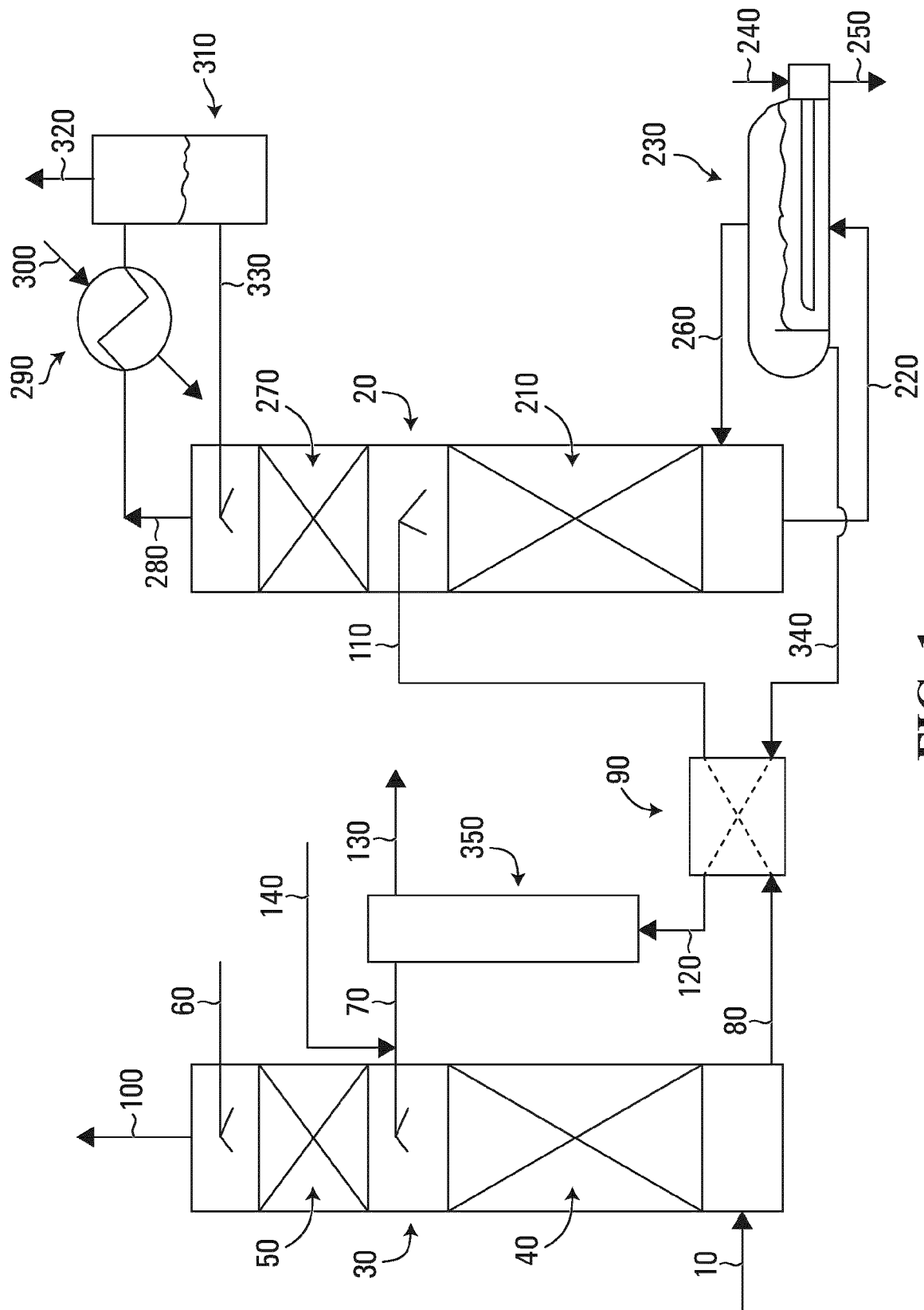
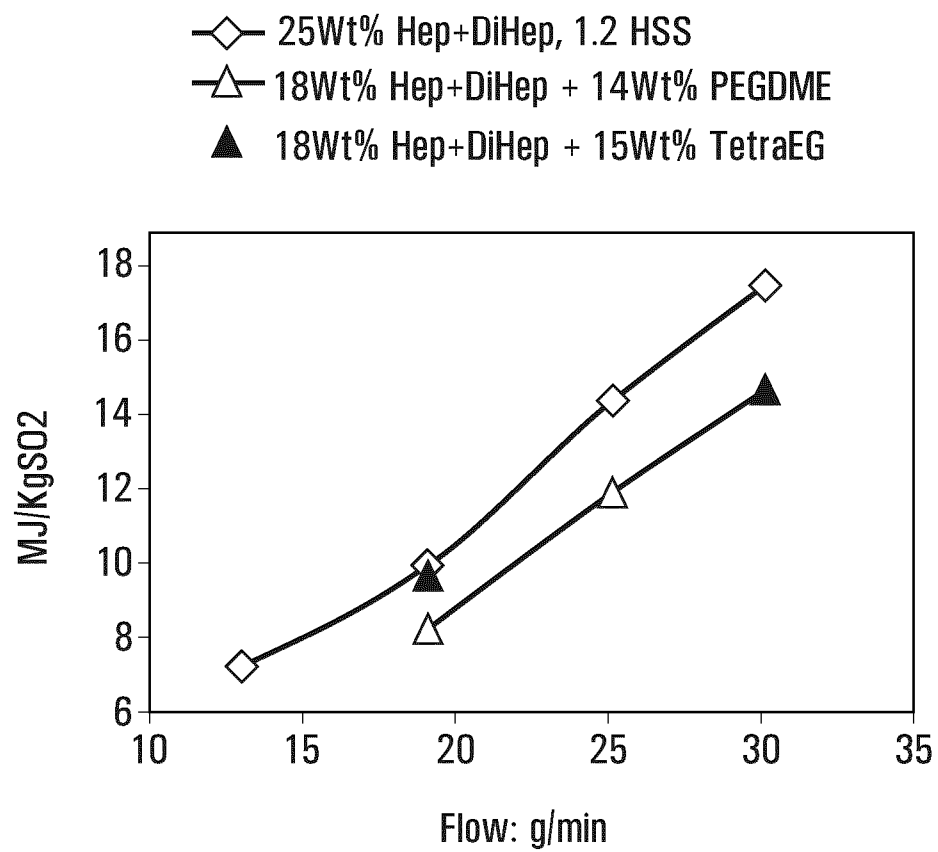
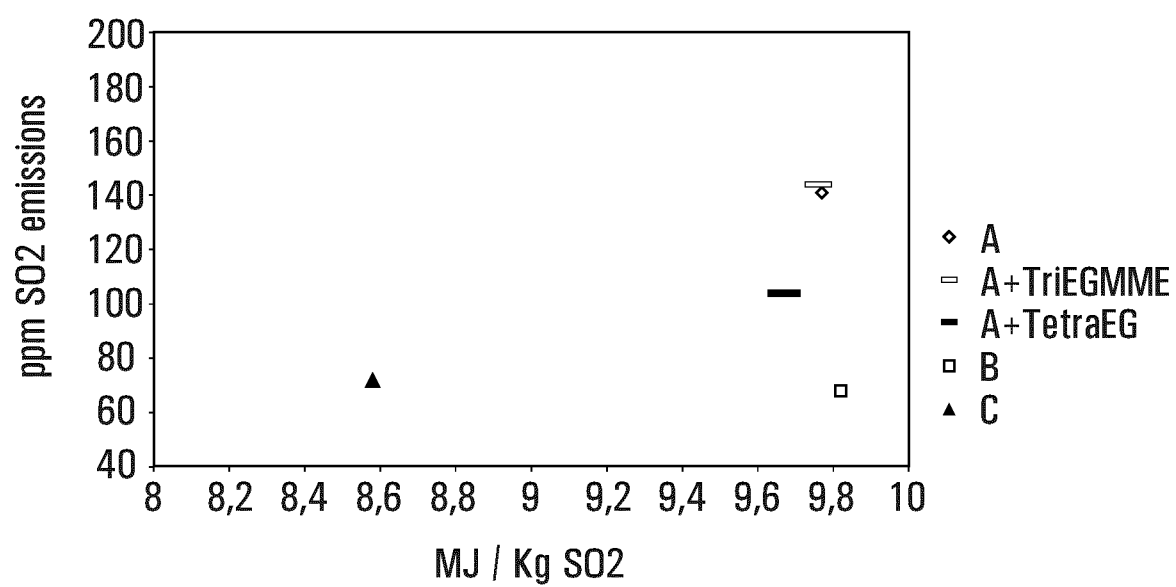


FIG. 1

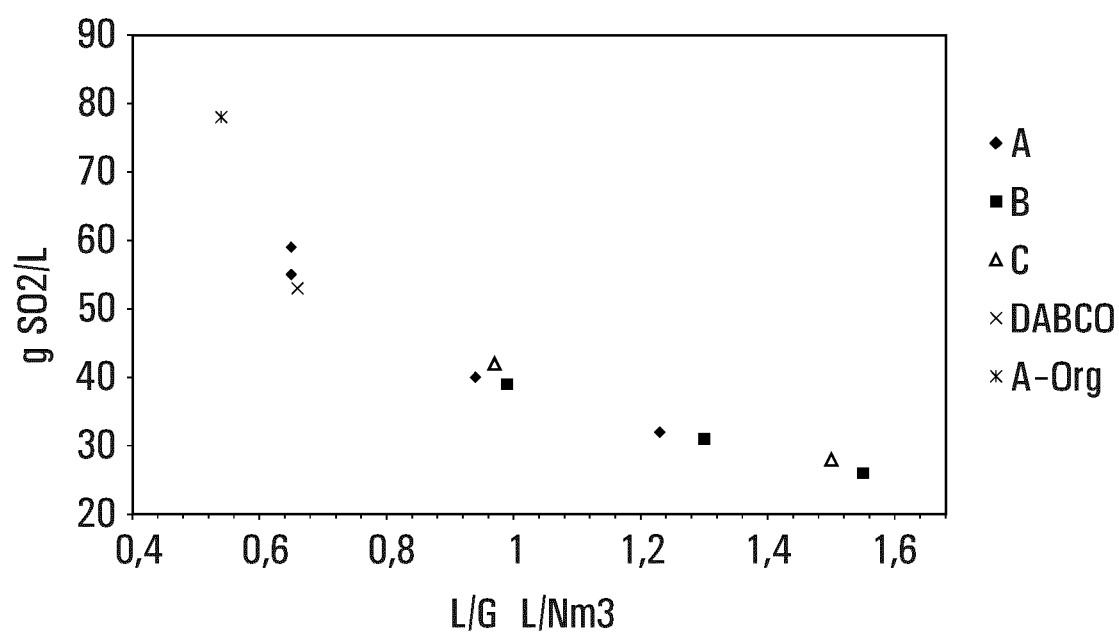
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**FIG. 2**

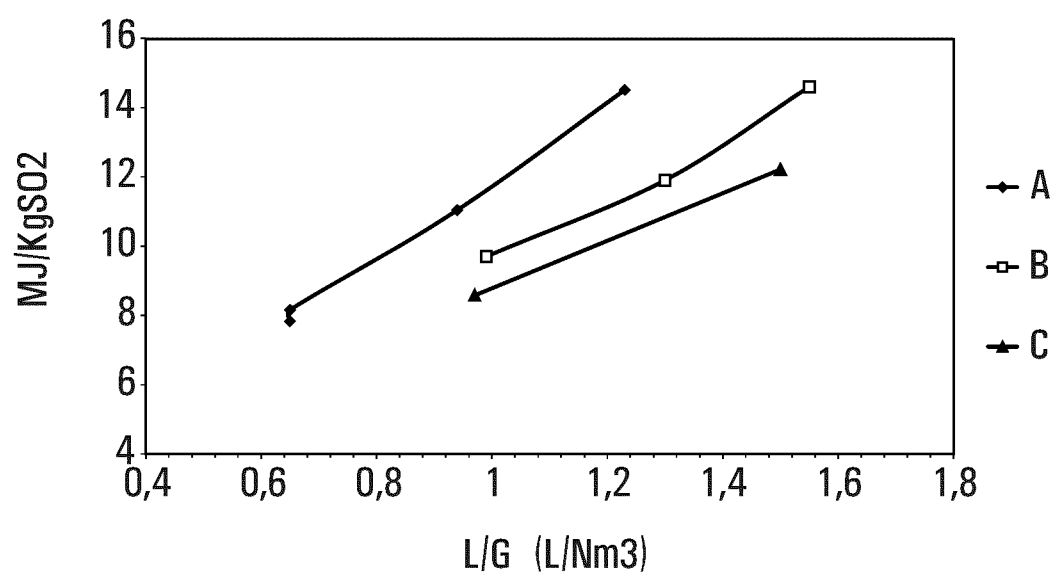
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**FIG. 3**

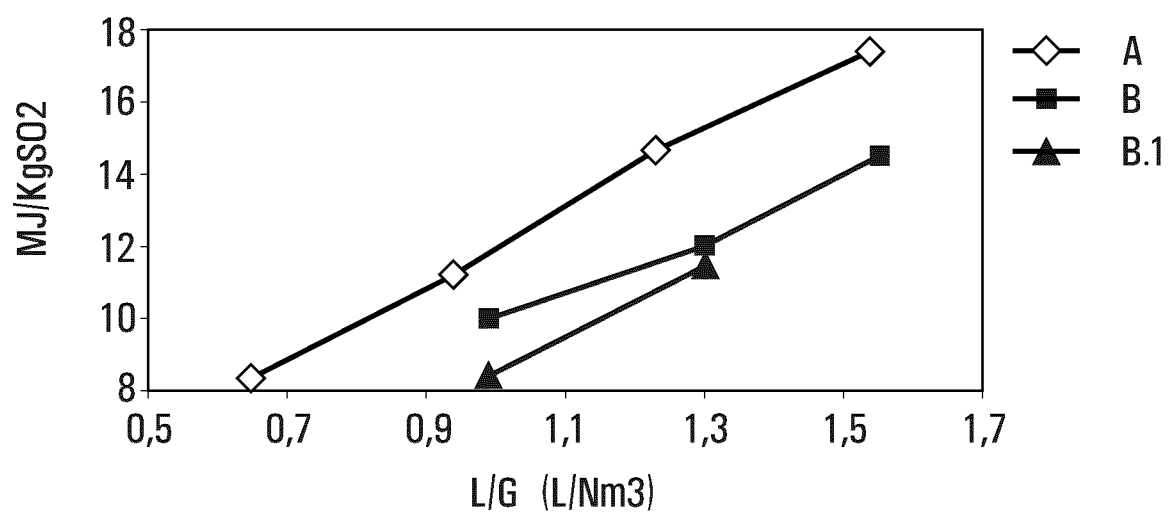
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**FIG. 4**

5/6

**FIG. 5**

6/6

**FIG. 6**

