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Various novel amine solutions both in aqueous and non-aqueous [monoethylene glycol (MEG)/triethylene glycol(TEG)] forms have been studied for hydrogen sulfide (H₂S) absorption. The study was conducted in a custom build experimental setup at temperatures relevant to subsea operation conditions and atmospheric pressure. Liquid phase absorbed H₂S, and amine concentrations were measured analytically to calculate H₂S loading (mole of H₂S/mole of amine). Maximum achieved H₂S loadings as the function of pK_a, gas partial pressure, temperature and amine concentration are presented. Effects of solvent type on absorbed H₂S have also been discussed. Several new solvents showed higher H₂S loading as compared to aqueous N-Methyldiethanolamine (MDEA) solution which is the current industrial benchmark compound for selective H₂S removal in natural gas sweetening process.

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Article

Study of Various Aqueous and Non-Aqueous Amine Blends for Hydrogen Sulfide Removal from Natural Gas

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Abstract: Various novel amine solutions both in aqueous and non-aqueous [monoethylene glycol (MEG)/triethylene glycol(TEG)] forms have been studied for hydrogen sulfide (H₂S) absorption. The study was conducted in a custom build experimental setup at temperatures relevant to subsea operation conditions and atmospheric pressure. Liquid phase absorbed H₂S, and amine concentrations were measured analytically to calculate H₂S loading (mole of H₂S/mole of amine). Maximum achieved H₂S loadings as the function of pK_a, gas partial pressure, temperature and amine concentration are presented. Effects of solvent type on absorbed H₂S have also been discussed. Several new solvents showed higher H₂S loading as compared to aqueous N-Methyldiethanolamine (MDEA) solution which is the current industrial benchmark compound for selective H₂S removal in natural gas sweetening process.

Keywords: H₂S absorption; amine solutions; glycols; desulfurization; aqueous and non-aqueous solutions

1. Introduction

Natural gas is considered one of the cleanest forms of fossil fuel. Its usage in industrial processes and human activities is increasing worldwide, providing 23.4% of total world energy requirement in 2017 [1]. Natural gas is half of the price of crude oil and produces 29% less carbon dioxide than oil per unit of energy output [2]. Methane is a major energy providing component in natural gas. However, it also contains other hydrocarbons and a variety of impurities like acid gasses (CO₂ and H₂S) and water. Besides reducing the gas energy value, the impurities can cause operational problems such as corrosion in the pipeline and other equipment [3]. Mercury, mercaptans and other sulfur components are also often found in natural gas and must be removed. Sulfur components can produce SO₂ during combustion which ultimately leads to acid rain. Therefore, it is necessary to remove acid gases, water vapors, and other impurities before the usage of natural gas.

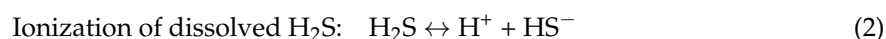
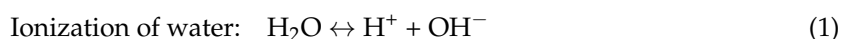
H₂S is an extremely poisonous component, and it can cause instant death when concentrations are over 500 parts per million volume (ppmv) [4,5]. H₂S exposure limits by the Norwegian Labour Inspection Authority are 5 ppmv for an eight-hour time-weighted average (TWA) and 10 ppmv for 15 min short-term exposure limit (STEL) [6]. The most commonly used method for H₂S removal is liquid scavenging. These processes usually employ non-regenerative chemicals such as triazine or aldehydes, and because of costs and operational issues (e.g., chemicals disposal), scavengers are not used for gases with high H₂S concentrations. Alkanolamines, in particularly N-Methyldiethanolamine (MDEA), are generally used for regenerative H₂S removal processes [7].

Natural gas is commonly saturated with water increasing the chances of solid gas hydrates formation with methane at high pressure and low temperatures potentially causing plugging in

gas transport pipelines. One common way to avoid hydrate formation and to achieve problem-free continuous gas transportation operations is to add hydrates inhibitors like monoethylene glycol (MEG) or triethylene glycol (TEG) in gas pipelines [8].

A system which could selectively remove H₂S and control hydrate simultaneously would not only reduce equipment footprint but also help to reduce the installation and operational costs for both subsea and platform operations. This type of system was initially proposed by Hutchinson [9]. The idea of combined H₂S and water removal was presented in 1939 by using amine glycol solution as a solvent. 2-ethanolamine (MEA) and diethylene glycol (DEG) solution in water solution was the tested solvent for the concept. McCartney [10,11] and Chapin [12] built upon Hutchinson concept and presented the idea of both absorption and regeneration process in two-stages. They discussed various arrangements to get higher efficiency and lower energy requirement. Later on, this process development discontinued due to lower selectivity of H₂S compared to CO₂, higher amine degradation and corrosion rate of MEA [7]. However, tertiary amine systems could be very interesting for this type of operations as they are known for their high selectivity to H₂S. Tertiary amine systems, like a blend of methyl diethanolamine (MDEA) with glycols (MEG/TEG), have, additionally, higher amine stability and reduce corrosion rates [7,13,14].

In the literature, there is limited data available for the tertiary amine-glycols blends and most of the data is available for aqueous Triethanolamine (TEA), diisopropanolamine (DIPA), and MDEA. TEA was the first commercially used alkanolamine for gas treating process [7]. It is now being replaced with other amines like monoethanolamine (MEA), diethanolamine (DEA), diisopropanolamine (DIPA), methyl diethanolamine (MDEA), 2-amino-2-methyl-1-L-propanol (AMP), ethyl diethanolamine (EDEA) and 2-(2-aminoethoxy) ethanol (DGA) due to its low capacity and high circulation rate [15]. MDEA based system offered advantages like selective hydrogen sulfide removal over carbon dioxide, low vapor pressure, higher thermal stability, less corrosion, lower heat of reactions and specific heat [7,13]. Equations (1)–(6) show the mechanism and overall reactions of H₂S with aqueous secondary and tertiary amines. These reactions are instantaneous and involve a proton transfer.



The solubility of H₂S in aqueous solutions of MDEA from 11.9 wt.% to 51 wt.% in the temperature range from 10 °C to 120 °C and H₂S partial pressure from 0.141 kPa–6900 kPa were studied by various authors [16–27]. All the previous studies of aqueous MDEA showed similar trends like increasing the partial pressure of H₂S (pH₂S) increases H₂S loading at given concentration and temperature, while the increase in amine concentration at a given temperature and pH₂S decreases H₂S loading. Surplus to MDEA data, TEA from 0.09–6.32 kPa H₂S and DIPA at a pressure range of 19–1554 kPa has been reported [28,29].

Xu et al. [24] also studied H₂S absorption in 30 wt.% MDEA in MEG and MEG–H₂O solutions over a range of partial pressures of H₂S from 0.34 to 38.8 kPa and found that increasing the water content in solution increases the H₂S loading at a given temperature (40 °C). Also, the increase in temperature decreases the H₂S loading for a given concentration (30 wt.% MDEA–65 wt.% MEG—5 wt.% H₂O). Most of the previous studies were conducted using static cell apparatus and higher liquid phase H₂S loading can be obtained by using total gas pressure (>101.3 kPa) with higher amine

concentration. Therefore, very few H₂S absorption studies are available for low amine concentrations at low temperatures and low acid gas inlet partial pressure range in literature.

The objective of this study is to identify blends where the solute (amine) can give higher H₂S removal capacity as compared to MDEA in the presence of glycol. The overall goal for this process is to absorb H₂S and water simultaneously at the subsea level in two-steps. In the first step, absorption can take place at the subsea level, potentially using a co-current contactor for absorption and flash drum to separate the natural gas from solvent at subsea levels. In the second step, loaded solution can be sent to a platform for regeneration and natural gas will be transported directly from subsea allowing a system where the natural gas will not enter the platform at all. The current work focuses on the identification of amine-glycol blends with high H₂S absorption capacity. The amines for this work were chosen systematically so that insight into the influence of its structure, like amine alkanol groups, alkyl chain length, and a hydroxyl group, can be obtained. In total twelve amines were studied, one secondary sterically hindered amines (diisopropylamine), one tertiary sterically hindered amine (N-tert-butyl-diethanolamine), and ten tertiary amines. The list of amines along their chemical structure used in the study is given in Figure 1.

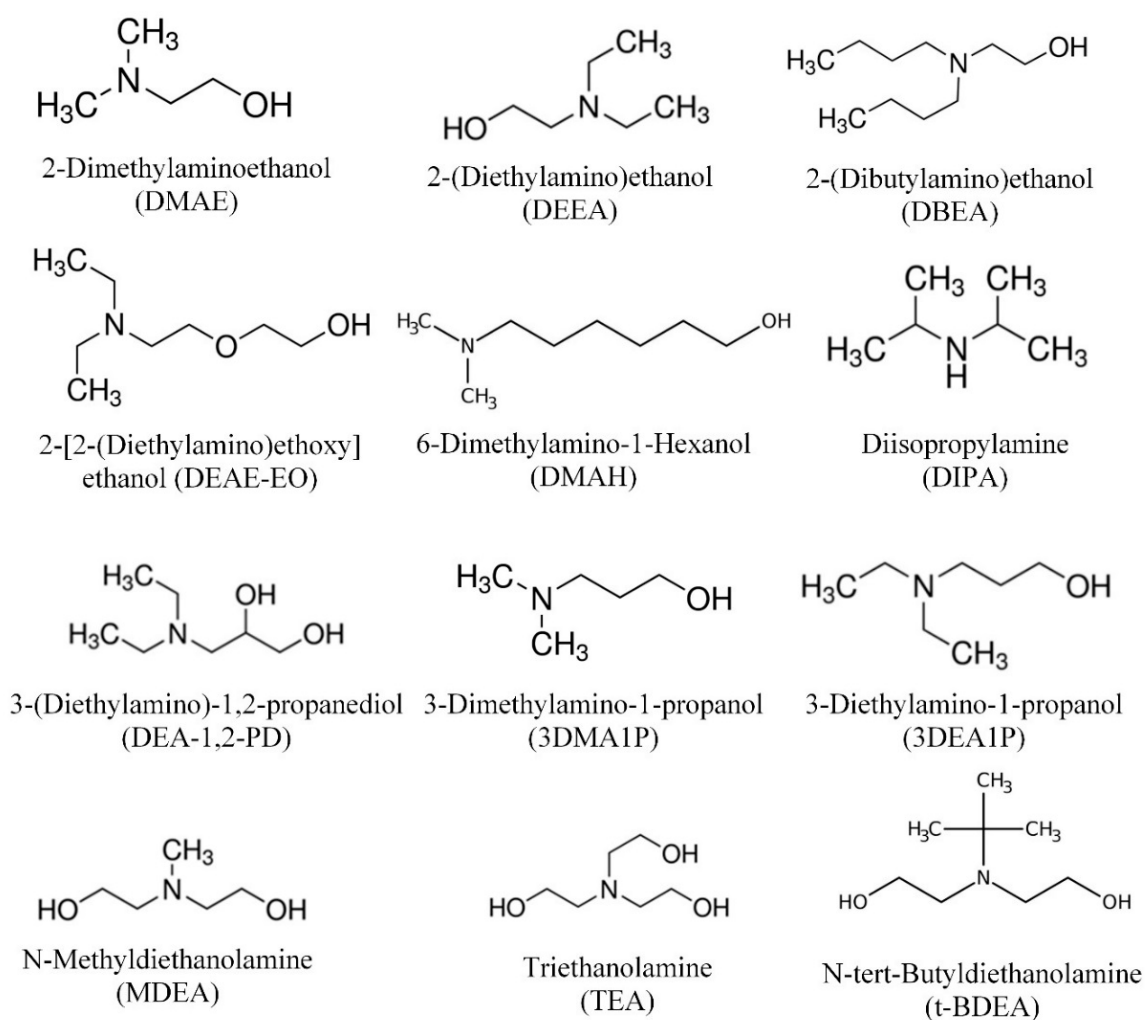


Figure 1. List of amines with chemical structures.

2. Material and Methodology

2.1. Materials

2-Dimethylaminoethanol (DMAE), 2-(Diethylamino) ethanol (DEEA), 2-(Dibutylamino) ethanol (DBAE), Diisopropylamine (DIPA), 3-Dimethylamino-1-propanol (3DEA-1P), N-Methyldiethanolamine (MDEA), Triethanolamine (TEA), Ethylene glycol (MEG), and Triethylene glycol (TEG) were bought from Sigma-Aldrich (Oslo, Norway), while 3-(Diethylamino)-1,2-propanediol (DEA-1,2-PD), 2-[2-(Diethylamino) ethoxy] ethanol (DEAE-EO), 6-Dimethylamino-1-Hexanol (DMAH), N-tert-Butyldiethanolamine (t-BDEA), and 3-Diethylamino-1-propanol (3DEA-1P) were bought from TCI Europe (Zwijndrecht, Belgium) in available maximum commercial purity. Additionally, premixed 1500 ppmv (0.15 vol.%) Hydrogen Sulphide (H₂S) in Nitrogen (N₂), 10,000 ppmv (1 vol.%) Hydrogen Sulphide (H₂S) in Nitrogen (N₂) and pure Nitrogen (N₂) (99.998 vol.%) were purchased from AGA Norway, Oslo. All chemicals were used without further purifications. Chemicals with their abbreviation, CAS number, purity, molecular weight, and pKa are given in Table 1 except hydrogen sulfide and deionized water.

Table 1. Name, abbreviation, CAS, purity (wt.%), and pKa of chemicals.

Chemical	CAS	Purity (wt.%)	Molecular Weight (g/mol)	pKa
2-Dimethylaminoethanol (DMAE)	108-01-0	≥99.5	89.14	9.49 [30]
2-(Diethylamino)ethanol (DEEA)	100-37-8	≥99.5	117.19	9.75 [31]
2-(Dibutylamino)ethanol (DBAE)	102-81-8	≥99.0	173.30	9.04 [32]
2-[2-(Diethylamino)ethoxy] ethanol (DEAE-EO)	140-82-9	>98.0	161.25	10.15 [31]
6-Dimethylamino-1-Hexanol (DMAH)	1862-07-3	>97.0	145.24	10.01 [31]
Diisopropylamine (DIPA)	108-18-9	≥99.0	101.19	8.84 [33]
3-(Diethylamino)-1,2-propanediol (DEA-1,2-PD)	621-56-7	>98.0	147.22	9.68 [31]
3-Dimethylamino-1-propanol (3DMA-1P)	3179-63-3	≥99.0	103.16	9.54 [30]
3-Diethylamino-1-propanol (3DEA-1P)	622-93-5	≥95.0	131.22	10.29 [30]
N-Methyldiethanolamine (MDEA)	105-59-9	≥99.0	119.16	8.65 [30]
Triethanolamine (TEA)	102-71-6	≥99.0	146.19	7.85 [30]
N-tert-Butyldiethanolamine (t-BDEA)	2160-93-2	≥97.0	161.24	9.06 [30]
Ethylene glycol (MEG)	107-21-1	≥99.5	62.07	14.44 [34]
Triethylene glycol (TEG)	112-27-6	≥99.8	150.17	14.50 [35]

All amine solutions were prepared by weighing the required amount of the amines using the Mettler Toledo MS6002S Scale, with an uncertainty of $\pm 10^{-5}$ kg. Aqueous solutions were made with deionized water produced by ICW-3000 Millipore water purification system, while for non-aqueous solutions MEG/TEG was used as a solvent. All amines were miscible in DI water, MEG and TEG except DBAE which made visible two phases with DI water but less visible two phases with MEG and TEG. DBAE solutions appeared homogeneous while stirring.

2.2. Methodology and Equipment

A custom-built apparatus, as shown in Figure 2, was used to screen amine solutions for hydrogen sulfide absorption study. The apparatus is designed to operate at atmospheric pressure and temperatures up to 80 °C and is similar to apparatus previously used for CO₂ absorption and desorption studies by Ma'mun et al. and Hartono et al. [31]. The apparatus consisted of the water-jacketed reactor with volume of ~200 cm³ (NTNU, Trondheim, Norway) with a magnetic stirrer, Alicat MCS series Mass flow controllers (Tucson, AZ, USA), thermocouple (Omega Engineering Limited, Nærum, Denmark), Hubor[®] water bath (Huber Kältemaschinenbau AG, Offenburg,

Germany), and sodium hydroxide (NaOH) vessel for caustic wash. LabVIEW (National Instruments Norway, Drammen, Norway) was used to control and record gases flowrates and both reactor and water bath temperatures. The apparatus and H₂S gas bottles were installed in a closed fume cabinet equipped with an H₂S sensor, alarm and fail-safe system; which shut down the whole apparatus automatically in case of any H₂S leakage (limit >10 ppmv) or electrical failure. Personal protective equipment and personal H₂S sensor were used during experiments.

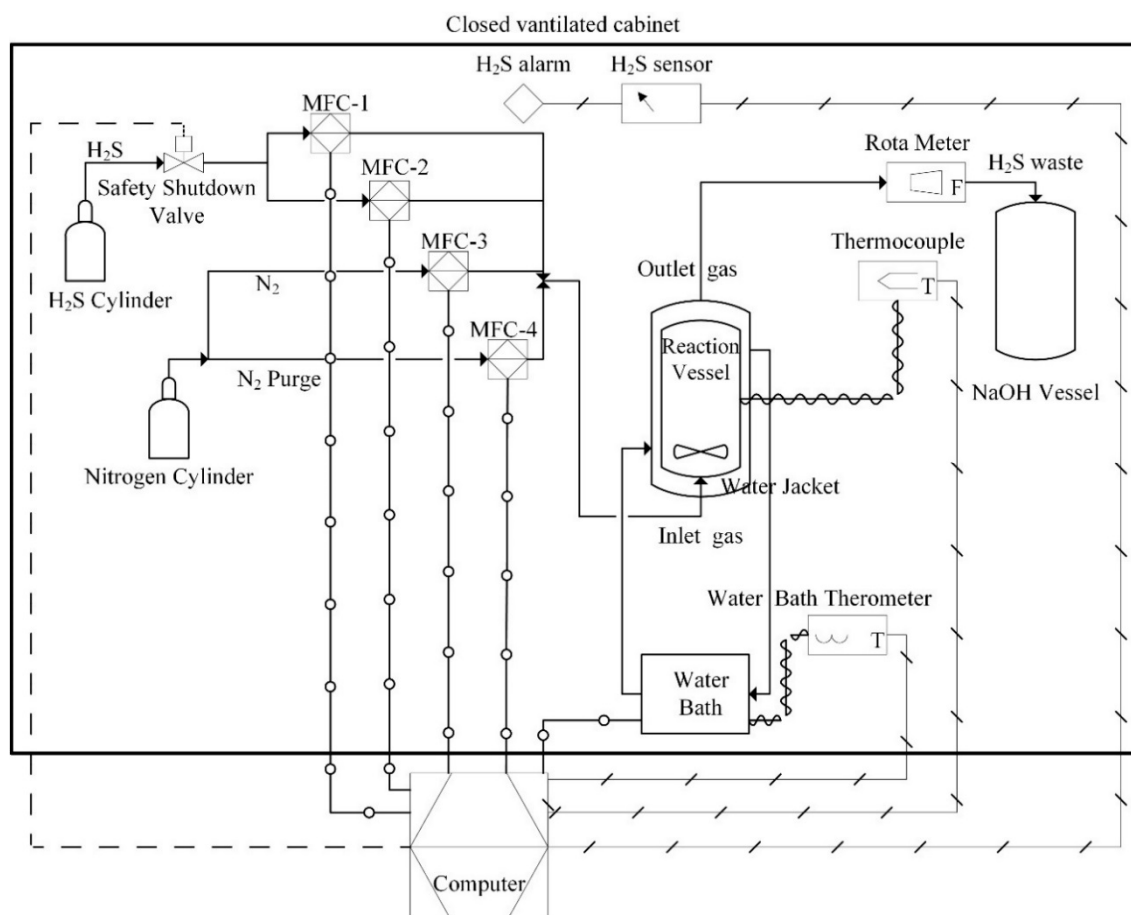


Figure 2. Schematic flow diagram of the screening apparatus.

Since the overall goal is to develop a solvent system that could be used at the subsea level, where the total gas pressure is high and H₂S content is from 50 ppm and up, higher partial pressure of H₂S up to 1 vol.% was used to achieve similar H₂S quantity during these experiments at atmospheric pressure.

At the start of each experiment 150 g of the solution was filled in the reactor and cooled/heated to the required experiment temperature after purging it with nitrogen to remove any air present. Pre-mixed H₂S and N₂ were used to achieve the required inlet hydrogen sulfide partial pressure (pH₂S) with the help of MFCs. The reactor was continuously stirred with a magnetic stirrer at isothermal and isobaric condition during the whole experiment. Hubor[®] water bath was used to maintain the temperature constant. A thermocouple was placed in the liquid phase and used for continuous monitoring of reactor temperature. The exit gas from the reactor was sent to a series of three 10 wt.% NaOH solution vessels in order to remove residual H₂S present in it. All experiments were run for 120 min to give sample time to reach close to equilibrium between acid gas and amine solution. To ensure that 120 min is enough, experiments with 20 wt.% MDEA were performed until 240 min at 5 °C with sampling after every 15 min. The data showed that H₂S stopped absorbing after 45 min. This is in line with Lemoine et al. [20] and confirms that 120 min is enough time to reach close to equilibrium. Also, several parallel experiments for both aqueous and non-aqueous solutions of various

amines were run and repeatability the data were confirmed. Different solutions were tested at different temperatures (5 °C, 25 °C and 40 °C) and inlet H₂S partial pressures (0.03 kPa, 0.5 kPa, 0.75 kPa and 1 kPa). For inlet pH₂S = 0.5 kPa to 1 kPa, 10,000 ppm H₂S gas mixture at total flow rate of 200 mL/min and for inlet pH₂S = 0.03kPa, 1500 ppm H₂S gas mixture at total flow rate of 1000 mL/min were used. The uncertainty of the inlet partial pressure of H₂S was estimated to be 2% including both the uncertainty of the ready H₂S gas mixture and the pre-calibrated mass flow controllers.

After the experiment, liquid samples were stored at <4 °C in the fridge and later on delivered to the analytical lab (St. Olav's Hospital Laboratory, Trondheim, Norway) for total sulfur analysis with inductivity coupled plasma mass spectrometry (ICP-MS). The samples were transported in ice box along with ice to keep sample temperature <5 °C. To ensure no amine loss during the experiments, amine concentration was determined by with Mettler Toledo G20 compact titrator [36] using a liquid sample of 0.2 mL that was diluted with 50 mL deionized water and titrated with 0.1 mol/L H₂SO₄. Each liquid sample was analyzed twice for total sulfur and amine concentration. The standard deviations between the duplicates of each solution were <2.5% for total sulfur and <1.5% for amine concentration. The differences in the amine concentration were less than 2% found in initial and final amine concentrations for all the solutions indicating that there was no significant amine loss during the experiments. The hydrogen sulfide loadings calculated by Equation (7), given in this work are based on the analyzed values for H₂S and amine in the liquid phase.

$$\alpha_{H_2S} = \frac{\text{mole of } H_2S}{\text{mole of amine}} \quad (7)$$

3. Results

This screening apparatus was validated with a benchmarking 30 wt.% aqueous monoethanolamine (MEA) for CO₂ absorption before using it for H₂S absorption. Inlet CO₂ partial pressure was 10 kPa and absorption was done at 40 °C until 95% CO₂ absorption. Rich loading was found 0.54 mol CO₂/mol MEA after titration which was in good agreement with Hartono et al. [31] with an average deviation of 1.9%. 23.8 wt.% aqueous MDEA has been mostly used to study H₂S absorption. Therefore, the same amine concentration was used to verify the screening equipment and experimental parameters at 40 °C and pH₂S = 1 kPa. The liquid phase of H₂S loading was measure 0.14 (mol/mol) with the deviation of 4.6% from Jou et al. [16]. The experimental data are shown in with experimental uncertainties at the end is shown in Table 2.

Table 2. Experimental data.

Amine	Initial Amine (wt.%)	Solvent	H ₂ S Loading (α)	Inlet pH ₂ S (kPa)	Temperature (°C)
MDEA	20%	Water	0.015	0.03	5
DEEA	20%	Water	0.008	0.03	5
DBEA	20%	Water	0.011	0.03	5
DIPA	20%	Water	0.012	0.03	5
TEA	20%	Water	0.013	0.03	5
t-BDEA	20%	Water	0.009	0.03	5
MDEA	20%	MEG	0.010	0.03	5
MDEA	20%	TEG	0.006	0.03	5
MDEA	20%	Water	0.213	0.5	5
DEEA	20%	Water	0.237	0.5	5
3DEA-1P	20%	Water	0.258	0.5	5
DMAE	20%	Water	0.281	0.5	5
DEAE-EO	20%	Water	0.416	0.5	5
DBAE	20%	Water	0.032	0.5	5
DBAE	20%	MEG	0.089	0.5	5
3DEA-1P	20%	Water	0.271	0.75	5
DEAE-EO	20%	Water	0.389	0.75	5
DBAE	20%	Water	0.040	0.75	5
MDEA	20%	Water	0.254	1	5
MDEA	20%	Water	0.189	1	40

Table 2. Cont.

Amine	Initial Amine (wt.%)	Solvent	H ₂ S Loading (α)	Inlet pH ₂ S (kPa)	Temperature (°C)
DEEA	20%	Water	0.240	1	5
DEEA	20%	Water	0.249	1	40
3DEA-1P	20%	Water	0.355	1	5
3DEA-1P	20%	Water	0.272	1	25
3DEA-1P	20%	Water	0.248	1	40
3DEA-1P	30%	Water	0.120	1	5
3DEA-1P	50%	Water	0.061	1	5
DMAE	20%	Water	0.344	1	5
DMAE	20%	Water	0.260	1	40
DEAE-EO	20%	Water	0.378	1	5
DEAE-EO	20%	Water	0.413	1	25
DEAE-EO	20%	Water	0.385	1	40
DEAE-EO	50%	Water	0.094	1	5
DEAE-EO	30%	Water	0.231	1	5
DBAE	20%	Water	0.139	1	5
DBAE	20%	Water	0.043	1	25
DBAE	20%	Water	0.052	1	40
DBAE	30%	Water	0.049	1	5
DBAE	50%	Water	0.023	1	5
DIPA	20%	Water	0.185	1	5
TEA	20%	Water	0.165	1	5
t-BDEA	20%	Water	0.407	1	5
DEA-1,2-PD	20%	Water	0.306	1	5
DMAH	20%	Water	0.368	1	5
3DMA1P	20%	Water	0.299	1	5
MDEA	20%	MEG	0.189	1	5
DEEA	20%	MEG	0.155	1	5
3DEA-1P	20%	MEG	0.193	1	5
DMAE	20%	MEG	0.104	1	5
DEAE-EO	20%	MEG	0.280	1	5
DBAE	20%	MEG	0.121	1	5
DBAE	20%	MEG	0.061	1	40
MDEA	20%	TEG	0.040	1	5
DEEA	20%	TEG	0.035	1	5
3DEA-1P	20%	TEG	0.080	1	5
DMAE	20%	TEG	0.025	1	5
DEAE-EO	20%	TEG	0.073	1	5
DBAE	20%	TEG	0.049	1	5

$u(\text{pH}_2\text{S}) = \pm 2\%$; $u(T) = \pm 0.1\text{ }^\circ\text{C}$; $u(C_{\text{Amine}}) = \pm 1.5\%$; $u(C_{\text{H}_2\text{S}}) = \pm 2.5\%$

3.1. Effect of pKa

Effect of pKa on H₂S loading in 20 wt.% aqueous amine solutions at T = 5 °C ± 0.1 °C and pH₂S = 1 kPa is shown in Figure 3. In the reaction between H₂S and aqueous amine solution, H₂S acts as weak acid whereas aqueous amine acts as a strong base, therefore, an increase in pKa increases the hydration of H₂S subsequently increasing the H₂S loading. This is also evident in tertiary amines aqueous solutions with DEEA, t-BDEA, and DBAE acting like outliers. DEEA shows lower absorption capacity than its closest pKa tertiary amine (DEA-1,2-PD), which can be due to short molecular chain of DEEA. t-BDEA. Sterically hindered amine shows the highest loading of all amines whereas DBAE shows the lowest loading, and it makes two phases with almost all the H₂S absorbed in the upper phase, i.e., amine (solute). If the amount of H₂S absorbed only in the amine phase (solute) is used to calculate H₂S loading in DBAE aqueous solutions, these solutions also start to follow the trend. The amount of H₂S absorbed in DBAE amine phase is $\approx 4.6 \pm 0.2$ times of absorbed H₂S in the whole solution both in aqueous and non-aqueous solutions.

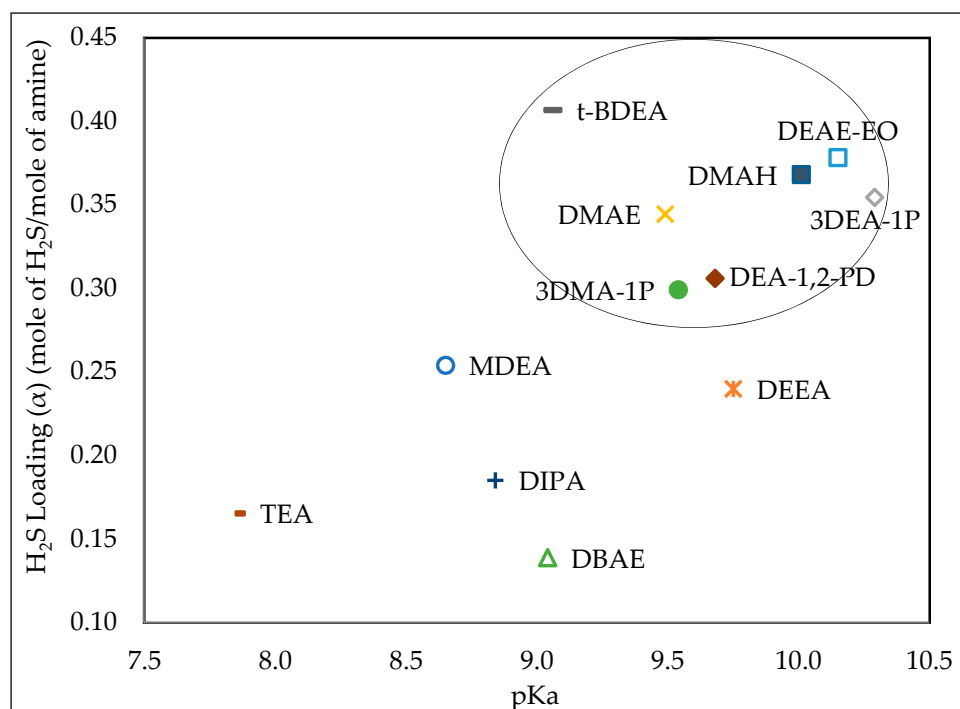


Figure 3. Effect of pKa on H₂S loading in aqueous amine solutions; T = 5 °C ± 0.1 °C; pH₂S = 1 kPa; amine concentration = 20 wt.% (unloaded); DBAE solutions make two phases.

Aqueous solutions of amines highlighted in the circle in Figure 3 presented higher H₂S loading as compared to MDEA and can be potential amines for further studies. t-BDEA showed the highest H₂S loading, but in-house data show it also degraded a lot in the presence of CO₂ and caused higher corrosion rates leading to damages in steel pipelines and equipment as compared to MDEA [37].

When looking into the amine structure, the results show that an increase in alkyl group decreases the H₂S loading in an amino-ethanol group, i.e., DMAE > DEAE (DEEA) > DBAE. It can be due to reduction in activity of nitrogen group due to increase in chain length of alkyl group in ethanol amine, a similar trend was previously observed in carbon dioxide capture studies [38,39]. Structure wise it would have been interesting to test 2-Dipropylaminoethanol (DPAE). Unfortunately, we were unable to purchase the chemical since it is commercially unavailable in Norway as it is being used in the weapon industry. A reverse trend was seen in an amino-propanol group where an increase in the alkyl group increases the H₂S loading, i.e., 3DMA-1P < 3DEA-1P. Hydroxyl group attracts electrons therefore, addition of more hydroxyl group reduces the activity of nitrogen atom of amine resulting in decreased H₂S loading in aqueous amine solutions, i.e., DMAE > MDEA > TEA and 3DEA-1P > 3DEA-1,2-PD. Also, an increase in the length of chain for hydroxyl group from -N- decreases the H₂S loading as seen when comparing DEAE-EO and DMAH (DEAE-EO shows higher capacity). Moreover, by adding the ethoxy group in DEEA, (DEAE-EO) increases the H₂S loading significantly.

Effects of pKa on H₂S loading in 20 wt.% amine solutions in MEG and TEG at T = 5 °C ± 0.1 °C; pH₂S = 1 kPa are shown in Figures 4 and 5 respectively. In each of the figures, a weak trend with respect to pKa is observed. H₂S loading in (DEAE-EO)-MEG solutions is higher than all other amine-MEG solutions and is even higher than aqueous MDEA solution shown in Figure 3. Increase in alkyl group in amine-TEG solutions and amine-MEG solutions are in line with each other, but not with the trend seen in aqueous solutions. However, adding an ethoxy group in DEEA has a similar effect in all three solutions.

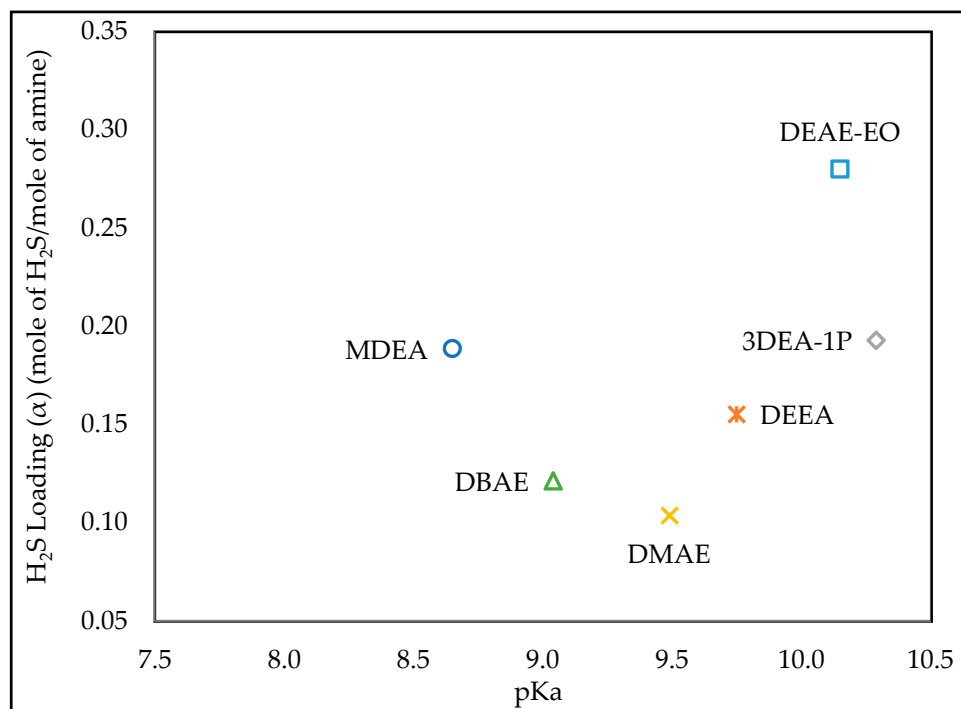


Figure 4. Effect of pKa on H₂S loading in amine-MEG solutions; T = 5 °C ± 0.1 °C; pH₂S = 1 kPa; amine concentration = 20 wt.% (unloaded); DBAE solutions makes two phases.

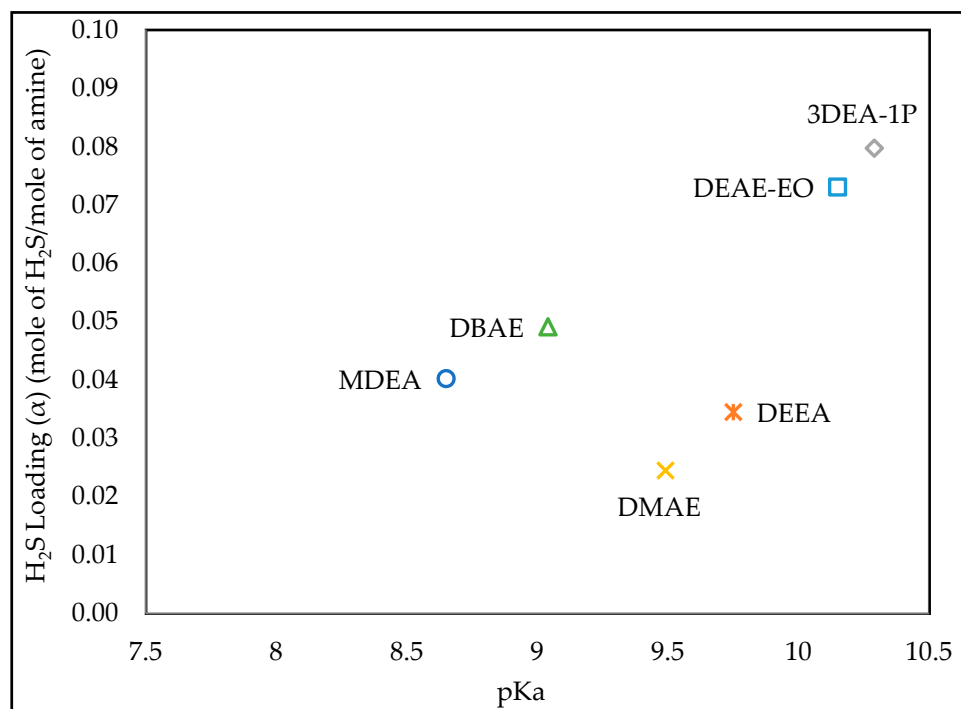


Figure 5. Effect of pKa on H₂S loading in amine-TEG solutions; T = 5 °C ± 0.1 °C; pH₂S = 1 kPa; amine concentration = 20 wt.% (unloaded); DBAE solutions makes two phases.

3.2. Effect of Solvent

Each aqueous amine solution gives more H₂S absorption capacity than its non-aqueous counterpart when compared on weight bases and having same system temperature, inlet partial pressure of gas and residence time of gas in the reactor as shown in Figure 6. Change of solvent from water to ethylene glycol or tri-ethylene glycol has a similar effect on all the amine solutions.

Replacing the solvent from water to monoethylene glycol decrease the H₂S loading significantly, the maximum decrease was observed in DEEA solutions while minimum has observed DBAE solutions. H₂S absorption decreased more rapidly when TEG had used as a solvent compared to MEG or water. Visual inspection also showed TEG solutions become more viscous as compared to MEG and H₂O solutions in respective amines. Furthermore, MEG shows more reactivity than TEG due to the autoprotolyses. However, the H₂S absorption capacity in TEG solutions is expected to increase significantly if water is present even at relatively low amounts [40].

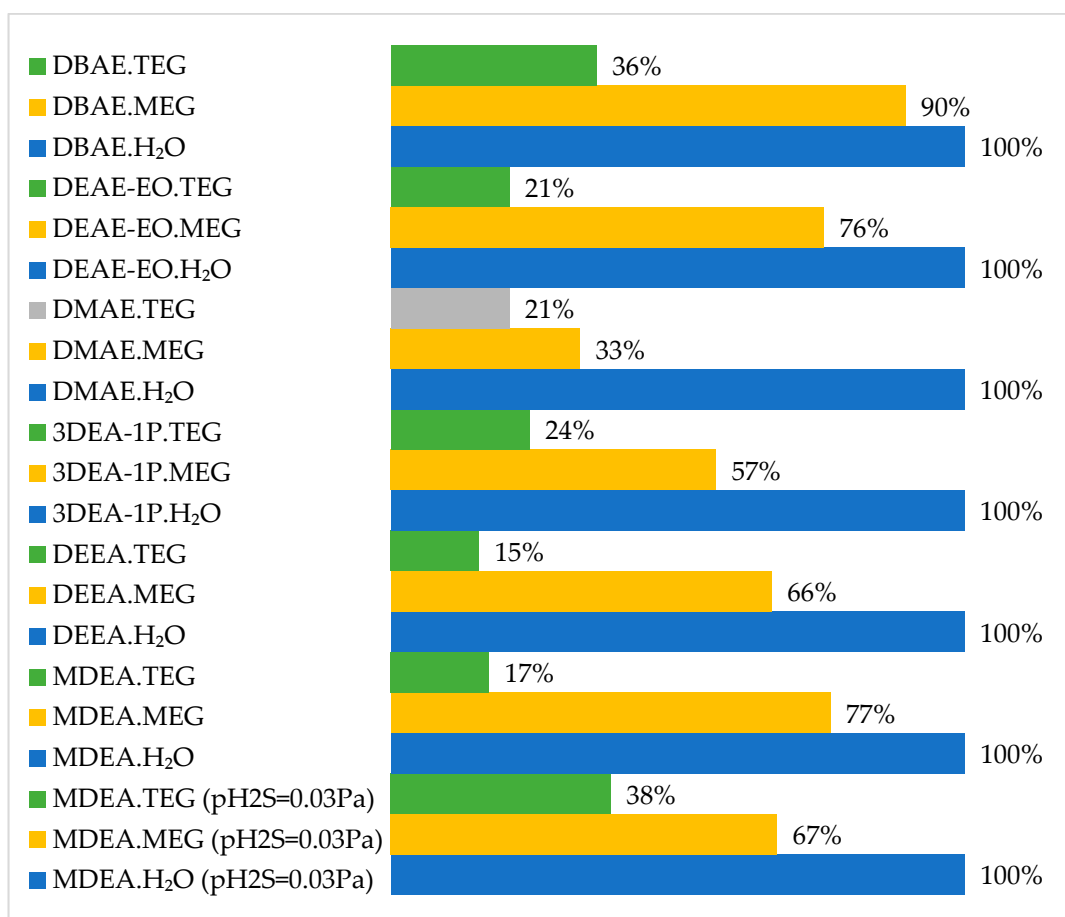


Figure 6. Effect of solvent on absorbed H₂S; T = 5 °C ± 0.1 °C; p_{H₂S} = 1 kPa; mine concentration = 20 wt.% (unloaded); DBAE solutions makes two phases.

3.3. Effect of H₂S Partial Pressure

Hydrogen sulfide loading as the function of inlet H₂S partial pressure (p_{H₂S}) at T = 5 °C ± 0.1 °C for 20 wt.% amine solutions is shown in Figure 7. The rise in inlet H₂S partial pressure (p_{H₂S}) increases the H₂S loading at given temperature and amine concentration for both aqueous and non-aqueous solutions except DEAE-EO by providing more reaction sites for reaction between H₂S and amine solutions. The same trend was seen in previous studies. However, in aqueous DEAE-EO solution, H₂S loading starts to decrease with increases in p_{H₂S} from 0.5 kPa to 1.0 kPa for an unknown reason. It is not possible to explain the behavior with the current data.

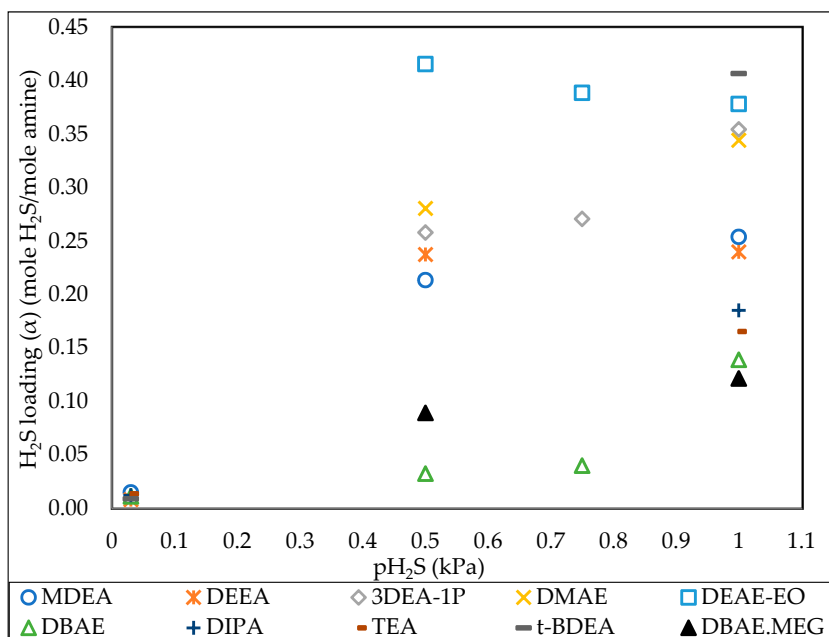


Figure 7. Effect of inlet H₂S partial pressure on H₂S loading; T = 5 °C ± 0.1 °C; amine concentration = 20 wt.% (unloaded); DBAE solutions makes two phases.

3.4. Effect of Temperature

The effect of temperature on H₂S loading on various 20 wt.% amine solutions at pH₂S = 1 kPa is shown in Figure 8. As the screening temperature increases from 5 °C to 40 °C H₂S loading decreases for all solutions except DEAE-EO and DEEA. The decrease in loading is as expected since the final loading in the experiments is almost in equilibria with the gas phase [16,20]. For DEEA, the loading difference between 5 °C and 40 °C is 0.01 mol H₂S/mol DEEA indicating that loading capacity is not as dependent on temperature as for some of the other amines. In the case of DEAE-EO, the changes are larger: The loading difference between 5 °C and 40 °C is 1.8% which is within our analytical uncertainty. However, the reason for the increase in loading seen at 25 °C, is unknown. We believe this is due to uncertainties in the analysis of H₂S and amine concentrations in the liquid samples.

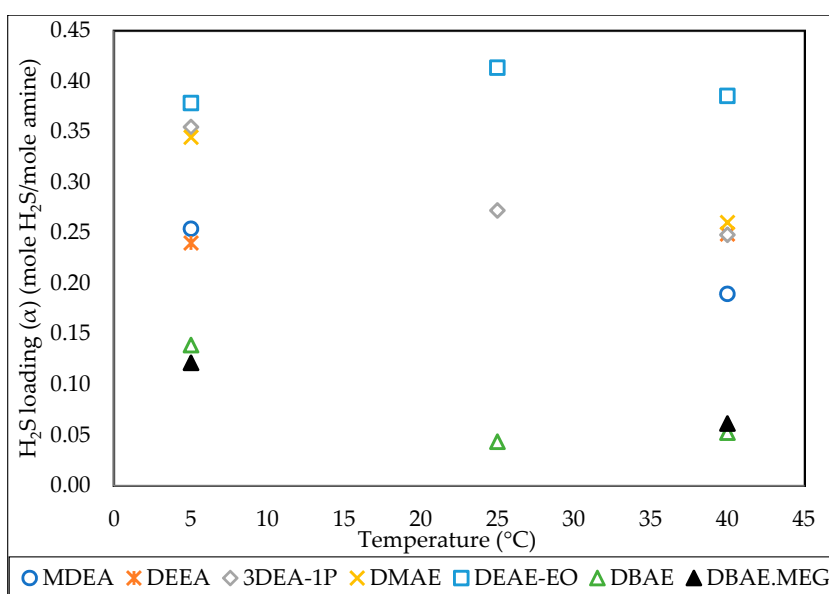


Figure 8. Effect of temperature on H₂S loading; pH₂S = 1 kPa; amine concentration = 20 wt.% (unloaded); DBAE solutions makes two phases.

3.5. Effect of Amine Concentration

Figure 9 shows the effect of amine concentration on hydrogen sulfide loading in aqueous solutions at 5 °C and inlet H₂S partial pressure of 1 kPa. The increase in amine concentration from 20 wt.% to 50 wt.% at given temperature and pressure decreases the H₂S absorption (mole/mol) subsequently decreasing H₂S loading. The trends are similar to those reported for MDEA as seen in the figure. In case of MDEA, the absorption capacity decreases by 40–50% when the MDEA concentration increases from 2.5 mol/kg to 4.2 mol/kg and it is similar to the reduction seen for DBEA. For 3DEA-1P and DEAE-EO a higher reduction in the absorption capacity is seen. Overall, the results indicate that increase in amine concentration changes the vapor-liquid equilibria behavior of the system [16,27,41,42].

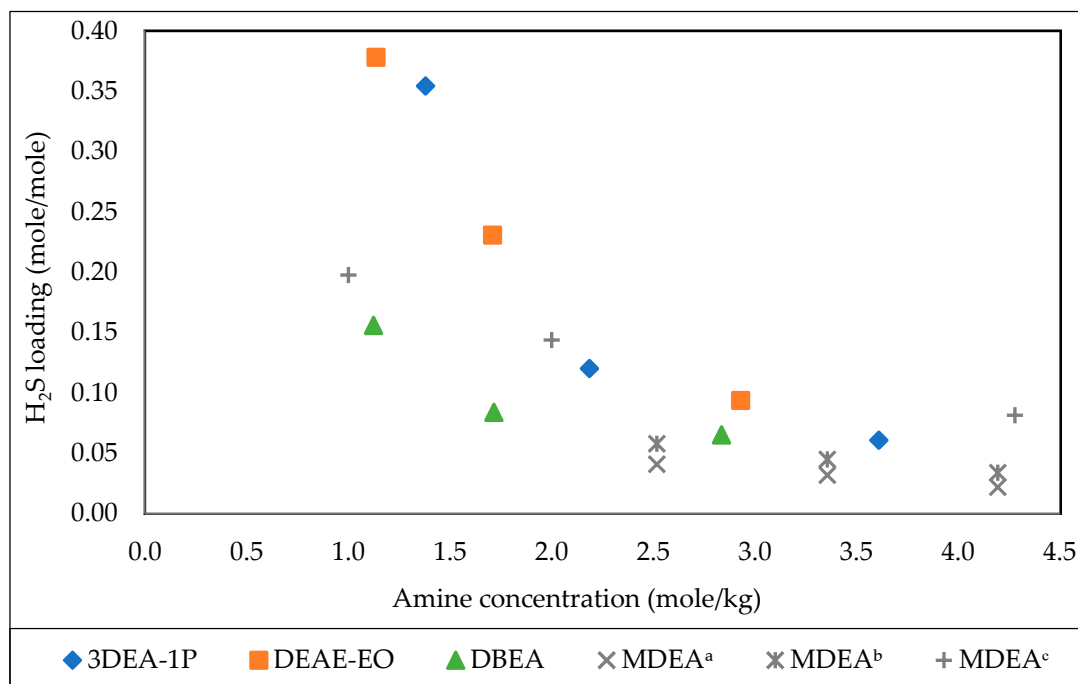


Figure 9. Effect of amine concentration on H₂S loading; T = 5 °C ± 0.1 °C; p_{H₂S} = 1 kPa for all amines except MDEA; DBAE solutions makes two phases.; MDEA is at T = 40 °C and MDEA^a p_{H₂S} = 0.3 kPa [27]; MDEA^b p_{H₂S} = 0.5 kPa [27], MDEA^c p_{H₂S} = 1 kPa [16].

The data at 50 wt% allows us to compare the absorption capacity of 3DEA-1P, DEAE-EO and DBEA in aqueous and MEG solutions with similar mole fraction (mole amine/mole solution). The mole fraction of 3DEA-1P in 3DEA-1P.MEG solution (0.13) is similar to that of aqueous 50 wt% DEA-1P (0.15). Likewise, DEAE-EO and DBEA have similar mole fraction for 50 wt% aqueous solutions and 20 wt.% MEG solutions. For these three amines, the absorption capacity is 60–80% higher in the presence of MEG as compared to water (Figure 10). Further studies will be required to explain the performance differences between water and MEG based solvents.

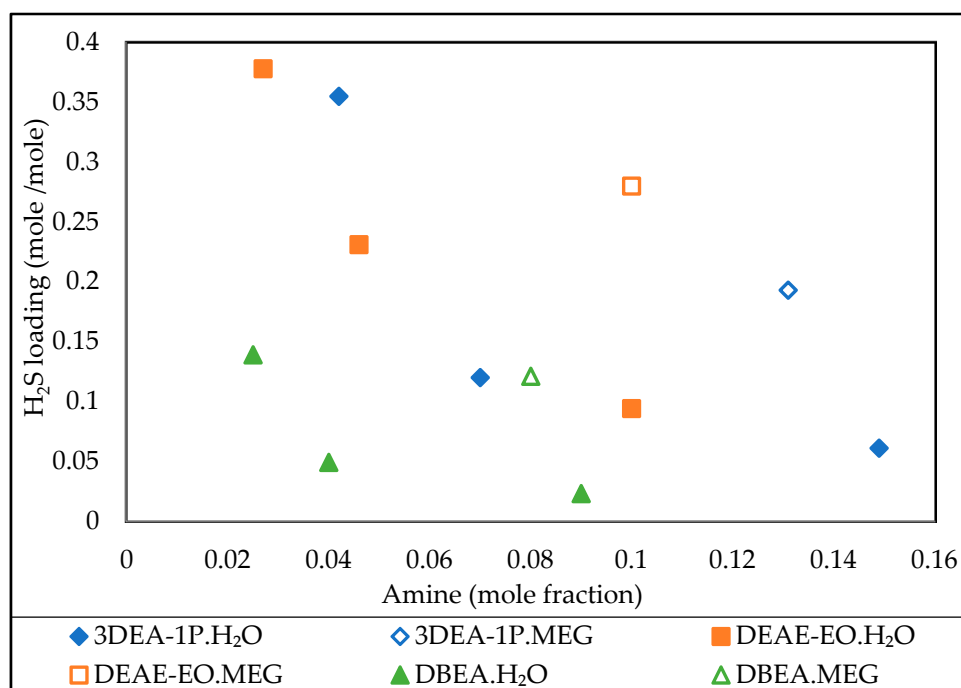


Figure 10. Effect of amine (mole fraction) on H₂S loading; T = 5 °C ± 0.1 °C; p_{H₂S} = 1 kPa; DBAE solutions makes two phases.

4. Conclusions

In this study, various new aqueous and non-aqueous amine blends have been tested for H₂S absorption. The results show that an increase in hydroxyl group and addition of ethoxy group in amines increases the H₂S absorption in aqueous amine solutions. In general, the H₂S absorption increases also with increasing pK_a. Also, increase in alkyl group enhances the H₂S loading in aqueous ethanol amines and vice versa for aqueous propanol amines. Several of the tested amines show higher H₂S absorption capacity compared to MDEA in aqueous solutions. Even though replacing water with TEG or MEG significantly decreased the H₂S loading in all tested solvents, the non-aqueous solution of (DEAE-EO)-MEG showed higher loading than aqueous MDEA at same weight concentration.

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References

1. BP Statistical Review of World Energy. 67th Edition. 2018. Available online: <https://www.bp.com/content/dam/bp/en/corporate/pdf/energy-economics/statistical-review/bp-stats-review-2018-full-report.pdf> (accessed on 19 October 2018).
2. Subramaniam, R.; Yasa, S.; Bertrand, T.; Fontenot, B.; Dupuis, T.F.; Hernandez, R. Advanced simulation of H₂S scavenging process with triazine at different depths of gas well. *J. Nat. Gas Sci. Eng.* **2018**, *49*, 417–427. [CrossRef]
3. El-Gendy, N.S.; Speight, J.G. *Handbook of Refinery Desulfurization*; Taylor & Francis: Boca Raton, FL, USA, 2015.

4. Doujaiji, B.; Al-Tawfiq, J.A. Hydrogen sulfide exposure in an adult male. *Ann. Saudi Med.* **2010**, *30*, 76–80. [[CrossRef](#)] [[PubMed](#)]
5. Amosa, M.; Mohammed, I.; Yaro, S. Sulphide scavengers in oil and gas industry—A review. *Nafta* **2010**, *61*, 85–92.
6. *Grunnlag for Fastsettelse av Administrativ Norm for for Hydrogensulfid (H₂S)*; Arbeidstilsynet Statens Hus: Trondheim, Norway, 2011; Volume 7468.
7. Kohl, A.L.; Nielsen, R.B. Chapter 2—Alkanolamines for Hydrogen Sulfide and Carbon Dioxide Removal. In *Gas Purification*, 5th ed.; Gulf Professional Publishing: Houston, TX, USA, 1997; pp. 40–186.
8. Kohl, A.L.; Nielsen, R.B. Chapter 11—Absorption of Water Vapor by Dehydrating Solutions. In *Gas Purification*, 5th ed.; Kohl, A.L., Nielsen, R.B., Eds.; Gulf Professional Publishing: Houston, TX, USA, 1997; pp. 946–1021.
9. Hutchinson, A.J. Process for Treating Gases. U.S. Patent Application No. 2177068 A, 24 October 1939.
10. McCartney, E.R. Gas Purification and Dehydration Process. U.S. Patent Application No. 2435089, 27 January 1948.
11. McCartney, E.R. Extraction of acidic impurities and moisture from gases. U.S. Patent Application No. 2547278, 3 April 1951.
12. Chapin, W.F. Purification and Dehydration of Gases. U.S. Patent Application No. 2518752, 2 August 1950.
13. Shoukat, U.; Fytianos, G.; Knuutila, H.K. Thermal Stability and Corrosion Studies of Amines for Combined Acid Gas Removal and Hydrate Control for Subsea Gas Treatment Systems. In Proceedings of the 2016 Techno-Ocean (Techno-Ocean), Kobe, Japan, 6–8 October 2016; pp. 176–180.
14. Eimer, D. *Simultaneous Removal of Water and Hydrogen Sulphide from Natural Gas*; Department of Chemical Engineering, Norwegian University of Science and Technology: Trondheim, Norway, 1994.
15. Mathias, P.M.; Jaspersen, L.V.; VonNiederhausen, D.; Bearden, M.D.; Koech, P.K.; Freeman, C.J.; Heldebrant, D.J. Assessing anhydrous tertiary alkanolamines for high-pressure gas purifications. *Ind. Eng. Chem. Res.* **2013**, *52*, 17562–17572. [[CrossRef](#)]
16. Jou, F.Y.; Mather, A.E.; Otto, F.D. Solubility of hydrogen sulfide and carbon dioxide in aqueous methyldiethanolamine solutions. *Ind. Eng. Chem. Process Des. Dev.* **1982**, *21*, 539–544. [[CrossRef](#)]
17. Macgregor, R.J.; Mather, A.E. Equilibrium solubility of H₂S and CO₂ and their mixtures in a mixed solvent. *Can. J. Chem. Eng.* **1991**, *69*, 1357–1366. [[CrossRef](#)]
18. Li, M.H.; Shen, K.P. Solubility of hydrogen sulfide in aqueous mixtures of monoethanolamine with N-methyldiethanolamine. *J. Chem. Eng. Data* **1993**, *38*, 105–108. [[CrossRef](#)]
19. Kuranov, G.; Rumpf, B.; Smirnova, N.; Maurer, G. Solubility of single gases carbon dioxide and hydrogen sulfide in aqueous solutions of N-methyldiethanolamine in the temperature range 313–413 K at pressures up to 5 MPa. *Ind. Eng. Chem. Res.* **1996**, *35*, 1959–1966. [[CrossRef](#)]
20. Lemoine, B.; Li, Y.G.; Cadours, R.; Bouallou, C.; Richon, D. Partial vapor pressure of CO₂ and H₂S over aqueous methyldiethanolamine solutions. *Fluid Phase Equilib.* **2000**, *172*, 261–277. [[CrossRef](#)]
21. Pérez-Salado Kamps, Á.; Balaban, A.; Jödecke, M.; Kuranov, G.; Smirnova, N.A.; Maurer, G. Solubility of single gases carbon dioxide and hydrogen sulfide in aqueous solutions of N-methyldiethanolamine at temperatures from 313 to 393 K and pressures up to 7.6 MPa: New experimental data and model extension. *Ind. Eng. Chem. Res.* **2001**, *40*, 696–706.
22. Sidi-Boumedine, R.; Horstmann, S.; Fischer, K.; Provost, E.; Fürst, W.; Gmehling, J. Experimental determination of hydrogen sulfide solubility data in aqueous alkanolamine solutions. *Fluid Phase Equilib.* **2004**, *218*, 149–155. [[CrossRef](#)]
23. Huttenhuis, P.J.G.; Agrawal, N.J.; Hogendoorn, J.A.; Versteeg, G.F. Gas solubility of H₂S and CO₂ in aqueous solutions of N-methyldiethanolamine. *J. Pet. Sci. Eng.* **2007**, *55*, 122–134. [[CrossRef](#)]
24. Xu, H.J.; Zhang, C.F.; Zheng, Z.S. Solubility of hydrogen sulfide and carbon dioxide in a solution of methyldiethanolamine mixed with ethylene glycol. *Ind. Eng. Chem. Res.* **2002**, *41*, 6175–6180. [[CrossRef](#)]
25. Sadegh, N.; Thomsen, K.; Solbraa, E.; Johannessen, E.; Rudolfsen, G.I.; Berg, O.J. Solubility of hydrogen sulfide in aqueous solutions of N-methyldiethanolamine at high pressures. *Fluid Phase Equilib.* **2015**, *393*, 33–39. [[CrossRef](#)]
26. Maddox, R.N.; Bhairi, A.H.; Diers, J.R.; Thomas, P.A. Equilibrium solubility of carbon dioxide or hydrogen sulfide in aqueous solutions of monoethanolamine, diglycolamine, diethanolamine and methyldiethanolamine. In *GPA Research Report: Project 104*; GPA: Tulsa, OK, USA, 1987.

27. Tian, X.; Wang, L.; Fu, D.; Li, C. Absorption and Removal Efficiency of Low-Partial-Pressure H₂S in a Monoethanolamine-Activated N-Methyldiethanolamine Aqueous Solution. *Energy Fuels* **2019**, *33*, 629–635. [[CrossRef](#)]
28. Jagushte, M.V.; Mahajani, V.V. Low pressure equilibrium between H₂S and alkanolamine revisited. *Ind. J. Chem. Technol.* **1999**, *6*, 125–133.
29. Mazloumi, S.H.; Haghtalab, A.; Jalili, A.H.; Shokouhi, M. Solubility of H₂S in aqueous diisopropanolamine + piperazine solutions: New experimental data and modeling with the electrolyte cubic square-well equation of state. *J. Chem. Eng. Data* **2012**, *57*, 2625–2631. [[CrossRef](#)]
30. Chowdhury, F.A.; Yamada, H.; Higashii, T.; Goto, K.; Onoda, M. CO₂ capture by tertiary amine absorbents: A performance comparison study. *Ind. Eng. Chem. Res.* **2013**, *52*, 8323–8331. [[CrossRef](#)]
31. Hartono, A.; Vevelstad, S.J.; Ciftja, A.; Knuutila, H.K. Screening of strong bicarbonate forming solvents for CO₂ capture. *Int. J. Greenh. Gas Control* **2017**, *58*, 201–211. [[CrossRef](#)]
32. ScolarTM, S. *Advanced Chemistry Development (ACD/Labs) Software V11.02 2016*; ACD/Labs: Toronto, ON, Canada, 2016.
33. Hamborg, E.S.; Versteeg, G.F. Dissociation Constants and Thermodynamic Properties of Amines and Alkanolamines from (293 to 353) K. *J. Chem. Eng. Data* **2009**, *54*, 1318–1328. [[CrossRef](#)]
34. Woolley, E.M.; Tomkins, J.; Hepler, L.G. Ionization constants for very weak organic acids in aqueous solution and apparent ionization constants for water in aqueous organic mixtures. *J. Solut. Chem.* **1972**, *1*, 341–351. [[CrossRef](#)]
35. Jacob, P. Potential Membrane Based Treatment of Triethylene Glycol Wastewater from Gas Separation Plant. *J. Water Sustain.* **2014**, *4*, 123–136.
36. Ma'mun, S.; Jakobsen, J.P.; Svendsen, H.F.; Juliussen, O. Experimental and Modeling Study of the Solubility of Carbon Dioxide in Aqueous 30 Mass% 2-((2-Aminoethyl)amino)ethanol Solution. *Ind. Eng. Chem. Res.* **2006**, *45*, 2505–2512. [[CrossRef](#)]
37. Shoukat, U.; Baumeister, E.; Pinto, D.D.D.; Knuutila, H.K. Thermal stability and corrosion of tertiary amines in aqueous amine and amine-glycol-water solutions for combined acid gas and water removal. *J. Nat. Gas Sci. Eng.* **2019**, *62*, 26–37. [[CrossRef](#)]
38. Bernhardsen, I.M.; Krokvik, I.R.T.; Jens, K.-J.; Knuutila, H.K. Performance of MAPA Promoted Tertiary Amine Systems for CO₂ Absorption: Influence of Alkyl Chain Length and Hydroxyl Groups. *Energy Procedia* **2017**, *114*, 1682–1688. [[CrossRef](#)]
39. El Hadri, N.; Quang, D.V.; Goetheer, E.L.V.; Zahra, M.R.M.A. Aqueous amine solution characterization for post-combustion CO₂ capture process. *Appl. Energy* **2017**, *185*, 1433–1449. [[CrossRef](#)]
40. Eimer, D. *Gas Treating: Absorption Theory and Practice*; Wiley: Hoboken, NJ, USA, 2014.
41. Fu, D.; Chen, L.; Qin, L. Experiment and model for the viscosity of carbonated MDEA–MEA aqueous solutions. *Fluid Phase Equilib.* **2012**, *319*, 42–47. [[CrossRef](#)]
42. Fu, D.; Wang, L.; Zhang, P.; Mi, C. Solubility and viscosity for CO₂ capture process using MEA promoted DEAE aqueous solution. *J. Chem. Thermodyn.* **2016**, *95*, 136–141. [[CrossRef](#)]

