Comparative Analysis of the Efficiency Penalty in Power Plants of Different Amine-Based Solvents for CO₂ Capture

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ABSTRACT: Amine solvents are one of the main options for post-combustion CO₂ capture applications. The main drawback of the carbon capture processes is the required energy to regenerate the solvent once it has reacted with the CO₂. When applied to a power plant, the energy requirement has an important impact on the net efficiency of the overall system. Several solvents, i.e., monoethanolamine (MEA), diethanolamine (DEA), methyl diethanolamine (MDEA), and many others have been proposed to overcome this drawback. Regeneration temperature and heat duty reduction are considered to be the significant objectives. Moreover, enhancement of the amine's concentration and its working capacity without the impact on the other variables are important. In this work, different types of amines with a wide range of heat duty and regeneration temperatures under the same set of assumptions are calculated and compared. The effect of both variables on the energy penalty caused by carbon capture is measured. A review of amines and their effects on the net efficiency of the overall system (power plant, chemical absorption, CO₂ compression) are conducted and analyzed. As expected, the impact of heat duty is greater than the modification of regeneration temperature. The effect of reducing 1 GJ/ton CO₂ in the heat duty is similar to the effect of reducing the regeneration temperature from 40 °C to 25 °C.
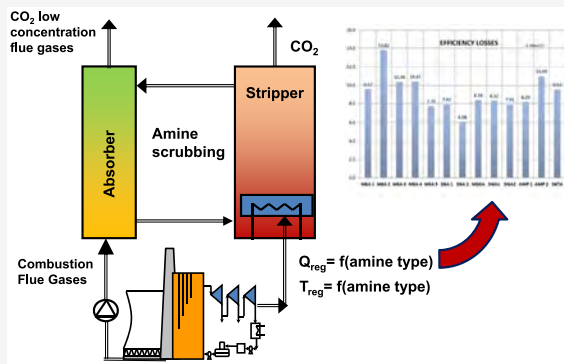
1. INTRODUCTION

The global climate policy is focused on the deployment of renewable technologies to fight against climate change caused by CO₂ emissions. Nevertheless, some heavy industries are still dependent on the consumption of fossil fuels rather than electricity for their operation. Moreover, renewables are generally capital-intensive and, in some scenarios, their availability and accessibility reduces their competitiveness with fossil fuels.[1] As a result, fossil fuels seemingly would still play an important role in the future energy mix of the world; therefore, nations should reconsider their climate policies and turn their attention toward CO₂ capture and storage (CCS) technologies if they are inclined to attain a carbon-free and, more importantly, reliable energy system.

Chemical absorption based on amine solvents is the main technology for post-combustion CO₂ capture.[2] The technology has some drawbacks, such as giant energy consumption for regeneration of the CO₂-loaded amine solution, degradation of the amine solvent, and equipment corrosion.[3] But the energy consumption of regeneration is considered to be the major challenge, because it consumes more than 70% of the total operating costs of a CO₂ capture plant. Consequently, the reduction in heat duty of regeneration should be the main priority for further development in amine-based CO₂ capture technology.[4]

Several works in the literature have demonstrated, through techno-economic analysis, the improvement or advantages of using one specific solvent instead of, mainly, monoethanolamine (MEA). Nevertheless, MEA is the most-common, most-developed, and most-demonstrated solvent for carbon capture. For example, an optimization and economic analysis for an amine-based acid gas capture unit using MEA and methyl diethanolamine (MDEA) has been reported.[5] In that report, Aspen Plus software was utilized to demonstrate that the use of an MEA/MDEA mixture, in lieu of MEA, leads to a surge in CO₂ removal rate and a reduction in energy penalty. For industrial applications, it has been highlighted, for a cement plant, that the total equipment cost and capital expenditure and the energy penalty of the AMP-PZ-MEA blend (where AMP represents 2-amino-2-methyl-1-propanol and PZ represents piperazine) is lower, compared to the MEA.[6] Furthermore, it was shown that the effect of a reduction in the percentage of CO₂ captured in an industrial CCS plant integrated to the natural gas fired power
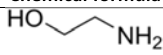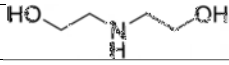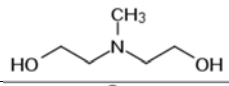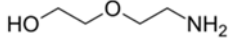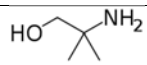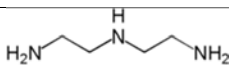
Table 1. Amines Used in This Work

Acronym	Name	Chemical formula
MEA	Monoethanolamine 2-aminoethanol	
DEA	Diethanolamine bis(2-Hydroxyethyl)amine	
MDEA	Methyl diethanolamine Bis(2-hydroxyethyl) methyl amine	
DGA	Diglycolamine 2-(2-aminoethoxy)ethanol	
AMP	Aminomethylpropanol 2-amino-2-methyl-1-propanol	
DETA	Diethyelenetriamine 2,2'-Diaminodiethylamine	
NH3	Ammonia	NH_3

plant. The influence of the CCS plant on the efficiency of the power plant was analyzed. The efficiency of power plants declined from 33.22% to 32.86% by capturing 5% of the flue gas.[7]

It is more common in the literature to present comparisons between several solvents to prove that the improvement can be attributed to some specific variables. For energy regeneration, it has been shown that DEA had superiority over MEA for several configurations studied, in terms of total equivalent work and reboiler duty;[8] amine-carbonate blend with 30 wt % (MEA) experimentally needed 10% lower energy regeneration than MEA;[9] or the use of triblend amines (MEA-MDEA-PZ) was an excellent approach for increasing CO_2 desorption activity and extremely declining the energy penalty of amine-based CO_2 capture systems.[10] CO_2 absorption rate,[10,11] and high absorption and desorption efficiencies[12] are other variables that have been used for comparison.

Despite all of the literature, there is a lack of literature to compare, energetically, the performance of different solvents under common assumptions. Some relevant attempts have been made using Aspen and multiobjective genetic algorithm optimization to select the best process design and operating conditions for MEA, diethanolamine (DEA), and AMP.[13] Generally, most of the studies are specific for pilot plants and use ASPEN software for simulations. MEA with a heat consumption of ~3.6−3.7 GJ/tCO_2 is the most popular option used in these works[8,14,15,16]. Different capture rates were reported in these studies, from 79.9%[8] to 86%[14,15] Studies also shows that there is room for improvements and the previous value of 3.6 GJ/t CO_2 could be reduced to 3.1 GJ/t CO_2.[16] They did not show any effect of this reduction on hypothetical power plant efficiency. The use of other solvents is not as numerous as MEA; for example, blended amines with a combination of mixtures of AMP and MDEA, using DEA and MEA as activating agents, has been also simulated for a 600 MWe conventional coal-fired power plant[17] or MEA, PZ, and PZ−MDEA blends for the flue gas from the Norcem Brevik cement plant.[18] Results showed that the mixture of 5 wt % DEA and 25 wt % AMP performed better than the other mixtures by obtaining heat consumptions of 3.03 GJ/tCO_2. Similarly, ammonia, as a carbon dioxide solvent, has been compared and evaluated under optimum operating conditions that minimize the impact on the thermal performance of the regeneration, depending on the CO_2 loading of the lean solution and the ammonia concentration.[19]

In this work, a comparative analysis, in terms of the efficiency penalty in power plants, is conducted for different amine-based solvents aimed at capturing CO_2. First, a sensitivity analysis of the efficiency penalty, which is dependent on regeneration temperature and stripper heat duty, is presented. Then, a comparison of the impact of several amines with different energy requirements on the efficiency of the power plants, based on literature data, is presented. Note that, apart from the significant energy requirement, it is not the only variable to be considered when the amine is selected for CO_2 capture. Operability, cost, maintenance, corrosive risk, etc. should be also considered.

2. AMINE-BASED SOLVENTS FOR CO₂ CAPTURE

Several types of amines have been chosen for simulations (Table 1). Amines are organic compounds that can act as bases by protonation of the lone pair of electrons in the nitrogen atom. In the case of CO_2 capture, a molecule of H_2O is involved in the reaction leading to the formation of substituted ammonium bicarbonates:

$$R_1R_2R_3N + H_2O + CO_2 \rightarrow R_1R_2R_3NH^+HCO_3^-$$
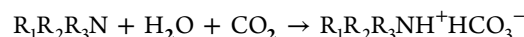
Besides this general mechanism of reaction, ammonia and amines containing a hydrogen atom bonded to the nitrogen one—that is, primary and secondary amines—can react with CO_2, forming substituted ammonium carbamates. The reaction proceeds in two steps:

$$R_1R_2NH + CO_2 \rightarrow R_1R_2NH^+HCOO^-$$

$$R_1R_2NH^+HCOO^- + R_1R_2NH \rightarrow R_1R_2NCOO^- + R_1R_2NH_2^+$$
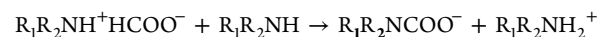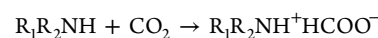
The first reaction involves a molecule of amine plus a molecule of CO_2 to form a zwitterion that, subsequently, forms the substituted ammonium carbamate.

It can be seen from the reaction equations that absorption as bicarbonate requires one molecule of amine per molecule of CO_2, while the formation of ammonium carbamate requires two molecules of amine per molecule of CO_2. Despite the different loading of amine required, primary and secondary amines are mainly used because of the different kinetics of the reactions, with bicarbonate formation being much slower that carbamate formation. Work done in the capture of CO_2 by different amines show that the reaction constants can vary from 7500 m³/s/kmol for MEA[20] to 1200 m³/s/kmol for DEA[21] and to 3.5 for MDEA.[22] A mixed primary-secondary amine, such as DETA, has

a reaction constant of 47 740 m^3/s/kmol.[23] Other amines not used in this work, such as piperazine, have been proposed.[24,25] The use of other amines can be found in a review.[26]

Tables 2 and 3 shows the amines used in this work. Apart from ammonia, there are three primary amines (MEA, DGA, and AMP), a secondary amine (DEA), a tertiary amine (MDEA), and a mixed primary−secondary amine (DETA).

Primary amines seem to be more capable to capture CO_2, while they have the highest reaction enthalpies, meaning that the regeneration operation will be conducted under strong

Table 2. Review of Regeneration Temperature and Heat Duties of MEA and DEA

temperature (°C)	heat duty (MJ/kg CO_2)	details	data source
		MEA	
116.1	4.0	30 wt %, 0.25 lean loading, 0.478 rich loading, 1.5 bar	ref 8
116.3	**3.9**	**30 wt %, 0.25 lean loading, 0.481 rich loading, 1.5 bar**	ref 10
115.2	4.0	30 wt %, 0.25 lean loading, 0.477 rich loading, 1.5 bar	ref 10
115.0	8.7	0.30 rich loading	ref 29
115.0	6.5	0.35 rich loading	ref 29
115.0	4.5	0.45 rich loading	ref 29
130.0	12.0	0.21 lean loading, 0.5 rich loading (4 kmol/m^3), 1 atm	ref 30
130.0	**5.4**	**0.22 rich loading, 0.5 rich loading (4 mol/m^3), 1 atm**	ref 30
130.0	4.8	0.26 rich loading, 0.5 rich loading (4 mol/m^3), 1 atm	ref 30
130.0	**3.8**	**0.32 lean loading, 0.5 rich loading (5 kmol/m^3), 1 atm**	ref 30
120.0	**4.3**	**30 wt %, 0.16 lean loading, 0.42 rich loading, 2 bar**	ref 31
112.0	4.55	30 wt %	ref 32
112.0	3.36	30 wt %, process improvement	ref 32
120.0	3.98	30.3 wt %	ref 33
126.7	4.01	0.414 lean loading, 29.3 wt % MEA, 2.23 bar	ref 16
115.0	3.75	150 kPa	ref 16
133.0	3.45	275 kPa	ref 16
114.0	**3.09**	**0.30 lean loading, 0.477 rich loading, 7 mol/kgMEA (30 wt %), 1.7 bar**	ref 34
71.6	16.0	20 wt %, 30 kPa	ref 27
83.8	11.0	20 wt %, 50 kPa	ref 27
94.3	9.0	20 wt %, 75 kPa	ref 27
114.1	6.75	20 wt %, 150 kPa	ref 27
123.0	5.1	20 wt %, 200 kPa	ref 27
130.0	4.7	20 wt %, 250 kPa	ref 27
136.0	4.4	20 wt %, 300 kPa	ref 27
		DEA	
70.4	4.2	40 wt %, 30 kPa	ref 27
82.3	4.0	40 wt %, 50 kPa	ref 27
92.0	3.8	40 wt %, 75 kPa	ref 27
110.9	**3.4**	**40 wt %, 150 kPa**	ref 27
114.9	3.3	40 wt %, 175 kPa	ref 27
117.1	3.2	40 wt %, 200 kPa	ref 27
127.6	3.05	40 wt %, 300 kPa	ref 27
−	3.55	30 wt %	ref 35
−	3.15	40 wt %	ref 35
−	2.9	50 wt %	ref 35
−	2.8	0.447 rich loading (mol CO_2/mol DEA)	ref 35

Table 3. Review of Regeneration Temperature and Heat Duties of MDEA, DGA, AMP, DETA, and NH$_3$

temperature (°C)	heat duty (MJ/kg CO_2)	details	data source
		MDEA	
130.0	1.2	0.06 lean loading, 0.5 rich loading (4 kmol/m^3 solvent), 1 atm	ref 30
130.0	**2.8**	**0.04 lean loading, 0.5 rich loading (4 kmol/m^3 solvent), 1 atm**	ref 30
130.0	7.5	0.03 lean loading, 0.5 rich loading (4 kmol/m^3 solvent), 1 atm	ref 30
120.0	4.36	3/27 wt % DEA/MDEA	ref 36
112.9	3.82	30 wt %	ref 36
		DGA	
85.8	4.5	60 wt %, 50 kPa	ref 37
96.5	**4.1**	**60 wt %, 75 kPa**	ref 37
110.9	3.2	60 wt %, 150 kPa	ref 27
116.0	2.9	60 wt %, 200 kPa	ref 27
120.7	**2.8**	**60 wt %, 250 kPa**	ref 27
124.1	2.7	60 wt %, 300 kPa	ref 27
		AMP	
115.9	3.18	30 wt %	ref 36
95.2	3.6	5/25 wt % DEA/AMP 0.41 lean loading, 0.54 rich loading	ref 17
114.7	3.05	5/25 wt % DEA/AMP 0.07 lean loading, 0.545 rich loading[17]	ref 17
95.7	3.6	10/20 wt % DEA/AMP 0.43 lean loading, 0.575 rich loading	ref 17
112.3	3.05	10/20 wt % DEA/AMP 0.11 lean loading, 0.548 rich loading	ref 17
95.2	**3.8**	**15/15 wt % DEA/AMP 0.44 lean loading, 0.568 rich loading**	ref 17
114.6	3.1	15/15 wt % DEA/AMP 0.07 lean loading, 0.541 rich loading	ref 17
115.0	9.2	5/25 wt % PZ/AMP 0.04 lean loading 0.26 rich loading	ref 38
112.0	**4.9**	**5/25 wt % PZ/AMP 0.14 lean loading, 0.28 rich loading**	ref 38
110.0	4.6	5/25 wt % PZ/AMP 0.14 lean loading, 0.28 rich loading	ref 38
		DETA	
100.0	**4.8**	**2 kmol DETA/m^3, 0.4 cyclic loading**	ref 39
100.0	2.2	2 kmol DETA/m^3, 0.56 cyclic loading	ref 39
		NH$_3$	
90.0	5.75	0.23 lean loading	ref 40
131.0	2.53	0.33 lean loading, 0.66 rich loading, 7.3 wt % NH$_3$, 10 bar	ref 41
149.7	1.86	10.2 wt % NH$_3$, 0.236 lean loading, 0.41 rich loading, 12 bar	ref 42
145.5	3.27	0.225 lean loading, 0.41 rich loading, 10 bar	ref 43
145.5	2.46	0.225 lean loading, 0.41 rich loading, process improvement, 10 bar	ref 43
110.0−139.0	4−4.2	1.9−5.8 wt % NH$_3$, 0.21−0.41 lean loading	ref 44

operating conditions. Several options for MEA are summarized in Table 2. The heat duty ranges between 3.1 GJ/ton CO_2 and 16 GJ/ton CO_2, but it is usual does not exceed 5 GJ/ton CO_2.[27] Efforts to reduce this penalty have led to the use of other types of amines, such as DEA with a heat duty of 2.8−4.2 GJ/ton CO_2 and regeneration temperatures of 70−130 °C. MDEA and DGA also show low regeneration energy requirements, with values of 2.8 GJ/ton CO_2 with usual stripper temperatures of 120−130 °C, Table 3. Furthermore, DETA and AMP, that is a sterically

hindered primary amine that produces unstable carbamates leading to easier regeneration than other primary amine carbamates although with reduced speed capture,[28] exhibit low regeneration temperatures with heat duties in the range of 3.2−4.9 GJ/ton CO_2.

3. MODELING AND SIMULATION

Simulations have been performed considering a Rankine cycle power plant that generates a gross of 500 MW$_e$ with a net efficiency of 43.42%. The boiler of the case-study power plant produces 353.0 kg/s of steam at 300 bar and 600 °C. There is a single reheat at 620 °C. At nominal load, boiler steam output is 1073 MW$_{th}$. Boiler efficiency is assumed as 92%. Overall cycle data, electricity output, and efficiency estimations are summarized in Table 4. Figure 1 shows the flow diagram of

Table 4. Overall Cycle Data, Electricity Output, and Efficiency Estimations

| parameter	value	parameter	value
boiler energy input	1073 MW$_{th}$	CO_2 produced	106.6 kg/s
gross power output	500.9 MW$_e$	gross power plant efficiency	46.70%
net power output	465.8 MW$_e$	net power plant efficiency	43.42%
thermal energy LPH	169.9 MW$_{th}$	thermal energy LPH	121.9 MW$_{th}$

the power cycle. Live steam is expanded through nine turbine stages to a condenser pressure of 0.05 bar. Two high-pressure (HP) turbine stages operate between 600 °C and 300 bar at the inlet, and 320 °C and 50 bar at the outlet. Three medium-pressure (MP) turbine stages operate between 620 °C and 45 bar at the inlet and 386 °C and 10 bar at the outlet. Finally, five low-pressure (LP) turbine stages operate up to the condenser pressure. Four HP heaters, a deaerator, and five LP heaters are present. Table 5 shows the $m-P-T$ data corresponding to the flows shown in Figure 1.

Coal consumption and CO_2 emissions have been calculated through a governmental report,[45] using the average of coal low heating value (LHV) for Spain (22.40 GJ/t coal) and its emission factor (99.40 kg CO_2/GJ$_{PCI}$). This represents a CO_2 emission of 384.5 TCO$_2$/h (106.8 kg/s) at nominal power. This value is fixed for the comparative analysis and does not influence the solvent comparison.

Simulation of the power cycle is accomplished by means of EES (Engineering Equation Solver), and the following simplifying assumptions have been used:

- For the different stages of the HP, MP, and LP turbines: A constant isentropic efficiency, equal to 0.90 for HP stages, 0.92 for MP stages, and 0.90 for LP turbines, have been assumed. Since no partial load operation is evaluated, constant pressure bleeding in HP and IP turbines are considered. For the LP turbine, since the steam bleeding is influenced by the regeneration temperatures in the stripper, small pressure variations are allowed as it is explained.

- Regarding the pressure decrease in feedwater heaters, 2% inlet pressure at both steam and water sides is assumed. When amine scrubbing is considered, the extraction pressure is set by the steam temperature required in the stripper. The pressure is set whose saturation temperature is that of regeneration and a pressure decrease of 2% is taken into account.

- A terminal temperature difference (TTD) and the temperature drain cooler approach (TDCA) have been chosen for modeling the HP and LP heaters. They remain constant for the simulations: TTD values ranged between 3 °C and 4 °C, and the values for TDCA are between 5 °C and 6 °C, except in the second low-pressure heater (LPH2), which was set to 10 °C.
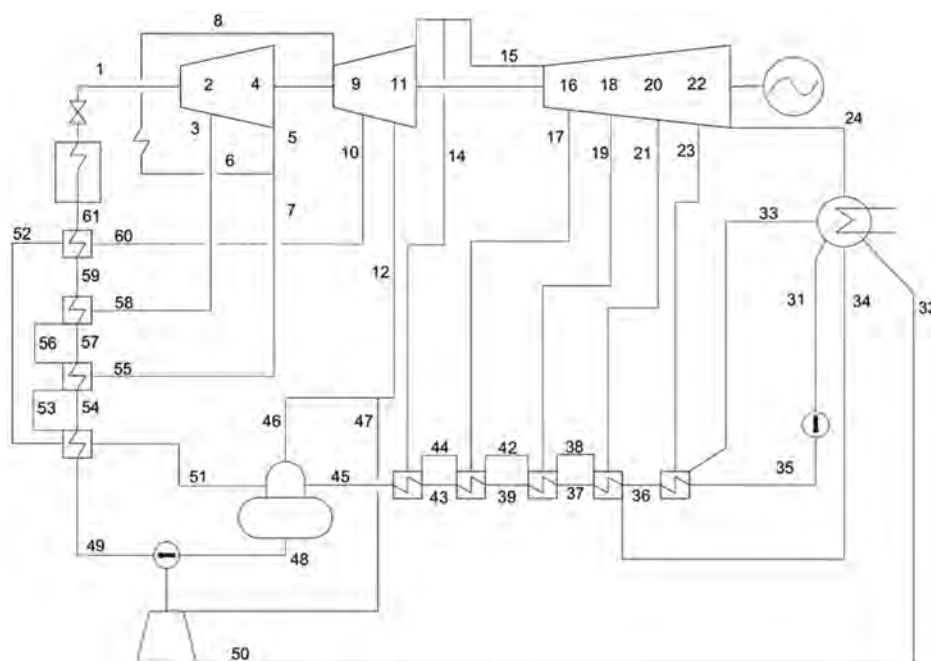
Figure 1. Layout of the Rankine steam cycle considered for simulations.

Table 5. $m-P-T$ Data Corresponding to the Rankine Steam Cycle Shown in Figure 1

stream	m (kg/s)	p (bar)	t (°C)	stream	m (kg/s)	p (bar)	t (°C)
1	353.00	300.00	600.0	33	16.02	0.29	39.4
2	328.56	75.00	375.2	34	49.17	1.18	76.1
3	24.44	75.00	375.2	35	281.48	20.91	33.4
4	328.56	50.00	320.7	36	281.48	20.49	66.1
5	328.56	50.00	320.7	37	281.48	20.08	101.3
6	304.01	50.00	320.7	38	34.08	3.43	107.3
7	24.56	50.00	320.7	39	281.48	19.68	135.9
8	304.01	45.00	620.0	41	281.48	19.68	135.9
9	296.44	29.00	545.7	42	18.85	5.88	140.9
10	7.57	29.00	545.7	43	281.48	19.29	154.8
11	266.30	18.00	470.7	44	10.08	9.90	160.8
12	30.14	18.00	470.7	45	281.48	18.90	175.9
13	266.30	10.00	386.1	46	14.96	18.00	470.7
14	10.08	10.00	386.1	47	15.18	18.00	470.7
15	256.22	10.00	386.1	48	353.00	18.52	208.5
16	247.44	6.00	320.5	49	353.00	330.00	215.3
17	8.77	6.00	320.5	50	15.18	0.05	32.1
18	232.22	3.50	257.4	51	56.56	27.85	221.3
19	15.23	3.50	257.4	52	7.57	28.42	293.5
20	217.12	1.20	150.0	53	48.99	49.00	232.9
21	15.09	1.20	150.0	54	353.00	326.70	226.9
22	201.10	0.30	69.1	55	24.56	50.00	320.7
23	16.02	0.30	69.1	56	24.44	73.50	265.7
24	201.10	0.05	32.9	57	353.00	323.43	259.7
				58	24.44	75.00	375.2
31	281.48	0.05	32.9	59	353.00	320.20	287.5
32	15.18	0.05	32.1	60	7.57	29.00	545.7

The absorber pressure is assumed to be 1 bar for all cases. To supply the energy for the rich CO_2 solvent regeneration, one LP steam bleeding is used. There are two main variables that determine the analysis. The required regeneration temperature fixes the pressure of the steam bleeding, and the required regeneration energy fixes the amount of steam mass flow needed in the steam turbine extraction. Steam bleeding pressures range from 0.3 bar of the last LP extraction before the condenser, to 10 bar of input to the LP turbine. Then, the regeneration temperature, which is the steam saturation temperature at these pressures, could range from 68.9 °C to 179.5 °C. These values agree with regeneration requirements of the selected amines. Higher regeneration temperatures require an increase in steam bleeding pressure and, consequently, a reduction in the power produced by the turbine.

In this study, it is assumed that the third steam bleeding changes the pressure to adapt the stripper necessities. The fourth steam bleeding is not considered; it is a close bleeding. This due to the thermal energy of the CO_2 compression intercooler that is used to preheat the water after the condenser. Temperature of the water after these heat exchangers is usually in the range of 89−132 °C, which is well below the maximum temperature of the CO_2 after compressors (143−160 °C). These conditions make it technically feasible for a heat exchange with enough temperature difference. A new mixing heat exchanger is located instead of the LPH2, as shown in Figure 2. Water from the condenser, after being heated in CO_2 compression intercoolers, is injected in this heat exchanger with the condensed water from the stripper and the condensed water from the third low-pressure heater (LPH3). When the regeneration temperature increases, the stripper pressure must be higher than the second low-pressure steam bleeding. In order to avoid any potential

problem or unfair results comparison for all cases, the pressure of the second low-pressure steam bleeding is calculated as the average of the first (10 bar) and stripper (depending on the regeneration temperature) low-pressure steam bleeding.

Figure 2 illustrates the original layout of low-pressure heaters and two modified layouts to take into account the carbon capture installation and its integration in the power plant. In some cases, where the CO_2 intercooling heat exceeds the capacity of the water to be heated, the remaining energy is integrated into the mixing heat exchanger after checking for any temperature limitation. This criterion is assumed to compare under the same assumptions (maximum heat recovery from CO_2 intercooling) the different alternatives. For most of the cases, the first layout is analyzed when the required heat duty is <3.5 MJ/kg CO_2. For heat requirements of >3.5 MJ/kgCO_2 and high temperatures, both the pressure of the bleeding and the water through the low-pressure heaters limit the temperature difference in heat exchangers and the second steam bleeding also must be closed.

The CO_2 compression scheme is simulated after that described in ref 46. Four CO_2 compressor stages with intermediate intercooling and a pressure ratio of 3.23 are used to increase the pressure up to 120 bar. Each intercooler is divided into two heat exchangers. The first one reduces the CO_2 temperature from the compressor outlet, in the range of 143−160 °C to 60 °C. This is a useful heat stream that is integrated in the low-pressure part of the steam cycle, as shown in Figure 2. The second heat exchanger stage reduces the CO_2 temperature from 60 °C to 30 °C to diminish the CO_2 compressor power requirements. This heat stream is discarded and is released to the ambient environment. The isentropic efficiency of each CO_2 compressor is assumed to be 0.8.
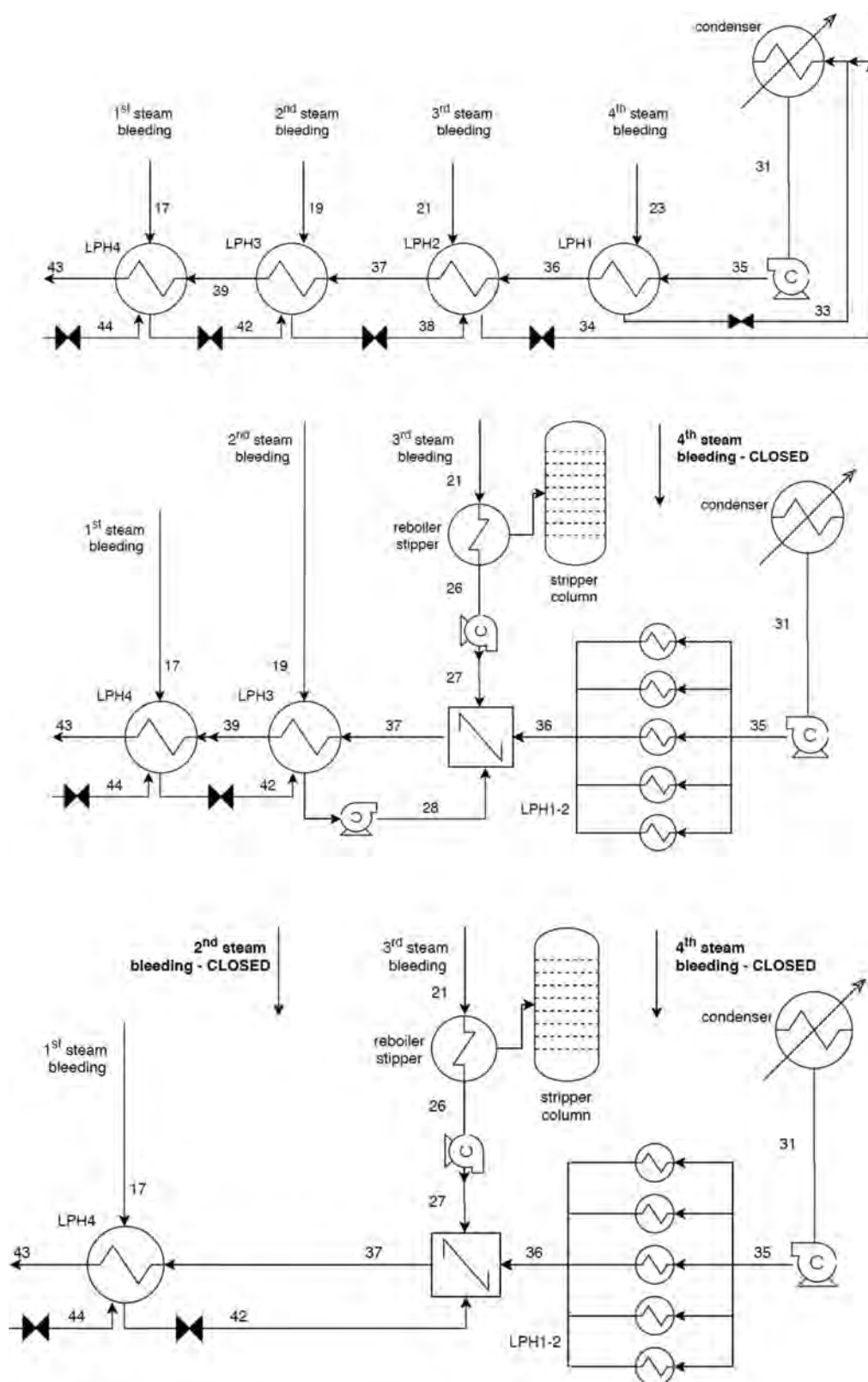
Figure 2. Original low-pressure heaters layout and modified layout with the carbon capture installation.

4. RESULTS AND DISCUSSION

4.1. Effect of the Regeneration Temperature and Stripper Heat Duty. First, a sensitivity analysis of the impact of regeneration temperature and heat duty shows how both variables influence the net global efficiency of the CCS power plant. The regeneration temperature is varied from 90 °C to 160 °C and the heat duty is varied from 1.5 to 5.5 GJ/ton CO_2. Figure 3 illustrates the influence of the regeneration temperature at different heat duties, and Figure 4 shows the influence of the heat duty at different regeneration temperatures.
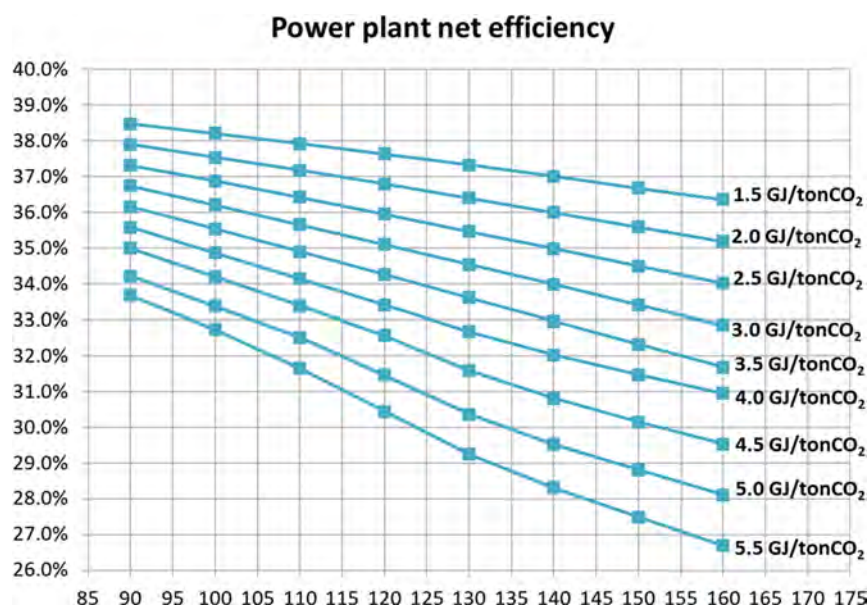
Power plant net efficiency

Figure 3. Influence of the regeneration temperature for different heat duties in the power plant net efficiency.
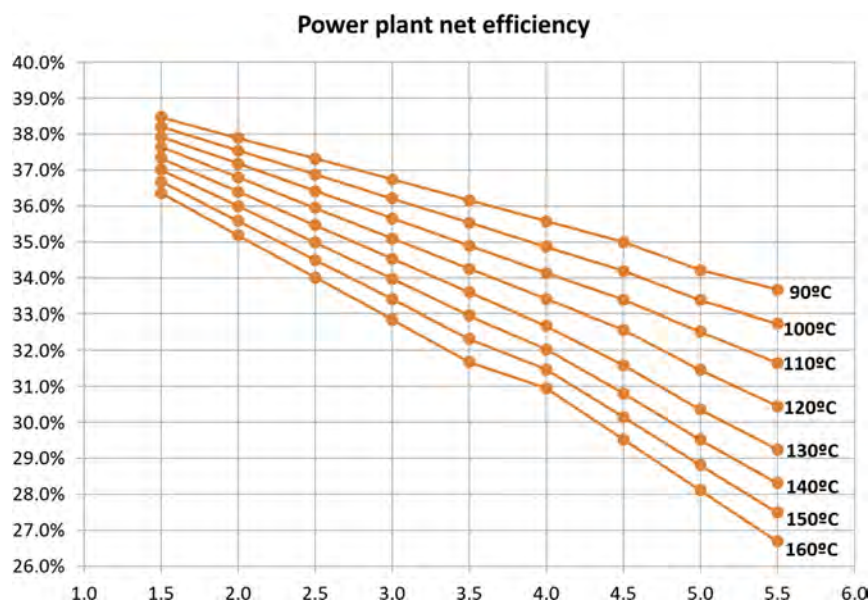
Power plant net efficiency

Figure 4. Influence of the heat duty for different regeneration temperatures in the power plant net efficiency.

With high regeneration temperatures, the steam bleeding pressure increases and, consequently, reduces the steam turbine power and net efficiency. It may also alter the design of the low-pressure heaters. The condensed steam from stripper recycled to the steam cycle has high temperatures and influences the layout of the low-pressure heaters.

For illustration purposes, when the heat duty and regeneration temperature are 3.0 MJ/kg CO_2 and 130 °C, the heat required in the stripper ascends to 287.8 MW$_{th}$, which is fulfilled with 123.16 kg/s of steam at 2.7 bar. Steam bleedings (first, second, and third) are completely open. If the heat duty increases up to 5.5 MJ/kg CO_2 and 150 °C, the heat required in the stripper increases to 479.7 MW$_{th}$, which is fulfilled with 203.68 kg/s of steam at 4.8 bar. For this reason, the second steam bleeding is not necessary and the contribution of the first steam bleeding is almost negligible. In this case, the temperature

of stream 27 is high enough to make the heating of the condensate before deaerator unnecessary.

For a heat duty of 2.0 GJ/ton CO_2, the net efficiency range is 35.19%−37.90%, and the values for regeneration temperatures are 160−190 °C. For a heat duty of 5.0 GJ/ton CO_2, the net efficiency range is slightly higher between 28.11% and 34.22%. The difference in efficiency penalty is close to 10 points, ranging from 7 points to ~17 points, when the heat duty to the stripper changes from 1.5 to 5.5 GJ/ton CO_2 for high regeneration temperatures (160 °C). Moreover, the value is ~5 points for regeneration temperatures of 90 °C. Taking into consideration stripper temperatures, the efficiency range for 90 °C is 33.68% and 38.48%, and for 160 °C is 26.69%−36.36%. The difference in efficiency is 9.7 points for the highest heat duty (5.5 GJ/ CO_2) and 4.8 points for the lowest heat duty of 1.5 GJ/ton CO_2. It is clear that the heat duty has more influence on the efficiency penalty in the power plants than the regeneration temperature,
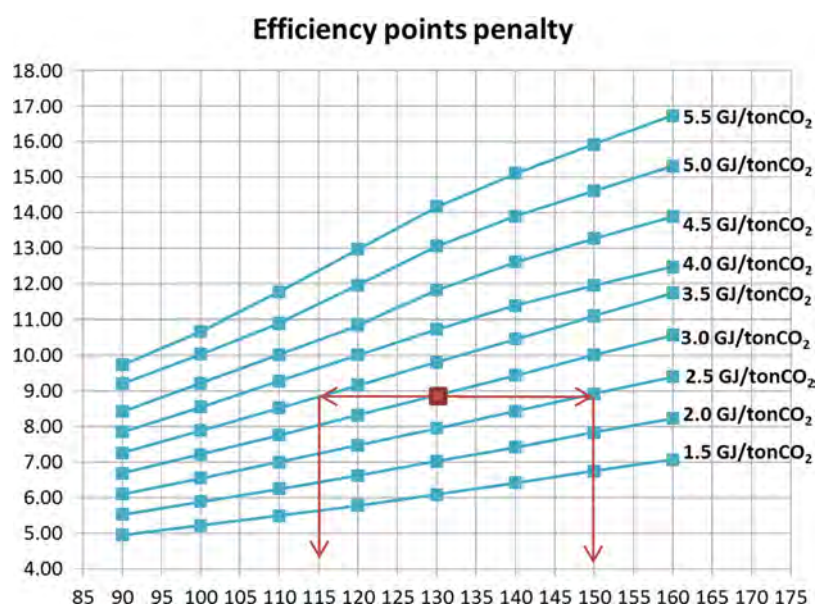
Figure 5. Net power plant efficiency, as a function of the regeneration temperature and heat duty in the amine regeneration.

Table 6. Solvents Selected To Perform the Energetic Comparison

	temperature (°C)	heat duty (MJ/kg CO_2)	details	data source
			MEA	
MEA1	116.3	3.9	30 wt %, 0.25 lean loading, 0,481 rich loading, 1.5 bar	ref 10
MEA2	130.0	5.4	0.22 rich loading, 0.5 rich loading (4 kmol/m³), 1 atm	ref 30
MEA3	130.0	3.8	0.32 lean loading, 0.5 rich loading (5 kmol/m³), 1 atm	ref 30
MEA4	120.0	4.3	30 wt %, 0.16 lean loading, 0.42 rich loading, 2 bar	ref 31
MEA5	114.0	3.1	0.30 lean loading, 0.477 rich loading, 7 mol/kg MEA (30 wt %)	ref 34
			DEA	
DEA	110.9	3.4	40 wt %, 150 kPa	ref 27
			MDEA	
MDEA1	130.0	2.8	0.04 lean loading, 0.5 rich loading (4 kmol/m³ solvent), 1 atm	ref 30
			DGA	
DGA1	96.5	4.1	60 wt %, 75 kPa	ref 37
DGA2	120.7	2.8	60 wt %, 250 kPa	ref 27
			AMP	
AMP1	95.2	3.8	15/15 wt % DEA/AMP 0.44 lean loading, 0.568 rich loading	ref 17
AMP2	112.0	4.9	5/25 wt % PZ/AMP 0.14 lean loading, 0.28 rich loading	ref 38
			DETA	
DETA	100.0	4.8	2 kmol DETA/m³, 0,4 cyclic loading	ref 39

but this last influence is also remarkable. The effect on the net efficiency of increasing the heat duty by 1 GJ/ton CO_2 is similar to the reduction between 40 °C (for 1.5 GJ/ton CO_2) and 20 °C (for 4.5 GJ/ton CO_2) of the regeneration temperature.

As Figure 5 shows, for a heat duty of 3.0 GJ/ton CO_2 at a regeneration temperature of 130 °C, the penalty into the power plant is 8.9 efficiency points. A similar penalty is also caused by a solvent with a heat duty of 3.5 GJ/ton CO_2 and a regeneration temperature of ~115 °C, and with a solvent of a heat duty of 2.5 GJ/ton CO_2 and a regeneration temperature of >150 °C. In this last case, even with lower heat duty, if the regeneration temperature is high, the integration possibilities decrease, and some of the low-grade heat must be discarded to condensers in some cases.

The effect of high regeneration temperatures is more critical when combined with high heat duties. When regeneration requirements are high, the steam bleeding to stripper increases, and a small fraction of the steam drives the LP turbine and goes

into the condenser. Since the condenser stream is small, the most of the CO_2 intercooler heat must be used together with the mixing heat exchanger as the stream temperature of 36 °C is limited by the CO_2 maximum temperature after compressors. This causes that the stream temperature of 37 °C would be high enough to make the steam bleeding (19) and even (17) useless (see Figure 1), because of the high condensate temperature. This effect is observed with a heat duty of >4.0 GJ/ton CO_2 when the temperature is >150 °C. When the heat duty increases to >5.0 GJ/ton CO_2, the steam bleeding is closed for the all temperature ranges.

When the regeneration temperature is high, the temperature of stream 37 should be limited to maintain a suitable temperature difference in CO_2 intercoolers and mixing heat exchangers. Then, part of the intercooling heat must be thrown away. In these cases, efficiency decreases because of the discarded energy. It happens for a heat duty of 5.5 GJ/ton CO_2 and temperatures of >120 °C, for a heat duty of 5.0 GJ/ton
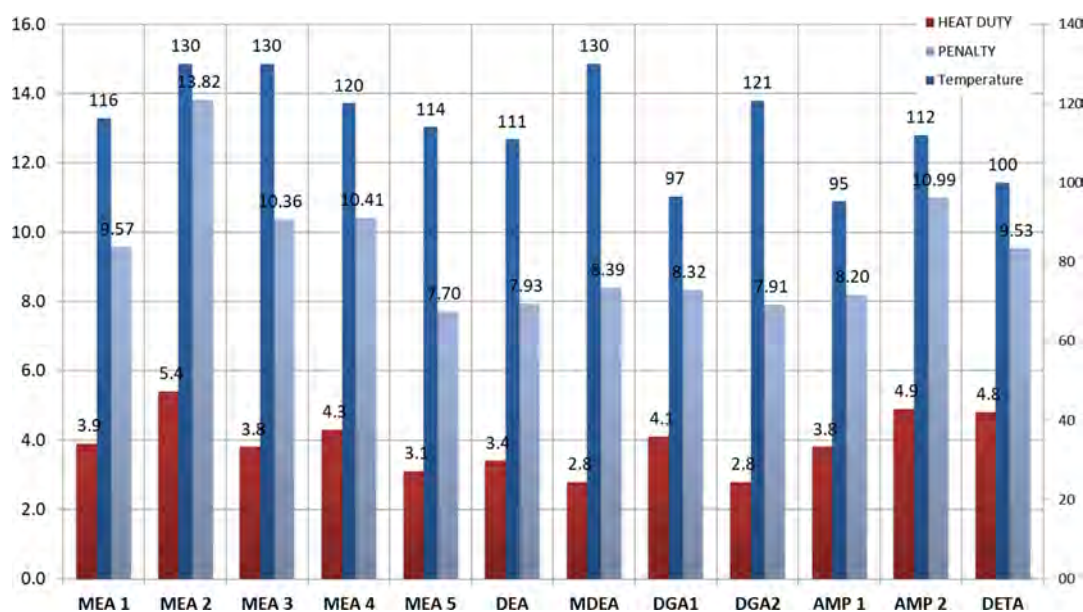
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Figure 6. Net power plant efficiency as a function of the regeneration temperature and heat duty in the amine regeneration.

CO_2 and temperatures of >130 °C, and for heat duties of 4.0–4.5 GJ/ton CO_2 and temperatures of >140 °C. It is illustrated in Figure 5 as the slope of the high heat duties increases more than in the case of low heat duties.

4.2. Energy Comparison of Different Solvents for CO_2 Capture. Table 6 illustrates the comparison of the selected solvents shown in bold in Tables 2 and 3. There are five solvents based on MEA: one on DEA and MDEA, two on DGA, two on AMP, and one on DETA. Different regeneration energy requirements and temperatures have been selected. Ammonia has not been taken into consideration; it requires different energy integration and additional assumptions should be made to perform the comparison.

Simulations have been performed by using the following parameters as inputs: the stripper heat duty, regeneration temperature, amine concentration in the dissolution, and lean and rich amine loading. Results include the impact on the steam cycle streams and, as a consequence, the calculation of the power plant net efficiency and efficiency point losses. The general trend of the results is in agreement with the previous figures. In this case, the quantification of the impact in efficiency is the most important outcome.

For MEA, regeneration temperatures are between 110 °C and 130 °C and heat duties range from 3.1 GJ/ton CO_2 to 5.4 GJ/ton CO_2. Results are in agreement with some references.[47−50] In this solvent, as Figure 6 illustrates, the effect on efficiency loss varies between 7.7 and 13.82 points. Differences between MEA1 and MEA3 are mainly caused by temperature and the effect on efficiency is relatively small (10.36 points for 130 °C and 9.57 points for 116 °C). However, for MEA2 and MEA3, the deviation is caused by the difference in heat duty and a gap of 3.46 efficiency points is observed.

In the case of DEA, the results are significantly good with low efficiency penalty, similar to MEA with the lowest heat duty and temperature. The cases of MDEA and DGA also show good results with efficiency losses of ∼8 points. In these cases, the heat duty is low, compared to that for MEA (2.8 GJ/ton CO_2), and, in the case of comparable requirements, the regeneration temperature (96.5 °C) is much lower than the 130 °C observed for MEA.

The cases selected for AMP that are blended with DEA and PZ in different proportions[1738] do not have evidence of improved results, even if a low stripper temperature is necessary for AMP1. Finally, the case of DETA is in agreement with the general trend and shows better results than AMP2 (similar heat duty) caused by a lower temperature required.

5. CONCLUSIONS

There are many possible solvents for CO_2 capture based on amine scrubbing. A common feature is the high energy requirements in the regeneration step. Two main variables define the energy necessities: the heat duty and the regeneration temperature. A literature review shows different values for these variables, and it is difficult to know and isolate the independent effect of each variable on the net effect of energy requirements of the carbon capture.

A comparison under the same set of assumptions of the effect in net power plant efficiency of heat duty and regenerations temperatures in the stripper of amine scrubbing for CO_2 capture has been presented. Results show that efficiency points penalty ranges between 7 for a heat duty of 1.5 GJ/ton CO_2 and regeneration temperature of 90 °C to ∼17 points for a heat duty of 5.5 GJ/ton CO_2 and regeneration temperature of 160 °C. The effect on efficiency is more important for higher heat duties, and steam cycle modification should be done to use as much waste energy from CO_2 intercooling and steam turbine bleeding as possible. The effect of a net efficiency increase of 1 GJ/ton CO_2 on the heat duty is similar to the reduction in the regeneration temperature between 40 °C (for 1.5 GJ/ton CO_2) and 20 °C (for 4.5 GJ/ton CO_2).

For the amines, the impact in efficiency points varies between 7.70 for MEA5 and 13.82 for MEA2. This is mainly caused by the difference in heat duty (3.1 GJ/ton CO_2 vs 5.4 GJ/ton CO_2), as well as regeneration temperature (114 °C vs 130 °C). There are several options (MEA5, DEA, MDEA, DGA1, DGA2, AMD1) that show that achieving an efficiency reduction of ∼8 points is possible. It is important to highlight that these conclusions only show the results of an energy analysis. It is evident that, in the utilization of these amines for carbon

capture, there are other limitations that can affect a complete techno-economic feasibility study.

■ ASSOCIATED CONTENT

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Notes

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