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Energy optimizing of CO_2 capture process from emissions of gas plant refinery boiler flue gasses based on MEA chemical reaction

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Carbon dioxide is the most important greenhouse gas in the atmosphere. In fact, carbon dioxide exists naturally and constitutes about 4% of atmosphere and plants use it in the process of photosynthesis in order to produce oxygen. Almost all additional carbon dioxide emissions in atmosphere are produced by fossil fuels combustion needed for energy and transportation.There are several ways to capture carbon dioxide from flue gasses; one of these methods is an amine solvent absorption method-MEA as a first solvent- in post combustion method. In this study, energy optimizing of carbon dioxide capture process from flue gases emitted from boiler of gas refinery in Iran is studied. Firstly, the process is modeled and described and mass and energy balance of each process will present. The reaction in the process of absorption in order of appearance are water ionization, H2S ionization, Hydrolysis and ionization of unsolved carbon dioxide and Proton formation of alkanolamines and carbomate other adverse reactions may also produce unwanted products and waste the material and energy. Process energy sources include blower and reboiler; it can be observed that after providing consumed energy, reboiler is the main source of process energy and energy optimization must focus on it. Finally, three parameters including solvent flow rate, tower height and pressure will be examined in order to optimize the energy consumption of reboiler.

Keywords: Energy Optimizing, Gas Plant, Boiler Flue, MEA Chemical

INTRODUCTION

The primary greenhouse gases in Earth's atmosphere are water vapor, carbon dioxide, methane, nitrous oxide, and ozone. Without greenhouse gases, the average temperature of Earth's surface would be about −18 °C (0 °F)[1] , carbon dioxide in Earth's atmosphere is a trace gas, currently having an average concentration of 402 parts per million by volume[2,3,4]

Concentrations of carbon dioxide are generally higher in urban area [5], combustion of fossil fuels and deforestation have caused the atmospheric concentration of carbon dioxide to increase by about 43% since the beginning of the age of industrialization[6].

Although both methane and carbon dioxide pose significant threats [7], but Worldwide commitments to meet carbon dioxide emissions reduction targets are challenging the process industries to reduce carbon dioxide emissions, Kyoto protocol for example [8].

Carbon dioxide is used in many industries such as metal industry to enhance their hardness [9], food industry, in carbonated drinks, brewing and flash drying method. Also other Industrial use of carbon dioxide includes enhanced oil recovery, welding, chemical raw materials, inert gas, firefighting and it is used as supercritical fluid in extraction with solvent [10]. Carbon dioxide is also a critical component in medical oxygen, since it is useful at low concentrations as a respiratory stimulant; Also carbon dioxide is produced by refinery flare in heavy duty and very important for external cost calculation [11].

The largest potential market for carbon dioxide is enhanced oil recovery operations.

The main sources of carbon dioxide to be used in EOR operations include:

- Natural resources: carbon dioxide wells
- Industrial byproduct: sweetening natural gas, synthesis gas production
- Chimney emissions: fossil fuel power plants, industrial furnaces, cement plants

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- emissions of engines, and emissions from lime kilns[12]

Conventional fossil fuels continue to be the dominant sources of primary energy in the process industries [13], the most economic sources of carbon dioxide include carbon dioxide wells, byproduct of sweetening natural gas and purifying synthesis gas.

also gas flaring is one of the most GHG emitting sources in the oil and gas industries[14] but any source of combustion-For example Boiler-Could be considered as GHG source.

The price of crude oil in 1999 (12 to 16 dollars per barrel) cannot provide needed economic justification in order to recover carbon dioxide from flue gases to be used in enhanced oil recovery operations, while, the current 50$ price per barrel provides economic justification for this method.

Flue gases from the past are important sources of carbon dioxide supply, especially for countries that don't access to carbon dioxide as byproduct of different processes. In its simplest form, the fuel is burned in order to produce flue gas then, carbon dioxide is separated from flue gas using a dilute solution of mono ethanol amine, while required heat is provided by heat of combustion. This process is widely used and is associated with high energy consumption. Smaller and mounted units which benefit from "Econamine FG" method can be used where there is source of flue gas and it is desirable less energy to be used. Currently, eleven of these units can be used in order to recover the carbon dioxide for food and drink [15].

DESCRIPTION OF PROCESS

Figure 1 shows a scheme of whole process. The general shape of process is similar to sweetening processes. The exception is that at low pressure gas enters into absorbing tower.

Among the different technologies, post-combustion chemical absorption in amine-based solvents is the most mature at this stage of development [16]. Monoethanolamine (MEA) has been used widely as solvent for carbon dioxide capture due to its high reactivity and low price for production [17].

A commonly applied carbon dioxide capture technology is based on absorption by amine solutions [18]; using Amines leading to sustainability and safety issues [19]. The process uses MEA with concentration of 30% and inhibitor for carbon dioxide absorption. Using inhibitors will make possible using carbon steel for most process equipment. Also, experience has shown that loss of solvent is reduced under these conditions.

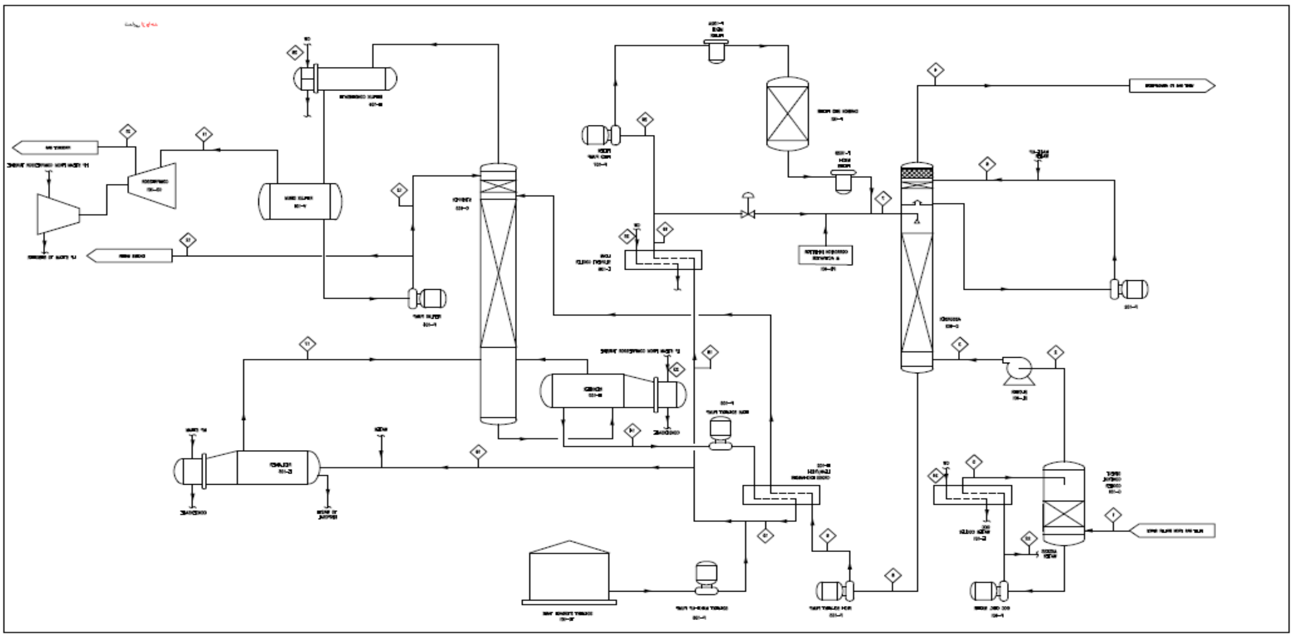

Fig.1. Overview of process.

Firstly, the flue gas (flow 1) enters direct contact cooler and is cooled there using water (flow 5) to a temperature of 50°C (dew point of water at atmospheric pressure) (flow 2), then, gas pressure in increased using a blower in order to overcome the pressure drop in absorption tower (flow 3).

The flue gas is filled in an absorption tower with a diluted solvent (flow 7) in pressure conditions a little more than atmospheric pressure and a temperature of 45°C and is contacted. Heated gas by reaction heat enters into washout section at the top of tower and solvent steam is washed and recovered by down flow of water and is added to circulating solvent and cooled gas is discharged down to atmospheric (flow 4). Solvent absorbs 80-95% of carbon dioxide through a chemical reaction (flow 8).

The emissions of tower bottom that include rich solvent (flow 8), flow towards recovery tower with higher temperature (the approach temperature of 11 ^{oC}) after pressure increase (flow 9) and passing through a heat exchanger (flow 10) (that transfer heat between rich and dilute solvent) which increases recovery efficiency. In recovery tower reboiler, barg transfers water with a solvent to steam phase though heating by fuel or transferring heat with low steam pressure (here is used, 40 to 60 psig and about 3 to 4). Part of steam flow is cooled and liquidated due to increase in solvent temperature (apparent heat) and providing needed heat for recovery (latent heat) another part plays the carrier liquid for separation of carbon dioxide from solvent (flow 13) in recovery operation.

If we have amines with normal concentrations, reboiler temperature in process must match boiling point of water that is determined based on pressure. Higher temperatures of tower shell destroy the amine. In this case, the pressure of recovery tower is limited to about 1.5 barg (20 psig) so that shell temperature does not exceed 280 degrees F (138 ºC). Since there is not available complete information on shell temperature, process temperature is usually limited to 260 ° F and steam temperature is limited to 285 °F.

Steams over recovery tower are cooled to 45 °C (122 °F) so that more steam stripper becomes liquid and more or all of them are returned to tower (flow 13). Some of water returned to tower (reflux) is fractioned and discharged in order to balance the amount of water or prevent the formation of compounds such as ammonia and surfactants which can have an important role in destruction of amines. Compounds such as ammonia amount don't excess one percent by weight with the aim of keeping most of amine solvent, since this compound leads to re-absorption of carbon dioxide, corrosion in reflux return flow and need for additional power to recovery. Surfactants increase foaming and often lead to tower instability against flooding. Recovery tower steam with a purity of 98.2% (flow 11) enters the compressor and is compressed up to 100 bar pressure (flow 21). We need refrigerator with a temperature of -15°C for more carbon dioxide purity of 99.9% that is not economically justified, but according to existing research and units, using drying unit after compressor by glycol is recommended.

Return flow enters into tower usually two to four trays above the feed entrance point in order to prevent the rise of Amine with steam flow and prevent the decay of Amine discharge part of return flow. Return flow doe not usually effect on heat balance in fractional distillation towers except hydrocarbons, and can easily change the rate of return flow with no negative effects on stability of tower.

Diluted solution (flow 14) is cooled after heat exchange with rich solvent in heat exchanger (flow 15). Diluted solvent is pumped and approximately 3% flow is entered into reclaimer (flow 16) in order to separate salts formed from amines and steamed amine in reclaimer reenters the recovery tower (flow 17). The fluid is then re- cooled in order to reach optimal temperature for absorption tower that is 45 ºC (flow 19). Part of diluted solution is separated (flow 20) in order to separate particles suspended in flue gas or corrosion products through filtration. Filtration of particles is done usually through a fixed bed of granular activated carbon in order to separate the amine degradation products that are active at surface and lead to foaming. Cooling liquid of this project is cooling water.

Pilot-scale experiments and modeling usually suggest that doing recovery operation in vacuum pressure leads to a reduction of 5 to 10% in energy consumption per unit of separated carbon dioxide and leads to more compact[20, 12].

Table 1 provides reboiler information that is provided in order to recycle carbon dioxide from emissions of flues:

Normally from six reboiler of each gas refiner, 3 of them are in the service. It must be mentioned

that 7 items are examined in design process.

1. Reboilers A + E
2. Reboilers C + E
3. Reboilers A + C
4. Reboilers A + C + E
5. Reboiler A
6. Reboiler C
7. Reboiler E

Dimensions of equipment and high pressure of steam were significant for case 1 to 4 items (amount of required steam more than 200 t/h) after simulating mentioned items. As a result, assuming that two average units are cost-effective than a large unit, we studied unit for reboiler A as design case. It is worth noting that according to Meriz study [12, 15] units have been studied to 1000 t/d for large unit.

MASS AND ENERGY BALANCES

Finally, mass and energy balance and composition of materials percent were identified after simulation. Table 2 shows the composition of flows percent and Table 3 shows the characteristics of flow. Finally, all simulated equipment can be seen in Table 4.

PROCESS OPTIMIZATION

A comparison of the effectiveness of different policies shows that the energy efficiency is very important in carbon dioxide emissions[21],The thermal energy required for regeneration of carbon dioxide -rich adsorbents or absorbents is usually regarded as the most important criterion to evaluate different materials and processes for application in commercial-scale carbon dioxide capture systems [22], The process will require energy in two sections; one case is in stripper reboiler (recovery) and the other in blower that directs flue gas into absorption tower. The energy required for blower is not vary variable and is set by flue gas flow rate and pressure drop of absorption tower. It must be noted that required reboiler energy is strongly influenced by process prevailing circumstances. Optimal performance point of process is related to the time when we need at least a certain amount of energy in order to separate carbon dioxide for reboiler [23].

Process variables that effect on heat load and energy required for reboiler and are examined here include solvent flow rate, column height and recovery column pressure.

Different operating conditions are compared with each other at a constant rate of separation from carbon dioxide. The separation rate is among design variables and is not considered as a variable for optimization.

Reboiler is placed in focal point of discussion for minimization of energy consumption in process, to do so we need to identify the effective factors on Reboiler energy. Equations 5.1 and 5.2 represent the Reboiler energy components that have been obtained using a simple energy balance on recovery tower.

$$Q_r = Q_{des,CO2} + Q_{steam\ generation} + Q_{sensible\ heat} \tag{5.1}$$

heat provided by consumed steam in Reboiler is used for CO_2 + MEA reaction in reverse direction, steam generation in Reboiler (Qsteam generation), and heating the solvent from inlet temperature to temperature of Reboiler (Qsensible heat).

$$\begin{aligned} Q_r = -n_{CO2}\Delta H_{abs,\ CO2} + (V - n_{H2O})\Delta H_{vap,\ H2O} + \\ + Lc_p\,(T_{bottom} - T_{top}) \end{aligned} \tag{5.2}$$

Qr shows reboiler energy, ΔH_{abs} carbon dioxide shows carbon dioxide absorption heat, $\Delta Hvap\ H_2O$ shows heat for water steam. Also, NCO_2 displays the number of carbon dioxide moles that are separated in recovery tower, nH_2O shows the number of water moles that are cooled in column, V shows molar flow rate of steam in reboiler, L shows liquid molar flow rate in reboiler, it is assumed that is equal to molar flow rate of solvent feed, cp shows specific heat of liquid and Tbottom - Ttop shows the temperature difference between top and bottom of recovery tower. Equation 5.2 is based on the fact that water is cooled in cooling tower. In this case, reboiler energy will be divided by the number of carbon dioxide moles that are recovered and then will be reported (equation 5.3).

Table 1. Reboiler information that is provided in order to recycle carbon dioxide from emissions of flues

Stack Tag	Composition									Exit Gas Temperature	Exit Gas Velocity	Flow Rate
	O_2	CO	CO_2	NO	NO_2	NO_X	SO_2	H_2S	HC			
	%	ppm	%	ppm	ppm	ppm	ppm	ppm	ppm	(C)	(m/s)	m^3/hr
spy-02-121-U-103(A)	4.18	0.00	9.53	96.30	3.00	93.00	1.00	0.00	0.00	142.3	18.5	423452.7
spy-02-121-U-103 (C)	4.82	0.00	9.18	122.40	6.30	193.40	2.90	0.00	0.00	132	19	426700.71
spy-02-121-U-103(E)	5.02	0.00	9.60	73.00	5.30	161.40	3.00	0.00	0.00	127	18.6	416422.87

Table2. The composition of flows percent in simulation

Stream Mole Fraction (%)	O_2	CO_2	NO	NO_2	SO_2	MEA	H_2O
1	4.18	9.53	0.00963	0.0096	0	0	86.2707
2	26.7176	60.898	0.0615521	0.0613507	0.000634	0	12.2609
3	26.7176	60.898	0.0615521	0.0613507	0.000634	0	12.2609
4	90.4892	0.0733441	0.208446	0.207316	0.0011122	0.0003408	9.02027
5	0	0.0030928	0	0	0	0	99.9968
6	0.0017052	0.028302	0	0	0	0.194226	99.7757
7	0	1.39577	0	0	0	11.2093	87.3949
8	0.0003464	4.9744	0	0	0	10.7355	84.2897
9	0.0003464	4.9744	0	0	0	10.7355	84.2897
10	0.0003464	4.9744	0	0	0	10.7355	84.2897
11	0.0092264	95.5191	0	0.0002306	0.0001127	0	4.47126
12	0	0.0825883	0	0	0	0.0044799	99.9129
13	0	0.0825883	0	0	0	0.0044799	99.9129
14	0	1.44687	0	0	0	11.1926	87.3606
15	0	1.44687	0	0	0	11.1926	87.3606
16	0	1.44687	0	0	0	11.1926	87.3606
17	0	1.44666	0	0	0	11.1926	87.3607
18	0	1.44687	0	0	0	11.1926	87.3606
19	0	1.44687	0	0	0	11.1926	87.3606
20	0	1.44687	0	0	0	11.1926	87.3606
21	0.0096042	99.4441	0	0.0002401	0.0001165	0	0.545907
22	0	0.0030931	0	0	0	0	99.9968
23	0	0	0	0	0	0	100
24	0	0	0	0	0	0	100
25	0	0	0	0	0	0	100
26	0	0	0	0	0	0	100

Table 3. Characteristics of streams in simulator

Property	Temperature	Pressure	Mass Density	Mass Flow	Vapor Volumetric Flow	Mass Enthalpy	Ideal Gas Cp/Cv Ratio	Dynamic Viscosity	Thermal Conductivity	Surface Tension
Units	°C	bar	kg/m^3	kg/h	m^3/h	kJ/kg	-	cP	W/(m*°C)	dyne/cm
Stream Number										
1	127	1	0.637358	269891	423453	-11510.5	1.31187	0.0154305	0.027145	-
2	49.085	0.97	1.36612	75323.1	55136.4	-7137.18	1.30778	0.0173358	0.0215086	-
3	64.8087	1.13	1.51738	75323.1	49640.4	-7122.38	1.30278	0.01807	0.0226866	-
4	43.6681	1	1.16974	18193.9	15553.8	-692.367	1.38726	0.0209383	0.0269405	-
5	45	1.03	989.486	2.22E+06	2239.53	-15781	1.3238	0.587057	0.631773	68.862
6	43.6867	1.03	990.505	941540	950.566	-15707.2	1.32057	0.610186	0.625964	69.012
7	45.037	1.03	1038.2	745000	717.59	-12231.6	1.20189	1.45024	0.459597	63.2492
8	80.7757	1.13	1094.68	802196	732.816	-12013.8	1.19762	0.80007	0.435728	55.1067
9	80.7906	2.5	1094.69	802196	732.808	-12013.6	1.19762	0.800085	0.435734	55.1041
10	112	2.4	117.079	802196	6851.75	-11865.6	1.19209	-	-	-
11	45	2.18333	3.57274	53923.1	15092.9	-9009.57	1.2815	0.0160178	0.0184522	-
12	45	2.18333	989.896	1989.35	2.00966	-15766.7	1.32368	0.585474	0.629833	68.8053
13	45.0329	5.15769	989.924	44929.3	45.3866	-15766.3	1.32368	0.585479	0.629871	68.7989
14	128.898	2.38333	985.914	769364	780.357	-11908.1	1.18712	0.348223	0.481586	48.4941
15	128.908	3	985.919	769364	780.352	-11908	1.18711	0.348239	0.481589	48.4905
16	90.1063	2.9	1014.74	23080.9	22.7457	-12062.3	1.1937	0.582943	0.47837	55.5018
17	158.607	2.62857	1.73059	23080.8	13336.9	-10055.9	1.18236	0.0145548	0.0310161	-
18	90.1063	2.9	1014.74	746283	735.445	-12062.3	1.1937	0.582943	0.47837	55.5018
19	45	2.4	1039.37	746283	718.012	-12232.7	1.202	1.45519	0.459068	63.2265
20	45	2.4	1039.37	74628.3	71.8012	-12232.7	1.202	1.45519	0.459068	63.2265
21	222.162	105.364	120.918	53019.7	438.477	-8805.55	1.2302	0.0258943	0.0375597	-
22	v	1.03	989.478	194568	196.637	-15780.9	1.3238	0.586832	0.631796	68.858
23	147.986	4.5	2.38235	90823.8	38123.7	-13204.4	1.31566	0.0115658	0.0290625	-
24	35	5.2	993.32	1.20E+07	12065.7	-15823	1.32465	0.71605	0.619709	70.8294
25	35	5.5	993.324	3.04E+06	3061.01	-15823	1.32465	0.716093	0.619709	70.8294
26	35	5.2	993.32	2.90E+06	2923.75	-15823	1.32465	0.71605	0.619709	70.8294

Table 4. Lists of equipment in simulation

Description	Tag No.	type	Qty	Description
Reflux Drum	V-101	Horizontal Vessel	1	ID=1.8(m);L=5.6(m)
Blower	BL-101	Centrifugal	1	Duty=310(Kw);Flow Rate=55136(m3/h)
Compressor	Co-101	Centrifugal	1	Duty=7300(Kw);Flow Rate=15100(m3/h)
Direct Contact Cooler Water cooler	E-101	Plate	1	Duty=145(Mw); Surface=1220m2
Lean/Rich Cross Exchanger	E-102	Plate	1	Duty=33(Mw);Surface=4700m2
Reflux Condenser	E-104	AES	1	Duty=34(Mw);ID=1.2(m);L=3(m);Surface=578m2
Reboiler	E-103	AKU	3	Duty=54(Mw);ID=1.3(m);L=4.3(m);Surface=3x973m2
Reclaimer	E-105	AKU	1	Duty=13(Mw);ID=0.65(m);L=4(m);Surface=1x74m2
Lean Solvent Cooler	E-106	Plate	1	Duty=35(Mw);Surface=3100m2
Direct Contact Cooler	C-101	Column	1	ID=6.5(m);L=20(m);Design Pressure=3.5Barg
Absorber	C-102	Column	1	ID=5(m);L=18(m);Design Pressure=3.5Barg
Stripper	C-103	Column	1	ID=4(m);L=15(m);Design Pressure=3.5Barg
Direct Contac Cooler Pump	P-101	Centrifugal Pump	2	Duty=185.2(Kw);Flow Rate=2500(m3/h)
Water Cooler Pump	P-102	Centrifugal Pump	2	Duty=70.4(Kw);Flow Rate=950(m3/h)
Rich Solvent Pump	P-103	Centrifugal Pump	2	Duty=41(Kw);Flow Rate=805(m3/h)
Reflux Pump	P-104	Centrifugal Pump	2	Duty=5(Kw);Flow Rate=45(m3/h)
Lean Solvent Pump	P-105	Centrifugal Pump	2	Duty=19.3(Kw);Flow Rate=780(m3/h)
Solvent make-up Pump	P-106	Centrifugal Pump	2	Duty=5.9(Kw);Flow Rate=80(m3/h)
Filter Feed Pump	P-107	Centrifugal Pump	2	Duty=4.6(Kw);Flow Rate=72(m3/h)
Corrosion Inhibitor & activator	PK-101	By Vendor	1	-
Carbon Bed filter	F-101	charcoal filter	2	Flow Rate=72(m3/h)
Mechanical Filter	F-102A	cartridge filter	2	Flow Rate=72(m3/h)
Mechanical Filter	F-102B	cartridge filter	2	Flow Rate=72(m3/h)
Solvent Storage tank	TK-101	Fixed roof	1	ID=10(m);L=8(m)

$$\frac{Qr}{nCO2} = -\Delta H_{abs,\ CO2} + \frac{(V - nH2O)}{nCO2} \Delta H_{vap,\ H2O} + \frac{Lcp}{nCO2} (T_{bottom} - T_{top}) \qquad (5.3)$$

An energy balance on Reboiler results in equation 5.4.

$$Q_r = V\ \Delta H_{vap,\ H2O} \qquad (5.4)$$

Combining equations 5.2 and 5.4 will lead into Equation 5.5 and provides us with the notion that in this tower, sensible heat and reverse heat of reaction CO_2 + MEA will be supplied through heat from cooling steam.

$$n_{H2O}\ \Delta H_{vap,\ H2O} = -n_{CO2} \Delta H_{abs,\ CO2} + + Lc_p\ (T_{bottom} - T_{top}) \qquad (5.5)$$

Equations 5.2 and 5.5 have been written by some negligible, since it is assumed that only water is evaporated in reboiler while we face the excretion of carbon dioxide.

In sensible heat section, bottom tower temperature is in fact equal to liquid flash temperature at end of tower. Up tower temperature is also dependent on the amount of heat that is transferred in cross heat exchanger. Here, the distance between temperature and balance of heat exchanger is cross constant and is considered 20^{o}F (11^{o}C). Low temperature of tower is 121 ^{o}C and high temperature of tower is 110 ^{oC} for a stripped tower (recovery) that acts in Bellingham pressure conditions (180 kPa in reboiler).

In this section, required energy of Reboiler on number of separated carbon dioxide moles is reported.

Optimizing solvent flow rate

Energy required for Reboiler is very sensitive and variable to changes of solvent flow rate. Absorption tower efficiency is reduced when solvent flow rates are very low. This means that capacity of solvent has been filled very fast; as a result absorption is reduced. The way that is recommended for stabilizing carbon dioxide absorption is to keep down the amount of carbon dioxide dissolved in diluted solvent. This means that a lot of energy must be used in order to produce steam in Reboiler in order to hold down loading of output diluted solution from absorption tower. If high flow rate of solvent is used and even if diluted solvent with high rate of loading is used, separation of carbon dioxide can also be achieved. This will reduce the needed energy for recovery tower reboiler. On the other hand, too high flow rates of solvent will lead to overcoming the sensitive heat; as a result, reboiler heat will rise again. Examining the results and plotting them in diagram, optimal flow rate can be achieved.

Table 5 shows optimized ratio of L/G_{mass} and Optimized loading of rich and diluted solvent.

Table 5.optimized ratio of L/G_{mass} and optimized loading of rich and diluted solvent

Absorber L/Gmass	Optimization Lean loading	optimization rich loading	Q reboiler
3.13	0.098	0.473	1.12
3.15	0.1	0.474	1.1
3.2	0.1	0.475	1.07
3.23	0.106	0.481	1.056
3.27	0.105	0.48	1.07
3.3	0.104	0.479	1.08
3.34	0.102	0.478	1.1
3.38	0.101	0.477	1.11
3.4	0.1	0.476	1.13
3.5	0.1	0.475	1.15
3.56	0.097	0.473	1.18
3.64	0.095	0.47	1.21
3.75	0.093	0.468	1.25

Curve slope of reboiler energy is different in Figure 2 at both sides of minimum point. Heat is increased on the left side of minimum point more than right side. At more flow rates of solvent, reboiler energy is spent to produce steam, while in high flow rates of solvent, sensible heat is very important. Since the separation of carbon dioxide is kept constant, part of the formula that is related to absorption heat, is considered constant against solvent flow rate changes.

Reboiler energy is divided by the number of separated carbon dioxide moles and also by common industrial rate with MJ / Kmol unit.

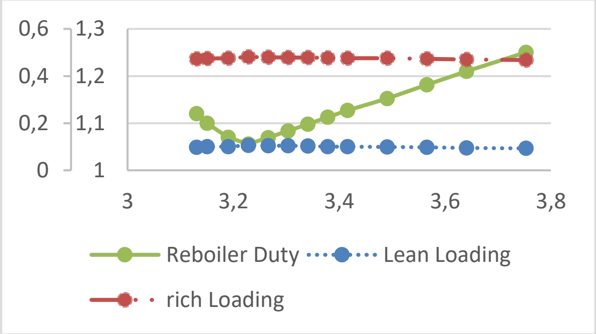

Fig 2. Solvent optimization flow rate

The effect of column height on optimization

Here, we study the effects of adsorption and recovery column height on process performance. Clearly we can say that the process performance becomes better through increasing the height of towers. Since taller columns have more space for mass transfer. The only point where there is uncertainty refers to wet part of total filled chamber (with packing).

Pinch is commonly used in designing recovery tower and therefore, more space will not have any effect on its performance. While, heat transfer rate of absorption tower is not so high and as a result, increased space will affect more on performance. Problems facing the recovery column height are related to increased costs and increased the pressure drop.

Absorption tower height effects, concentration and separation of carbon dioxide

The first case study is done about height of absorption tower. In this case, the diameter of tower can is considered constant.

All points of plotted curves are calculated at optimized solvent rate and tower height amounts have been normalized through dividing by Bellingham tower base height.

Figure 3 shows that minimum energy is obtained through increasing the tower height and height increase will continue until energy does not change and is plotted in diagram.

What looks interesting is that sufficient and plenty amount of filler (packing) the separation of carbon dioxide will not effect on reboiler energy.

Reboiler energy is separated on number of carbon dioxide moles and is divided into common industrial scale and the number of trays is used as a measure of tower height.

The effect of recovery tower height

In this case study, stabilizing the diameter of tower, tower height effects on process performance is evaluated.

As Figure 4 shows the minimum energy is obtained through increasing the tower height and the height will increase until the energy does not change and is depicted in diagram.

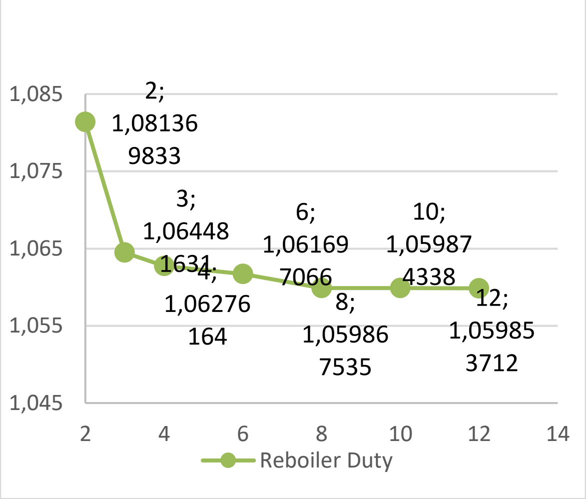

Fig. 3. The effect of absorption tower height on reboiler energy

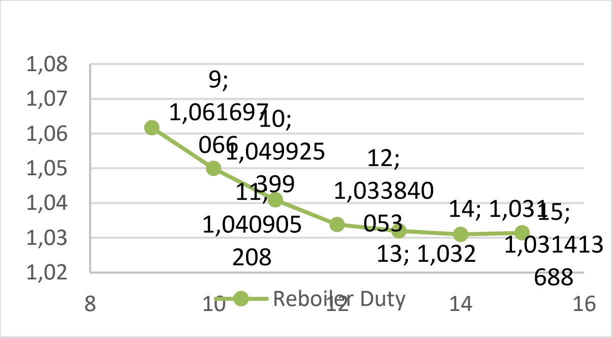

Fig.4. The effect of recovery tower height on reboiler energy

The effect of recovery tower pressure

Recovery tower pressure is kept above atmospheric pressure, since it reduces the amount of required heat in order to produce stripping steam. Reboiler required heat is provided through cooled steam. Required steam is supplied from power plant that is upstream carbon dioxide separation unit. This steam cannot be used for power generation. In fact, the required energy can be called lost work since this can produce work in the case of steam expansion in turbine.

With this approach, the difference between high or low pressure steams can be realized: high pressure steam in the case of expansion can create more energy and greater loss will follow compared to lower pressure steam.

As previously described, the reboiler energy is reduced by pressure, since the relative amounts of carbon dioxide and water with temperature in gas phase are increased. Writing Clausius–Clapeyron relation for both carbon dioxide and water and their mix will produce two Equations of 5.6 and 5.7.

$$\frac{d\ln P_{CO2}^{sat}}{d\ln P_{H2O}^{sat}} = \frac{-\Delta H_{abs,CO2}}{-\Delta H_{vap,H2O}} \approx 2 \qquad (5.6)$$

$$P_{CO2}^{sat} \ \alpha \ P_{H2O}^{sat\ 2} \qquad (5.7)$$

This means that carbon dioxide steam pressure increases by temperature with higher speed compared to water pressure. Higher pressure means a higher temperature and P_H$_2$O ^ sat / P_CO$_2$ ^ ratio is high. Increasing the total pressure, the less stripping steam will be needed to drive carbon dioxide gas. At lower pressure, lower levels of carbon dioxide are needed to saturate carbon dioxide and water mixture: as a result, more stripping steam will be required to perform the separation. Number 2 in Equation 5.6 shows that required energy is increased sharply at lower pressure.

When the pressure is low, the gas density is reduced; as a result, the gas superficial velocity is increased and the increase results in flooding. We change the tower diameter change speed in order to avoid these accidents and to stabilize gas speed at 80 percent. Column pressure drop is changed proportional to reboiler pressure.

Reboiler energy drops rapidly at low pressures, while is tends to asymptote at higher pressures. Asymptote displays the minimum amount of required energy while no stripping steam is produced and pure steam is full of carbon dioxide.

It seems that in conditions of using aqueous solvent of mono ethanol amine, reducing the recovery tower pressure does not seem very good. This case requires much more work in order to compress the carbon dioxide. Work at pressure higher than 2 atmospheres is not recommended due to the risk of mono ethanol amine polymeric decomposition.

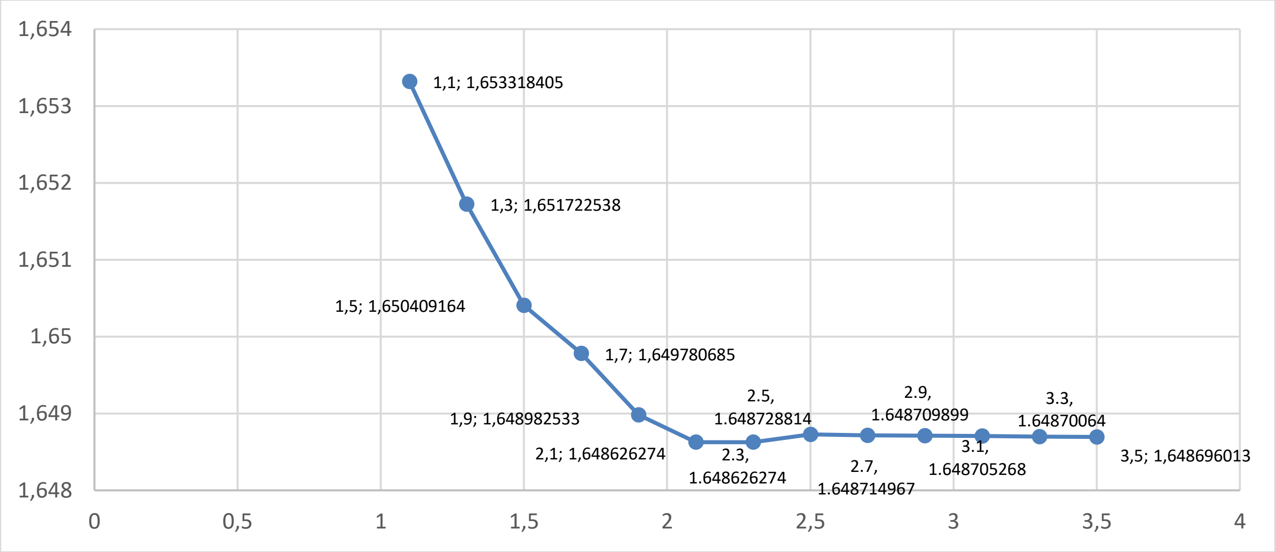

Fig 5. The effect of recovery tower pressure on reboiler energy.

The high temperature of carbon dioxide absorption is a case that does not make mono ethanol amine suitable for recovery tower with low pressure. Other solvents with lower absorption energy are more suitable in this case.

CONCLUSION

Optimizing the energy consumption of two processes was studied, one of them in reboiler stripper column (recovery), and other blower. As described in Table III, reboiler energy consumption is 310 kW compared to blower 54 MW power. It is clear that energy efficiency measures must be done on Reboiler.

Three parameters of solvent flow rate, tower height and tower pressure are studied in order to optimize reboiler energy consumption.

Required energy for reboiler is very sensitive to changes of solvent flow rate and absorption tower efficiency is reduced at very low rates of solvent. Also, it can be said that the process performance becomes better through increasing the height of towers; in the case of using mono ethanol amine aqueous solvent, reduced pressure of recovery tower does not seem very appropriate.

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