



Verification of a solvent optimization approach for postcombustion CO₂ capture using commercial alkanolamines



Zhiwei Li^{a,*}, Shiaoguo Chen^a, David Hopkinson^b, David Luebke^{b,1}

^a Carbon Capture Scientific, LLC, 4000 Brownsville Road, P.O. Box 188, South Park, PA 15129, United States

^b National Energy Technology Laboratory, 626 Cochran Mill Road, P.O. Box 10940, Pittsburgh, PA 15236, United States

ARTICLE INFO

Article history:

Received 7 April 2015

Received in revised form

20 September 2015

Accepted 3 November 2015

Available online 7 December 2015

Keywords:

CO₂ capture

Solvent optimization

Approach verification

Alkanolamine

Postcombustion flue gas

Conventional absorption/desorption process

ABSTRACT

This paper verified a phase equilibrium approach for optimization of conceptual solvents by using process simulations of commercial solvents N-methyl-diethanolamine (MDEA) and 2-amino-2-methyl-1-propanol (AMP) aqueous solution, for a conventional absorption/desorption based postcombustion CO₂ capture process. The simulated total heat/total equivalent work for the investigated tertiary/hindered amines has the same trends as those based on the phase equilibrium approach for conceptual solvents with the same heat of reactions. Moreover, the simulated CO₂ working capacities for the commercial solvents agree well with those obtained with the phase equilibrium approach for the corresponding conceptual solvents, verifying the phase equilibrium approach. Results of parametric tests using the AMP aqueous solution illustrate that there is an optimal lean loading for the lean solution and an optimal temperature for the stripper inlet solvent to achieve the least total equivalent work/total heat.

© 2015 Elsevier Ltd. All rights reserved.

1. Introduction

Amine scrubbing is currently the leading technology for post-combustion CO₂ capture (Rochelle, 2009). However, high energy consumption in solvent-based CO₂ capture continues to be a major hurdle for commercialization. The U.S. Department of Energy (DOE)/National Energy Technology Laboratory (NETL) (DOE/NETL, 2013) estimates that the deployment of current postcombustion CO₂ capture technology, aqueous monoethanolamine (MEA) solution based chemical absorption process, in a new pulverized coal power plant would decrease the plant efficiency by 30%. Clearly, an economical and energy efficient CO₂ capture process is a prerequisite for global CO₂ emission control from fossil energy sources to mitigate global warming. However, through optimization of solvent properties and the absorption/desorption process, it may be possible to reduce the energy penalty from CO₂ capture (Oyekan and Rochelle, 2006; Hoff et al., 2006).

Energy consumption for solvent regeneration, represented by reboiler heat duty, is an important parameter for design and

operation of a cost-effective CO₂ capture process. Some commonly used solvents are aqueous solutions of alkanolamines, such as MEA, diethanolamine (DEA), and methyl-diethanolamine (MDEA) (Kohl and Nielsen, 1997). MEA is the most widely used because it has a fast rate of reaction with CO₂, reasonable degradation resistance, and low solvent cost. However, high desorption energy consumption, vaporization losses and equipment corrosion issue are disadvantages of MEA. Sterically hindered and tertiary amines, 2-amino-2-methyl-1-propanol (AMP) and MDEA are receiving increased attention recently due to their loading up to 1 mol of CO₂/mol of amine and relatively low energy consumption for solvent regeneration, leading to significant savings in process costs (Sartori et al., 1994). MEA solvent requires high reboiler heat duty, from 3800 to 5400 kJ/kg CO₂ (Sakwattanapong et al., 2005). The AMP with DEA activated solvent has a modest heat requirement of 3030 kJ/kg CO₂ (Adeosun and Abu-Zahra, 2013). To compare among the CO₂ capture processes, Oyekan and Rochelle (2006) applied the concept of “equivalent work,” which is the equivalent loss of electricity in the power plant due to the steam extraction and the required power demand due to the CO₂ compression.

Effects of various stripper configurations on energy consumption have been studied (Oyekan and Rochelle, 2006; Jassim and Rochelle, 2006; Freguia and Rochelle, 2003) and both reboiler duty and the total equivalent work were reduced compared to a simple stripper by using configurations including multipressure stripping,

* Corresponding author.

E-mail address: zhiweili@carboncapturescientific.com (Z. Li).

¹ Present address: Liquid Ion Solutions, 1817 Parkview Blvd, Pittsburgh, PA 15217, United States.

Nomenclature

C_p	specific heat of aqueous solvent [kJ/(kmol K)]
ΔH_R	the heat of reaction between the solvent and CO ₂ (kJ/mol)
ΔH_w	latent heat of water vaporization (kJ/mol)
L	flow rate of the lean solvent exiting from cross heat exchange (kmol/h)
P_{H_2O}	water partial pressures (bar)
P_{CO_2}	CO ₂ partial pressures (bar)
Q_R	heat of reaction (kJ/mol CO ₂)
Q_S	sensible heat (kJ/mol CO ₂)
Q_T	total heat demands (kJ/mol CO ₂)
Q_w	stripping heat (kJ/mol CO ₂)
R	the ideal gas constant; $R=8.314$ kJ/(kmol K)
T_1	temperature of the rich solvent entering to cross heat exchange (K)
T_2	temperature of the lean solvent exiting from cross heat exchange (K)
ΔT	temperature approach of the cross heat exchanger (K)
W_T	total equivalent work (kWh)
x	mole fraction of a solvent in aqueous solution
Y_{CO_2}	CO ₂ product yield (kmol/h)
<i>Greek symbols</i>	
α	CO ₂ loading in the aqueous amine solution (mol/mol amine)
$\Delta\alpha$	solvent working capacity (mol/mol amine)
η	Carnot cycle efficiency

vacuum stripping or stripping with vapor recompression. Because the reboiler heat duty relates inversely to lean-CO₂ loading/rich-CO₂ loading and alkanolamine concentration (Sakwattanapong et al., 2005), the reboiler energy saving can be also achieved by optimizing the lean solvent loading, the solvent concentration and the stripper operating pressure for MEA-based CO₂ capture process but the optimal heat duty is still 3000 kJ/kg CO₂ (Abu-Zahra et al., 2007).

Heat of reaction has been identified as a key property for optimization of the CO₂ capture process (Hoff et al., 2006; Hopkinson et al., 2014). A phase equilibrium approach (Hopkinson et al., 2014) was developed to describe a conceptual solvent, which is characterized by the heat of reaction between CO₂ and the solvent. Optimization of a conceptual solvent is achieved through quantification of the impact of heat of reaction on the sensible and stripping heat. The conceptual solvent was optimized for the least total equivalent work for a conventional absorption based post-combustion CO₂ capture process. Results show that the least total equivalent work is about 0.1034 kWh/kg CO₂ with a heat of reaction of 71 kJ/mol CO₂ for tertiary or sterically hindered amines under typical solvent regeneration conditions of 2 atm operating pressure.

This paper verifies the phase equilibrium approach for optimization of conceptual solvents using commercial solvents, including a tertiary amine (MDEA aqueous solvent), and a sterically hindered amine (AMP aqueous solvent). In this study, the energy performance of the absorption/desorption process using MDEA or AMP was obtained through process simulations. The thermal performance of MEA aqueous solvent was also simulated using a similar optimization approach as a reference case. The findings obtained from this work provide new insights and guidance for identifying energy efficient solvents and develop a strategy for cost-effective postcombustion CO₂ capture.

2. Methodology

2.1. Overview

The verification of the phase equilibrium approach was conducted with process simulations (Hopkinson et al., 2014). Specifically, the thermal performance of a conventional absorption/desorption based CO₂ capture process was simulated using ProTreat[®] software (Otimeas Treating Inc, 2009). ProTreat was developed for simulating processes for acid gas removal from a variety of high and low pressure gas streams by absorption into solutions including single/blended amines or physical solvents. The ProTreat simulator is a mass and heat transfer rate based model for both absorption and regeneration columns. The software was validated by comparison of vapor-liquid-equilibrium data from simulation and experimental data for various amines and at various concentrations under various temperatures.

2.2. Phase equilibrium approach

The phase equilibrium approach was developed for optimizing a conceptual solvent by minimizing the total equivalent work for absorption/stripping based post-combustion CO₂ capture (Hopkinson et al., 2014). The conceptual solvent was characterized according to its heat of reaction, which was further used to estimate sensible and stripping heat through a phase equilibrium model that links equilibrium CO₂ partial pressure with solvent properties (such as heat of reaction and working capacity) and operating conditions (such as temperature). For hindered/tertiary amines, the phase equilibrium model can be described by Hopkinson et al. (2014),

$$P_{CO_2}^* = e^{17.163+0.1035\Delta H_R - \frac{\Delta H_R}{RT}} \cdot \frac{\alpha^2}{(1-\alpha)} \quad (1)$$

where $P_{CO_2}^*$ is CO₂ equilibrium partial pressure with aqueous amine solutions; α is CO₂ loading in the aqueous amine solution, expressed as moles of chemically combined CO₂ per mole of amine; ΔH_R is the heat of reaction between the solvent and CO₂; R is the ideal gas constant; T is absolute temperature.

In absorption/stripping-based CO₂ capture processes, the total heat consumption (Q_T) includes three components: heat of reaction or heat of absorption (Q_R), sensible heat (Q_S), and stripping heat (Q_w):

$$Q_T = Q_R + Q_S + Q_w \quad (2)$$

In the phase equilibrium approach, the heat of reaction is known, which refers to the total heat released when CO₂ is absorbed from a low CO₂ loading (0–0.4 mol CO₂/mol amine) at 40 °C with typical amine concentrations. The sensible heat refers to the additional energy required after the heat recovery to achieve the desired regeneration temperature. This additional energy is quantified by a temperature approach (ΔT), which is usually known for commercial heat exchangers. The sensible heat is estimated by,

$$Q_S = \frac{C_p \Delta T}{x \Delta \alpha} \quad (3)$$

where C_p is the specific heat of the aqueous solvent; ΔT is the heat exchanger's temperature approach; x is the mole fraction of a solvent in aqueous solution; $\Delta\alpha$ is the working capacity of the conceptual solvent, defined by $\Delta\alpha = \alpha_2 - \alpha_1$, where α_1 and α_2 are the CO₂ loading (mol CO₂/mol amine) in the lean and rich solutions, respectively.

In the phase equilibrium approach, gas-liquid phase equilibrium for CO₂ is assumed to be achieved for both lean and rich solutions in both absorber and stripper. Therefore, both lean and rich CO₂ loadings can be estimated using Eq. (1) where CO₂ partial pressures in

Download English Version:

<https://daneshyari.com/en/article/1742948>

Download Persian Version:

<https://daneshyari.com/article/1742948>

[Daneshyari.com](https://daneshyari.com)