

CARBON DIOXIDE ABSORPTION INTO AQUEOUS BLENDS OF N-METHYLDIETHANOLAMINE AND 2-ETHYLAMINOETHANOL

Władysław Moniuk*, Ryszard Pohorecki, Piotr Machniewski

Warsaw University of Technology, Faculty of Chemical and Process Engineering, Waryńskiego 1, 00-645 Warszawa, Poland

Measurements of the absorption rate of carbon dioxide into aqueous solutions of N-methyldiethanolamine (MDEA) and 2-ethylaminoethanol (EAE) have been carried out. On this basis a mathematical model of the performance of an absorption column operated with aqueous solution of a blend of the above amines at elevated temperatures and pressures have been proposed. The results of simulations obtained by means of this model are described. The work is a part of a wider program, aimed at the development of a new process.

Keywords: alkanoloamines, carbon dioxide, absorption, simulation, packed column

1. INTRODUCTION

The removal of CO₂ from synthesis and flue gases has been gaining increasing interest in the recent years. This is caused by both technological and ecological reasons. The most widespread method of CO₂ removal is that of absorption accompanied by a chemical reaction. Among possible reacting solvents, aqueous amine solutions seem to be the most promising. Both single amines and amine blends have been so far proposed for industrial use. The use of mixed amines in gas purification processes is of increasing interest today. The mixed amine systems, which combine higher equilibrium capacity of the tertiary amine with the higher reaction rate of the primary or secondary amine, have been suggested for gas purification processes (Horng and Li, 2002; Liao and Li, 2002; Mandal et al., 2001, 2003; Sun et al., 2005; Ume et al., 2012; Xiao et al., 2000). In many cases the absorption process has to be carried out at elevated temperatures and pressures (e.g. in the case of synthesis gases).

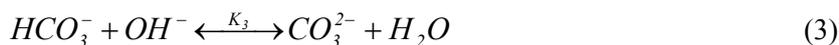
The aim of this work was to check the suitability of selected amines for flue and synthesis gas purification. To this end experimental investigations of the rate of absorption into selected amines, and simulations of the column performance using these amines were carried out. The work is a part of a wider experimental program aimed at the development of a new process.

2. REACTION MECHANISMS

2.1. MDEA [CH₃CH₂(CH₂CH₂OH)₂N]

MDEA is a tertiary amine. Reaction mechanisms of primary and secondary amines are different from those of tertiary amines. The following reactions occur during the CO₂ absorption into aqueous MDEA solutions (Donaldson and Nguyen, 1980; Rinker et al., 1995).

*Corresponding author, e-mail: w.moniuk@ichip.pw.edu.pl



The rate of reactions (1) and (2) can be expressed as:

$$r_1 = k_{21}[CO_2][R_3N] - \frac{k_{21}}{K_1}[R_3NH^+][HCO_3^-] \quad (6)$$

$$r_2 = k_{22}[CO_2][OH^-] - \frac{k_{22}}{K_2}[HCO_3^-] \quad (7)$$

where

$$K_1 = \frac{[R_3NH^+][HCO_3^-]}{[CO_2][R_3N]} \quad (8)$$

$$K_2 = \frac{[HCO_3^-]}{[CO_2][OH^-]} \quad (9)$$

In reactions (3-5) only proton exchange occurs and these reactions can be treated as very fast ("instantaneous"). Equilibrium constants for these reactions are as follows:

$$K_3 = \frac{[CO_3^{2-}]}{[HCO_3^-][OH^-]} \quad (10)$$

$$K_4 = \frac{[R_3N]}{[R_3NH^+][OH^-]} \quad (11)$$

$$K_5 = [OH^-][H_3O^+] \quad (12)$$

Rinker et al. (1995) measured the rate of carbon dioxide absorption into aqueous MDEA solutions in a wetted-sphere absorber. For the interpretation of the obtained results, three different mathematical models, based on Higbie's penetration theory, were used. The first model is the most general model. It includes reactions (1-5) and treats them as reversible reactions. The second model is almost the same as the first model except that it neglects reaction (2). The third model neglects reactions (2-5) and treats reaction (1) as an irreversible pseudo-first-order reaction.

A comparison of the predicted enhancement factor E from models 1-3 is presented in Fig. 1 (Rinker et al., 1995). As it is seen, for high values of interfacial concentration of CO₂ (~ 10⁻² kmol/m³) the values of the enhancement factor E for these 1-3 models, are very close. For the conditions considered in this work $c_i = (0.7-1.2) \cdot 10^{-2}$ kmol/m³, and therefore the simplest, the third model, can be used.

Many studies have been performed on the kinetics of the reaction (1) between MDEA and CO₂.

In Fig. 2 the dependence of the reaction rate constant, k_{21} , on the amine concentration is presented ($t = 20$ °C). As it is seen, the discrepancies between the results obtained by various authors are very large. The present measurements (Fig. 5) are in agreement with the earlier data of Moniuk and Pohorecki (2000). These data are used in the modelling.

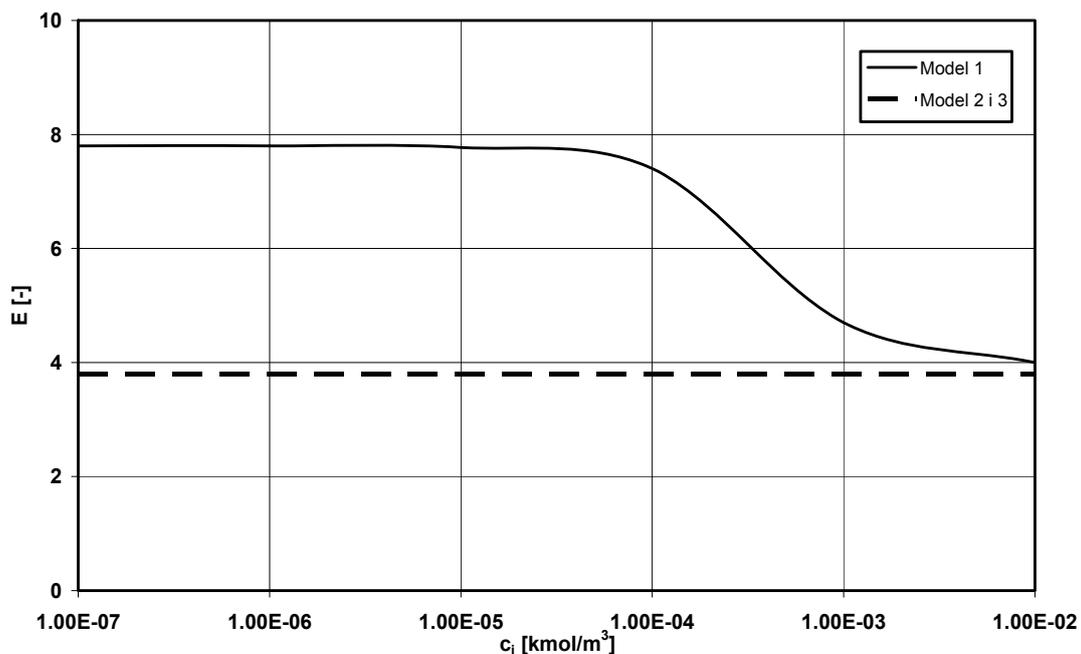


Fig. 1. A comparison of the enhancement factor values, E , obtained for models 1-3 (Rinker et al., 1995)

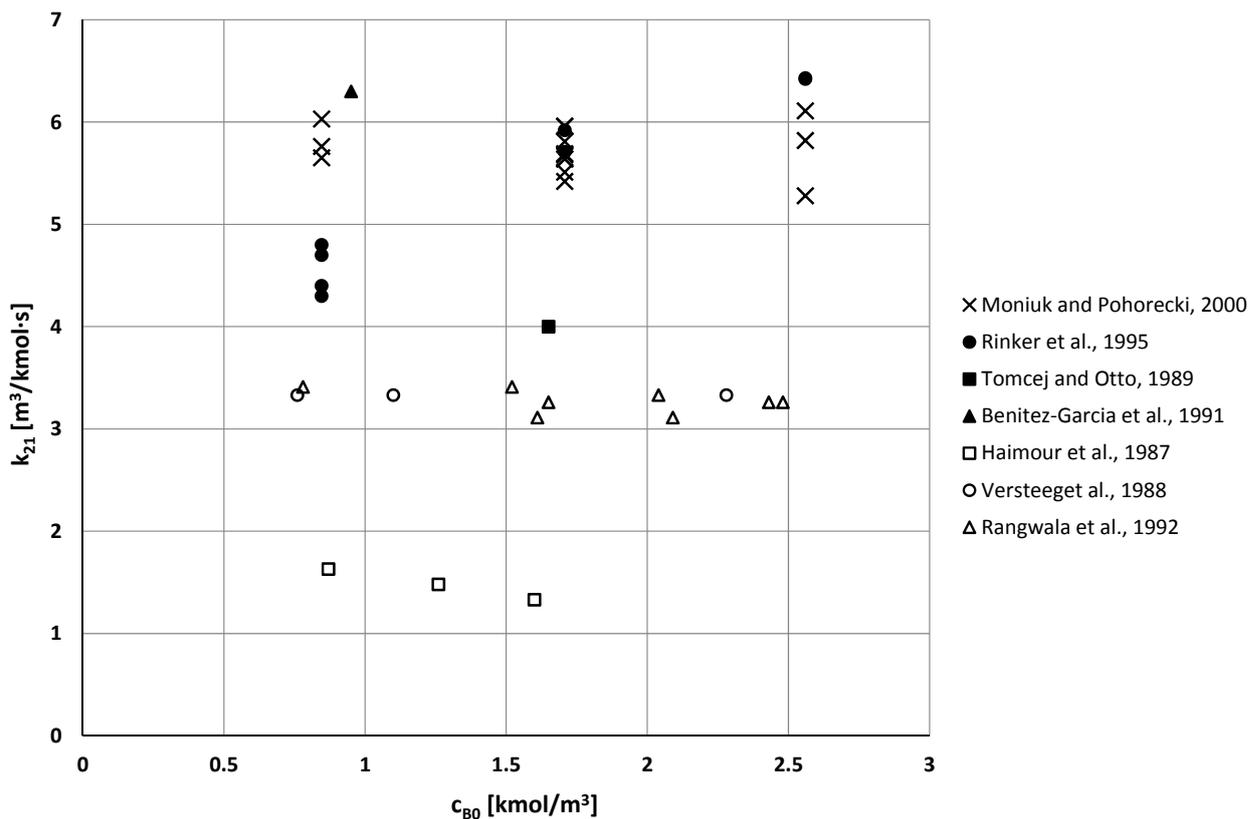


Fig. 2. The dependence of the reaction rate constant, k_{21} , on the amine concentration c_{B0} ($t = 20^\circ\text{C}$)

2.2. EAE [$\text{C}_2\text{H}_5\text{NHC}_2\text{H}_4\text{OH}$]

EAE is a secondary amine. The reactions of CO_2 with primary and secondary alkanolamines in aqueous solutions are believed to take place according to the zwitterion mechanism (Danckwerts, 1979). In this

mechanism CO₂ reacts with amine to form a zwitterion, which further reacts with any base present in the system.



If [z] is the concentration of zwitterion (in quasi-steady state) then the rate of CO₂ reaction can be expressed as:

$$r = k_2[CO_2][Am] - k_{-2}[z] = [z]\sum k_B[B] \quad (15)$$

The term $\sum k_B[B]$ indicates the contribution of the various bases present in the system to the rate of removal of protons from the zwitterion.

In our case the base (B) may be water or OH⁻ ions. From Eq. (15)

$$\frac{r}{[CO_2][Am]} = \frac{k_2}{1 + \frac{k_{-2}}{\sum k_B[B]}} \quad (16)$$

If $\frac{k_{-2}}{\sum k_B[B]} \ll 1$ we have simple second-order kinetics, as in the case of monoethanolamine (MEA).

If $\frac{k_{-2}}{\sum k_B[B]} \gg 1$ we get

$$\frac{r}{[CO_2][Am]} = \frac{k_2}{k_{-2}} \sum k_B[B] \quad (17)$$

and we have third-order kinetics as in the case of diethanolamine (DEA).

There is also an alternative mechanism (termolecular mechanism) proposed by Vaidya and Kenig (2010), but the above (zwitterion) mechanism is more frequently used.

3. EXPERIMENTAL APPARATUS AND PROCEDURE

Measurements of the rate of CO₂ absorption into aqueous solutions of MDEA and EAE were carried out in a stirred reactor (Autoclave Engineers Inc.) with capacity 1·10⁻³ m³. The scheme of the experimental apparatus is shown in Fig. 3.

At the beginning of the experiment, a mixture of gases (CO₂ and N₂) was supplied through the tree-way valve (10) to the CO₂ analyser (12). During the main part of the experiment the experimental conditions (pressure, temperature and stirrer speed) were kept constant, and the mixture of gases was supplied through the three – way valve to the gas sparger (8) below the stirrer (9). CO₂ concentration at the gas outlet from the reactor was measured by the CO₂ analyser (12).

The measurements were carried out under 1-2 bar pressure at the temperature range: 20 – 70 °C. The other parameters were as follows: stirred speed 700 rpm; concentration of CO₂ in the inlet gas 46-50 %vol.; amines concentration: MDEA: 10, 20 and 30 wt.%; EAE: 1 and 2 wt.%.

The MDEA was obtained from Sigma – Aldrich with a purity > 98,5 wt.%, whereas the EAE was obtained from Aldrich with a purity > 98wt.%.

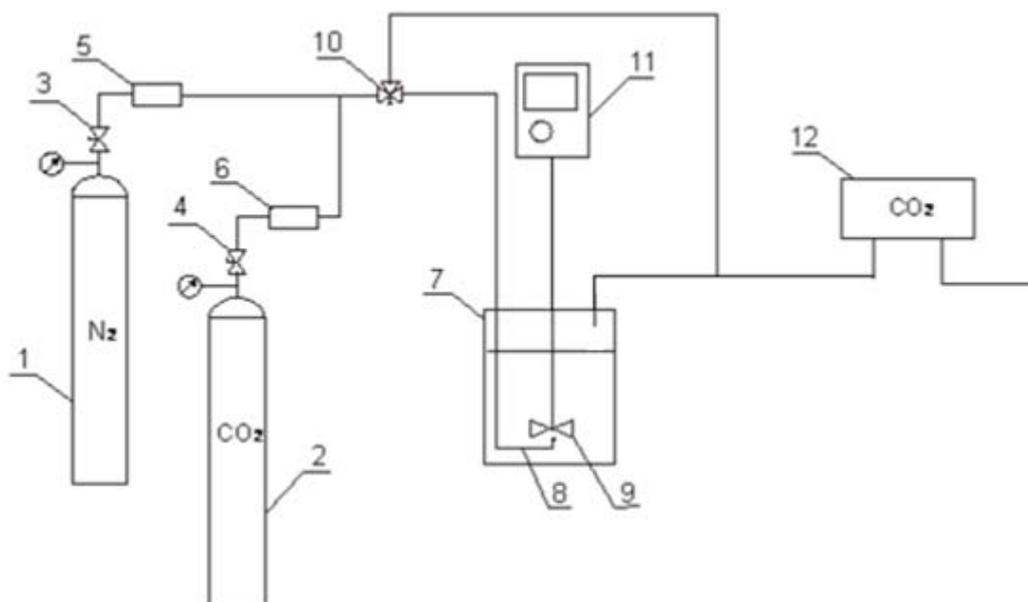


Fig. 3. Scheme of the experimental apparatus

1, 2 – gas cylinders; 3, 4 – pressure reducing valves; 5, 6 – gas flow meters; 7 – stirred reactor; 8 – sparger; 9 – stirrer; 10 – three – way valve; 11 – temperature, pressure and stirrer speed regulators; 12 – CO₂ analyser (DCS, Model 300)

On the basis of the CO₂ concentrations measured at the gas inlet to the reactor and the gas outlet from the reactor, the rate of CO₂ absorption, R , was calculated. On the other hand we have:

$$R = N \cdot a \quad (18)$$

In the process of absorption with chemical reaction, the molar flux of absorbed CO₂ is equal to

$$N = k_L^* \cdot c_{Ai} \quad (19)$$

The concentration of CO₂ at the interface was calculated from the Henry's law

$$p_{Ai} = H \cdot c_{Ai} \quad (20)$$

The partial pressure of CO₂ at the interface, p_{Ai} , was calculated from the relation

$$N = k_g (p_{Ao} - p_{Ai}) = k_L^* \cdot c_{Ai} \quad (21)$$

The values of the mass transfer coefficient in the gas phase were taken from the Versteeg et al. (1987) data.

For the gas flow in the reactor we assumed plug flow. In this case the average value of the partial pressure of CO₂ in the gas phase is equal to

$$p_{Ao} = \frac{(p_{Ao(inlet)} - p_{Ao(outlet)})}{\ln \frac{p_{Ao(inlet)}}{p_{Ao(outlet)}}} \quad (22)$$

4. RESULTS

From the CO₂ concentrations measured in the gas phase, the rate of CO₂ absorption, R and the mass transfer coefficient with chemical reaction, k_L^* were calculated.

Making use of the values of the physical mass transfer in the liquid phase, k_L , determined experimentally in our earlier work (Moniuk et al., 1997) the values of the enhancement factor, E , defined as:

$$E = \frac{k_L^*}{k_L} \quad (23)$$

were calculated.

On the other hand, enhancement factor E can be determined from the van Krevelen and Hoftijzer diagram (Ramm, 1976)

$$E = f(Ha, \beta) \quad (24)$$

where

$$Ha = \frac{\sqrt{k_{21} \cdot c_{B0} \cdot D_A}}{k_c} \quad (25)$$

$$\beta = \frac{c_{B0} \cdot D_B}{b \cdot c_{Ai} \cdot D_A} \quad (26)$$

For $Ha \geq 2$ Porter's relations can be used:

$$E = 1 + \beta \cdot \left[1 - \exp\left(\frac{-(Ha - 1)}{\beta}\right) \right] \quad (27)$$

For a fast pseudo-first-order reaction, when:

$$k_L \leq \sqrt{k_{21} \cdot c_{B0} \cdot D_A} \quad (28)$$

$$Ha \leq \frac{c_{B0} \cdot D_B}{b \cdot c_{Ai} \cdot D_A} \quad (29)$$

the enhancement factor is equal to the Hatta number value:

$$E = H \cdot a \quad (30)$$

Enhancement factor E is thus a function of the Hatta number and the parameter β . In our case, depending on the values of enhancement factor E and parameter β , Eqs. (24) or (27) or (30) could be used to determine the values of the enhancement factor, and finally the reaction rate constant, k_{21} or k_2 values were calculated by comparison with Eq. (23).

4.1. System: CO_2 – MDEA (tertiary amine)

In Fig. 4 the dependence of the reaction rate (1) constant k_{21} on the amine concentration for various temperatures is presented.

In Fig. 5 the Arrhenius plot ($\log k_{21} = f(1/T)$) is presented. As it is seen the values k_{21} practically do not depend on the amine concentration. The straight line can be described by the relation:

$$\log k_{21} = 10.25 - (2792.3/T) \quad (31)$$

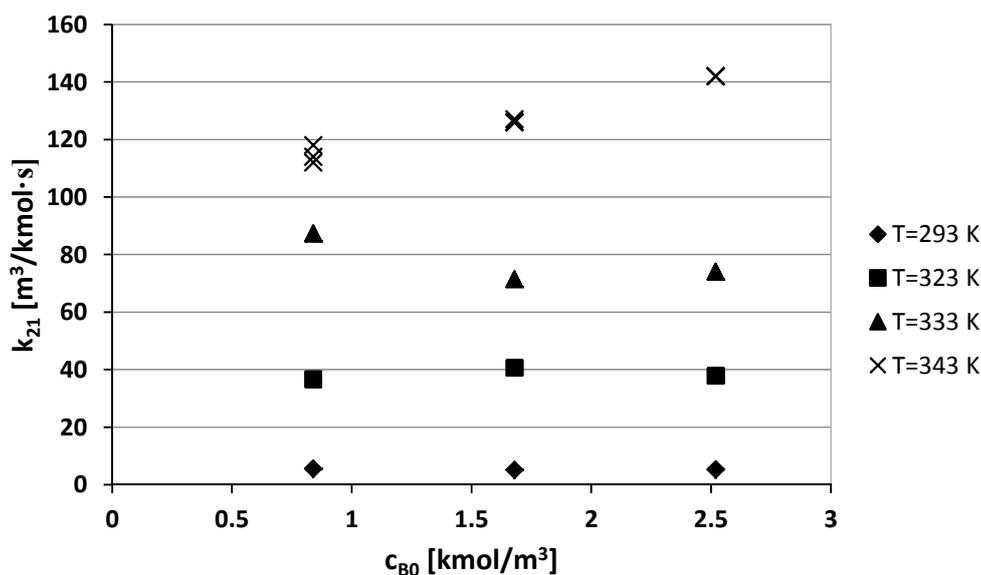


Fig. 4. The dependence of the reaction (1) rate constant k_{21} on the amine (MDEA) concentration for various temperatures

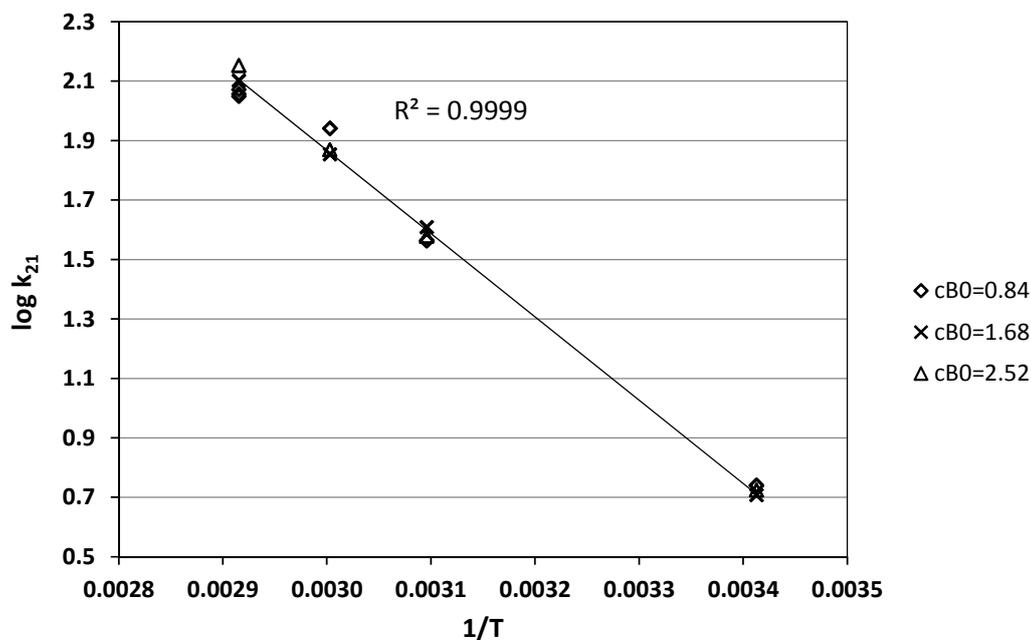


Fig. 5. The Arrhenius plot $\log k_{21} = f(1/T)$

4.2. System: CO_2 – EAE (secondary amine)

The values of the reaction (13) rate constant k_2 were calculated in a similar way as the reaction (1) constant, k_{21} . In Fig. 6 the dependence of the reaction (13) rate constant k_2 on the amine concentration for various temperatures is presented. In Fig. 7 the Arrhenius plot ($\log k_2 = f(1/T)$) is presented. The straight line can be described by the relation:

$$\log k_2 = 10.052 - (1929.8 / T) \quad (32)$$

The obtained data on the reaction rate constants k_{21} (CO_2 –MDEA system) and k_2 (CO_2 –EAE system) were used for simulation of CO_2 absorption into aqueous solutions of a blend of both amines.

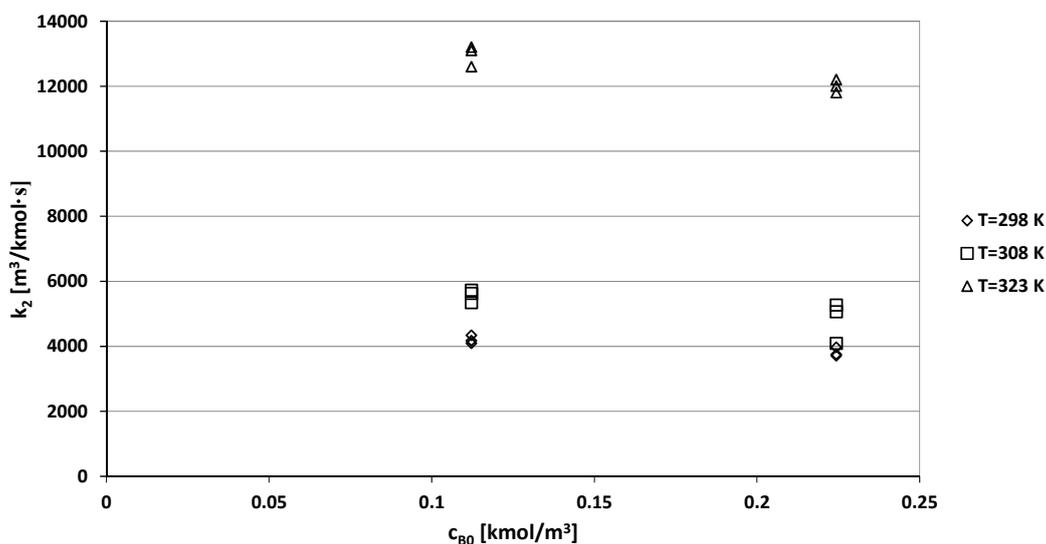


Fig. 6. The dependence of the reaction (13) rate constant, k_2 , on the amine (EAE) concentration for various temperatures

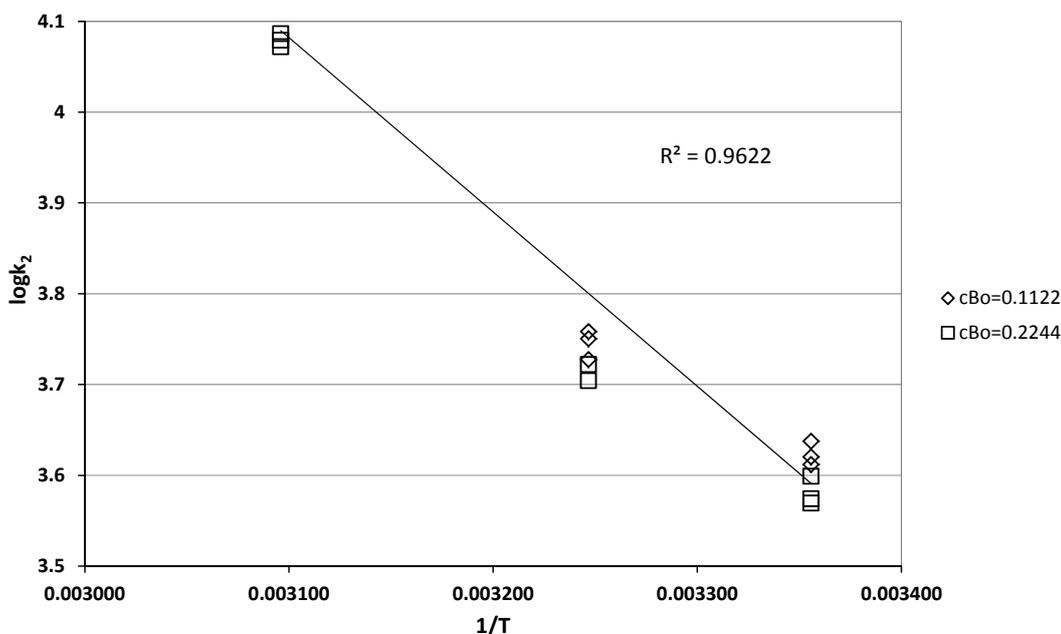


Fig. 7. The Arrhenius plot $\log k_2 = f(1/T)$

5. MATHEMATICAL MODEL OF CO₂ ABSORPTION INTO AQUEOUS ALKANOLAMINE SOLUTIONS

In order to estimate the performance of an absorption process in a solvent containing the mixture of both amines, a rate-based mathematical model of a packed column has been used. For the process of absorption with a chemical reaction of the type:



The differential equation for the rate of absorption in a counter-current packed column can be written as follows (Danckwerts, 1970):

$$R_A \cdot a \cdot dh = d(G_{OM} \cdot x_A) \quad (34)$$

For a dilute gas phase we have:

$$d(G_{OM} \cdot x) \cong \frac{G_{OM}}{1-x_A} dx_A \quad (35)$$

Hence, from Eq. (34) and (35) we have:

$$\frac{dx_A}{dh} = \frac{1-x_A}{G_{OM}} R_A \cdot a \quad (36)$$

In the heat balance one should consider heat exchange between phases, heat effects of absorption, reaction and solvent evaporation (or condensation), as well as heat losses to the environment.

In the case of heat exchange between phases, one can neglect the heat transfer resistance in the liquid phase (Sherwood and Pigford, 1952). Thus we have:

$$dN_t = \alpha_g \cdot a (T_g - T_l) dh \quad (37)$$

From the heat balance

$$dN_t = d(G_{OM} \cdot C_p \cdot T_g) \cong G_{OM} \cdot C_p \cdot dT_g \quad (38)$$

Hence

$$\frac{dT_g}{dh} = \frac{\alpha_g \cdot a}{C_p \cdot G_{OM}} (T_g - T_l) \quad (39)$$

The heat transfer coefficient can be determined from heat and mass transfer analogy

$$\frac{Nu_g}{Sh_g} = \frac{A \cdot Re_g^B \cdot Pr_g^C}{A \cdot Re_g^B \cdot Sc_g^C} = \left(\frac{Pr_g}{Sc_g} \right)^C \quad (40)$$

Assuming $C = 0.33$ (Sherwood and Pigford, 1952) and making use of the definitions of Prandtl and Schmidt numbers we have:

$$\alpha_g = k_g \left(\frac{\lambda_g}{D_g} \right)^{0.66} \cdot (C_p \cdot \rho_g)^{0.33} \quad (41)$$

Hence

$$\frac{dT_g}{dh} = \frac{k_g \cdot a}{G_{OM}} \left(\frac{\lambda_g}{C_p \cdot D_g} \right)^{0.66} \cdot \rho_g^{0.33} \cdot (T_g - T_l) \quad (42)$$

In the case of solvent evaporation, one can neglect the mass transfer resistance in the liquid phase. Thus we have:

$$dN = k_{gw} \cdot a \cdot (p_w - p_{wr}) dh = k_g \cdot a \cdot P \cdot (x_w - x_{wr}) dh \quad (43)$$

and

$$dN = d(G_{OM} \cdot x_w) \cong d\left(\frac{G_M}{1-x_w} \cdot x_w\right) = G_M d\left(\frac{x_w}{1-x_w}\right) = G_M \frac{dx_w}{(1-x_w)^2} = G_{OM} \frac{dx_w}{1-x_w} \quad (44)$$

Hence

$$\frac{dx_w}{dh} = \frac{k_{gw} \cdot a \cdot P \cdot (1 - x_w)}{G_{OM}} (x_w - x_{wr}) \quad (45)$$

The heat effect of solvent evaporation is equal to the product of the heat of vaporization (q_w) and the amount of evaporated solvent. The heat effect of absorption and reaction is equal to the product of the overall heat of absorption and reaction (q_A) and the amount of absorbed component. In these conditions the heat balance can be written as follows (Hobler, 1968)

$$\frac{dT_l}{dh} = \frac{G_{OM}}{L_o \cdot c_l} \cdot \left(-C_p \frac{dT_g}{dh} + q_w \frac{dx_{H_2O}}{dh} + q_r \frac{dx_{CO_2}}{dh} \right) + \frac{q_s}{L_o \cdot c_l} \quad (46)$$

The simulations of CO₂ absorption into aqueous alkanoloamines solutions in the packed column consisted in simultaneous numerical integrations of the following equations:

- Eqs. (36) and (45) describing the concentration profiles of the absorbed component x_A and the solvent vapour x_w in the gas phase;
- Eqs (42) and (46) describing the temperature profiles in the gas (T_g) and liquid (T_l) phases;
- Eqs (47) and (48) describing the profiles of gas (G_o) and liquid (L_o) flow rates

For the dilute gas phase

$$\frac{dG_o}{dh} = M_A G_{oM} \frac{dx_A}{dh} + M_W G_{oM} \frac{dx_w}{dh} \quad (47)$$

For the counter-current flow

$$\frac{dL_o}{dh} = \frac{dG_o}{dh} \quad (48)$$

6. RESULTS OF SIMULATIONS

A computer program for the numerical integration of Eqs. (38), (42), (45), (46), (47) and (48) along the column height has been developed in MATLAB environment. The rate of CO₂ absorption in packed column can be calculated from the relation (Pohorecki et al., 1978, 1990; Moniuk 2006)

$$R_A = (x_A \cdot P - p_{Ar}) \frac{H \cdot \sqrt{D_A \cdot k_l + k_l^2}}{1 + \frac{H \cdot R \cdot T}{k_g} \cdot \sqrt{D_A \cdot k_l + k_l^2}} \quad (49)$$

The pseudo-first-order reaction rate constant, k_l is equal to:

$$k_l = k_{21(MDEA)} [MDEA] + k_{2[EAE]} [EAE] \quad (50)$$

The simulations were carried out for the following industrially important conditions (Pohorecki et al., 1979, 1990; Moniuk 2006):

- height of column: 20-30 m
- column diameter: 2 m
- packing of column: steel Pall rings
- pressure: 20 - 25 bar
- liquid temperature at the inlet to the column (top): 70 - 110 °C
- gas temperature at the inlet to the column (bottom): 70 - 110 °C
- mole fraction of CO₂ in the gas inlet: $x_A = 0.18 - 0.20$
- liquid flow: 345 m³/h

- gas flow: 79000 m³/h (in normal conditions)
- amines concentration: MDEA: 30 wt.%, EAE: 2 wt.%

Mass transfer coefficient in the gas phase, k_g , was calculated from Gildenblat correlation (Pohorecki et al., 1978; Ramm, 1976); mass transfer coefficient in the liquid phase k_L was calculated from Sherwood and Holloway correlation (Hobler, 1962; Pohorecki et al., 1978). The overall heat of absorption and reaction in aqueous MDEA solution is equal to $q_R = 48.6$ kJ/mol (Critchfield and Rochelle, 1987).

The heat of solvent vaporization was calculated from Watson relation (Pohorecki et al., 1978)

$$q_w = 40.62 \cdot \left(2.3615 - \frac{T}{274.1} \right)^{0.38} \quad (51)$$

Some simulation results are presented in Figs. 8 – 11.

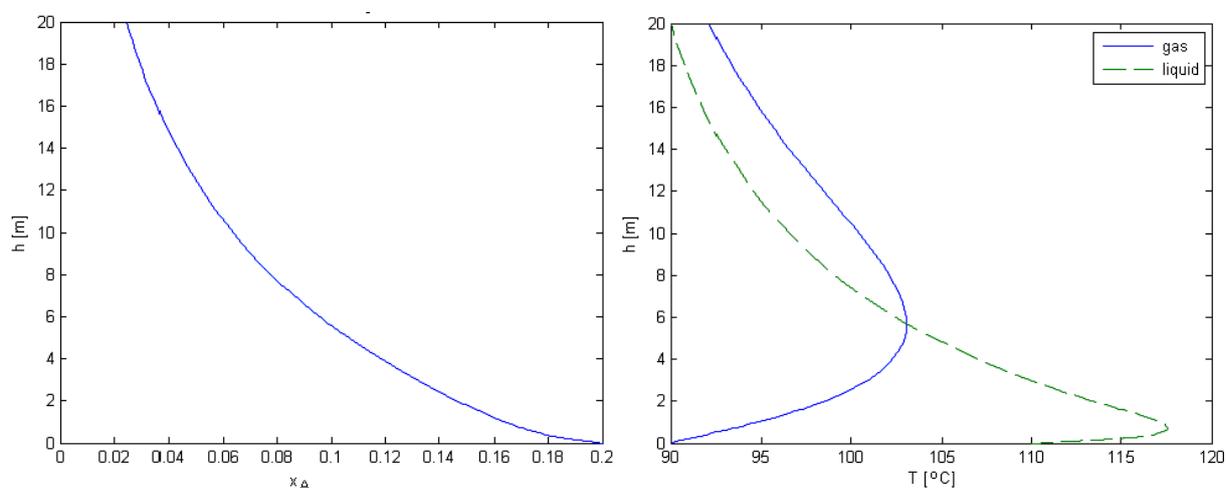


Fig. 8. Profiles of CO₂ mole fraction in the gas phase, (x_A) and gas and liquid temperatures (T) along the column height, h ; $P=20$ bar, $T_g=90$, $T_L=90$, 30wt.% MDEA + 2wt.% EAE

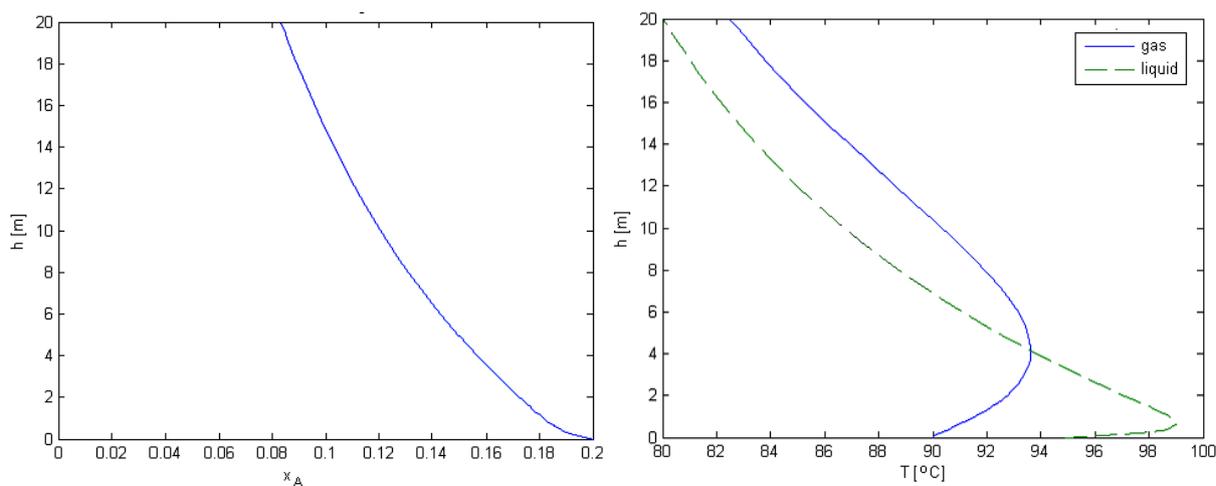


Fig. 9. Profiles of CO₂ mole fraction in the gas phase, (x_A) and gas and liquid temperatures (T) along the column height, h ; $P=20$ bar, $T_g=90$, $T_L=80$, 30wt.% MDEA + 0wt.% EAE

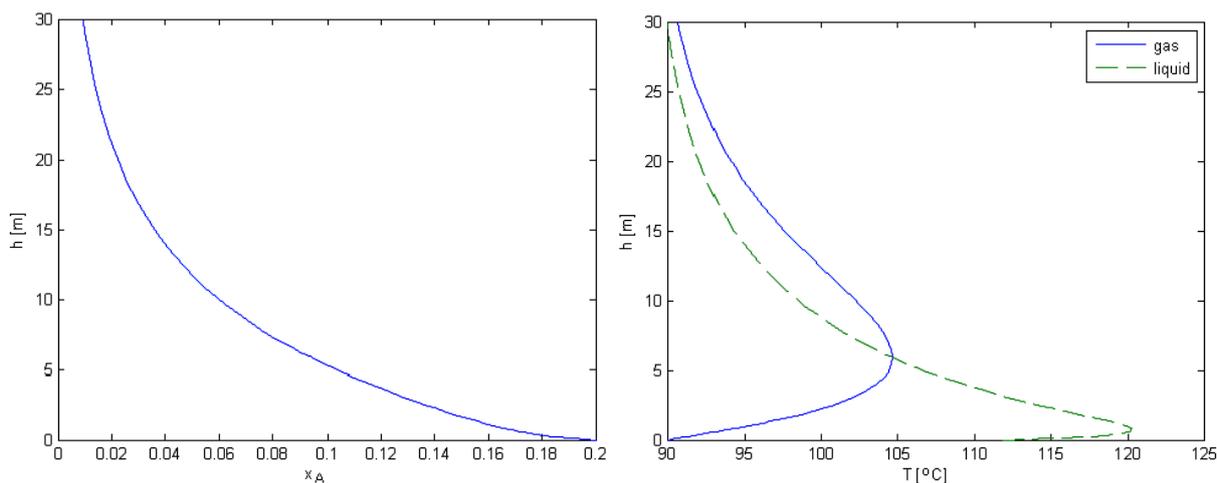


Fig. 10. Profiles of CO₂ mole fraction in the gas phase, (x_A) and gas and liquid temperatures (T) along the column height, h ; $P = 20$ bar, $T_g = 90$, $T_L = 90$, 30wt.% MDEA + 2wt.% EAE

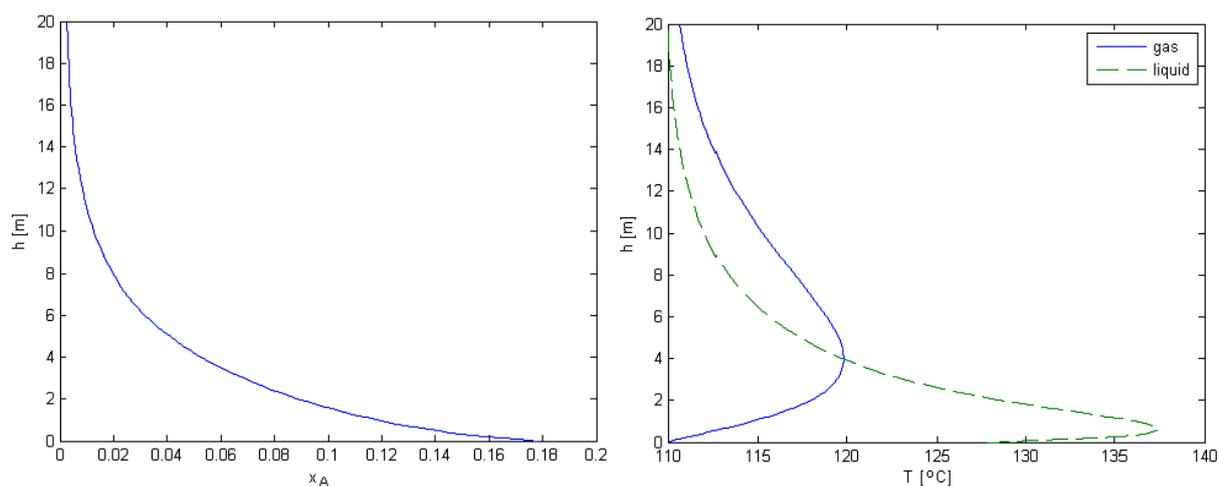


Fig. 11. Profiles of CO₂ mole fraction in the gas phase, (x_A) and gas and liquid temperatures (T) along the column height, h ; $P = 25$ bar, $T_g = 110$, $T_L = 110$, 30wt.% MDEA + 2wt.% EAE

In order to determine the most suitable conditions for an industrial process, similar calculations have been carried out for a range of gas and liquid temperatures.

It has been found that absorption efficiency of CO₂ practically does not depend on the gas temperature but significantly depends on the liquid temperature. It is therefore important to note, that there is a significant temperature increase at the bottom of the column where the rate of CO₂ absorption is the highest. It is the result of the high value of the overall heat of absorption and reaction of CO₂ in aqueous MDEA solutions.

Absorption efficiency significantly depends on EAE concentration. In similar conditions where $c_{EAE} = 0$ wt.% (Fig. 9) the mole fraction of CO₂ at the top of the column decreases only to 8.4% (Fig. 9). For the column height of 30 m (Fig. 10) CO₂ concentration decreases to 1%. In conditions similar to the industrial BENFIELD process (CO₂ absorption into aqueous potash solutions with DEA, $P = 25$ bar, t_L at the bottom of the column = 110 °C, $x_A = 0.178$) the mole fraction of CO₂ at the top of the column decreases to 0.018% (Fig. 11)

It has been concluded that the new solvent (aqueous 30% MDEA solutions with 2% EAE) may be promising in the process of gas purification in ammonia production. This solvent has been registered for a patent (Polish Patent Application P391333).

7. CONCLUSIONS

In order to develop a new process of synthesis and flue gases purification the measurements of the rate of CO₂ absorption into aqueous of MDEA and EAE were carried out in a stirred reactor and the reaction rate constants between CO₂ and alkanoloamines were determined.

On this basis a mathematical model for the performance of an absorption column operated with aqueous solution of a blend of the above amines at elevated temperatures and pressures has been developed. The results of simulations obtained by means of this model are described.

The work is a part of a wider program, aimed of the development of a new technological process.

SYMBOLS

<i>A, B, D</i>	components
<i>Am</i>	amine
<i>a</i>	interfacial area per unit volume of packing, m ² /m ³
<i>B</i>	base
<i>b</i>	stoichiometric coefficient
<i>C_P</i>	molar heat capacity of the gas, kJ/(kmol·K)
<i>c</i>	molar concentration, kmol/m ³
<i>c_{Bo}</i>	amine concentration, kmol/m ³
<i>c_i</i>	heat capacity of the solution, kJ/(kg·K)
<i>DEA</i>	diethanolamine
<i>D_A</i>	diffusivity of the absorbed component in the liquid phase, m ² /s
<i>D_B</i>	diffusivity of the liquid component in the liquid phase, m ² /s
<i>D_g</i>	diffusivity of the absorbed component in the gas phase, m ² /s
<i>d</i>	stoichiometric coefficient
<i>EAE</i>	2 ethylaminoethanol
<i>E</i>	enhancement factor
<i>G_o</i>	mass gas flow rate per unit column area, kg/(m ² ·s)
<i>G_M</i>	molar inert gas flow rate per unit column area, kg/(m ² ·s)
<i>G_{OM}</i>	molar gas flow rate per unit column area, kg/(m ² ·s)
<i>h</i>	packing height, m
<i>H</i>	Henry's constant, kmol/(m ³ ·bar)
<i>Ha</i>	Hatta number
<i>K_i</i>	equilibrium reaction (i) constant
<i>k_L</i>	liquid-film mass transfer coefficient, m/s
<i>k_L[*]</i>	liquid-film mass transfer coefficient with chemical reaction, m/s
<i>k_g</i>	gas-film mass transfer coefficient, kmol/(m ² s bar)
<i>k₁</i>	pseudo-first-order reaction rate constant, 1/s
<i>k₂</i>	reaction (13) rate constant (from left side to right side), m ³ /(kmol·s)
<i>k₋₂</i>	reaction (13) rate constant (from right side to left side), m ³ /(kmol·s)
<i>k₂₁</i>	reaction (1) rate constant, m ³ /(kmol·s)
<i>k₂₂</i>	reaction (2) rate constant, m ³ /(kmol·s)
<i>L_o</i>	mass liquid flow rate per unit column area, kg/(m ² ·s)

<i>MEA</i>	monoethanolamine
<i>MDEA</i>	N-methyldiethanolamine
<i>N</i>	molar flux, kmol/(m ² ·s)
<i>N_t</i>	heat flux, kJ/(m ² ·s)
<i>Nu</i>	Nusselt number
<i>P</i>	total pressure, bar
<i>Pr</i>	Prandtl number
<i>R</i>	rate of absorption, kmol/(m ² ·s)
\bar{R}	gas constant, J/(mol·s)
<i>r</i>	rate of chemical reaction, kmol/(m ³ ·s)
<i>Sc</i>	Schmidt number
<i>Sh</i>	Sherwood number
<i>T</i>	absolute temperature, K
<i>t</i>	temperature, °C
<i>x</i>	mole fraction
<i>z</i>	zwitterion

Greek symbols

α	heat transfer coefficient, kJ/(m ² ·s·K)
β	parameter defined by Eq. (25)
λ	thermal conductivity, W/(m·K)
ρ	density, kg/m ³

Subscripts

<i>A</i>	absorbed component
<i>g</i>	gas
<i>i</i>	interface
<i>l</i>	liquid
<i>w</i>	water

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