

The axial mixing influence on the number stages of countercurrent absorbers in absorbing NH₃-CO₂-H₂O mixture in NaCl-H₂O system

Milovan Jotanovic, Vladan Micic, Goran Tadic

Faculty of Technology Zvornik, University of East Sarajevo

e-mail: tfzv@rstel.net

This paper investigates the influence of axial or longitudinal mixing of stages on the number stages in a countercurrent absorber in absorbing vapour-gas mixture of NH₃-CO₂-H₂O in the multicomponent system NaCl-H₂O. The number stages in ideal mixing of phases N_T and the number of stages which involves axial mixing N_D were determined. The authors of this paper conclude that the influence of axial mixing does not affect the number stages in the countercurrent absorber. Otherwise, absorption in this concrete system is practically present in the process of soda ash production.

Keywords: Soda ash production, absorption, NaCl-H₂O system, NH₃-CO₂-H₂O system, absorber with countercurrent grating stages

Introduction

During the process of soda ash production, in a saturated solution of sodium chloride, ammonia and partially carbon dioxide are absorbed in the absorption plant, and then the multicomponent solution is saturated with carbon dioxide in the carbonization plant. The basic purpose of the absorption plant is to obtain a certain quantity of ammonized solution in the form of continuous material flow with the following parameters: the content of NH₃-85.1 kg/m³; the content of NaCl-310 kg/m³; temperature "28-32 °C". Vapour-gas mixture of NH₃-CO₂-H₂O which comes from the plant for ammonium desorption, is used for the saturation of NaCl solution (NaCl-H₂O system) by means of the absorption of NH₃ and CO₂. The amount of steam in this mixture depends on the temperature of the mixture. The decrease of the steam which arrives for absorption enables the increase of NH₃ and NaCl concentration in the multicomponent solution. Vapour-gas mixture which arrives for absorption usually contains around 52% of NH₃, 25% of CO₂, and the rest is steam with the temperature of 60°C. The absorption of vapour-gas mixture containing NH₃, CO₂ and steam in the above mentioned proportion takes place in a countercurrent barbotage apparatus. The basic part of these absorbers is the stage with bells. The stage with bells is a practical form of mass transfer units. The number stages or the number of mass transfer units depends on the material balance and the equilibrium conditions during the absorption of NH₃ and CO₂ in the multicomponent system of NaCl-H₂O.

The latest construction solution for the absorber is a countercurrent absorber with grating stages (Zaiisev et al., 1986). The absorber consists of four countercurrent grating stages with the free section of 0.15-0.25 m²/m².

The authors of this paper investigated the effect of axial (longitudinal) mixing of liquid and gas phases on the number stages in the countercurrent absorber with the mentioned construction. Considering the non-ideal characteristics of the the phase flow, it is necessary to determine the number stages in the absorber during the absorption of

vapour-gas mixture in the solution of NaCl-H₂O of the same concentration and parameters as in the factories where soda ash is produced.

Material balance of the absorber

In the absorber, the absorption of NH₃ and CO₂ from the vapour-gas mixture G (kg/h) takes place in the solution which initially is the system NaCl-H₂O L (kg/h). The gas and the liquid phases flow through the absorber countercurrently. At contact elements (stages), mass and heat transfer occurs. The number stages, four of them in practice, allows the exchange of the total mass of NH₃ and CO₂ from the gas phase to the liquid phase. Figure 1 shows the scheme of the countercurrent absorber, whereas Table 1 gives the material balance of the absorber.

FIGURE 1. SCHEME OF COUNTERCURRENT ABSORBER

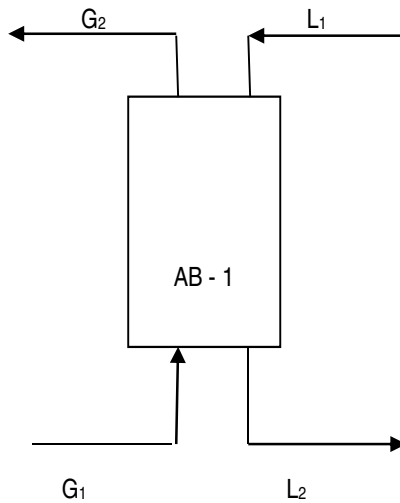


TABLE 1. MATERIAL BALANCE OF THE ABSORBER IN KG/1T Na₂CO₃

Components	G ₁ , kg	y, kg/kg	L ₁ , kg	x, kg/kg	G ₂ , kg	y, kg/kg	L ₂ , kg	x, kg/kg
NH ₃	191.6	0.4947	79.3	0.0127	4.5	0.1018	266.4	0.0405
CO ₂	110.2	0.2845	36.3	0.0058	35.8	0.8099	110.7	0.0168
NaCl	-	-	1575.0	0.2530	-	-	1575.1	0.2398
NaSO ₄	-	-	34.9	0.0056	-	-	34.9	0.0053
H ₂ O	85.6	0.2210	4499.5	0.7228	3.9	0.0882	4581.1	0.6974
Total	387.3	1.0000	6225.0	1.0000	44.2	1.0000	6568.1	1.0000

It must be noted that this absorber, denoted as AB-1, is not the only apparatus within the absorption plant with the task to perform the absorption of NH₃ and CO₂. There are also

the absorber AB-2 and three scrubbers- however, they have very different constructions for stages as contact elements and are not discussed in this paper.

Investigation of equilibrium during the absorption of NH₃ and CO₂ in NaCl-H₂O system

A well defined equilibrium during the absorption of NH₃ and CO₂ from vapour-gas mixture in the solution of multicomponent system NaCl-H₂O is of major importance in determining the number stages in the absorber. Calculation of number stages is performed from the top downwards and it is done only for ammonia in order to achieve simplicity.

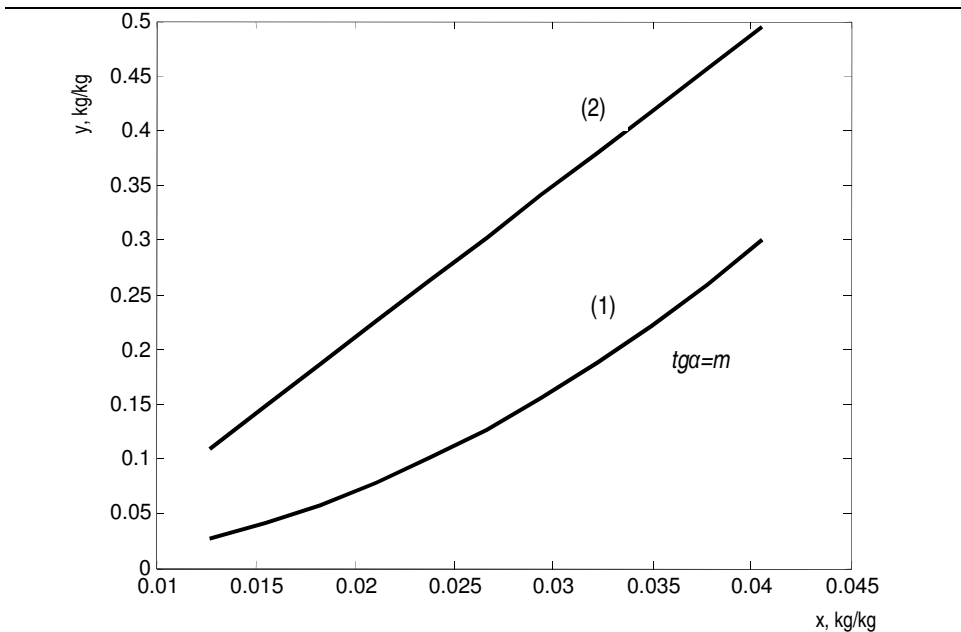
The equilibrium ammonia content in the gas over each stage is determined according to the following equation of interphase equilibrium (Zaiisev et al., 1986):

$$y^* = 188.63x^2 - 0.2562x \quad (1)$$

Where : y^* -equilibrium ammonia content in the gas, kg/kg; x -the ammonia content in the liquid, kg/kg.

Using the information for material balance (Table1) and the equation for interphase equilibrium (1), we can present a graph of the absorption for this system, which is defined by operating (direction) and equilibrium curve (Figure 2).

FIGURE 2. OPERATING AND EQUILIBRIUM LINE FOR ABSORPTION VAPOUR MIXTURE NH₃ - CO₂ - H₂O IN MULTICOMPONENT SYSTEM NaCl - H₂O



In Figure 2, both operating and equilibrium curves are determined by the following mathematical equations:

- operating line,

$$y_2 = \frac{G_1}{G_2} y_1 + \frac{L_2}{G_2} x_2 - \frac{L_1}{G_2} x_1 \quad (2)$$

- equilibrium line, whose functional dependence has already been stated,

$$y = 188.63x^2 - 0.2562x \quad (1)$$

Following the graph of the absorption shown in Figure 2, we can determine other significant parameters of equilibrium in this system:

- the degree of absorption,

$$\varphi = \frac{y_1 - y_2}{y_1 - y_2^*} = \frac{0.4947 - 0.10963}{0.4947 - 0.03384} = 0.8355 \quad (3)$$

- specific consumption of the liquid phase (Figure 2),

$$\ell = \text{tg } \alpha = 13.75 \quad (4)$$

- phase distribution coefficient (Figure 2) is determined as mean slope value of the curve showed by equation (1) toward numerical method for calculation mean value of function in interval $x_1 - x_2$

$$m = 9.7951 \quad (5)$$

- absorption factor,

$$A = \frac{\ell}{m} = \frac{13.75}{9.7951} = 1.40 \quad (6)$$

- phase flow ratio,

$$Q = \frac{L}{G}, \quad \frac{L_1}{G_1} = 16.95, \quad \frac{L_2}{G_2} = 140.8. \quad (7)$$

Calculation of the number stages

The stage or the mass transfer unit where phases are in the state of mutual equilibrium is termed theoretical stage. There is a significant number of written documents that give calculations of the number of theoretical stages using analytical, graphical, graphoanalytical, graphonumerical, and numerical methods. When choosing a method of calculation of the number of theoretical stages, one must take into consideration limiting conditions, available information, and required accuracy of calculation.

TABLE 2. CALCULATION OF NUMBER STAGES

	Number stages			
	1	2	3	4
Equilibrium NH ₃ content in the gas, kg/kg	0.02735	0.03947	0.09120	0.19647
Quantity, kg/h:				
- of the absorbed NH ₃	646	2046	2893	2439
- of the absorbed CO ₂	257	814	1150	970
Flow, kg/h:				
-of the liquid after the stage	260581	264331	269638	274111
- of the gas in the stage	3027	6779	12084	16557
Ammonia content, kg/kg:				
- in the liquid after the stage	0.01516	0.02268	0.03296	0.04132
- in the gas which reaches the stage	0.2754	0.4248	0.4777	0.4960

Calculation of the required number stages in the countercurrent absorber is performed by solving the system of equations for interphase transfer and material balance for each stage. Calculation starts from the top stage towards the bottom of the absorber. In order to achieve simplicity, it is only done for ammonia. It uses the equation (1) and the equations of material balance according to the methodology used in suitable scientific literature (Zaïisev et al., 1986). The results of the calculation are given in Table 2.

Calculation was repeated until the concentration of ammonia in the liquid phase at one stage was lower than the appointed NH₃ concentration at the entrance into the absorber. As it is seen in Table 2, the ammonia content in the liquid phase at the exit from the absorber is 0.04132 kg/kg which is higher than the appointed concentration (0.0406 kg/kg). Therefore, we can conclude that the absorber has to contain four stages.

Real apparatuses in the plants for soda ash production really contain four stages and they are in fact countercurrent grating stages.

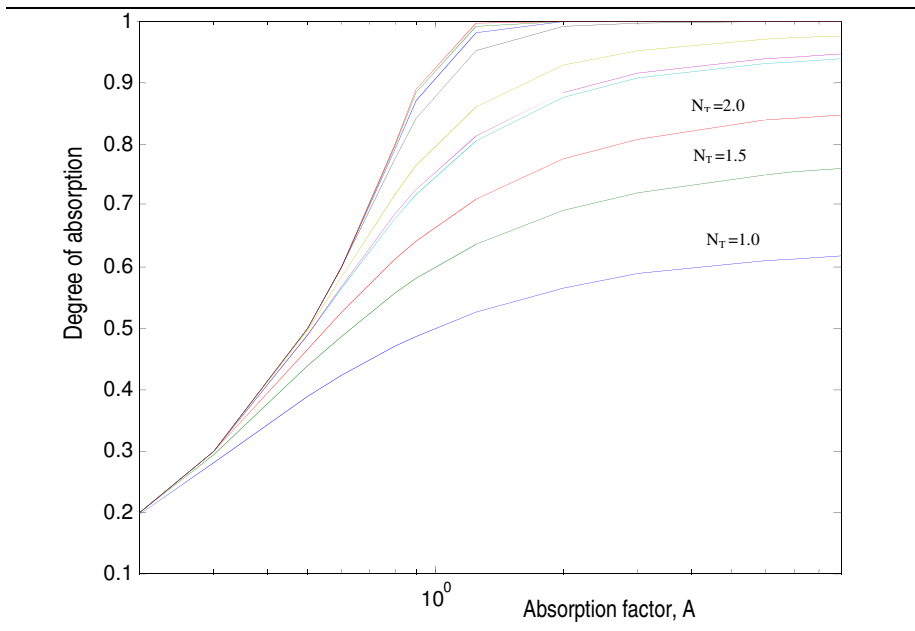
According to Ramm methodology (Ramm, 1966) the effectiveness of the operation of the absorber and calculation of number stages is performed based on the absorption coefficient φ and the absorption factor A . Correlation between these two indicators of the effectiveness of the absorption of gases and the number stages is determined by the following equation:

$$N_T = \frac{1}{1 - \frac{1}{A}} \ln \frac{1 - \varphi}{1 - \varphi A} \quad (8)$$

where N_T - number stages of the absorber in ideal mixing. If we include absolute values for parameters φ and A , stated in this paper, we obtain the following,

$$N_T = 3.14$$

FIGURE 3. DEPENDENCE BETWEEN N_T AND φ IN THE COUNTERCURRENT ABSORBER



This number is rounded to 4, which proves, using another methodology, that the absorber should contain 4 stages. The authors investigate and present further in the paper whether the effect of reversible mixing affects the number stages.

The impact of axial mixing of phases on the number stages in the absorber

The authors of this paper investigated the effect of axial (longitudinal) mixing of phases on number stages in the absorber in the absorption of vapour-gas mixture NH₃-CO₂-H₂O in the multicomponent system NaCl-H₂O.

The two-phased flow liquid-gas, as in this case the absorption of gas mixture in solution, is characterized by a suitable model based on ideal or nonideal phase flow (Levenspiel, 1962). Scientific reports recognize two well-known models of ideal phase flow- the model with ideal repressing and the model with ideal mixing. According to Mecklenburgh and Hartland (Mecklenburgh and Hartland, 1983) the flows which do not comply with ideal models- i.e. non-ideal flows, can be classified as segregational and non-segregational. The non-segregational flow is often termed as the regime of axial, i.e. longitudinal mixing. Axial mixing of the phases has a significant effect on the absorption, which is reflected in many investigations of different authors (see e.g., Miyauchi and Vermeulen, 1963; Cairns and Prausnitz, 1960; Levenspiel and Smith, 1957; Wiemann and Mewes, 2005a; 2005b).

In the investigations concerning material balance of the components in the gaseous phase for the elementary part of the absorber, Ramm (1966) has established a system of two differential equations whose solution defines the effect of longitudinal mixing on number stages in the absorber. Analytic solution of this system of equations was given by Miyauchi and Vermeulen (Miyauchi and Vermeulen, 1963). Since it is also too complex for practical usage, the authors of this paper used a simplified method of calculation which involves determining the real number stages N_p . The degree of absorption is calculated based on this method (Ramm, 1966).

The real number stages is determined according to the following formula,

$$\frac{1}{N_p} = \frac{1}{N_T} - \frac{1}{N_D} \quad (9)$$

where N_T - number stages in ideal mixing which was previously calculated in this paper, N_D -number stages that involves longitudinal mixing,

$$N_D = \frac{\ell n A}{1 - \frac{1}{A}} \Phi + Pe'_p \quad (10)$$

where Pe'_p - Pecle's pseudo criterion calculated based on the same criterion for the liquid and the gaseous phase according to the following formula:

$$\frac{1}{Pe'_p} = \frac{1}{A \cdot f_g \cdot Pe'_g} + \frac{1}{f_L \cdot Pe'_L} \quad (11)$$

Coefficients Φ , f_g and f_L in previous equations are calculated according to empirical functions,

$$\phi = 1 - \frac{0.05}{(Pe'_L)^{0.25}} \left(\frac{A}{N_T} \right)^{0.5} \quad (12)$$

$$f_g = \frac{N_T + 6.8A^{0.5}}{N_T + 6.8A^{1.5}} \quad (13)$$

$$f_L = \frac{N_T + 6.8A^{0.5}}{N_T + 6.8A^{-0.5}} \quad (14)$$

Using the previous dependences, relevant parameters which define the effect of axial (longitudinal) mixing on number stages of the absorber were calculated for the absorption of gas mixture NH₃-CO₂-H₂O in the multicomponent system NaCl-H₂O. The parameters calculated are given in Table 3.

TABLE 3. RELEVANT PARAMETERS REQUIRED FOR DETERMINING N_D, CALCULATED FOR THE GASEOUS AND LIQUID PHASE WHOSE CONCENTRATION WAS GIVEN IN TABLE 1

Molecular diffusion coefficient in the gaseous phase, D _G m ² /h	0.187
The rate of gaseous phase, w ₀ m/s	1.15
Pecle criterion, Pe' _g	154973
Molecular diffusion coefficient in the liquid phase, D _L m ² /h	0.66 · 10 ⁻⁵
The rate of the liquid phase, w ₀ m/s	0.0047
Pecle criterion, Pe' _l	179 · 10 ⁵
Non-ideal flow coefficient	
f _g	0.859
f _L	1.151
Φ	0.9995
Number stages involving longitudinal mixing, N _b	154 974

The real number stages, taking in consideration longitudinal mixing, is calculated according to the equation (9) and is the following,

$$\frac{1}{N_p} = \frac{1}{3.14} - \frac{1}{154974}$$

$$N_p = 3.14$$

It is clear that longitudinal or axial mixing does not affect the number stages in the absorber in the absorption of vapour-gas mixture NH₃-CO₂-H₂O in the multicomponent system NaCl-H₂O.

Conclusion

The aim of this study was to investigate the impact of longitudinal or axial mixing on number stages in the countercurrent absorber in the absorption of vapour-gas mixture NH₃-CO₂-H₂O in the multicomponent system NaCl-H₂O. Number stages was determined for ideal phase flow according to the model of ideal repressing N_T = 3.14, the real number stages involving longitudinal mixing is also 3.14, whereas the number stages in industrial absorbers for the same system is 4.

Therefore, based on the results of the research, it can be concluded that longitudinal mixing does not affect the number stages in the countercurrent absorber in the absorption of vapour-gas mixture NH₃-CO₂-H₂O in the multicomponent system NaCl-H₂O.

References

- Cairns, E., Prausnitz, J., 1960. "Longitudinal mixing in packed beds," *Chemical Engineering Science*, Vol.12, Issue 1, pp.20-34
- Kongto, A., Limtrakul, S. et al., 2005. "Mathematical modeling and simulation for gas-liquid reactors," *Computers and chemical engineering*, Vol.29 (11), pp.2461-2473.
- Levenspiel, O., 1962. *Chemical reaction engineering*, John Wiley and Sons, New York.
- Levenspiel, O., Smith, W., 1957. "Notes on the diffusion-type model for longitudinal mixing of fluids in flow," *Chemical Engineering Science*, Vol.6, pp.227-33.
- Mecklenburgh, J., Hartland, S., 1983. *The theory of backmixing*, John Wiley and Sons, New York
- Mhaskar, R., 1974. "Effect of backmixing on the performance of bubble column reactors," *Chemical Engineering Science*, Vol.29 (4), pp.897-905.
- Miyauchi, T., Vermeulen, T., 1963. "Longitudinal dispersion in two-phase continuous flow operations," *Industrial and Engineering Chemistry Fundamentals*, Vol.2, pp.113-126
- Ramm, V., 1966. *Absorbtion of gases (Absorbisiya gazov)*, Chemistry (Khimiya), Moscow.
- Schubert, S., Grunewald, M., Agar, D., 2001. "Enhancement of carbon dioxide absorption into aqueous methyldiethanolamine using immobilised activators," *Chemical Engineering Science*, Vol.56 (21), pp.6211-6216.
- Schultes, M., 2003. "Raschig super-ring: A new fourth generation packing offers new advantages," *Chemical Engineering Research and design*, Vol.81 (1), pp.48-57.
- Wiemann, D., Mewes, D., 2005a. "Calculation of flow fields in two and three-phase bubble columns considering mass transfer," *Chemical Engineering Science*, Vol.60, (22), pp.6085-6093.
- Wiemann, D., Mewes, D., 2005b. "Prediction of backmixing and mass transfer in bubble columns using a multifluid model," *Ind. Eng. Chem. Res.*, Vol.44 (14), pp.4959-4967.
- Zaiisev, I., Tkach, G., Stoev, N., 1986. *Soda production (Proizvodstvo sodi)*, Chemistry (Khimia), Moscow.