

*Chapter 3*

## **Chapter 3**

*Interfacial areas and gas hold-ups  
in two and three phase bubble column reactors  
at elevated pressures from 0.1 to 1.85 MPa.*

**ABSTRACT**

Interfacial areas and gas hold-ups have been determined at pressures up to 1.85 MPa in a bubble column with a diameter of 85.5 mm and for superficial gas velocities between 1 and 10 cm/s. In some experiments the bubble column has been packed with glass cylinders of 5.0 x 4.0 mm. The interfacial areas have been determined by the chemical method using the model reaction between CO<sub>2</sub> and aqueous diethanolamine (DEA) and hold-ups by observation of height differences.

The interfacial areas in the packed bubble column are unaffected by pressure. The gas hold-ups as well as the interfacial areas in the bubble column increase with increasing operating pressure. The magnitude of the pressure influence depends on the superficial gas velocity. This positive influence of pressure on the gas hold-ups and the interfacial areas in the bubble column originates from the formation of smaller bubbles at the gas distributor.

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### 1. INTRODUCTION

In this part of our study on mass transfer phenomena in gas-liquid reactors at elevated pressures we present the results of our investigations on interfacial areas and gas hold-ups in a two phase bubble column operating at pressures up to 1.7 MPa. The same experimental procedure, as used in Chapter 2 for the mechanically agitated gas-liquid reactor, is applied to the determination of the mass transfer parameters in the bubble column: the interfacial areas are determined with the chemical method using the reaction between  $\text{CO}_2$  and aqueous diethanolamine (DEA) and the gas hold-ups by taking the difference in height between the gassed and the non-gassed dispersion level respectively. In addition we present some results obtained in a three phase packed bubble column. This packed bubble column has similar geometries as the packed column used by van Gelder and Westerterp (1988) for the determination of liquid phase residence time distribution and of hold-ups at elevated pressures.

Both the mechanically agitated gas-liquid reactor and the bubble column are widely applied in process industry for contacting gases with liquids. Bubble columns are easier to operate and have lower operating costs in comparison to mechanically agitated reactors. They can be used either as gas-liquid reactors or as absorbers or strippers. Generally the backmixing in both phases will be smaller, therefore bubble columns may be preferred either for selectivity or for absorption efficiency. This is caused by the lower degree of turbulence in bubble columns also leading to larger bubbles and smaller interfacial areas per volume of liquid in comparison to a mechanically agitated reactor. For operation at elevated pressures bubble columns have the additional advantage of a simple construction and no sealing problems.

A considerable amount of information is available on the hydrodynamics and mass transfer characteristics in bubble columns. Comprehensive reviews are given by Van Landeghem (1980) and Shah et al (1982). These reviews are restricted to research at atmospheric pressure, nothing is said of a possible influence of the operating pressure on the hydrodynamics and mass transfer parameters in bubble columns. This despite the fact that it has been known since the late thirties that some high pressure bubble columns operate at extremely high gas hold-ups, see Tarmy et al (1984a, 1984b).

The upward gas flow in bubble columns is normally characterized into three separate flow regimes, which occur in order of increasing gas flow rate; the bubbly flow, the churn turbulent and the slug flow regime respectively, see Shah et al (1982). In the bubbly flow regime, which occurs at superficial gas velocities below 5 cm/s, the bubble diameters are more or less uniform and the rising velocity of the bubbles lies between 0.18 and 0.30 m/s. At higher superficial gas velocities an unsteady flow pattern occurs and this flow regime, the churn turbulent regime, is characterized by large bubbles moving with high rising velocities in the presence of small bubbles. Increasing the superficial gas velocity further will lead to the formation of very large bubbles stabilized by the reactor walls, the slug flow regime.

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In Chapter 1 we concluded that in pressurized gas-liquid systems the initial bubble size at a single orifice decreases with increasing pressure. Several authors, see Kling(1962) and LaNauze and Harris(1974), attributed the reduction in size of the bubbles at higher pressures to the increased contribution of the momentum or the kinetic energy to the bubble formation process. Sagert and Quinn(1976) observed for some gases an increase in the coalescence time of the gas bubbles when they increased the pressure in their gas-liquid systems. Both phenomena, initially smaller bubbles and larger coalescence times, result in bubbles with lower rising velocities and give rise to larger interfacial areas and gas hold-ups in gas-liquid contactors.

In Chapter 2 we found no influence of pressure on the interfacial areas and the gas hold-ups in a mechanically agitated gas-liquid reactor. This may be due to the dominating influence of the agitation on the coalescence-breakup process of the bubbles in this reactor. Such an influence is absent in a bubble column, where the bubble diameters and gas hold-ups are largely determined by the type of gas distributor and the coalescence-breakup processes in the column. Therefore, if pressure affects the formation of the bubbles or the coalescence-breakup process, an influence of pressure on the interfacial area and the gas hold-up can be expected in a bubble column.

All literature data on mass transfer parameters in bubble columns at elevated pressures are summarized in Table 4 of Chapter 1. In Fig. 7 of Chapter 1 it has been shown that a twofold increase in the gas hold-up with a pressure increase from 0.1 to 2.0 MPa can be obtained. However, there are also studies in a bubble column known in which no influence of pressure on the gas hold-up and the bubble diameters has been found. A remarkable point is that all the studies, which report no influence of pressure, were executed in bubble columns with porous plates as gas distributors and with low superficial gas velocities between 0.1 and 3.5 cm/s.

Idogawa et al(1986) reported that with increasing pressure at a single orifice the critical gas velocity separating the multiple bubbling and jetting regions decreases. Such a kind of change was also observed for the flow regimes in a bubble column by Tarmy et al(1984a, 1984b) and Teurlings et al(1986). These authors reported that at higher pressures the transition from the bubbly flow towards the churn-turbulent regime occurs at higher gas velocities and higher gas hold-ups. Tarmy et al(1984a, 1984b) remarked that the persistence of the bubbly flow regime at higher gas hold-ups can be caused by the presence of very small, non-coalescing bubbles at higher pressures. This combination of larger gas hold-ups and smaller bubble diameters may lead to considerably larger interfacial areas at elevated pressures.

Only Idogawa et al(1985) determined simultaneously gas hold-ups and bubble diameters in a bubble column at elevated pressures. For air-water at a constant superficial gas velocity of 1 cm/s their correlations predict an increase in the gas hold-up of a factor 2.7 and a decrease in the average bubble diameter with a factor 0.6, when the pressure is increased from 0.1 up to 5 MPa: this results in an increase with a factor 4.5 in the interfacial area. Such an extreme large influence of the operating pressure on the interfacial area should be a strong incentive for further research of this phenomenon in bubble columns.

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### 2. EXPERIMENTAL

#### 2.1. Experimental method.

The interfacial areas are determined by the chemical method using the model reaction between  $\text{CO}_2$  and aqueous diethanolamine (DEA). We showed in Chapter 2 already that the absorption rate, as derived by Danckwerts (1970) for the penetration theory, becomes equal to

$$\phi_{\text{CO}_2} = m k_L a V_R E_A \Delta c_{\text{CO}_2, \text{G}} = m k_L a V_R \sqrt{1 + \text{Ha}^2} \Delta c_{\text{CO}_2, \text{G}} \quad (1)$$

If the conditions for the pseudo-first order reaction regime

$$2 < \text{Ha} \ll E_{A, \infty} \quad (2)$$

with

$$\text{Ha} = \frac{\sqrt{k_{1,p} c_{\text{BL}}^p D_A}}{k_L} \quad (3)$$

and

$$E_{A, \infty} = \sqrt{\frac{D_A}{D_B}} + \sqrt{\frac{D_B}{D_A}} \frac{c_{\text{BL}}}{m c_{A, \text{G}} v_B} \quad (4)$$

are met, the interfacial areas can be obtained from absorption rate measurements without knowledge of the exact value of the liquid phase mass transfer coefficient  $k_L$ .

The gas phase in the bubble column reactor can be assumed to be in plug flow and for this case the average driving force for mass transfer,  $\Delta c_{\text{CO}_2, \text{G}}$ , becomes equal to

$$\Delta c_{\text{CO}_2, \text{G}} = \frac{c_{\text{CO}_2, \text{G}, \text{in}} - c_{\text{CO}_2, \text{G}, \text{out}}}{\ln \frac{c_{\text{CO}_2, \text{G}, \text{in}}}{c_{\text{CO}_2, \text{G}, \text{out}}}} \quad (5)$$

This results in the following equation from which the interfacial areas in the bubble column can be calculated

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$$a = \frac{\phi_V \ln \frac{c_{\text{CO}_2, \text{G.in}}}{c_{\text{CO}_2, \text{G.out}}}}{m V_R \sqrt{k_{1,p} c_{\text{DEAL}}^p D_{\text{CO}_2}}} \quad (6)$$

The same commercial grade aqueous DEA solutions as in Chapter 2 are used for the determination of interfacial areas in the bubble column. An empirical correlation for the absorption rate constant  $m (k_{1,p} (c_{\text{BL}})^p D_{\text{CO}_2})^{0.5}$  at 298 K in the gas-liquid reaction system as a function of the free DEA concentration (0.2 - 2.0 mol/kg) is given as Eq. (7) in Table 1. Experimentally determined correlations for the solubility  $m$ , the viscosity  $\mu_L$  and the density  $\rho_L$  as a function of the total DEA concentration are also given as Eqs (8), (9) and (10) in the same Table.

For each absorption experiment the  $\text{CO}_2$  concentrations in the in- and outlet gas flows, the DEA concentration and the  $\text{CO}_2$ -liquid load in the inlet liquid flow are separately determined. The total DEA concentration is corrected for the  $\text{CO}_2$ -liquid load in the feed and for the amount of  $\text{CO}_2$  absorbed from the gas phase, as part of the DEA in the feed and in the reaction mixture itself is already converted. The resulting free DEA concentration is used for the calculation of the absorption rate constant from Eq. (7) and then for each absorption experiment the interfacial areas are calculated from Eq. (6).

Table 1. Data for the DEA-water system at 298 K.

$$m \sqrt{k_{1,p} c_{\text{DEAL}}^p D_{\text{CO}_2}} = 9.39 \cdot 10^{-4} [\text{DEA}]_{\text{free}}^{0.38} - 2.61 \cdot 10^{-4} \quad \text{m/s} \quad (7)$$

$$m = .791 - .044 [\text{DEA}] \quad (8)$$

$$\rho_L = 995.8 + 15.55 [\text{DEA}] - 1.141 [\text{DEA}]^2 \quad \text{kg/m}^3 \quad (9)$$

$$\mu_L = 10^{-3} \exp(-.1135 + .257 [\text{DEA}] + 4.694 \cdot 10^{-2} [\text{DEA}]^2) \quad \text{Ns/m}^2 \quad (10)$$

[DEA] in mol/kg and [DEA]\* in mol/l

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#### 2.2. Experimental installation and procedure.

The same experimental installation as described in detail in Chapter 2 is used for the absorption experiments in the bubble column reactors. Only the mechanically agitated reactor is replaced by a column also made of glass and manufactured by SFS (Zürich), see Fig. 1. It has a diameter of  $D = 85.5$  mm and a total height equal to 63.0 cm. The column can be operated at pressures up to 2.0 MPa and is thermostatted at a constant temperature of 298 K. The column is operated continuously with respect to both the gas and the liquid phase. A gas mixture of  $\text{CO}_2$  and  $\text{N}_2$  is introduced into the column via a gas distributor located at the bottom. A liquid inlet tube is installed at the top of the column just below the dispersion level, resulting in a countercurrent flow of gas and liquid through the column. The dispersion level is kept at a constant height of  $H/D = 6.1$ . In order to obtain a uniform bubble flow through the bubble column we have used in most of the experiments a cross-shaped gas distributor with 16 holes of 0.5 mm, see Fig. 1.

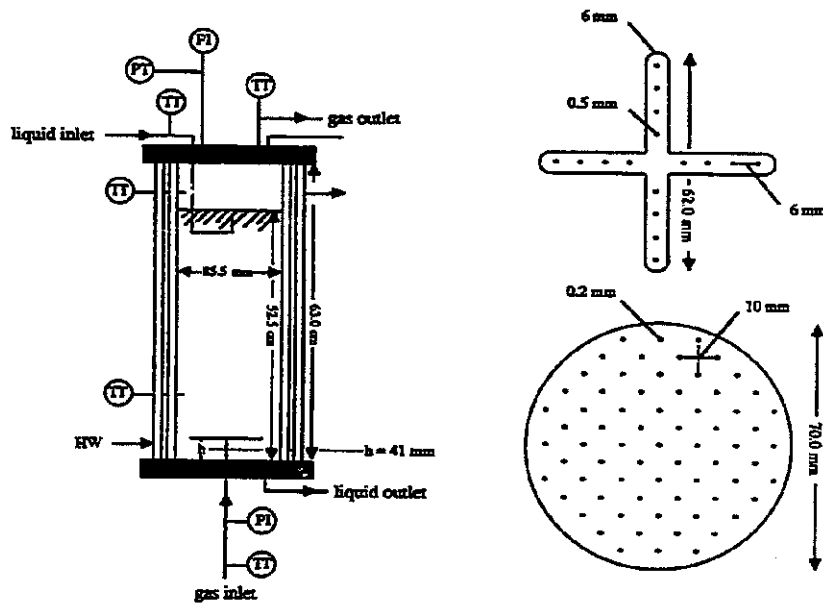


Figure 1: The experimental bubble column reactor with the two gas distributors.

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For the experiments with a packed bubble column we have filled the column with glass cylinders with a length and a diameter equal to 5.0 and 4.0 mm respectively. The density of the packing material is equal to  $2200 \text{ kg/m}^3$  column and the bed porosity equals 0.35. The packed column also operates continuously with respect to both the gas and liquid phase and the phases flow countercurrently. The cross-shaped gas distributor in the packed bubble column leads to serious maldistributions of the gas. Therefore, we have replaced it by a circular distributor with 69 holes of 0.2 mm for the experiments with the packed bubble column, see also Fig. 1.

The same experimental procedure as described previously, in Chapter 2 is followed for all the absorption measurements. Special care is taken to keep the DEA conversion below 20%. The gas hold-ups are determined by measuring the height difference between the gassed and non-gassed dispersion level. As the gas distributors are 41 mm high we have subtracted 41 mm from the dispersion height. So all the reported specific interfacial areas and gas hold-ups are based on the true reaction volume above the gas distributor.

#### 2.3. Liquid phase mass transfer coefficients and experimental conditions.

As stated before, we do not need to know the exact value of  $k_L$  for the determination of the interfacial area if the conditions of Eq. (2) for the pseudo-first order reaction regime are fulfilled. However, a rough idea of the magnitude of  $k_L$  and its dependence on the operating conditions is necessary in order to know whether  $Ha$  is larger than 2. This allows us to estimate the DEA concentrations in the liquid phase and  $\text{CO}_2$  concentrations in the gas phase at which the conditions of Eq. (2) are met, so that the interfacial areas can be determined. In Chapter 1 it has been concluded that the liquid phase mass transfer coefficient in gas-liquid reactors is not affected by the operating pressure. Therefore,  $k_L$  can be determined at atmospheric conditions and the results can be used for the choice of the experimental conditions at elevated pressures.

By means of the Danckwerts plot method we have determined  $k_L$  in a series of absorption experiments at one constant pressure of  $P = 0.15 \text{ MPa}$  and at four different superficial gas velocities of  $v_G = 1.0, 2.0, 4.0$  and  $6.0 \text{ cm/s}$  respectively. This method is based on a form of Eq. (1) rewritten to

$$\left[ \frac{\phi_{\text{CO}_2}}{V_R m \Delta c_{\text{CO}_2, G}} \right]^2 = (k_L a)^2 + a^2 k_{1,p} c_{\text{DEAL}}^p D_{\text{CO}_2} \quad (11)$$

By plotting the left hand side of Eq. (11) versus  $k_{1,p} (c_{\text{DEAL}})^p D_{\text{CO}_2}$  the values for  $k_L$  and  $a$  can be obtained by combination of the slope and the intercept on the vertical axis of the straight line through the experimental results.



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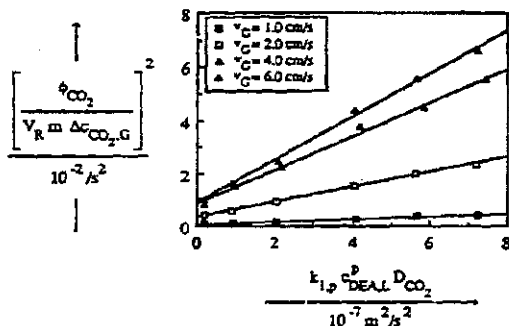


Figure 2: Danckwerts plots for the bubble column at a pressure of 0.15 MPa.

The absorption experiments were performed in six different aqueous DEA solutions with DEA concentrations between 0.10 and 1.10 mol/kg. This results in a fortyfold variation in  $k_{L,p} (C_{DEAL})^D D_{CO_2}$ . The Danckwerts plots as obtained from these experiments are shown in Fig. 2. A major drawback of this method is the variation in the liquid phase properties of the different solutions, which may affect the hydrodynamics in the reactor. This may give rise to experimental errors, especially in  $CO_2$ -amine systems in which the variation in the reaction rate can only be realised by changing the amine concentrations and not by the addition of a catalyst. However, the tenfold change in the DEA concentration from 0.10 to 1.10 mol/kg causes only a minor change in the liquid phase viscosity from  $0.92 \cdot 10^{-3}$  to  $1.25 \cdot 10^{-3}$  N s/m<sup>2</sup> and can probably be neglected. This is confirmed by the Danckwerts plots in Fig. 2, which show straight lines over the whole range of operating conditions and indicate that  $k_L$  and  $a$  are constant within the applied range of viscosities.

The values of  $k_L$  and  $a$ , which are obtained from Fig. 2, are summarized in Table 2. It can be seen that  $k_L$  is practically constant over the range of superficial gas velocities applied: it has an average value equal to  $3.5 \cdot 10^{-4}$  m/s. The same independence of the liquid phase mass transfer coefficient has been reported recently by Schumpe et al(1987). These authors have calculated values for  $k_L$  from individual results on  $a$  and  $k_L a$ , as obtained in a bubble column with a diameter of  $D = 9.1$  cm with the gas-liquid system  $O_2$ -aqueous sodium sulphite(0.8 N  $Na_2SO_3$ ). They have found a constant  $k_L$  value of about  $3.2 \cdot 10^{-4}$  m/s, which agrees rather good with the average  $k_L$  from this study.

The interfacial areas increase with an increasing superficial gas velocity up to 4 cm/s, but above that start to decrease. The decrease in the interfacial area indicates a transition from bubbly flow to churn-turbulent flow regime around  $v_G = 5$  cm/s, which is likely to occur around this superficial gas velocity according to Shah et al(1982). A decrease of the interfacial area after the transition from one

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Table 2. Results of the Danckwerts plots for the bubble column at a pressure of 0.15 MPa.

Data: [ DEA ] = 0.10 - 1.10 mol/kg.

$v_G \cdot 10^2$ m/s	$k_L \cdot 10^4$ m/s	a 1/m
1.0	3.54	71
2.0	3.38	170
4.0	3.39	283
6.0	3.78	251

flow regime to another has already been reported by Schumpe and Deckwer(1980) and will also be discussed furtheron in this study.

A comparison with literature correlations, as given for the liquid phase mass transfer coefficients in bubble columns by Calderbank and Moo-Young(1961) and Akita and Yoshida(1974), is made in Fig. 3. The data for  $k_L$  from the present work coincide with the relation of Calderbank and Moo-Young(1961) and differ a factor 2.5 from the relation of Akita and Yoshida(1974). Therefore, it can be concluded that the use of an average value of  $k_L = 3.5 \cdot 10^{-4}$  m/s leads to conservative values of the DEA concentration at which the condition  $Ha > 2$  in Eq. (2) is fulfilled.

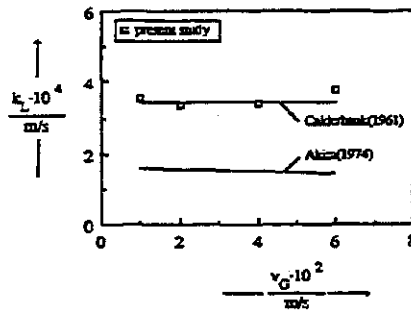


Figure 3: Comparison between literature correlations for  $k_L$  and data from the present study.

From Eqs (7) and (8) it can be calculated that with  $k_L = 3.5 \cdot 10^{-4}$  m/s the free DEA concentration should be equal to 1.6 mol/kg to obtain a Hatta number of 3.5. Together with the experimental condition of a maximum DEA conversion of 20% this leads to a minimum value for the total DEA concentration equal to 2.0 mol/kg for the amine solutions with which the interfacial areas in the bubble column can be determined. All interfacial areas for the bubble column will be determined with amine solutions with a total DEA concentration between 2.0 and 2.2 mol/kg. The  $\text{CO}_2$ -fractions, as supplied in the gas feed, are the same as in the mechanically agitated reactor and range from 1 to 2%. For values of  $D_{\text{DEA}}/D_{\text{CO}_2} = 0.5$  and of  $m = 0.7 - 0.8$  at a temperature of 298 K this results at all pressures in values of  $E_{A_m} > 60$ .

### 3. RESULTS

#### 3.1. Gas hold-ups in the bubble column.

##### 3.1.1. Experimental results.

Gas hold-up measurements have been performed with nitrogen in distilled water and in an aqueous DEA solution partially loaded with  $\text{CO}_2$  having a  $c_{\text{DEA,L}} = 2.16$  mol/kg and  $c_{\text{CO}_2\text{L}} = 0.190$  mol/kg. The results are plotted in Figs 4 and 5. The operating pressures of these measurements range from 0.15 to 1.85 MPa and the superficial gas velocities vary between 1.0 and 9.0 cm/s. Because of experimental restrictions it is not possible to work below pressures of 0.15 MPa and to use gas flows larger than  $2.0 \cdot 10^{-3}$   $\text{m}^3/\text{s}$ . The gas hold-ups can not be determined with sufficient accuracy at gas velocities above 10 cm/s and the reported results are therefore also restricted to gas velocities below 10 cm/s.

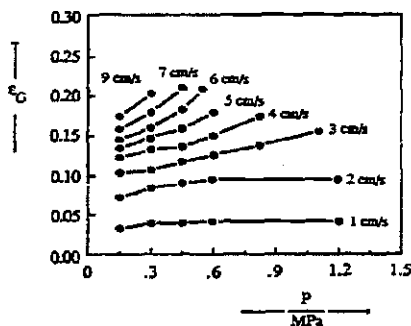


Figure 4: Gas holdup in water vs. the reactor pressure in the bubble column.

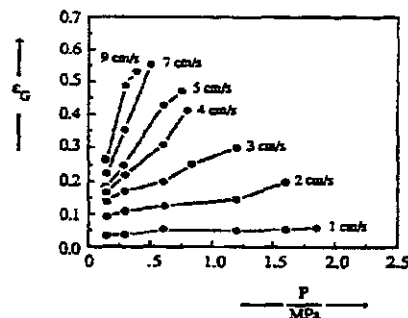


Figure 5: Gas holdup in a DEA solution partially loaded with  $\text{CO}_2$  vs. the reactor pressure in the bubble column.

It is evident from Figs 4 and 5 that there is a positive influence of the operating pressure on the gas hold-up. Both Figures show an increase in the gas hold-up with increasing pressure. It can be seen that both the relative and the absolute increase in the gas hold-up are larger in the aqueous DEA solution than in water. For water the relative increase of the gas hold-up can be calculated to be independent of the superficial gas velocity applied. For a pressure increase from 0.15 to 0.30 MPa it amounts to 10 - 20% and for an increase from 0.15 to 0.60 MPa it amounts at all gas velocities to 20 - 30%. Contrarily to this, for the aqueous DEA solution the relative increase of the gas hold-up depends on the gas velocities. For a pressure increase from 0.15 to 0.30 MPa it ranges from 20% at low superficial gas velocities up to 90% at high superficial gas velocities. Comparison of the gas hold-ups in the aqueous DEA solution at 0.15 and 0.60 MPa shows a relative increase of the gas hold-up ranging from 40 up to 150%.

### 3.1.2. Comparison with literature correlations.

The correlations of Akita and Yoshida(1973) and Hikita et al(1980) are recommended by Shah et al(1982) for the calculation of gas hold-ups in bubble columns with single or multi-nozzle gas distributors. These correlations - given in Table 3 - are based on atmospheric measurements. Hikita et al(1980) have used different gases with different densities to account for the effect of the gas density on the gas hold-up. It has been argued by Fijis et al(1985) that the effect of the operating pressure on the gas hold-up cannot be simulated by using gases of different densities. Also our results show a much larger dependence on the gas density than as suggested by the exponent 0.062 in the relation of Hikita et al(1980).

A comparison between the gas hold-ups at 0.15 MPa from this study and the relations in Table 3 is given in Fig. 6. It can be seen that the gas hold-ups in water agree fairly well with the correlations, whereas those in the aqueous DEA solution differ substantially compared to the gas hold-ups as predicted by Akita and Yoshida(1973) and Hikita et al(1980). Recently, also Öztürk et al(1987) have reported gas hold-ups in mixtures of liquids to be sometimes higher than as calculated from the aforementioned literature correlations, which have been derived on the basis of measurements in water and in pure organic liquids.

A comparison between our experimental gas hold-ups in water and those predicted by the correlation for gas hold-ups in a bubble column at elevated pressures, as presented by Idogawa et al(1985)

$$\frac{\epsilon_G}{(1 - \epsilon_G)} = 1.44 v_G^{0.58} \rho_G^{0.12} \sigma_L^{-0.16} \exp(-P) \quad (12)$$

SI units except  $P$  in MPa and  $\sigma_L$  in mN/m

is made in Fig. 7. This parity plot shows that the relation of Idogawa et al(1985) predicts too high

Table 3. Literature correlations on the gas holdup and the interfacial areas in bubble columns.

$$\frac{\epsilon_G}{(1-\epsilon_G)^4} = 0.20 \left[ \frac{gD^2 \rho_L}{\sigma_L} \right]^{0.125} \left[ \frac{gD^3}{v_L^2} \right]^{0.083} \left[ \frac{v_G}{\sqrt{gD}} \right] \quad \text{Akita and Yoshida(1973)}$$

$$aD = \frac{1}{3} \left[ \frac{gD^2 \rho_L}{\sigma_L} \right]^{0.50} \left[ \frac{gD^3}{v_L^2} \right]^{0.10} \epsilon_G^{1.13} \quad \text{Akita and Yoshida(1974)}$$

$$\epsilon_G = 0.672 f \left[ \frac{v_G \mu_L}{\sigma_L} \right]^{0.578} \left[ \frac{\mu_L^4 g}{\rho_L \sigma_L^3} \right]^{-0.131} \left[ \frac{\rho_G}{\rho_L} \right]^{0.062} \left[ \frac{\mu_G}{\mu_L} \right]^{0.107} \quad \text{Hikita et al(1980)}$$

$f = 1.0$  for non-electrolyte solutions

$f = 10^{0.0414I}$  for  $0 < I < 1.0 \text{ kg ion/m}^3$  and  $f = 1.1$  for  $I > 1.0 \text{ kg ion/m}^3$

$I =$  ionic strength of the solution

Akita and Yoshida investigated column diameters in the range of 0.15 to 0.6 m

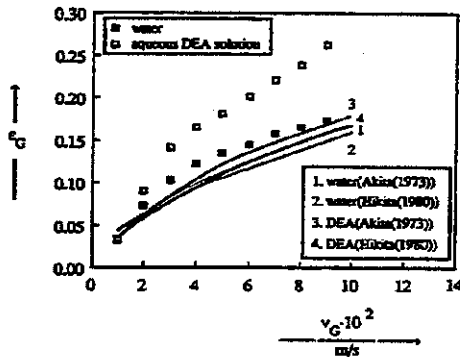


Figure 6: Comparison between literature correlations for  $\epsilon_G$  and the data of the present study at  $P = 0.15 \text{ MPa}$ .

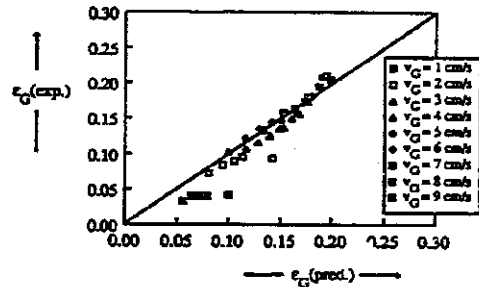


Figure 7: Parity plot between the experimental gas holdups in water and those predicted by the correlation of Ito et al (1985).

gas hold-ups in water at the low gas velocities of 1 and 2 cm/s respectively, but for gas velocities above 5 cm/s the agreement is satisfactory at all pressures. Contrarily to these findings for water, we have found that the relation of Ilogawa et al(1985) cannot be used to predict either the absolute gas hold-ups in the aqueous DEA solution or the dependence of the gas hold-up on the pressure and the superficial gas velocity respectively.

### 3.1.3. Influence of pressure on the flow regimes.

Tarmy et al(1984a, 1984b) have also found a positive influence of the operating pressure on the gas hold-up. Furthermore, these authors observe that at higher pressures the transition from the bubbly flow towards the churn-turbulent regime occurs at higher gas velocities and gas hold-ups. This transition in flow regime is normally evaluated in a plot of the drift flux  $q_d$  versus the gas hold-up, see Wallis(1969). The drift flux  $q_d$  is defined as the volumetric gas flux relative to a surface moving at the average velocity of the dispersion. For a semi-batch system with a non-flowing liquid the drift flux is equal to

$$q_d = v_G (1 - \epsilon_G) \quad (13)$$

Plots for the drift flux in water and in the aqueous DEA solution are presented in Figs 8 and 9. At 0.15 MPa the drift flux data in both liquids can be divided into two regions. In the first region corresponding to the bubbly flow regime the drift flux is relatively low and increases slowly with increasing gas hold-up. The second region corresponds to the churn-turbulent regime where the drift flux is larger and increases rapidly with increasing gas hold-up. In both liquids at a pressure of 0.15 MPa the transition from the bubbly flow regime towards the churn-turbulent regime occurs around  $\epsilon_G = 0.12 - 0.15$ . This is in accordance with literature data on bubble columns with multi-nozzle gas distributors at atmospheric pressure, see Shah et al(1982).

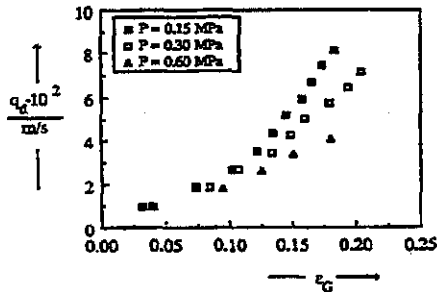


Figure 8: Drift flux vs. the gas holdup in water in the bubblic column.

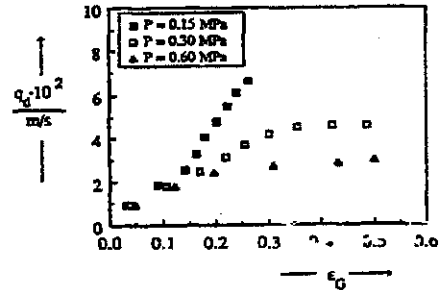


Figure 9: Drift flux vs. the gas holdup in a DEA solution partially loaded with  $\text{CO}_2$  in the bubblic column.

Also it can be seen in Figs 8 and 9 that both the absolute values of  $q_d$  as well as the increase in  $q_d$  with  $\epsilon_G$  are smaller at the higher pressures of 0.30 and 0.60 MPa. This indicates that at higher pressures the bubbly flow regime persists at higher gas hold-ups and gas velocities, which agrees with the findings of Tarmy et al(1984a, 1984b).

The higher gas hold-ups at elevated pressures can be attributed to the initially smaller bubbles at the gas distributor as concluded in Chapter 1. These smaller bubbles have lower rising velocities, which lead to higher gas hold-ups at constant superficial gas velocity. The persistence of the bubbly flow regime at these higher gas hold-ups also indicates that these smaller bubbles may be noncoalescing and therefore keep their identity throughout the whole column.

In a series of additional experiments we have determined gas hold-ups in water in the bubble column with a lower dispersion height of 27.0 cm ( $H/D = 3.2$ ). The results of these measurements are presented in Fig. 10. A comparison with the gas hold-ups of Fig. 4 in water at  $H/D = 6.1$  shows that at  $H/D = 3.2$  the gas hold-ups are higher, except at  $v_G = 1$  cm/s. This means that at higher gas velocities and thus at higher gas hold-ups the bubbles still coalesce in the upper part of the column. Also the influence of the operating pressure on the gas hold-up is stronger at  $H/D = 3.2$  even for  $v_G = 1$  cm/s. For an increase in the pressure from 0.15 to 0.30 MPa the relative increase in the gas hold-up ranges from 20% at  $v_G = 1$  cm/s up to 40% at  $v_G = 9$  cm/s. These results are a strong indication that the positive influence of the operating pressure on the gas hold-up originates from the formation of smaller bubbles at the gas distributor. Both the increase in the gas hold-up as well as the decrease in the bubble diameters cause an increase in the interfacial area with pressure as will be shown in the next part.

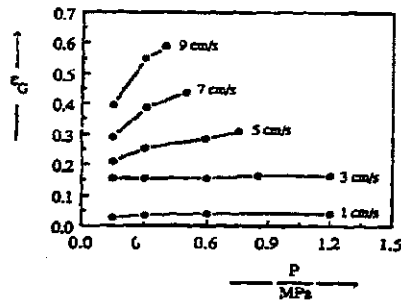


Figure 10: Gas holdup in water vs. the reactor pressure in the bubble column and with  $H/D = 3.2$ .

### 3.2. Interfacial areas in the bubble column.

#### 3.2.1. Experimental results.

The operating pressures and the superficial gas velocities, which are applied in the absorption experiments for the determination of interfacial areas, are the same as in the gas hold-up experiments. The results are given in Table 4 and partially plotted in Fig. 11. It can be seen from this Figure that also the interfacial areas are influenced by the operating pressure and increase with increasing pressure for all the superficial gas velocities. The relative increase of the interfacial areas is the largest for a pressure increase from 0.15 to 0.30 MPa and varies between 10 and 30%. At the higher superficial gas velocities it seems that the interfacial areas approach a maximum value at a pressure of 0.5 MPa.

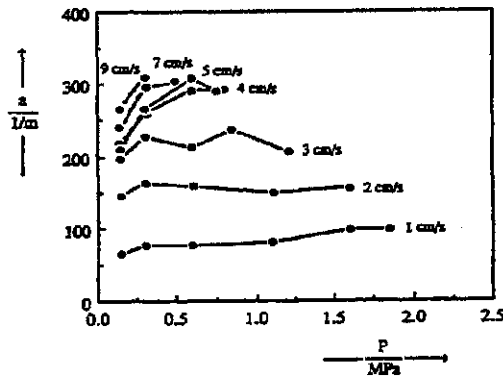


Figure 11: Interfacial area in a DEA solution partially loaded with  $\text{CO}_2$  vs. the reactor pressure in the bubble column.

The influence of pressure on the interfacial areas is smaller than on the gas hold-ups in an aqueous DEA solution of similar composition, compare Figs 5 and 10. As the Sauter mean bubble diameter  $d_{mb}$  is related to the gas hold-up and the interfacial area according to

$$d_{mb} = \frac{6\epsilon_G}{a} \quad (14)$$

this means that the bubble diameters increase with increasing pressure, see Table 4. However, such an increase in the bubble diameters with pressure has not been observed visually and is completely in contrast with all the literature results up till now, see Chapter 1. Besides, larger bubbles should



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Table 4. Interfacial areas and Sauter mean bubble diameters in the bubble column at different pressures.

Data: [ DEA ] = 2.0 - 2.2 mol/kg.

$v_G \cdot 10^2$	P	$\zeta_{CO_2}$	a	$d_{mb} \cdot 10^3$
m/s	MPa	%	1/m	m
1.0	0.15	96	66	3.1
	0.30	97	77	3.0
	0.60	97	77	3.8
	1.10	98	80	3.7
	1.60	99	98	3.3
	1.85	99	97	3.5
2.0	0.15	96	147	3.7
	0.30	98	163	3.9
	0.60	98	158	4.6
	1.10	97	150	5.6
	1.60	98	156	7.6
3.0	0.15	95	198	4.3
	0.30	97	228	4.5
	0.60	96	212	5.6
	0.85	98	238	6.3
	1.20	96	207	8.8
4.0	0.15	93	221	4.5
	0.30	96	263	5.0
	0.60	97	294	6.3
	0.80	97	292	8.6
5.0	0.15	86	211	5.1
	0.30	92	266	5.7
	0.60	95	309	8.4
	0.75	94	291	9.8
6.0	0.15	82	217	5.6
	0.30	89	286	6.3
	0.60	91	296	10.2
7.0	0.15	80	240	5.5
	0.30	88	314	6.8
	0.50	87	304	10.9
8.0	0.15	79	263	5.5
	0.30	84	305	8.3
9.0	0.15	76	268	5.9
	0.30	81	312	9.4

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also lead to lower gas hold-ups and interfacial areas, which is in contrast with the experimental results on  $\epsilon_G$ . Therefore, we can ask ourselves whether either the experimental results on the gas hold-ups are too high or the absorption experiments result in an underestimation of the interfacial area.

#### 3.2.2. Underestimation of the interfacial areas.

The use of Eq. (6) for the evaluation of interfacial areas from absorption experiments is in fact restricted to an ideally micromixed gas phase in plug flow. However, the gas phase in the bubble column is dispersed and subjected to finite coalescence and breakup rates. Moreover, the bubble sizes are not uniform. Schumpe and Deckwer(1980) have shown that in gas-liquid dispersions with non-uniform bubble sizes the chemical method can lead to an underestimation of the interfacial area. This underestimation is a direct result of the use of an overall conversion for the reactant in the gas phase, while the conversion of the gas phase reactant in a single bubble depends non-linearly on the specific area of the bubble. Besides that, the bubbles may have different residence times  $\tau_b$ . These two phenomena lead to deviations of the chemically determined interfacial area  $a_{chem}$  from the true geometric interfacial area  $a_{geo}$ . These deviations become larger the broader the distribution of  $\tau_b/d_b$ , and the higher the overall conversion of the reactant in the gas phase, see Schumpe and Deckwer(1980).

In Chapter 5 we modified the method of Schumpe and Deckwer(1980) in order to calculate  $a_{chem}/a_{geo}$  as a function of the overall conversion  $\Omega_A$  of the gas phase reactant and the bubble size distribution. Some calculated results for a bubble column and a first order reaction with respect to the gas phase reactant, being the reaction between  $CO_2$  and DEA in water, are given in Fig. 12. Bubble size distributions equal to a log-normal distribution with variances of  $\sigma_{ln}^2 = 0.10$  and  $0.20$ .

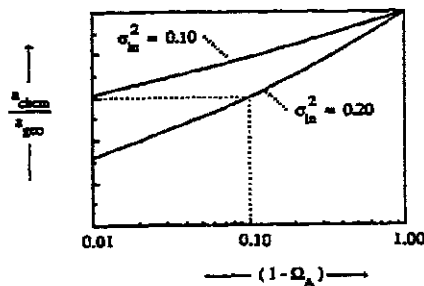


Figure 12: Ratio  $a_{chem}/a_{geo}$  for a log-normal distribution vs.  $(1 - \Omega_A)$  for a first order reaction and plug flow of the gas phase.

as found to be valid for a bubble column with single or multi-nozzle gas distributors by Akita and Yoshida(1974), have been used for the calculations. All bubbles are assumed to have the same residence time  $\tau_b$  and to move separately and completely segregated. Overall the gas phase is assumed to be in plug flow. The geometric interfacial area  $a_{geo}$  can be obtained from the Sauter mean bubble diameter, whereas the chemically determined interfacial area  $a_{chem}$  is evaluated from the overall conversion  $\Omega_A$  as obtained by a summation of the conversions per class of bubble diameters, see further Chapter 5.

It can be seen in Fig. 12 that  $a_{chem}/a_{geo}$  decreases with increasing overall conversions  $\Omega_A$  and variances  $\sigma_{ln}^2$ . The error in  $a_{chem}$  becomes larger than 20% for overall conversions  $\Omega_A$  larger than 0.90 and amounts about 35% at  $\Omega_A = 0.99$  and  $\sigma_{ln}^2 = 0.20$ . However, in Chapter 5 it is also reported that, if realistic coalescence and breakup rates between the bubbles are taken into account, the error in  $a_{chem}$  will reduce considerably and becomes less than 20% for overall conversions  $\Omega_A$  lower than 0.99, see also Fig. 12. Therefore, it can be concluded that the use of Eq. (6) for the evaluation of interfacial areas is justified and does not lead to the large underestimations of the interfacial area, which could account for the increase of the Sauter mean bubble diameter with pressure, as shown in Table 4.

Because also the absorption rate constants in  $(k_{1,p} (c_{DEAL})^p D_{CO_2})^{0.5}$  are known within an experimental accuracy of 5%, this means that part of the experimentally determined gas hold-ups does not contribute to the mass transfer of  $CO_2$ . This is probably caused by two phenomena, which have been observed visually in the aqueous DEA solutions at atmospheric and at elevated pressures:

1. The formation of a froth layer in the upper part of the dispersion. This froth layer is more pronounced at higher superficial gas velocities and is less effective in the mass transfer of  $CO_2$ .
2. The occurrence of tiny bubbles with diameters smaller than 1 mm and with long residence times. These tiny bubbles accumulate in the dispersion and are in equilibrium with the liquid phase. They do not contribute to the mass transfer of  $CO_2$ .

These two phenomena have also been observed in aqueous glycol mixtures by Bach and Pilhofer(1978) and in organic liquid mixtures Özurk et al(1987). The tiny bubbles originate from the gas distributor and are more pronounced at high kinetic energy contents of the gas phase, thus at higher pressures and at higher gas velocities, see Idogawa et al(1986). These findings indicate that part of the dispersed gas phase does not contribute or contributes less to the mass transfer of  $CO_2$ . However, sad to say it is not possible to determine the hold-up of these tiny bubbles within the experimental bubble column separately from the other bubbles.

### 3.2.3. Comparison with literature correlations.

A comparison between the interfacial areas at 0.15 MPa from this study and those calculated with the correlation of Akita and Yoshida(1974), see Table 3, is presented in Fig. 13. It can be seen that the interfacial areas as predicted by the relation of Akita and Yoshida(1974) are around a factor 2

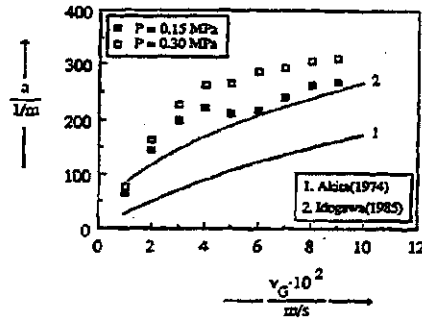


Figure 13: Comparison between literature correlations for interfacial areas and data of the present study.

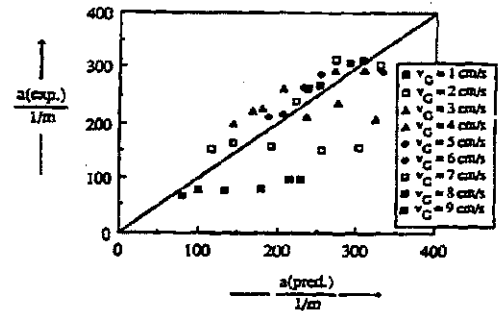


Figure 14: Parity plot between the experimental interfacial area and those predicted by the correlations of Idogawa et al(1985).

lower than the experimental values of this study. It is known that this relation of results in a conservative value for the interfacial areas, see Shah et al(1982). A much better agreement can be obtained with literature data on interfacial areas in sulfite solutions in bubble columns, see Schumpe and Deckwer(1980, 1982). These authors summarize data on interfacial areas, which have been presented by several investigators for different bubble columns over a wide range of operating conditions. The comparison between these literature data and the interfacial areas in aqueous DEA at 0.15 MPa is good. Therefore, it can be concluded that the interfacial areas at atmospheric pressures as determined by us are realistic.

The line representing the interfacial areas at 0.15 MPa as can be calculated from the relations of Idogawa et al(1985) for the gas hold-up, see Eq. (12), and for the Sauter mean bubble diameter

$$d_{mb} = 3.10 \cdot 10^{-3} \rho_L^{-0.028} \sigma_L^{0.088} \exp(-P) \quad (15)$$

SI units except P in MPa and  $\sigma_L$  in mN/m

is also plotted in Fig. 13. In contrast to the relation of Akita and Yoshida(1974) it can be seen that the interfacial areas according to the relations of Idogawa et al(1985) compare rather good to the experimental data for the interfacial areas at 0.15 MPa. A parity plot between the interfacial areas at all pressures and those predicted by the correlations of Idogawa et al(1985) is presented in Fig. 14. The parity plot shows that the literature relations of Idogawa et al(1985) predict too high interfacial areas for the lower superficial gas velocities at the higher operating pressures. Moreover, our interfacial areas are more or less independent of pressure for the lower superficial gas velocities, whereas Idogawa et al(1985) do predict a strong pressure influence. However, the agreement at superficial gas velocities higher than 3 cm/s is satisfactory for all experimental conditions.

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It can be seen in Fig. 13 that the interfacial areas at 0.15 MPa increase with an increasing superficial gas velocity up to 4 cm/s, then remain constant around a value of  $220 \text{ m}^{-1}$  for superficial gas velocities up to 6 cm/s and start to increase again for higher gas velocities. This has already been observed for the interfacial areas as obtained from the Danckwerts plots, see Table 2. The plateau value of the interfacial areas as a function of the superficial gas velocity is caused by the transition from the bubbly flow towards the churn-turbulent regime around superficial gas velocities of 5 cm/s, see Shah et al(1982). Schumpe et al(1979) have demonstrated experimentally that the transition in flow regimes is accompanied by a sharp decrease in the conversion of the gas phase reactant. This effect can also be observed for the  $\text{CO}_2$ -conversions at 0.15 MPa in Table 4.

The results for the gas hold-ups in both water and in an aqueous DEA solution at elevated pressures reveal that the transition from the bubbly flow towards the churn-turbulent regime is influenced by the operating pressure. The interfacial areas at a pressure of 0.30 MPa are therefore also plotted in Fig. 13 in order to verify whether this can also be concluded from the results on the interfacial areas. As can be seen in Fig. 13 the plateau has almost disappeared at a pressure of 0.30 MPa and indeed the transition between the two flow regimes is influenced by the operating pressure.

#### 3.3. Interfacial areas in the packed bubble column.

In addition to the experiments in the two phase bubble column we have determined gas hold-ups and interfacial areas in a three phase packed bubble column. We have filled the column with glass cylinders with a length and diameter equal to 5.0 and 4.0 mm respectively. This results in a packed bubble column with similar geometries as the one used by van Gelder and Westerterp(1988) for the determination of liquid phase residence time distribution and of hold-ups at elevated pressures.

Packed bubble columns are used e.g. in hydrogenation and oxidation processes for which long liquid phase residence times and low liquid phase backmixing are required. Literature data on the hydrodynamics and the mass transfer characteristics in packed bubble columns are scarce and are often obtained for larger columns with larger and other types of packing and for higher gas and liquid flows than applied in this study. Besides the aforementioned article of van Gelder and Westerterp(1988) no information for packed bubble columns at elevated pressures is available: they concluded that at constant volumetric gas flow rates, the hold-ups in a packed bubble column are unaffected by the operating pressure.

##### 3.3.1. Liquid phase mass transfer coefficients.

Liquid phase mass transfer coefficients  $k_L$  in the packed bubble column were determined by means of the Danckwerts plot method in a series of absorption experiments at 0.15 MPa and at four different superficial gas velocities of 0.5, 1.0, 2.0 and 3.0 cm/s respectively. These superficial gas velocities are based on the empty cross sectional area of the column and correspond to gas

velocities, based on the empty cross sectional area of the packed column, equal to 1.4, 2.9, 5.7 and 8.6 cm/s respectively. In order to compare the performance of the packed bubble column to the bubble column all results on interfacial area and gas hold-ups reported for the packed bubble column will be corrected for the bed porosity and based on the true volume available for gas and liquid.

The absorption experiments have been performed in five different aqueous DEA solutions with DEA concentrations between 0.10 and 0.85 mol/kg. The Danckwerts plots as obtained from these experiments are shown in Fig. 15 and the values for  $k_L$  and  $a$  as obtained are given in Table 5: it can be seen that  $k_L$  remains practically constant and has an average value of  $2.6 \cdot 10^{-4}$  m/s. This value agrees very well with the value of  $2.5 \cdot 10^{-4}$  m/s as obtained for  $k_L$  by Pexidr and Charpentier(1975) in a bubble column packed with 6.4 mm Raschig rings and with a column height and diameter of 1.75 m and 0.10 m respectively. From Eqs (7) and (8) it can be calculated that with  $k_L = 2.5 \cdot 10^{-4}$  m/s a free DEA concentration of 1.2 mol/kg results in  $Ha = 4.0$ . Together with a maximum DEA conversion of 20% this leads to a recommended total DEA concentration of 1.5 mol/kg in the amine solutions for the determination of interfacial areas.

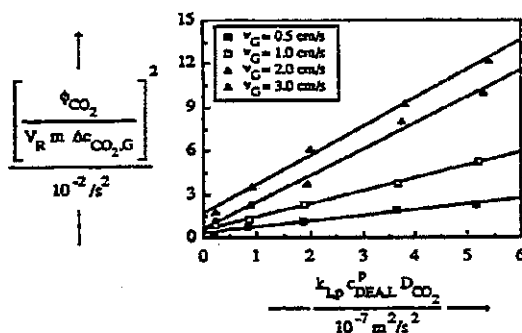


Figure 15: Danckwerts plots for the packed bubble column at a pressure of 0.15 MPa.

Table 5. Results of the Danckwerts plots for the packed bubble column at a pressure of 0.15 MPa.

Data: [ DEA ] = 0.10 - 0.85 mol/kg.

$v_G \cdot 10^2$ m/s	$k_L \cdot 10^4$ m/s	$a$ 1/m
0.5	2.98	200
1.0	2.49	297
2.0	2.07	423
3.0	2.94	443

### 3.3.2. Interfacial areas.

Absorption experiments in the packed bubble column have been performed for pressures between 0.25 and 1.30 MPa and at superficial gas velocities of 1.0, 2.0 and 3.0 cm/s. All experimental results are given in Table 6 and the results are plotted in Fig. 16: it is evident that the interfacial areas in the packed bubble column are not influenced by pressure up to operating pressures of 1.3 MPa. This is in agreement with the independence of the gas hold-ups on pressure in a packed bubble column as reported by van Gelder and Westertorp(1988).

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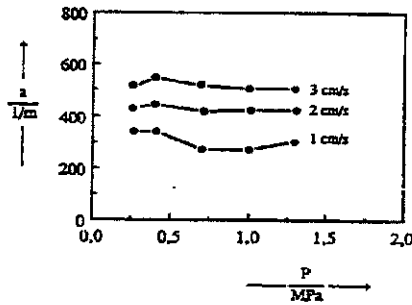


Figure 16: Interfacial area in a DEA solution partially loaded with CO<sub>2</sub> vs. the reactor pressure in the packed bubble column.

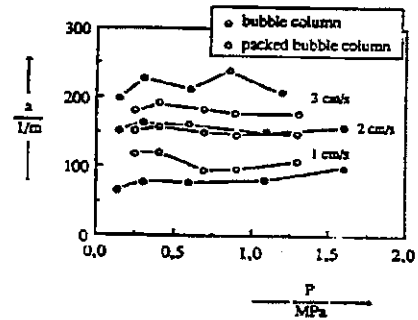


Figure 17: Comparison between the interfacial areas based on the reactor volume in the bubble column and in the packed bubble column.

Table 6. Interfacial areas in the packed bubble column at different pressures.

Data: [DEA] = 1.5 mol/kg.

$v_G \cdot 10^2$ m/s	P MPa	$\zeta_{CO_2}$ %	a 1/m
1.0	0.25	99	340
	0.40	99	340
	0.70	98	271
	1.00	98	274
	1.30	99	306
2.0	0.25	95	431
	0.40	95	449
	0.70	95	426
	1.00	95	423
	1.30	94	423
3.0	0.25	90	517
	0.40	92	551
	0.70	91	526
	1.00	90	506
	1.30	91	506

At the same operating conditions all interfacial areas in the packed bubble column are much higher than the interfacial areas in the bubble column as presented in Table 4. A comparison between the interfacial areas in the two bubble columns on the basis of the reactor volume of the column is shown in Fig. 17: it can be seen that at the lowest superficial gas velocity of 1.0 cm/s the interfacial areas in the packed bubble column are still higher. However, for the superficial gas velocity of 3.0 cm/s the interfacial areas in the packed bubble column as based on the total volume of the column become smaller than in the bubble column.

The higher interfacial areas at 1.0 cm/s can be attributed to lower rising velocities of the bubbles in the packed column and to the fact that larger bubbles are broken on the edges of the packing material, see also Pexidr and Charpentier(1975) and Böxkes and Hofmann(1975). It has also been observed that the bubbles tend to form chains of bubbles throughout the packed column. These chains of bubbles lead to a smaller increase of the interfacial area in the packed bubble column with increasing gas velocity as compared to the interfacial areas in the bubble column.

Results on interfacial areas obtained by the chemical method at high conversions of the gas phase reactant should be considered carefully. As the conversions of  $\text{CO}_2$  become larger than 80% the evaluation of interfacial areas becomes sensitive towards the assumed residence time distribution of the gas phase, see Chapter 2. Therefore, we have also performed absorption experiments in a shorter packed column with a packing height of 27.0 cm ( $H/D = 3.2$ ). This lower height results in  $\text{CO}_2$ -conversion levels around 70%. The interfacial areas at a superficial gas velocity of 2.0 cm/s are plotted in Fig. 18: the values are equal to those at 2.0 cm/s in the packed bubble column with a packing height at  $H/D = 6.2$ . This indicates that the interfacial areas as determined in the packed bubble column with a packing height at  $H/D = 6.2$  are realistic and that the assumed plug flow behaviour for the gas phase is correct.

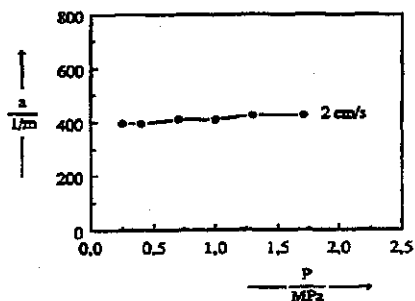


Figure 18: Interfacial area in a DEA solution partially loaded with  $\text{CO}_2$  vs. the reactor pressure in the packed bubble column and with  $H/D = 3.2$ .



#### 4. DISCUSSION AND CONCLUSIONS

The results from this study show that the gas hold-ups in the bubble column increase with increasing pressure. The positive influence of the operating pressure on the gas hold-ups in water is qualitatively in agreement with literature results, as reported for water and organic liquids in bubble columns at elevated pressures, see a.o. Tarmy et al(1984a, 1984b), Pijls et al(1985) and Teurlings et al(1986). The gas hold-ups in water can be predicted fairly well with the correlation of Idogawa et al(1985). For an aqueous DEA solution the increase in gas hold-up with pressure is even larger than for water. However, interfacial area determinations indicate that part of the gas hold-up in this DEA solution consists of a froth layer, which is less effective for mass transfer.

Plots of the drift flux  $q_d$  versus the gas hold-ups in the bubble column at elevated pressures indicate that the transition from the bubbly flow towards the churn-turbulent regime occurs at higher gas velocities and higher gas hold-ups. These findings are in agreement with the results of Tarmy et al(1984a, 1984b). It offers the interesting possibility of operating a bubble column at elevated pressures in the bubbly flow regime: at higher volumetric gas throughputs than attainable in a column operating at atmospheric pressure. Operating a bubble column in the bubbly flow regime may be required for high gas phase conversions, but may also be preferred, if low liquid phase backmixing is required, for selectivity. So, if selectivity is an important design objective, the liquid phase residence time distribution at higher pressures should be investigated in the development of a gas-liquid process in a bubble column operating at elevated pressures.

Interfacial areas in the bubble column as determined by the chemical method in an aqueous DEA solution show also an increase with increasing pressure. However, this increase is smaller than the increase in the gas hold-ups for the same solution. Comparison shows that the relative increase of the interfacial area in the aqueous DEA solutions with pressure is more in line with the relative increase of the gas hold-ups in water for the same range of operating pressures.

As no information in literature is available on interfacial areas in bubble columns at higher pressures no direct comparison is possible. The correlations of Idogawa et al(1985) on gas hold-ups and Sauter mean bubble diameters in bubble columns at elevated pressures can be used for a calculation of interfacial area. Although the gas hold-ups, as calculated from these relations, do not agree with the experimental gas hold-ups in the aqueous DEA solution, the calculated interfacial areas agree fairly well with the experimental interfacial area. However, in order to answer the question whether the relations of Idogawa et al(1985) can be used to predict the influence of pressure on gas hold-up and interfacial area in a bubble column, further research over a wider range of operating conditions and column geometries should be conducted.

In Chapter 1 we conclude that the initial bubble size at a single orifice decreases with increasing pressure. Gas hold-up measurements in the bubble column with a dispersion level at half of the original height indicate that the positive influence of pressure on the gas hold-up originates from the formation of smaller bubbles at the gas distributor. These smaller bubbles have lower rising

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velocities and lead to an increase in the gas hold-up and interfacial area with increasing pressures.

Comparison of the gas hold-ups and the interfacial areas in the bubble and the packed bubble column of this study and the same results for a mechanically agitated reactor at elevated pressures, see Chapter 2, confirms that the pressure influence originates from the formation of smaller bubbles at the gas distributor. Both the mechanically agitated reactor and the packed bubble column exhibit no influence of pressure on the gas hold-ups and interfacial areas, because in both gas-liquid contactors the coalescence-breakup process of the bubbles is dominated by the agitator or the packing. Any influence of pressure on the bubble formation at the gas distributor is therefore masked by the coalescence-breakup process in these contactors.

In a bubble column the gas hold-up and interfacial areas are partly dependent on the bubble formation process at the gas distributor and therefore this contactor exhibits a positive pressure effect. The effect of pressure may be masked also in long bubble columns in which the gas hold-ups and interfacial areas are mainly determined by the coalescence-breakup equilibrium in the column.

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#### NOTATION

$a$	specific interfacial area, $m^2/m^3$ dispersion
$c$	concentration, $mol/m^3$
$\Delta c$	driving force for mass transfer, $mol/m^3$
$d_b$	bubble diameter, m
$d_{mb}$	Sauter mean bubble diameter, m
$D$	vessel diameter, m
$D_A$	diffusivity of component A in the liquid phase, $m^2/s$
$D_B$	diffusivity of component B in the liquid phase, $m^2/s$
$E_A$	enhancement factor, Eq. (1)
$E_{A,\infty}$	enhancement factor for an infinitely fast reaction, Eq. (4)
$g$	gravitational acceleration, $m^2/s$
$h$	gas distributor height, m
$H$	dispersion height, m
$Ha$	Hatta number, Eq. (3)
$I$	ionic strength, $kg\ ion/m^3$
$k_L$	liquid phase mass transfer coefficient, $m/s$
$k_{n,p}$	reaction rate constant for a reaction of the order $(n,p)$ , $m^{3(p-n-1)}/mo^{(p+n-1)}s$
$m$	distribution coefficient, $c_L = m c_G$
$P$	pressure, Pa
$q_d$	drift flux, $m/s$
$v$	superficial velocity based on the empty cross-sectional area of the vessel, $m/s$

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V volume,  $m^3$

### Greek symbols

$\epsilon$  hold-up  
 $\phi$  flow rate,  $m^3/s$  or mol/s  
 $\mu$  dynamic viscosity,  $N\ s/m^2$   
 $\nu$  kinematic viscosity,  $m^2/s$   
 $\nu_B$  stoichiometric coefficient of component B ( $\nu_A = 1$ )  
 $\rho$  density,  $kg/m^3$   
 $\sigma$  surface tension, N/m  
 $\sigma_{ln}^2$  variance log-normal distribution  
 $\tau_b$  bubble residence time, s  
 $\zeta$  conversion  
 $\Omega$  overall conversion

### Subscripts/Superscripts

A component originally in the gas phase  
b bubble  
B component originally in the liquid phase  
chem chemical  
free free amine  
G gas  
geo geometric  
in inlet  
L liquid  
n reaction rate order for component A in the liquid phase  
out outlet  
p reaction rate order for component B in the liquid phase  
R reactor  
V volumetric

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