Some Considerations and Recalls on the Design of Gas Absorbers: Laboratory Apparatus to Simulate Absorber and to Predict the Effect of Chemical Reaction

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In publishing this note on a very convenient technique for the design of gas absorbers, we want to insist on the performances and the practical use of a method proposed by Danckwerts and Gillham (1966) that is not yet currently known that seems promised to a great future.

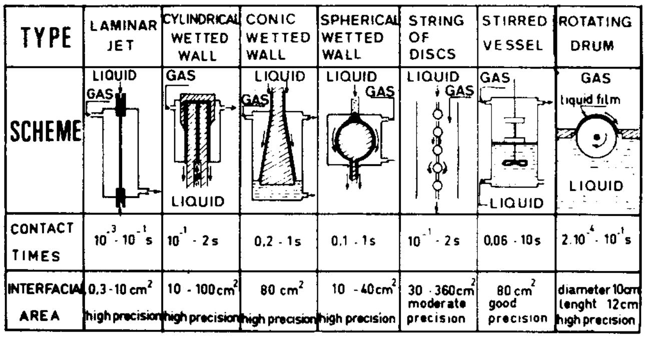
The average rate of absorption per unit interfacial area of a pure gas solute into a liquid with which it reacts is determined partly by physico-chemical factors (solubility, kinetics, diffusivity . . .) and partly by hydrodynamics factors (liquid flow rates, shape and size of the packing or of the plate, gas sparger, shape of the stirrer . . .). Various simple theoretical models of the absorption process have been used to calculate the effect of chemical reactions with simple kinetics, and they lead to conclusions about the effect of chemical reactions which are numerically in very close agreement. Therefore, for fixed value of the liquid mass transfer coefficient k_L, the different models (two films model, Higbie and Danckwerts surface-renewal models) give about the same value for the average rate of absorption.

It would follow that if a laboratory model of an industrial gas absorber is to be designed, the essential feature to be reproduced is the value of k_L. The rate of absorption in the laboratory model would then respond to chemical reaction in the same way as in industrial absorber, even though the means of agitating the liquid in the two systems might be quite different.

Various types of laboratory models may be used for simulating absorbers under different circumstances (falling films, rotating drums, jets, stirred reactor, string of the disks . . .) (Figure 1 and Table 1).

Following Danckwerts and Gillham (1966), assume the laboratory model is a stirred vessel and the industrial absorber is a packed column. Consider now absorption into a inert solution (physical absorption) and into a reacting solution and let the stirring rate in the vessel be constant N_1. The enhancement factor E_m in the laboratory apparatus will depend on the value of k_L (the liquid film coefficient in the absence of reaction) as well as on the nature of the reacting solution, (for instance, $E_m = (\sqrt{D_A\, k_2\, C_{B0}})/k_L$ for a pseudo first-order reaction or $E_m = 1 + \dfrac{D_B}{z\, D_A}\,\dfrac{C_{B0}}{C_A{}^*}$ for an instantaneous reaction [see Charpentier (1971)].

In the industrial packed column with the same solution, there will then be a particular liquid flow rate L_1 at which the value of k_L and therefore the enhancement factor E_c will be the same as in the stirred vessel. So if the superficial flow rate L in the column is systematically varied, the enhancement factor in the column varies, then the ratio E_m/E_c (for a fixed stirring speed in the vessel) varies and becomes equal to unity when L has such a value as to make the value of k_L in the industrial columns as that in the laboratory vessel (Figure 2). These experiments may be repeated with a number of different reacting solutions. If in each case the ratio E_m/E_c is plotted versus L, the



TYPE	LAMINAR JET	CYLINDRICAL WETTED WALL	CONIC WETTED WALL	SPHERICAL WETTED WALL	STRING OF DISCS	STIRRED VESSEL	ROTATING DRUM
SCHEME	LIQUID GAS	GAS LIQUID	LIQUID GAS	LIQUID GAS	LIQUID GAS	GAS LIQUID	GAS liquid film LIQUID
CONTACT TIMES	10^{-3} - 10^{-1} s	10^{-1} - 2 s	0,2 - 1 s	0,1 - 1 s	10^{-1} - 2 s	0,06 - 10 s	2.10^{-4} - 10^{-1} s
INTERFACIAL AREA	0,3 - 10 cm^2 high precision	10 - 100 cm^2 high precision	80 cm^2 high precision	10 - 40 cm^2 high precision	30 - 360 cm^2 moderate precision	80 cm^2 good precision	diameter 10cm lenght 12cm high precision

Fig. 1. Main laboratory models.

Table 1. Models for Simulating Absorbers

Laboratory model	Contact time, s	Industrial absorber	Contact time, s	Reference	Laboratory model to be used for simulation
Cylindrical wetted wall	0.1-2	Plate column	0.019-0.04	Sharma et al. (1960)	Laminar jet Rotating drum
Spherical wetted wall	0.1-1	Packed column	0.03-1	Danckwerts (1970)	Cylindrical wetted wall String of disks or of spheres Stirred vessel
Conical wetted wall	0.2-1	Venturi	0.004	Volgin et al. (1968)	Laminar jet
Laminar jet	0.001-0.1	Spray column	0.03-0.1	Mehta and Sharma (1970)	Laminar jet Rotating drum
String of disks or of spheres	0.1-2	Well stirred tank	0.01	Reith and Beek (1968) Mehta and Sharma (1971)	Laminar jet Rotating drum
Rotating drum	0.0002-0.1	Bubble column	0.01	Mashelkar and Sharma (1970)	Laminar jet Rotating drum
Stirred vessel	0.06-10				

various curves intersect the line $E_m/E_c = 1$ at a single point, corresponding to the flow rate L_1 at which the value of the liquid film coefficient in the absence of reaction k_L in the column is the same as that in the vessel at the fixed stirring speed.

The values of E_m for the various solutions can now be measured at different stirring speeds, and it is then possible to determine and tabulate the stirring speed corresponding to any flow rate in the industrial reactor and the corresponding rates of absorption in the inert solution (Figure 3).

If one wants now to predict the rate of absorption into some novel absorbent in the industrial apparatus at a given flow rate L_2, one puts the absorbent into the vessel, measures the absorption rate at the appropriate stirring speed N_2 (Figure 3), and thus (from the previously obtained results with the inert solution) determines $E_{m2} = E_{c2}$. The previously determined rate of absorption into the inert solution in the column will be multiplied by E_{m2} to give the rate of absorption into the novel solution in the column.

Once determined, the data relating the stirring speed in the laboratory vessel to the flow rate in the industrial apparatus will be permanently recorded for use by designers with the imperious condition that experiments on new liquids and gases must be realized with a vessel identical in design to that used to obtain the original data.

When the soluble gas is mixed with an insoluble gas, it must diffuse through the latter to reach the surface of the liquid, and there is an appreciable gas-side resistance. It may be necessary to adjust the value of k_G in the laboratory apparatus to make it the same as that in the industrial equipment. In the case of the stirred cell, the value of k_G might be regulated by a stirrer which agitates the gas phase. It is then possible as previously to relate the gas flow rate G in the industrial contactor to the stirring speed of gas phase.

In conclusion, for the simulation of an industrial apparatus by a laboratory apparatus, the liquid mass transfer coefficient (or the contact time) and the gas-phase mass transfer coefficient have to be the same in both equipments.

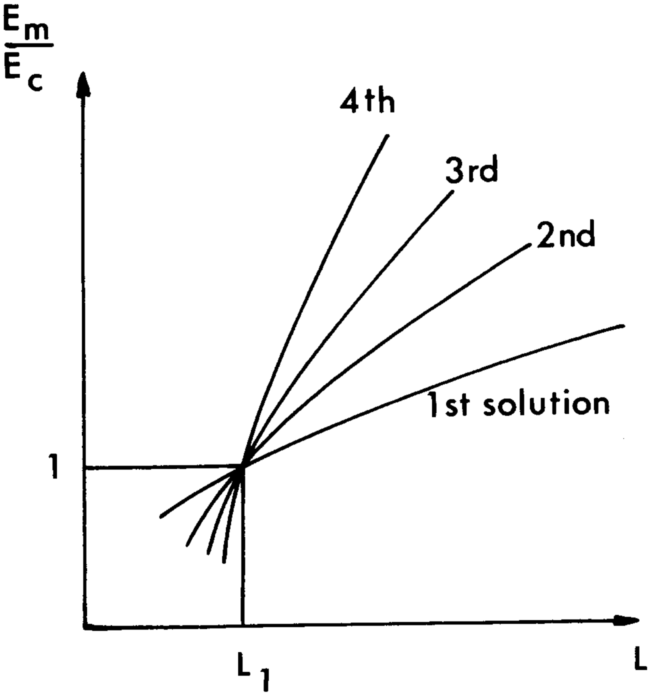


Fig. 2. Ideal relations between the ratio of enhancement factors and the column liquid flow rate.

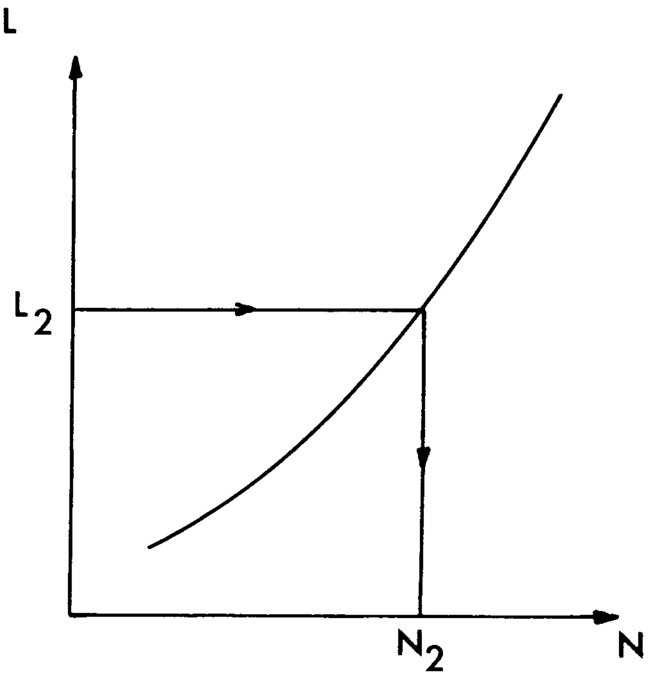


Fig. 3. Relation between column flow L and stirrer speed.

For practical use, on the next table we present the ranges of contact times for different industrial gas-liquid absorbers and laboratory models. For instance, in comparing the contact times, it may be seen that any wetted wall or a stirred vessel could simulate a packed column. This technique has been used successively by Jhaveri and Sharma (1969) for the absorption of oxygen by solutions of sodium dithionite and recently by Yano et al. (1973) for the liquid phase oxidation of propylene with palladium chloride and Shaffer et al. (1974) for the absorption of butenes in trifluoroacetic acid. In all cases, the chemical absorption rates measured in a stirred vessel model have allowed for the prediction of the chemical absorption rate of a packed column, the maximum discrepancy being less than 20%.

NOTATION

C_A^* = concentration of dissolved gas A at interface in equilibrium with gas at interface
C_{B0} = concentration of reactant B in bulk of liquid
D_A = diffusivity of dissolved gas A
D_B = diffusivity of reactant B
E_c = enhancement factor in the industrial absorber
E_m = enhancement factor in the laboratory apparatus
G = superficial velocity of gas
k_G = gas phase mass transfer coefficient
k_L = liquid film mass transfer coefficient
k_2 = second order rate-constant for reaction of A
L = superficial velocity of liquid
N = speed of revolution of impeller
z = number of moles of reactant B reacting with each mole of solute A

LITERATURE CITED

Charpentier, J. C., "Détermination des aires interfaciales et des coefficients de transfert dans les réacteurs gaz-liquide," Centre de Perfectionnement Ind. Chim. Cycle de Perfectionnement sur les réacteurs gaz-liquide, ENSIC, Nancy (1971).

Danckwerts, P. V., *Gas-liquid reactions*, McGraw Hill, London (1970).

———., and A. J. Gillham, "The design of gas absorbers, I-Method for predicting rates of absorption with chemical reaction in packed columns and tests with 1,5 in raschig rings," *Trans. Instn. Chem. Engrs.*, **44**, T42 (1966).

Jhaveri, A. S., and M. M. Sharma, "Absorption with fast chemical reaction," *Chem. Eng. Sci.*, **24**, 189 (1969).

Mashelkar, R. A., and M. M. Sharma, "Mass transfer in bubble and packed bubble columns", *Trans. Instn. Chem. Engrs.*, **48**, T162 (1970).

Mehta, K. C., and M. M. Sharma, "Mass transfer in spray columns," *Brit. Chem. Eng.*, **15**, 1440 (1970).

Mehta, V. D., and M. M. Sharma, "Mass transfer in mechanically agitated gas-liquid contractors," *Chem. Eng. Sci.*, **26**, 461 (1971).

Reith, T., and W. J. Beek, "Gas hold-up, interfacial areas and mass transfer coefficients in gas-liquid contactors," *Proc. 4th Euro. Symp. Chem. Reaction Eng.*, Brussels (1968).

Shaffer, D. L., J. H. Jones, and T. E. Daubert, "Simultaneous Absorption and Chemical Reaction of Butenes," *Ind. Eng. Chem. Process Design Develop.*, **13**, 14 (1974).

Sharma, M. M., R. M. Mashelkar, and V. D. Mehta, "Mass transfer in plate columns," *Brit. Chem. Eng.*, **14**, 70 (1969).

Volgin, B. P., T. F. Efimova, and M. S. Gofman, "Absorption of Sulfur Dioxide by Ammonium Sulfite-Bisulfite Solution in a Venturi Scrubber," *Intern. Chem. Eng.*, **8**, 113 (1968).

Yano, T., T. Suetaka, T. Umehara and T. Yamashita, "Liquid-Phase Oxidation of Propylene using a Packed Column," *ibid.*, **13**, 371 (1973).

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Entrainment in Rotary Cylinders

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To design a rotary cylinder, knowledge of the passage or holdup of solid and the heat transfer in the cylinder is required. However, an optimum design usually cannot be made without an understanding of the entrainment occurring inside the cylinder. Very limited knowledge on the entrainment has been available in the literature. Friedman and Marshall (1949) studied the holdup and entrainment in a rotary dryer. They report some entrainment data and correlations but did not obtain an entrainment equation. An equation is presented by Khodorov (1961). The equation, however, does not include important design variables such as the rotation speed and slope of the cylinder and the solid feed rate. Therefore, the equation is not adequate for design applications. An entrainment equation relating to the design variables is developed in this study.

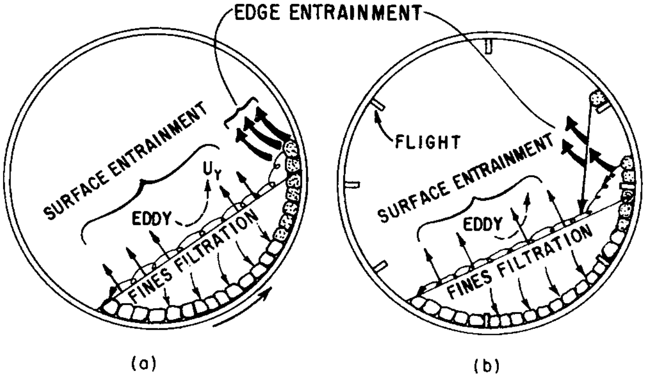
THEORETICAL CONCEPT

The path of solid in a rotary cylinder is such that a particle in the solid bed remains stationary with respect to the cylinder until it reaches a position close to the surface of the bed. From this point, it cascades down along the bed surface and comes to rest at some lower position. The cycle is then repeated (Saeman, 1951). The solid particles contacted with the cylinder wall will be carried up to a position slightly above the bed surface due to the friction between solid and wall. In case there are flights on the cylinder wall the carry-up will be pronounced. The carry-up gives the particles a better contact with the gas. Hence the entrainment at the higher edge of the bed surface, edge entrainment, should be more important than the entrainment occurring on the bed surface, surface entrainment. When the solid particles are cascading down on the surface, the finer particles filter down through the bed. This also causes the surface entrainment to be less significant.

The edge entrainment can be modeled by considering the solid material at the bottom layer of the bed as blocks which are carried up by the wall friction alone or with flights and then fall onto the bed surface as illustrated in Figure 1. In practical operations, the entrainment severity is maintained so low that the entrainment rate dW/dL may be assumed to be proportional to the falling rate of the blocks, the concentration of fines in the blocks C and the intensity of the eddy flow I_e. I_e is a function of the density, viscosity, and velocity of gas and the dimension of gas flow cross-sectional area. Noting that the block falling rate is proportional to the product of cylinder rotation speed N and the cylinder diameter D, we can write

$$\frac{dW}{dL} \propto ND\, CI_e \tag{1}$$

Since fines will filter through the larger solid particles, the fines will be concentrated in the bottom solid layer. Thus,



(a) (b)

Fig. 1. Entrainment mechanism: (a) no flights, (b) with flights.