Design of Commercial-Scale Gas-Liquid Contactors

A systematic design procedure using a liquid circulation unit for gas-liquid systems containing low-viscosity liquids is developed. The procedure can be used to predict and optimize the fluid mechanics and mass transfer in gas-liquid contactors for a specified performance. Batch liquid mass transfer data obtained from a commercial-scale aeration basin (12.2 m × 12.2 m × 6.7 m; gas flow rate range, 0.0023-0.01 m^3/s) are used to verify the model predictions and to test the design procedure.

Z. Otero Keil and T. W. F. Russell
University of Delaware
Chemical Engineering Department
Newark, DE 19716

Introduction

Gas-liquid contacting devices are an essential feature of the chemical processing industries. Liquid phase oxidation of hydrocarbons and production of single-cell protein are but two examples of important industrial gas-liquid systems. Gas-liquid contacting vessels in commercial use range in geometry from bubble columns (about 1 m dia., up to 15 m high) to aeration basins (10–70 m dia., 1–8 m high). A variety of sparging systems are used, ranging from a single orifice to many closely spaced orifices. The gas flow rates encountered are 5×10^{-4} m^3/s to about 0.05 m^3/s. In order to develop systematic design procedures, a quantitative understanding of the fluid mechanics and mass transfer processes is essential.

When gas is sparged into a pool of liquid, a two-phase region is formed above each orifice. The density difference between the two-phase region and the surrounding liquid induces a flow of liquid into the two-phase region. This liquid flows upward in the two-phase region. Near the liquid surface, it turns and flows radially away from the two-phase region until it meets liquid from an adjacent bubbling station or the vessel walls. At this point, the liquid flows downward into the vessel.

The two-phase region and the down-flowing liquid surrounding it form a liquid circulation unit. Mass transfer takes place only in the two-phase region. The concentration of the component transferred into the liquid in the two-phase region is diluted in the liquid outside the region before it flows radially outward, then down and back into the two-phase region. The characteristics and number of liquid circulation units in a gas-liquid system depend on the number of bubbling stations and the distance between them, the gas flow rate per bubbling station, and the total liquid height in the vessel.

The liquid circulation unit provides good liquid circulation and mixing. The volumetric ratio of liquid to gas flow rates can be greater than 100 depending on the system (Goosens, 1979; Kobus, 1968; Baines, 1961). Thus, in some low-viscosity liquid systems the liquid circulation unit can be used to replace mechanical agitators. For example, in systems with large aspect ratios and low-viscosity liquids, mechanical agitation does not significantly improve the mixing or mass transfer rates in a well-designed system (Bottom et al., 1980; Lehrer, 1968; Sideman et al., 1966; Braulick et al., 1965; Benedek and Heidiger, 1971; Cooper et al., 1944. Chardrasekharan and Calderbank (1981) and Bottom et al. (1980) have found that in low-viscosity liquid applications, mechanical agitation is only effective for vessels with small aspect ratios. The costs involved in the operation and maintenance of mechanical agitators may be large (Mitchell and Lev, 1970) and liquid circulation units offer an attractive alternative for mixing gas-liquid systems.

In this work, a systematic design procedure is developed for tank-type gas-liquid systems operating at industrial-scale conditions. The procedure can be used to design gas-liquid systems to meet specific requirements or to predict the performance of existing systems. The liquid circulation unit forms the basis for the design procedure. For the first time, the liquid circulation geometry and the effective two-phase region diameter have been defined using data from large-scale systems. The amount of mass transferred into the liquid phase is calculated as a function of gas flow rate, liquid height, and bubbling station spacing.

Background

Correlations obtained from laboratory-scale data are generally used in the design of gas-liquid systems. Such design practices result in scale-up problems and inefficient operation of gas-liquid contactors. The scale-up problems encountered in gas-liquid systems are mostly due to the different flow patterns that

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Correspondence concerning this paper should be addressed to T. W. F. Russell.

The present address of Z. Otero Keil is Keil Engineering Services, 19 Fremont Rd., Newark, DE 19711.

exist in different scales of operation (Otero et al., 1985; Rooney and Mignone, 1978; Kalbskopf, 1972). Hydrodynamic studies in small laboratory-scale systems cannot be used with confidence to predict the behavior of large-scale systems (Bjurstrom, 1985; Bull, 1983; Weiland, 1984; Deckwer, 1979). The dependence of the mass transfer coefficient on the power per unit volume and gas superficial velocity can vary with the scale of operation (Chardrasekharan and Calderbank, 1981; Oldshue, 1981).

Limited experiments in large-scale systems are sometimes used for design purposes. Data are usually obtained on a lumped basis. In a typical experiment, clear water is first deoxygenated, then air is sparged into the system at the desired flow rate. The concentration of oxygen in the liquid is measured as a function of time at various points in the vessel. A mass transfer coefficient is then obtained by fitting the concentration vs. time profiles using a well-mixed liquid phase mass balance.

$$\frac{dc_j}{dt} = k_L a (c_j^* - c_j) \tag{1}$$

The mass transfer coefficient in Eq. 1 is not representative of the microscale interactions between the phases. It is obtained by an empirical fit and is a function of system geometry, sparger configuration, gas flow rate, and liquid properties. Such a coefficient cannot be used with confidence to design or predict the performance of gas-liquid systems with different geometries and operating conditions.

Otero (1983) has introduced the liquid circulation unit as the design basis for gas-liquid systems. The liquid circulation unit comprises the two-phase region and the liquid mixing element shown in Figure 1. Liquid flows up within the two-phase region. Near the liquid surface, the liquid flows radially away from the two-phase region until it meets liquid from an adjacent bubbling station. It then flows back into the mixing element. Liquid velocity profiles in the mixing element confirm this liquid circulation pattern (Otero, 1983; Grevet et al., 1982). Figure 2 is a typical liquid velocity profile in a commercial-scale aeration basin of a size 12.2 m x 12.2 m x 6.7 m (Otero, 1983).

The fluid mechanics and mass transfer characteristics of the liquid circulation unit are a function of the system geometry and operating conditions. As a result, a variety of models for the two-phase region have been proposed. The two-phase region has been modeled as a cylinder (Ulbrecht and Baykhara, 1981; Rietema and Ottengraf, 1970) or an expanding plume (Chesters et al., 1980; Hussain and Siegel, 1975). Otero (1983) and Otero et al. (1985) have identified three distinct flow patterns for tank-type systems, as shown in Figure 3, and the conditions under which each pattern can exist.

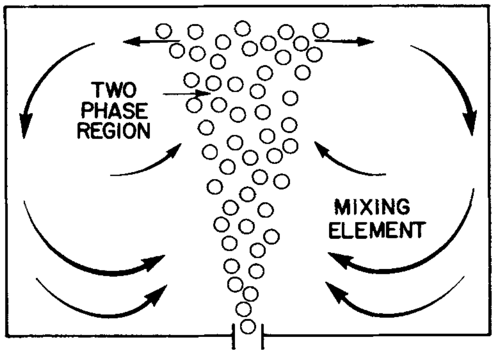

Figure 1. Liquid circulation unit flow pattern.

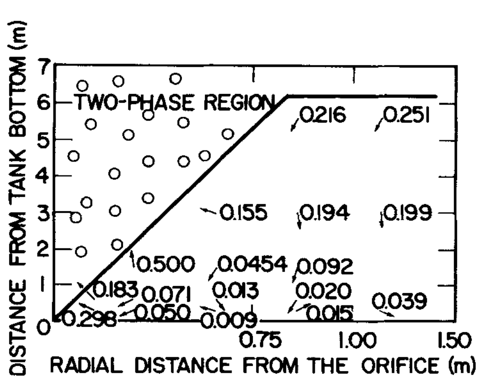

Figure 2. Liquid velocities in mixing element.
$Q_g = 0.0051\ m^3/s$, $D_s = 1.5$ m

Another important design parameter in gas-liquid systems is the gas bubble size, which must be known to compute mass transfer. In turbulent liquid systems the gas bubbles may reach a limiting size away from the orifice (for air/water d_{bm} = 0.0045 m; Bhavaraju et al., 1978; Jones, 1972). In general, the gas bubble size is a function of the turbulence in the system. The system turbulence is in turn a function of gas flow rate, vessel size, and liquid properties (Ueyama et al., 1980; Jackson and Shen, 1978; Lockett and Kirkpatrick, 1975; Freedman and Davidson, 1969; Urza and Jackson, 1975; Sideman et al., 1966; Towell et al., 1965; Leibson et al., 1956).

Experimental Apparatus and Procedures

The mass transfer measurements used to verify the mathematical models were carried out in a large aeration basin 12.2 m x 12.2 m x 6.7 m high. The basin is the property of Rexnord Corporation (Milwaukee, Wisconsin). The experiments were a cooperative effort involving E. I. DuPont de Nemours & Co., Rexnord, the Design Institute for Multiphase Processing, and the University of Delaware. The sparger system in the basin allowed bubbling station spacings of 1.5 m and 3.0 m. Figure 4 is a diagram of the sparger layout.

The basin was filled with City of Milwaukee tap water to a depth of 6 m. Cobalt chloride solution was added to the tank to obtain a cobalt ion concentration of 3×10^{-4} kg/m³. A water sample was analyzed for cobalt before and after each experiment.

The tank contents were saturated with oxygen prior to each test by sparging air until there was no change in the output of the oxygen probes. Then three samples were collected using a submersible pump and the average dissolved oxygen (DO) concentration was determined using a modified Winkler technique

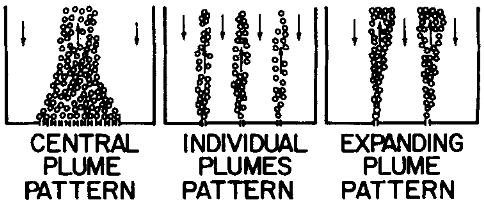

Figure 3. Flow patterns in gas-liquid system.

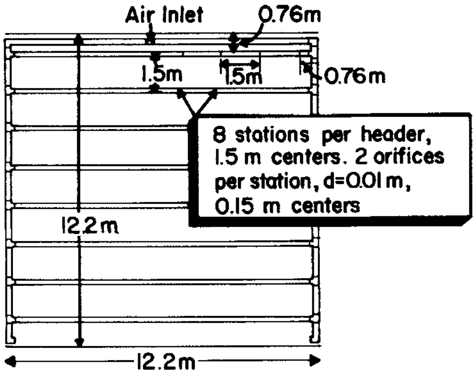

Figure 4. Sparger layout.

(Lakin, 1976). Following saturation, the tank contents were deoxygenated using technical grade sodium sulfite. The sulfite solution was premixed in a 1.9 m^3 tank and added to the basin at four different points. The sulfite flow rate and initial concentration in the basin were 0.006 m^3/s and 1.80 kg/m^3. After deoxygenation, the basin was aerated at the desired flow rate. The DO concentration was monitored at five basin locations using galvanic probes and, for several of the tests, samples were collected for Winkler analysis. These measurements showed that the liquid in the mixing element was well mixed for all of the experimental conditions used. Typical oxygen concentration profiles are shown in Figures 6 and 7.

Model Equations and Design Procedure

An energy balance in the two-phase region can be obtained by setting the energy input from the gas equal to the kinetic energy of the liquid and the turning losses associated with the circulating flow.

$$Q_{go} P_o \ln (P_o/P_s) = \tfrac{1}{2} \rho_L A_p (v_L^3 - v_s^3) + 2k\rho_L(\tfrac{1}{2} v_L^3 A_p + \tfrac{1}{2} v^3 A_s) \quad (2)$$

where $v_s = A_p v_L / A_s$ from a liquid mass balance. Equation 2 can then be solved for the average velocity of the liquid in the two-phase region.

$$v_L = \left[\frac{2Q_{go} P_o \ln \left(\frac{P_o}{P_s}\right)}{\rho_L \left(A_p - \frac{A_p^4}{A_s^3} + 2kA_p + 2k \frac{A_p^3}{A_s^2}\right)} \right]^{1/3} \quad (3)$$

The subscript o refers to conditions at the orifice; k is the number of velocity heads lost by the turning liquid (1.2–1.4 for most turbulent applications). The areas of the two-phase region, A_p, and the mixing element, A_s, are related by Eq. 4:

$$\frac{A_p}{A_s} = \frac{d_p^2}{D_s^2 - d_p^2} \quad (4)$$

The bubble size in the system can be calculated from Eq. 5 (Bhavaraju et al., 1978).

$$d_{bm} = 0.7 \frac{\sigma^{0.6}}{\left(\frac{P_G}{V_l}\right)^{0.4} \rho_L^{0.2}} \left(\frac{\mu L}{\mu g}\right)^{0.1} \quad (5)$$

where

$$P_G = Q_{go} P_o \ln \left(\frac{P_o}{P_s}\right) \quad (6)$$

Equation 5 is valid for bubble sizes down to the limiting bubble size.

Figure 5 is a diagram of the liquid circulation unit depicting the flow rates and concentrations in the two-phase region and mixing element. In the batch liquid case the liquid circulates through the two-phase region and the mixing element until it becomes saturated with the transferring component. In the continuous liquid flow case liquid is introduced and removed from the mixing element at a flow rate q. The appropriate mass balances are given below. It is assumed that the liquid is well mixed and the gas is in plug flow. The gas balance is made on a single bubble (Schaftlein and Russell, 1968).

The gas component balance is given by Eq. 7:

$$-k_G a' V_b P \left(y_j - c_j \frac{H_e}{P}\right) = v_b \delta \frac{\left(\frac{PV_b}{RT} y_j\right)}{\delta z} + \delta \frac{\left(\frac{PV_b}{RT} y_j\right)}{\delta t} \quad (7)$$

Equation 7 is valid for both batch and continuous liquid flow systems since the gas enters the sytem at the sparger and leaves at the liquid surface in both cases.

Liquid phase component balances

Batch Liquid System. A component balance on the liquid phase in the two-phase region yields

$$Q_L(c_e - c_j) + k_G a' N_b V_b P_A \left(\bar{y}_j - c_j \frac{H_e}{P_A}\right) \cdot V_l(1 - \phi) = \frac{dc_j}{dt} V_l(1 - \phi) \quad (8)$$

where $\bar{y}_j$ refers to an average gas phase concentration. The component balance around the mixing element yields

$$Q_L(c_j - c_e) = V_{II} \frac{dc_e}{dt} \quad (9)$$

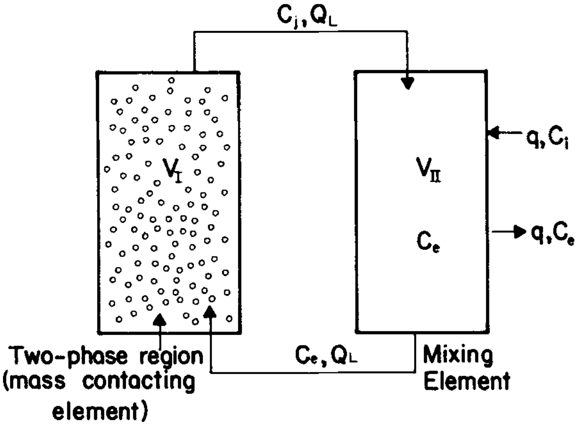

Figure 5. Liquid circulation unit.

Equation 7 can be used to obtain $\bar{y}_j$. The derivative with respect to time can be neglected since gas bubbles are continuously sparged into the system so that at any point there is no change in the bubble concentration with time. Rearrangement of Eq. 7 yields

$$\frac{dy_j}{dz} + y_j\left(\frac{d \ln P}{dz} + \gamma\right) = \frac{\gamma H_e}{P} c_j \tag{10}$$

where

$$\gamma = \frac{K_G a' RT}{V_b}$$

Equation 10 can be solved for $\bar{y}_j$ in terms of z, and integration of this expression over the two-phase region height yields

$$\bar{y}_j = \beta\left[H_e c_j + e^{-\epsilon}(y_o P_o - H_e c_j)\left(1 + \frac{\gamma}{\rho g \beta}\right)\right] \tag{11}$$

where

$$\beta = \frac{\ln\left(\frac{P_o}{P_s}\right)}{(P_o - P_s)}$$

and

$$\epsilon = \frac{\gamma P_o}{\rho g}$$

Equation 11 can be substituted into Eq. 8, and Eqs. 8 and 9 can be combined to yield an expression for c_e.

$$\frac{d^2 c_e}{dt^2} + \frac{dc_e}{dt}\left(\Phi + \frac{1}{\theta_{II}}\right) + \left(\frac{\Phi}{\theta_{II}} - \frac{1}{\theta_I \theta_{II}}\right) c_e - \frac{\alpha_c}{\theta_{II}} = 0 \tag{12}$$

where

$$\Phi = \frac{1}{\theta_I} + \frac{k_G a' N_b V_b \beta_1}{H_L}$$

$$\alpha_c = \frac{k_G a' N_b V_b P_A y_o \beta_1}{H_e H_L}$$

$$\beta_1 = \frac{1 - e^{-\gamma H_L}}{\gamma}$$

$$\theta_I = \frac{V_I(1 - \phi)}{Q_L} \quad \theta_{II} = \frac{V_{II}}{Q_L}$$

Equation 12 can be solved analytically for the concentration of the transferring component in the mixing element, c_e.

$$c_e = c_1 e^{r_1 t} - \frac{r_1}{r_2} c_1 e^{r_2 t} + \frac{\alpha_c}{\theta_{II} c} \quad \text{for } b^2 - 4c > 0 \tag{13}$$

and

$$c_e = c_2 e^{rt} - r c_2 t e^{rt} + \frac{\alpha_c}{\theta_{II} c} \quad \text{for } b^2 - 4c = 0 \tag{14}$$

where

$$b = \left(\phi + \frac{1}{\theta_{II}}\right), \quad c = \left(\frac{\phi}{\theta_{II}} - \frac{1}{\theta_I \theta_{II}}\right)$$

$$c_1 = \frac{c_o - \frac{\alpha_c}{\theta_{II} c}}{\left(1 - \frac{r_1}{r_2}\right)}$$

$$r_1 = \frac{-b + \sqrt{b^2 - 4c}}{2}, \quad r_2 = \frac{-b - \sqrt{b^2 - 4c}}{2}$$

$$c_2 = c_o - \frac{\alpha_c}{\theta_{II} c} \quad \text{and} \quad r = \frac{-b}{2}$$

The liquid concentration of the transferring component in the two-phase region, c_j, can be easily obtained from this analysis. However, in a mass transfer experiment it is c_e that is measured. The above equations completely describe the batch liquid system.

Continuous Liquid Flow System. The liquid phase component balances for the continuous liquid flow system are algebraic. Equation 15 is the component balance for the liquid in the two-phase region and Eq. 16 is the component balance for the liquid in the mixing element.

$$Q_L(c_e - c_j) + k_G a' N_b V_b P_A\left(\bar{y}_j - c_j \frac{H_e}{P}\right) V_I(1 - \phi) = 0 \tag{15}$$

$$Q_L c_j - q c_i = q c_e + Q_L c_e \tag{16}$$

Substituting Eq. 11 into Eq. 15 and solving Eqs. 15 and 16 for c_e and c_j yields

$$c_e = \frac{Q_L c_j + q c_i}{(q + Q_L)} \tag{17}$$

where

$$c_j = \frac{\frac{1}{\theta_I}\left(\frac{q c_i}{Q_L}\right) + k_G a' N_b V_b e^{-\epsilon}\left(1 - \frac{\gamma}{\rho g \beta}\right) y_o P_o}{\left[\frac{1}{\theta_I} - \frac{1}{\theta_I}\left(\frac{Q_L}{q + Q_L}\right) + k_G a' N_b V_b e^{-\epsilon} H_e\left(1 - \frac{\gamma}{\rho g \beta}\right)\right]}$$

Model parameters

The only parameters required in order to use these model equations are the mass transfer coefficient k_G, the gas bubble diameter d_b, the Henry's law constant H_e, the holdup ϕ, and the diameter of the two-phase region. All of these parameters can be calculated as follows.

The mass transfer coefficient can be calculated assuming that the gas phase resistance can be neglected, as is generally the case.

$$\frac{1}{k_G} = \frac{H_e}{k_L} \tag{18}$$

The liquid mass transfer coefficient can be obtained from penetration theory (Higbie, 1935).

$$k_L = 2\sqrt{\frac{Dv_b}{\pi d_b}} \tag{19}$$

where v_b is the bubble rise velocity and is given by Eq. 20:

$$v_b = \frac{(gd_b)^{1/2}}{2} \tag{20}$$

The gas bubble diameter can be calculated from Eq. 5, and the interfacial area available for mass transfer can be calculated from Eq. 21:

$$a = a'N_bV_b = \frac{6\phi}{d_b} \tag{21}$$

where ϕ, the gas holdup, is calculated from a mass balance at any point in the two-phase region and is given by Eq. 22. (Details are given in the supplementary materials deposited with NAPS. See the note at the end of the paper.)

$$\phi = \frac{U_s \frac{P_s}{P}}{(v_b + v_L)} \tag{22}$$

U_s is the superficial velocity of the gas in the two-phase region.

In order to calculate the holding times in the two-phase region, θ_I, and the mixing element, θ_{II}, the diameters of the two-phase region and the mixing element must be known. The holding times can be obtained using the diameters of the two-phase region and the mixing element. The ratio of the holding time in the mixing element to the holding time in the two-phase region is a measure of the amount of time that the liquid spends in each section of the liquid circulation unit. This ratio can be expressed as shown below.

$$\frac{\theta_{II}}{\theta_I} = \frac{D_s^2 - d_p^2}{d_p^2} \tag{23}$$

The ratio of the holding times can be obtained from comparisons of model predictions with mass transfer data. The holding time ratio links the fluid mechanics with the mass transfer characteristics of the system. Once the holding time ratio is known for a set of conditions, the diameter of the two-phase region can be easily obtained from Eq. 24:

$$d_p = \frac{D_s}{\sqrt{1 + \frac{\theta_{II}}{\theta_I}}} \tag{24}$$

Figure 6 is a plot of the concentration of oxygen in the mixing element as a function of time for the experimental conditions listed. The concentration profiles predicted using the model developed here are also shown in the figure for various values of the holding time ratio, θ_{II}/θ_I. Figure 7 is a plot of the concentration profiles for the same system at the two gas flow rates listed. The predicted concentration profiles for the best-fit value of the holding time ratio are also shown. Figure 7 indicates that a single value of the holding time ratio $\theta_{II}/\theta_I = 5$ can be used to predict the mass transfer performance of the air-water system used in this work for a range of air flow rates.

This analysis was used to obtain a holding time ratio for the aeration basin with a bubbling station spacing of 1.5 m. These data were then used to develop a plot of the effective two-phase region diameter as a function of the bubbling station spacing shown in Figure 8. This figure provides for the first time the relationship between the effective two-phase region diameter, d_p, and the bubbling station spacing, D_s. The two-phase region diameter is the key parameter in the design of gas-liquid systems. Figure 8 also accounts for the fact that in systems with bubbling station spacings less than 0.3 m, the cross-sectional area of the two-phase region and the mixing element are the

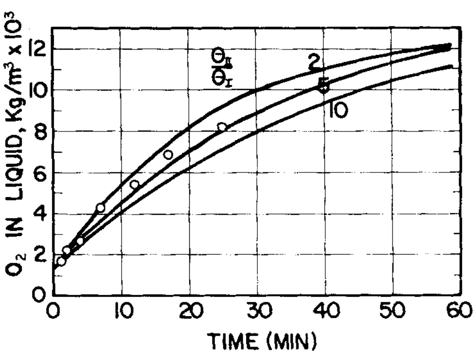

Figure 6. Batch aeration data.

$Q_g = 0.0037$ m³/s, $D_s = 1.5$ m, $T = 11.9$°C

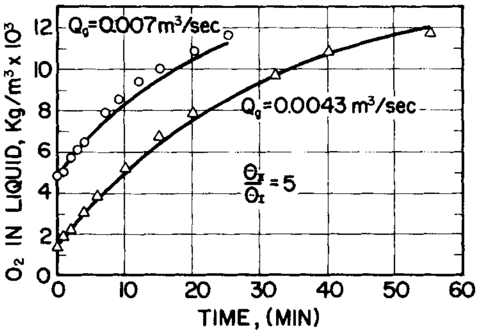

Figure 7. Batch aeration data.

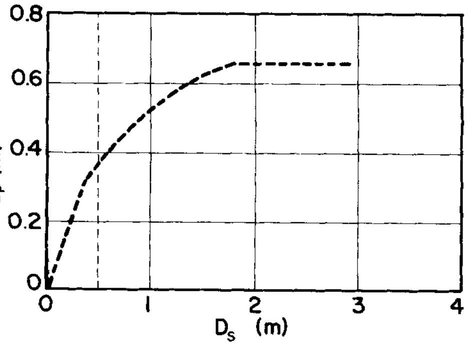

Figure 8. Effective diameter of two-phase region as a function of bubbling station spacing.

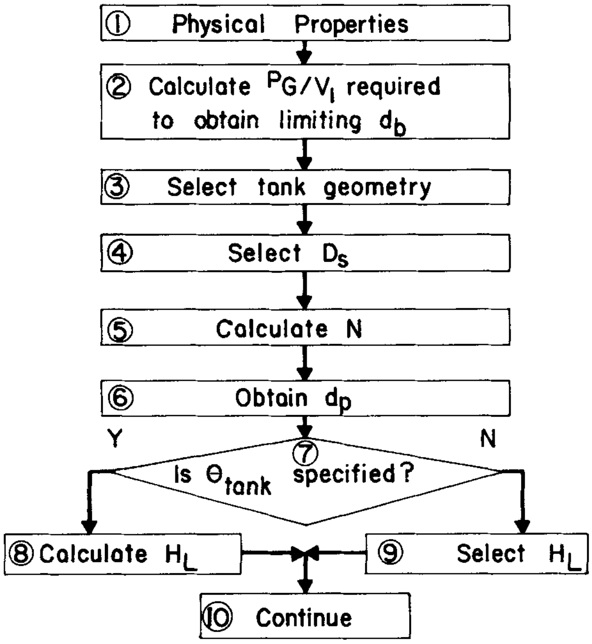

Figure 9. Design procedure flow chart.

same (Joshi and Shah, 1981). Thus, the holding time ratio is unity. For the case where D_s is of the order of the orifice bubble diameter, the two-phase region diameter is approximately the same as the bubbling device diameter. A central plume pattern is obtained as depicted in Figure 3.

Design procedure

The design procedure is illustrated by the flow chart in Figures 9 and 10 and is outlined below.

1. The power per unit two-phase region volume required to produce the limiting bubble size is first calculated from Eq. 5. In the case of air-water, the limiting bubble size is approximately 0.0045 m.

2. The tank geometry is selected (cylindrical or rectangular).

3. The spacing between bubbling stations, D_s, is selected.

4. The number of bubbling stations is calculated:

$$N = \frac{D_T^2}{D_s^2} \quad \text{(cylindrical vessel)} \qquad (25)$$

$$N = \frac{4(L_T W_T)}{\pi D_s^2} \quad \text{(rectangular vessel)} \qquad (26)$$

5. The two-phase region effective diameter is obtained from Figure 8.

6. The gas flow rate or the total liquid height is selected. There are many instances in which the tank holding time is constrained by process-related factors (e.g., reaction kinetics). This is the case in many waste treatment applications. The tank holding time and the liquid height are related as follows.

$$\theta_{tank} = \frac{A_T H_L}{Q_{L(tank)}} \qquad (27)$$

If a holding time is not specified, a liquid height can be selected (as a starting point, $H_L = 2D_s$ or $H_L = 2D_T$), depending on the tank geometry and bubbling station spacing.

7. The two-phase region volume is calculated.

8. The gas flow rate per bubbling station necessary to obtain the power per unit volume that will yield limiting bubble size is calculated using Eq. 6.

9. The liquid flow rate in the liquid circulation unit is calculated using the liquid velocity obtained from Eq. 3. This liquid flow rate must be large enough to provide adequate mixing. In

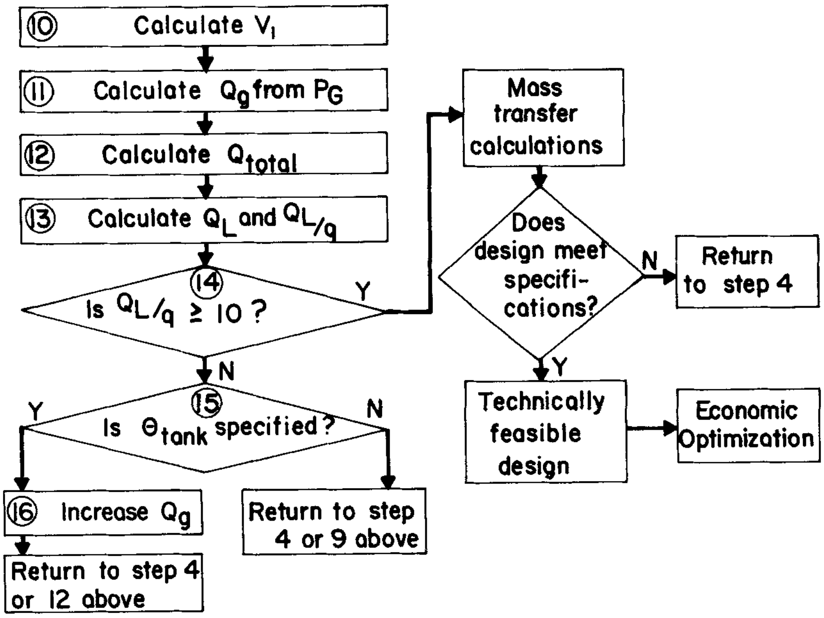

Figure 10. Design procedure flow chart.
Continued from Figure 9.

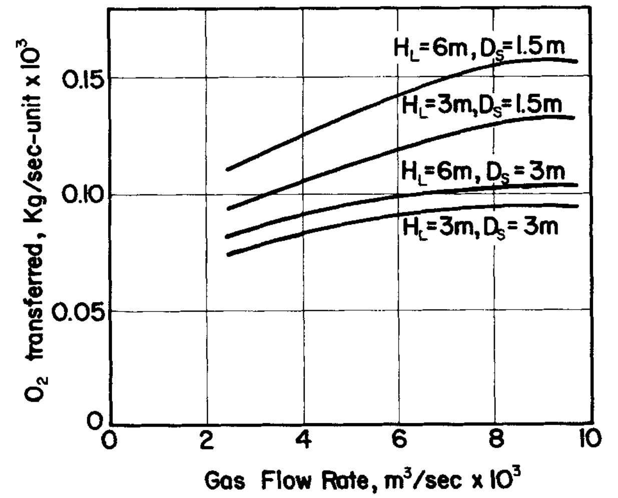

Figure 11. Effect of liquid height and bubbling station spacing on mass transfer.

Tank holding time, 1 h; process liquid flow rate per liquid circulation unit, 0.012 m^3/s

Table 1. Parameters for Design Example, D_s = 3m

Parameter	Value	Eq. No.
N	20	24
d_P	0.6 m	Fig. 8
P_G	200 W/m^3	5
H_L	6 m	26
Q_g	0.0078 m^3/s	6
A_p	0.3 m^3	—
A_s	7.0 m^3	—
v_L	0.9 m/s	3
$Q_L = v_L A_P$	0.27 m^3/s	—
q	0.012 m^3/s	—
Q_L/q	22.5	—
v_b	0.16 m/s	20
k_L	2.84×10^{-4} m/s	19
U_s	0.02 m/s	—
ϕ	0.02	22
a	23.8 m^{-1}	21
θ_I	6.7 s	12
θ_{II}	151 s	12
c_e	8×10^{-3} kg/m^3	17

general, ten turnovers through the two-phase region are adequate. Thus,

$$\frac{Q_L}{q} > 10$$

where

$$q = \frac{Q_{L(tank)}}{N} \tag{28}$$

If the condition of Eq. 28 is not met, an iteration from step 4 is required.

10. The appropriate mass transfer calculations are made using the development presented here. Design plots can be prepared for a particular system, Figure 11.

11. The results are checked to see if they meet the design specifications.

12. A technically feasible design can then be optimized via an economic evaluation.

Design Example

Problem

In a wastewater treatment basin the oxygen concentration of the water must be raised from 0 at the inlet to 6×10^{-3} kg/m^3 at the exit. A square tank with a length and width of 12 m is available. The water temperature and flow rate are 11°C and 0.24 m^3/s, respectively. A holding time of 1 h is necessary. A total of 1.4×10^{-3} kg/s of O_2 must be transferred into the water.

Solution

Table 1 is a listing of the parameters calculated using the design procedure outlined above. The equation used for each calculation is also listed. Calculation details are presented in the Supplementary Materials. Table 2 lists the feasible design parameters. The system as outlined in Table 2 will yield an oxygen concentration in the mixing element, c_e, of 8×10^{-3} kg/m^3. Thus, the design as specified will surpass the mass transfer specifications ($c_e = 6 \times 10^{-3}$ kg/m^3). If this initial design had not met the mass transfer specifications, a different bubbling station spacing could have been selected (step 3 in design procedure). The bubbling station spacing is one of the simplest design parameters to alter, and Figure 8 allows estimation of the two-phase region diameter as a function of bubbling station spacing. This kind of analysis can be used to optimize a design before an economic evaluation is carried out.

Discussion

This design procedure can be used to prepare a simple plot showing the amount of component transferred per liquid circulation unit vs. gas flow rate as a function of total liquid height and bubbling station spacing. Figure 11 is such a plot for the air-water system discussed here. Figure 11 can be used to obtain the key design variables for a specific mass transfer per liquid circulation unit. The figure shows that increasing the liquid height and decreasing the bubbling station spacing increases the amount of oxygen transferred per bubbling station unit. Plots of oxygen transferred per unit per time as a function of gas flow rate and liquid depth are used routinely to calculate system efficiencies from wastewater mass transfer data. The problem with such plots is that the bubbling station spacing is not specified. Figure 11 includes the bubbling station spacing as a parameter.

Table 2. Technically Feasible Design

Parameter	Value
Tank geometry	square
Tank dimensions	$L_T = W_T = 12$m
Liquid temperature	11°C
Liquid flow rate	0.24 m^3/s
Tank holding time	1 h
Liquid height	6 m
No. of bubbling stations	20
Bubbling station spacing	3 m
Gas flow rate per bubbling station	0.008 m^3/s (20°C, atm press.)

The total mass transfer in the system is obtained by multiplying the amount of component transferred in a single liquid circulation unit by the number of liquid circulation units.

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Notation

a = interfacial area per unit volume, m^{-1}
a' = interfacial area per bubble volume, m^{-1}
A_p = two-phase region area, m^2
A_s = mixing element area, m^2
A_T = area of tank, m^2
c_e = concentration in mixing element, kg/m^3
c_i = concentration of liquid flowing into circulation unit, kg/m^3
c_j = concentration in liquid phase of two-phase region, kg/m^3
c_j^* = concentration of saturated liquid, kg/m^3
d_b = bubble diameter, m
d_{bm} = bubble diameter calculated from Eq. 5, m
d_p = two-phase region diameter, m
D = diffusion coefficient, m^2/s
D_s = bubbling station spacing, m
D_T = tank diameter, m
g = acceleration of gravity, m/s^2
H_e = Henry's law constant, $N/m \cdot kg$
H_L = liquid height, m
k = number of velocity heads lost due to turning
k_G = overall mass transfer coefficient, $kg \cdot m^2/N \cdot s$
k_L = liquid phase mass transfer coefficient, m/s
L_T = square tank length, m
N = number of bubbling stations
N_b = number of bubbles
P = pressure, N/m^2
P_A = average pressure, N/m^2
P_G = power input by gas, W
P_o = pressure above orifice, N/m^2
P_s = pressure at liquid surface, N/m^2
q = liquid flow into mixing element for continuous liquid system, m^3/s
Q_g = gas flow rate, m^3/s
Q_{go} = gas flow rate at orifice, m^3/s
Q_L = liquid flow rate, m^3/s
R = universal gas constant
t = time, s
T = temperature, °C or K
U_s = gas superficial velocity in two-phase region, m/s
V_o = volume of bubble, m^3
v_b = bubble rise velocity, m/s
v_L = liquid velocity in two-phase region, m/s
V_I = volume of two-phase region, m^3
V_{II} = volume of mixing element, m^3
W_T = square tank width, m
y_j = gas phase mole fraction
$\bar{y}_j$ = average gas phase mole fraction
z = height, m

Greek letters

ρ = density, kg/m^3
ϕ = gas holdup
σ = surface tension, N/m^2
θ_I = holding time in two-phase region, s
θ_{II} = holding time in mixing element, s
θ_{tank} = tank holding time, s
μ_g = gas viscosity, $kg/m \cdot s$
μ_L = liquid viscosity, $kg/m \cdot s$

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Errata

In Table 1 of the R&D Note titled "P-V-T Relations of Propane in the Range 373.15–423.15 K and 0.2–4.0 MPa" (July, 1982, p. 695), the following column headings should be added across the table after 19 entries ending with 0.253 . . . :

Temp.: 403.15 K		Temp.: 413.15 K		Temp.: 423.15 K	
Pressure MPa	Compressibility Factor	Pressure MPa	Compressibility Factor	Pressure MPa	Compressibility Factor

In the same table, the compressibility factor at 1.074 MPa and 423.15 K is 0.9528, not 0.9697.

The price of the book titled "Mixing in the Process Industries" was given incorrectly as $87.95 in the review (Aug. 1986, p. 1403). It is $45.00.

In the paper titled "Diffusion and Mass Transfer in Supercritical Fluids," by P. G. Debenedetti and R. C. Reid [**32**(12), p. 2034, 1986], the following corrections are made:

p. 2035: First term of Eq. 11 should read "$\frac{2}{Re}(\nabla^{+})^{2}\underset{\sim}{v}^{+}$".

p. 2042: Line 12 should read "at any given temperature . . ."

p. 2044: In the Notation, Gz' should read "$x\mathcal{D}/\langle v\rangle\, b^{2}$."