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Revision Memo

12/01 Correction in Figure 2.


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SCOPE

This section covers the design of mixers used for gas-liquid contacting. This contacting is achieved by dispersing the gas in the liquid to increase the interfacial area for mass transfer. The overall objective of the contacting is the accomplishment of a chemical reaction or a gas stripping or absorption operation. Primary emphasis in this section is on the use of mechanical agitators.

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BACKGROUND

The efficient carrying out of a mass transfer operation is dependent upon establishing a large interface between the gas and liquid and then maintaining this interface long enough for the desired mass transfer to occur. The rate of mass transfer can be conveniently expressed in terms of:

$$r_A / V = k_L a \Delta C_A \quad \text{Eq. (1)}$$

where:

r_A	=	Rate of transfer of component A, lb/sec (kg/s) or moles/sec (k mol/s)
V	=	Volume of system, ft ³ (m ³)
k_L	=	Liquid phase mass transfer coefficient, ft/sec (m/s)
a	=	Interfacial area per unit volume, ft ⁻¹ (m ⁻¹)
ΔC_A	=	$(C_{A_i} - C_A)$ Concentration driving force (difference between concentration of Component A at the liquid-gas interface and the bulk liquid), lb/ft ³ (kg/m ³) or moles/ft ³ (moles/m ³)

This form is particularly appropriate when the gas is of low solubility in the liquid and "liquid film resistance" controls the rate of transfer. More complex forms which use an overall mass transfer coefficient, which includes the effects of gas film resistance, must be used otherwise. Also, if chemical reactions are involved, they are not rate limiting. The approach given here, however, illustrates the required calculation steps. Standard textbooks on mass transfer operations or chemical reaction engineering or Reactor & Fluid Dynamics Section of EMRE should be consulted for additional information.

The nature of the mixing or agitation primarily affects the interfacial area per unit volume, a . The liquid phase mass transfer coefficient, k_L , is primarily a function of the physical properties of the fluid. The interfacial area is determined by the size of the gas bubbles formed and how long they remain in the mixing vessel. The size of the bubbles is normally expressed in terms of their Sauter mean diameter, d_{SM} , which is defined below. How long the bubbles remain is expressed in terms of gas hold-up, H , the fraction of the total fluid volume (gas plus liquid) which is occupied by gas bubbles.

There are two approaches available for designing mixing systems for gas-liquid contracting. The first basically uses **Eq. (1)** and existing literature correlations for a and k_L . This approach explicitly uses agitation conditions (power, gas sparging rate, etc.) and fluid physical properties. It is particularly useful for assessing proposed changes in existing operations and in scaling between laboratory, pilot plant, and commercial results. This first approach can be termed "design based on mass transfer calculations." The second approach is entirely empirical and relates agitation conditions to the general description of the type of agitation desired for a given operation. It is useful for selecting the impeller type and sizing the mixing system. This second approach can be termed "design based on empirical scale of agitation."

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DEFINITIONS

Film resistance - Concept in a multi-phase mass transfer model which assumes the overall transfer rate to be controlled by transport through non-mixed films in the contacting phases on either side of the interface.

Gassed - The mixing condition prevailing when the presence of gas around an impeller reduces the power dissipation to the system.

Hold-up - Fraction of total fluid volume which is occupied by the phase under consideration.

Sauter mean diameter - Surface volume mean diameter defined for multiple bubble systems as:

$$\sum_j n_j d_j^3 / \sum_j n_j d_j^2$$

where: n_j = The number of bubbles of diameter d_j

Superficial velocity - The velocity of an individual gas or liquid phase at process conditions through a unit based solely on the cross-section of the unit without regard to the other phases present.

EQUIPMENT TYPES AND APPLICATIONS

Any type of equipment, which produces a large interfacial area between gas and liquid phases, can be used for gas-liquid mixing, contacting, or reaction. In addition to mechanically agitated vessels, possible equipment types include inline mixers and bubble, packed, and sieve columns. Mechanically agitated vessels produce backmixed contacting. Staging using separate vessels is required if other contacting patterns are desired. The other equipment types mentioned can in certain arrangements provide co-current plug flow or counter-current staged contacting.

Mechanically agitated vessels for gas-liquid mixing are commonly used in alkylation; Paramins, chlorobutyl, and neoacids production; and wastewater treatment. Arrangements based on inline mixers are also used in wastewater treatment. The calculation procedure presented in this section is directed toward the use of mechanical agitators. The same approach applies for other mixing devices. If the gas bubble sizes and interfacial areas generated by these other mixing devices are known, the mechanical agitation design procedure can be easily adapted for use.

Mechanical mixers for gas-liquid contacting normally use flat blade disk turbine agitators and the calculation procedures in this section are based on this style of agitator. Recently developed Chemineer CD6 impeller is also an efficient impeller for gas dispersion and can provide higher gas holding capability under certain conditions. The gas to be dispersed is usually added to the vessel through a ring sparger located below the agitator. For even gas distribution between sparger holes, the number and size of holes should be selected so that the gas kinetic energy through the holes is at least 10 times the gas kinetic energy through the pipe that forms the sparger ring. A method to design perforated distributors is given in [DP XIV](#).

DESIGN BASED ON MASS TRANSFER CALCULATIONS

To use **Eq. (1)** for design, the interfacial area, a , and mass transfer coefficient, k_L , must be calculated.

CALCULATION OF INTERFACIAL AREA

(Note that units for **Eqs. (2) to (6)** are in cgs and defined along with the equations.)

The interfacial area is dependent upon the bubble size and gas hold-up in the mixing vessel as given by:

$$\alpha = \frac{6H}{d_{SM}} \quad \text{Eq. (2)}$$

where: α = Interfacial area per unit volume, cm^{-1}
 H = Gas phase hold-up (volume of gas/volume of gas + liquid), dimensionless
 d_{SM} = Sauter mean bubble diameter, cm

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DESIGN BASED ON MASS TRANSFER CALCULATIONS (Cont)

The hold-up and bubble diameter in mechanical agitated vessels are given by:

$$d_{SM} = 4.15 \left[\frac{\sigma^{0.6}}{(P/V_L)^{0.4} \rho_L^{0.2}} \right] H^{1/2} + 0.09 \quad \text{Eq. (3)}$$

$$H = \left[\frac{18 v_s \mu_L H}{g \Delta \rho d_{SM}^2} \right]^{-1/2} + 0.0216 \left[\frac{(P/V_L)^{0.4} \mu_L^{0.2}}{\sigma^{0.6}} \right] \left[\frac{18 v_s \mu_L}{g \Delta \rho d_{SM}^2} \right]^{-1/2} \quad \text{Eq. (4)}$$

where: d_{SM} = Mean bubble diameter, cm
 σ = Surface tension between gas and liquid, dynes/cm
 P = Mixing power dissipation, ergs/sec
 V_L = Liquid volume, cm³
 ρ_L = Density of liquid, g/cm³
 v_s = Superficial gas velocity, cm/sec
 μ_L = Viscosity of liquid, poises
 g = Acceleration of gravity, cm/sec²
 $\Delta \rho$ = Density difference between gas and liquid phases, g/cm³

(Note that cgs units are used for convenience in the dimensionless groups. Also note that 1 HP/1000 gal = 1.97 x 10³ erg/sec. cm³.)

Eqs. (3) and (4) must be solved simultaneously and will require the solution of an equation of cubic form. These correlations are based on the gas phase being sparged into the mixing vessel. Gas dispersion from surface entrainment due to vortexing, etc., is not included. The mixing power dissipation must be corrected for gassing in accordance with **Figure 20** of **Section XIII-A**.

CALCULATION OF THE MASS TRANSFER COEFFICIENT

Two bubble diameter regimes must be considered when calculating the mass transfer coefficient. The first regime applies to bubbles having average diameters greater than 2.5 mm. Such bubbles are typically produced when pure liquids are aerated in mixing vessels or in sieve plate columns. In general, these large bubbles are termed to have unhindered interfacial flow and oscillation. The second regime applies to bubbles of average diameter of less than 2.5 mm. Such small bubbles are typically produced when aqueous solutions or impure liquids are aerated in mixing vessels or when most liquids are aerated with plates having very small perforations. For cases in which the liquid purity is not known, the small bubble regime should be assumed.

The correlations for the two regimes are:

unhindered interfacial flow

$$k_L = 0.42 \left(\frac{\mu_L}{D_L \rho_L} \right)^{-1/2} \left(\frac{\Delta \rho \mu_L g}{\rho_L^2} \right)^{1/3} \quad \text{Eq. (5)}$$

hindered interfacial flow

$$k_L = 0.31 \left(\frac{\mu_L}{D_L \rho_L} \right)^{-2/3} \left(\frac{\Delta \rho \mu_L g}{\rho_L^2} \right)^{1/3} \quad \text{Eq. (6)}$$

where: k_L = Liquid phase mass transfer coefficient, cm/sec
 D_L = Diffusivity of transferring solute in liquid, cm²/sec

If the diffusivity, D_L , needed for use in **Eq. (5)** or (6) is not known, it can be estimated from data or methods given in the Perry's Chemical Engineers' Handbook (Section 14 in 4th Edition or Section 3 in 5th Edition). Note that the calculation of the mass transfer coefficients for a given regime involves only physical properties and is independent of agitation conditions.