

Suspending Solids and Dispersing Gases in Mixing Vessels

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General principles of suspending solids in mixing vessels and the overall role of variables in gas absorption rates are presented in three categories of application

This discussion is concerned with three kinds of mixing operations:

Suspending solids in liquids

Dispersing gases in liquids

Suspending solids and dispersing gases simultaneously in liquids

Two different aspects of each of these operations will be considered:

Physical uniformity and physical appearance of the mixture

Mass transfer through an interface, either with or without chemical reaction in the liquid phase

Suspending Solids

The analysis of mixing processes is helped considerably if the description of the job to be done is accurately stated. Given below are five criteria which can help in more accurately defining the job to be produced in the tank.

Per cent suspension is the percentage of solids, either of the total solids compared to the total amount in the tank or of the weight of any particular particle size fraction compared to the total weight of this particle size fraction in the tank.

Per cent solids is the percentage of the total solids or of any particle size fraction, compared to the total weight of liquid and corresponding solid.

(1) Complete Uniformity

This implies that the per cent suspension at any point is 100%. The upper layer of liquid in the tank is the most difficult to bring to 100% suspension. It is difficult to get particles with settling velocities above 6 ft/min suspended uniformly in the upper 2% of the tank volume, since the primarily horizontal flow pattern at this point cannot keep high-settling velocity solids in suspension.

(2) Complete Off-Bottom Suspension

This is defined as all particles moving up off the tank bottom. It does not have any further requirement for a

particular per cent suspension at any other point in the tank.

(3) Complete Motion on Tank Bottom

This means that all particles are either suspended off the tank bottom or rolling around on the tank bottom. No statement is made concerning the per cent uniformity in the vessel.

(4) Filleting Permitted but No Progressive Fillet Buildup

A fillet is a stationary or stagnant deposit of solids most commonly at the outside periphery of the bottom where it joins the tank wall, but it could exist at any other part of the tank bottom depending upon the fluid flow pattern.

It is frequently less costly to let some of the solids settle out into fillets than it is to provide additional horsepower to eliminate the fillets, or to fabricate the tank to the contour of these fillets. These settled solids fillets should be nonprogressive. Fillets that are progressive can build up with time and eventually "sand in" the impeller if they do not remain stationary at some satisfactory point.

(5) Height of Suspension

The liquid height in the tank to which solids are suspended may be used to describe the operation. It is most commonly expressed as the per cent solids of each of the various particle size fractions at various liquid heights off bottom. This can also be expressed as the particle size distribution in samples taken at various points.

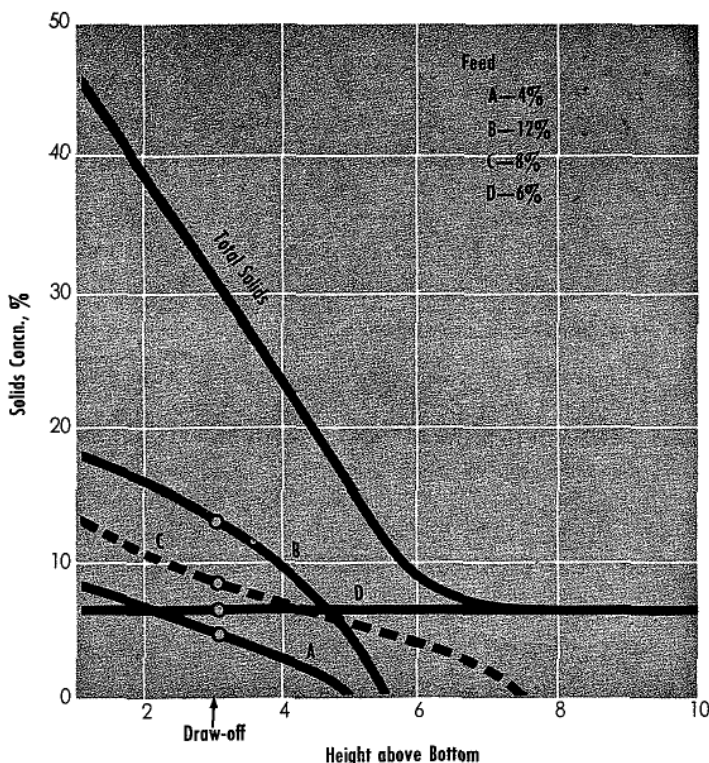


Figure 1. Concentration profiles in storage tank for particular draw-off position, partial solids uniformity

Figure 1 shows a concentration gradient in a tank that has complete off-bottom suspension, but is far from being completely uniform. The various fractions in Figure 1 are defined as:

Fraction A	-20 +42 mesh
Fraction B	-42 +80 mesh
Fraction C	-80 +150 mesh
Fraction D	-150 mesh and smaller

Continuous flow. When a slurry is continuously added and withdrawn from a single-stage mixer, the main requirement for steady-state operation is that the fillets must be nonprogressive and the discharge composition must equal the inlet composition. The discharge occurs at some particular draw-off point in the tank, so that it is the only point in the tank that must equal the inlet composition.

Figure 1 shows that the particle size distribution in the tank is quite different from the feed composition at power levels that do not provide complete uniformity.

In carrying out batch laboratory experiments to determine results in a continuous flow system with less than complete uniformity, the laboratory tank should not be charged with an initial weight of solids equal to the per cent of solids in the slurry. It must be charged with the proper per cent solids of each of the particle size fractions, so that the tank for batch observation is similar to the composition to be expected in the full-scale tank with continuous flow.

Mass transfer and leaching in continuous flow. In a tank that is completely uniform, the particle size distribution and concentration in the tank will be the same as it is in the feed and discharge stream. However, if the tank does not have complete uniformity, then leaching predictions must be based on the retention time characteristics that each of the selected particle size fractions has.

In Figure 1, the coarse fraction, -20 to +42 mesh, is approximately 4% in the feed and discharge. Its average per cent solids in the tank is only 2%, so that the average residence time of these solids is only one half of the calculated average retention time for the entire slurry in the tank.

In analyzing the mass transfer or the degree of leaching to be obtained, the actual residence time of each of the particle size fractions should be used to obtain the total performance of the system.

Figure 2 illustrates a continuous cocurrent multistage column for the contacting of liquids or solids. Superimposed on the flow velocity of the liquid phase is the additional settling velocity of the solid particles.

If sufficient agitation is provided for complete uniformity of the particles in each of the stages, then the residence time of the solids will be the same as the average retention time of the total slurry passing through the unit. If complete uniformity does not exist, then the average concentration of solids in the vessel will be lower than the feed and discharge, and the solids residence time will be less than the calculated slurry average retention time.

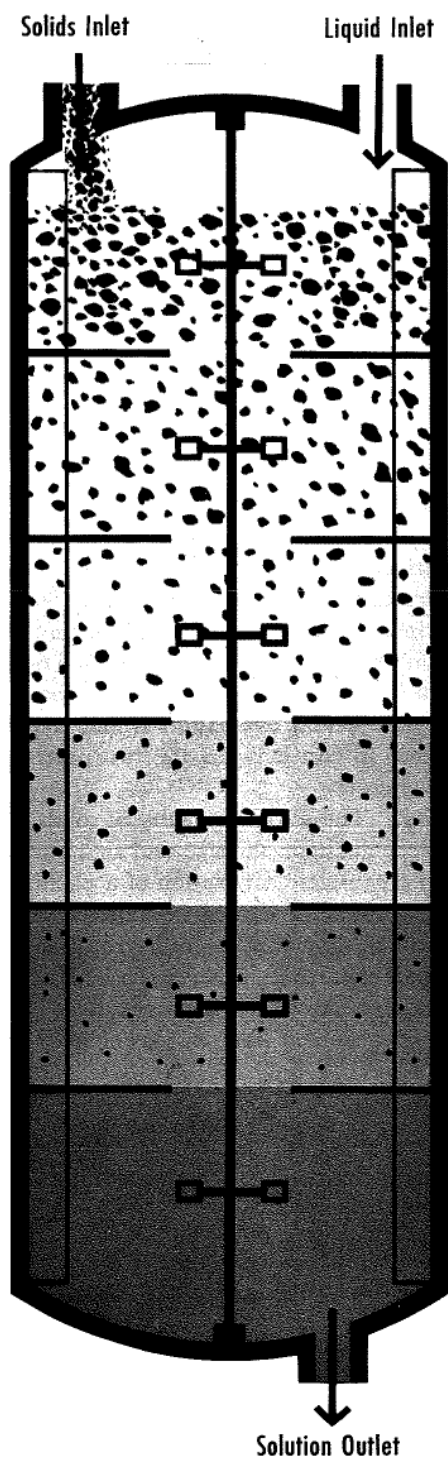


Figure 2. Schematic drawing of dissolver

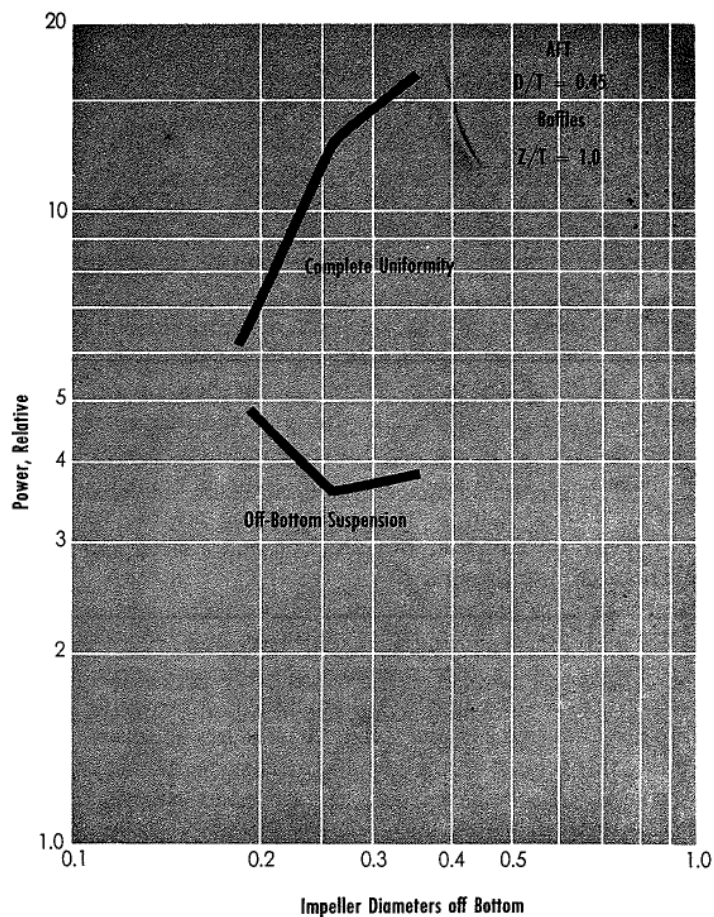


Figure 3. Effect of impeller off-bottom position on solids off-bottom suspension

Effect of off-bottom position. To illustrate the complexity of generalizing on the effect of mixing variables on solid suspensions, Figure 3 shows the effect of off-bottom position, C/D , for an 8-in. diameter impeller in a 17-in. diameter tank. The solids were 30% alunum with an average settling velocity of 5.0 ft/min. When the process criterion selected was off-bottom suspension, raising the impeller required more horsepower. When complete uniformity was the criterion, raising the impeller required less horsepower. The questions of optimum impeller off-bottom position, and many other similar geometric variables cannot be answered as generalities. The process requirement must be defined, as well as the settling velocity of the solids, before a statement can be made.

Several equations have been published in various sources for solid suspension. The definition of suspension must be carefully examined in each of these equations, since each gives somewhat different results (7, 8).

Height of suspension. In large slurry storage tanks involving 5 to 70 wt % solids, it is not always practical or necessary to have complete suspension of all the particles throughout the vessel at all conditions. It is common for the particle size distribution in slurries of coal and taconite to be such that the settling velocity is quite similar for all of the different particle size fractions.

At low mixer power levels, below complete uniformity,

there can be a zone in the tank where the per cent solids is quite uniform, but supernatant water will be on the top. The existence of this top layer of water does not affect the retention characteristics and blending uniformity of the slurry in suspension in the bottom portion. It might be argued that the tank and mixer are not doing their job by having this layer of water. Actually this extra volume is not contributing any effect to the process. However, at lower per cent solids, or at lower slurry levels, the supernatant water can be removed and incorporated into the slurry.

In terms of a general concept, a given mixer with a particular horsepower, impeller diameter, and impeller speed can suspend a certain volume of solids at a particular per cent solids regardless of whether it is providing complete uniformity in one tank, or whether it is suspending this volume of solids with partial suspension in a larger tank. In the larger tank, the fillets in the corners must be nonprogressive, and clear liquid may be on the surface.

Effect of impeller diameter, D/T ratio. To provide complete uniformity through a slurry, usually the larger the impeller size-to-tank size ratio (D/T) used, the less power is required. Large, slow-speed impellers require a lower horsepower for a given pumping capacity, and solid suspension is typically governed by circulation rate in the tank.

Figure 4 shows a typical curve for illustration purposes only. The actual slope and boundaries of these curves vary depending upon the concentration and settling velocity of the solids, and also on the particular definition of uniformity required.

The bigger the impeller and the lower the horsepower, the more economical the mixer might seem; however, the torque required to turn these large impellers is often

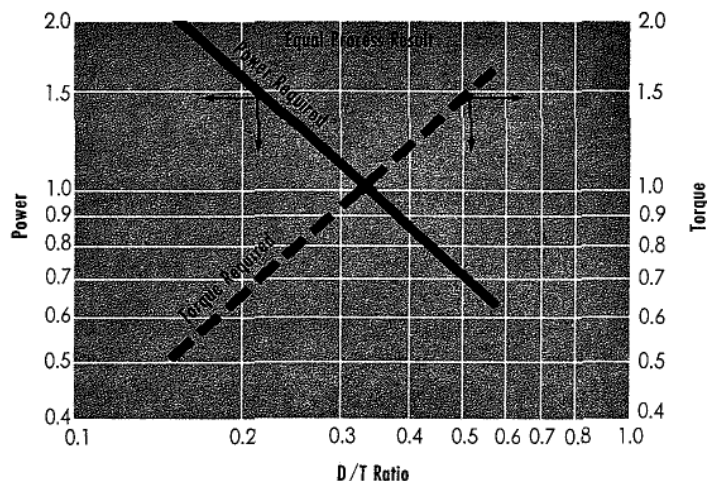


Figure 4. Change in mixer power and mixer torque for a typical low viscosity, solid-suspension application

TABLE I. TYPICAL COST ANALYSIS

Cost for 75-hp mixer compared to 60 hp^a

	1 Year, 8 hr/day	1 Year, 24 hr/day	5 Years, 24 hr/day	10 Years, 24 hr/day
Initial saving	\$1500	+\$1500	+\$1500	+\$1500
Power cost ^b	-\$ 350	-\$1050	-\$4500	-\$7000
	+\$1150	+\$450	-\$3000	-\$5500
Saving for 75 hp		Saving	Loss	Loss for 75 hp

^a A 75-hp mixer has smaller turbines and a smaller size speed reducer than a 60-hp unit. Therefore, a 75-hp mixer costs \$1500 less than a 60-hp mixer in this example.

^b Electric power cost is \$70/year/hp for a 24-hr/day operation.

greater even though the power required is less. The initial cost of the mixer is governed largely by the torque required for the drive.

Table I illustrates a typical cost analysis. For a short period of time, the 75-hp less expensive mixer is more desirable, but for longer periods of time, the 60-hp unit, even though it is initially more expensive, requires less total cost. The actual evaluation would depend upon the difference in cost between the two units and the allocated power costs.

Settling velocity. The settling velocity of solids in a particular fluid depends upon the size and shape of the solid particle. If a complete correlation of a particular type of solid over a range of settling velocities is given, curves similar to those shown in Figure 5 are obtained. Three different kinds of particles, spheres, crushed galena, and crushed silica, are shown. There are two properties which, if determined for a particular solid,

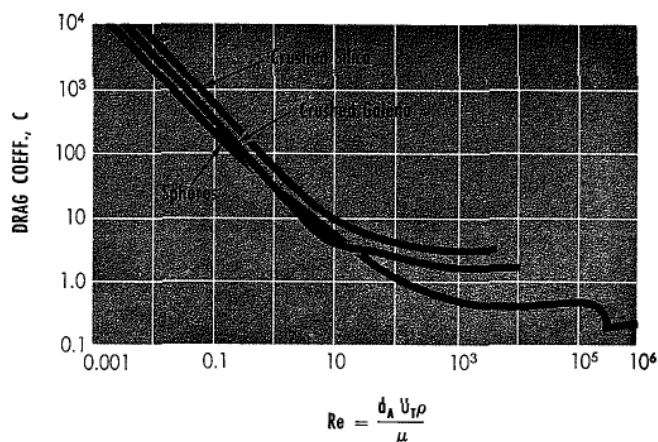


Figure 5. Drag coefficient of solid particles for three particular types of solids

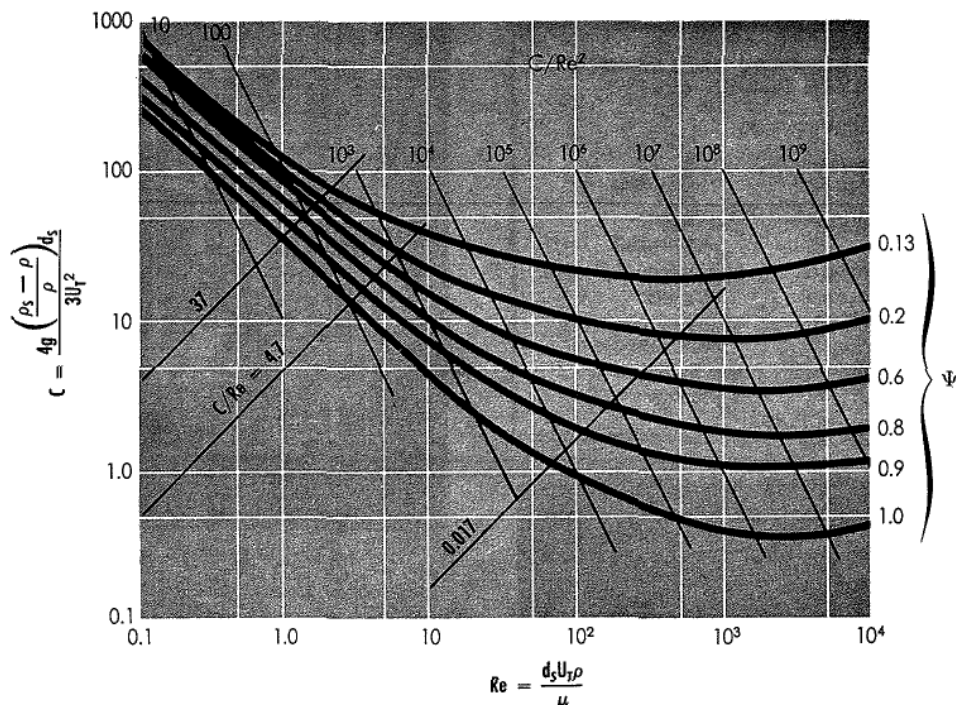


Figure 6. Drag coefficient for solid particles related to actual particle size diameter and sphericity

allow the use of general sizing curves and the prediction of settling velocities over a wide variety of conditions.

There are reliable data on the settling velocity of spheres. Two other parameters are needed to describe particle shape (2). One of these is the screen-size diameter of the particle, d_A , in a ratio to the diameter of a sphere having the same volume as the particle, d_s . As a general rule, the screen size, d_A , approximates the second largest dimension of the particle.

The next parameter is the sphericity, ψ , which is the surface area of the particle divided by the surface area

of a sphere of the same volume as the particle. Figure 6 illustrates the correlation.

For a given type of solid, crushed by the same equipment, it is possible experimentally to determine the sphericity and the diameter ratio, d_A/d_s . Two or three different-sized particles are used, and after measuring the settling velocity, proper cross-plotting yields the sphericity and the diameter ratio, d_A/d_s , as a unique relationship.

It is desirable to have one or two particle sizes in the turbulent region and one or two particle sizes in the viscous region to obtain these values.

To find the settling velocity of a particle of known d_A/d_s and ψ from the basic plot shown in Figure 6, curves of constant C/Re^2 can be used to facilitate what would normally be a trial and error solution. For the purpose of establishing the sphericity and d_A/d_s ratio from a known settling velocity and d_A , it is convenient to use the curves of constant C/Re calculated from the relationship

$$C/Re = \frac{490(\rho_s - \rho)}{3U_T^3}$$

The settling velocity data for any given particle size allow calculation of an infinite number of combinations of the ratio of d_A/d_s and the sphericity, ψ . Several of these values were plotted in Figure 7 for that particle size. Plotting these for each of the two or three particle sizes used will give an intersection at the best value of sphericity and d_A/d_s ratio.

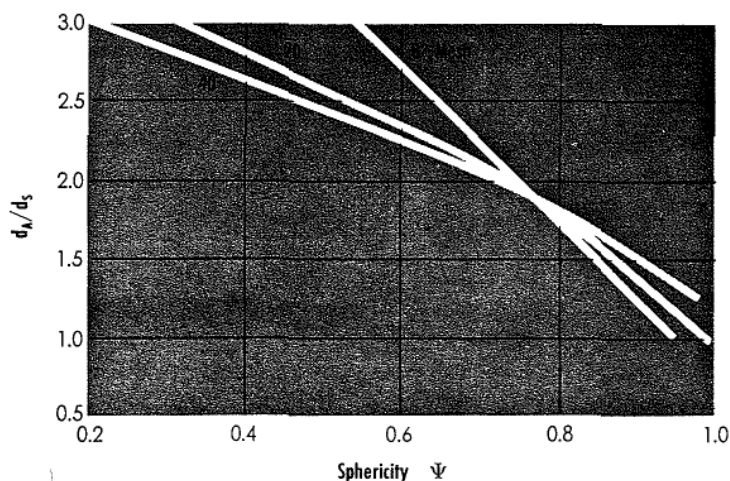


Figure 7. Cross plot of various d_A/d_s ratios vs. sphericity which will satisfy settling velocity data from three different mesh-size particles. Intersection gives unique value for both properties

TABLE II. LIQUID SURFACE APPEARANCE AT VARIOUS MIXER POWER LEVELS

Constant gas rate		
Mixer power	Surface appearance	Dispersion
Low	"Blurps"	Gas rises unhindered
increases	Uniform	Gas distributed to walls
↓	Uniform	Mixer flow pattern predominates
High	Uniform	Gas holdup plateau

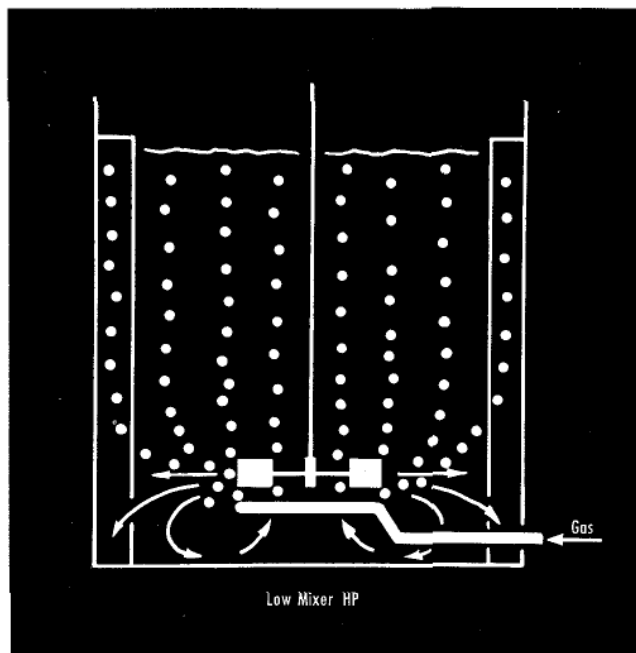


Figure 8. Typical flow pattern of gas bubbles for relatively low mixer horsepower compared to gas rate

Gas-Liquid Dispersion

Flooding. The question of what speed is needed for an impeller to disperse a given quantity of gas does not have a unique answer. For example, assume that a certain volume of gas is being bubbled through a tank from a sparge ring and the impeller speed is increased from zero up to a higher value. With the gas streaming out of the sparge ring and the impeller at zero speed, the gas flow pattern predominates. The horsepower transmitted to the liquid by the expanding gas, hp_G , can be calculated from pressure change in the expanded gas.

Referring to Figure 8 and Table II, as the mixer speed is increased to a point where the power input from the mixing impeller is approximately equal to the horsepower transmitted by the gas to the liquid, the mixer

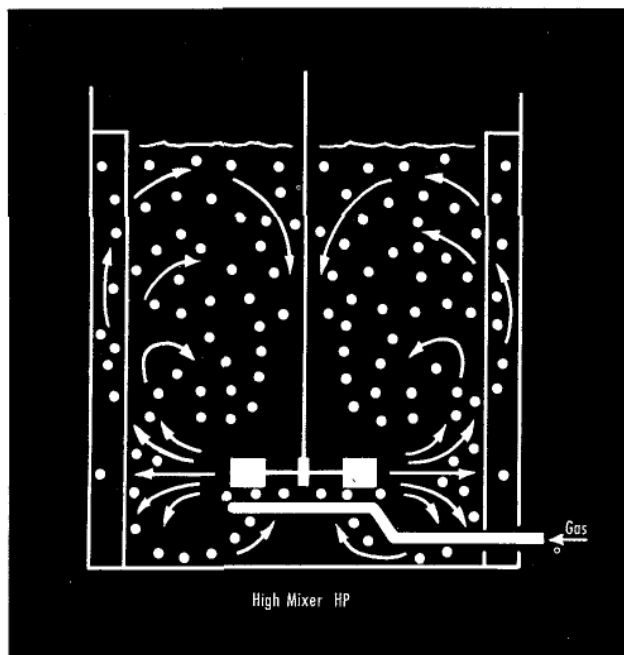


Figure 9. Typical gas bubble flow pattern for relatively high mixer horsepower for given gas rate

can disperse the gas stream, causing the majority of the gas to escape with relatively "Small" bubbles at the surface. There can be some gas bubbles that erupt as geysers. The liquid motion in the tank is still determined predominantly by the upward gas velocity.

When the mixer power level becomes several times higher than the gas horsepower, the gas is dispersed out to the side walls of the tank and then on up to the surface. The mixer flow pattern now predominates, giving the typical flow pattern of the radial flow turbine illustrated in Figure 9.

The mixer power level must be increased higher than this to drive the gas down into the bottom areas of the tank and give a still more intimate dispersion of the gas.

At extremely high power levels from the impeller, the gas holdup reaches a plateau of about 20 to 30% by volume when in a nonfoaming liquid.

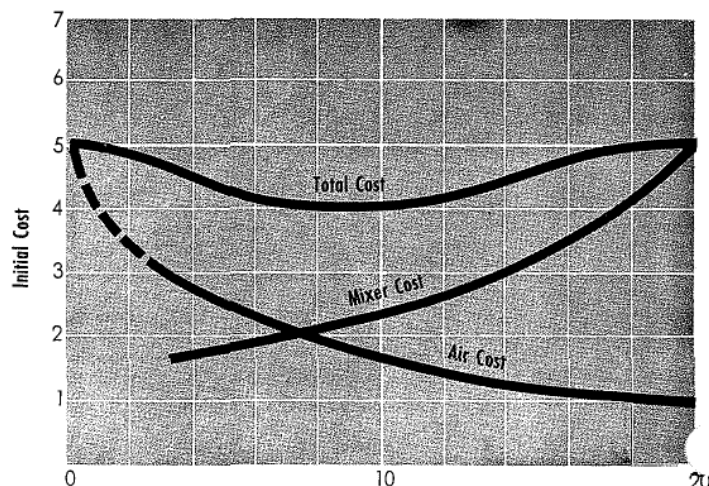


Figure 10. Total cost of air system and mixer for various ratios of mixer horsepower and air horsepower

TABLE III. EFFECT OF TANK SHAPE ON MIXER AND AIR HORSEPOWER REQUIREMENTS

Constant tank and air volume and constant absorption rate

Liquid level ungassed, ft	Air hp	Mixer hp
10	5	62
20	10	30
40	14	13

Any one of these conditions can be utilized in practical applications. Any increment of impeller power increases the absorption rate from the gas stream and is a practical condition. At about equal gas power and mixer power (5), the maximum absorption rate per unit total gas and mixer horsepower is often achieved, but at relatively low absorption rates per unit volume of liquid (Figure 10). At higher mixer power levels, there are much higher gas absorption rates per unit volume of liquid, but at a somewhat decreased absorption rate per unit of total horsepower of gas and mixer.

Mass transfer rate. The rate of mass transfer is given as

$$\text{Rate} \propto K_L a (\Delta \text{conc.})$$

The driving force is calculated from the total pressure at the bottom of the tank, the partial pressure of the gas phase at the top of the tank, and the dissolved gas concentration in the liquid.

For a given liquid volume, it is normally possible to get a given mass transfer quantity at a lower total power level in relatively tall, small-diameter tanks. Table III illustrates the parameters involved. The final decision, of course, rests on the cost comparison for the mixer, air compressor, and tank construction.

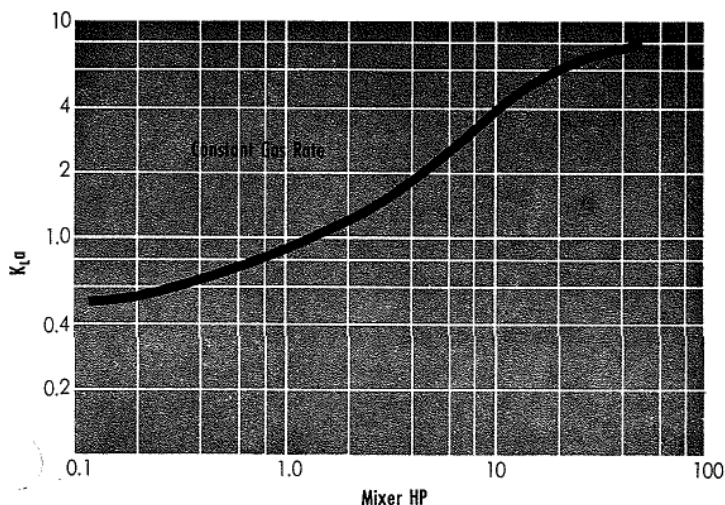


Figure 11. Typical graph for change in absorption coefficient from low mixer horsepower to high mixer horsepower levels at constant gas rate

The smaller diameter, taller tank has a higher superficial gas velocity than the larger diameter, shorter tank. If the mixer power required for the absorption rate is lower than that required for physical dispersion in the tall, small-diameter tank, then the above comparison in Table III cannot be used.

On scale-up, if equal volumes of gas per minute per unit volume of liquid are used, higher superficial velocities result in full-size equipment. This can mean a saving of impeller horsepower level as long as minimum horsepower levels for adequate gas dispersion are not violated.

The disk on an impeller usually helps to keep gas from going up to the center of the impeller and gets it out into the higher fluid shear zone of the unit.

Figure 11 shows that there is a change in the slope of the gas absorption coefficient *vs.* mixer power at constant gas rate from a lower slope to a higher slope (4, 6). This change in slope usually occurs at a mixer power level two or three times higher than the power input to the liquid from the gas stream.

For most gas absorption requirements, there are different combinations of mixer horsepower and gas horsepower that will give the same absorption rate (Figure 12).

Scale-up

Scale-up involves consideration of dimensions, velocities, and fluid forces. If the final parameter of interest is any one of these three variables, then the use of dimensionless or dimensional relationships involving these three quantities is all that is needed. For example, the power consumed by a mixing impeller is related to the fluid forces in the mixing vessel, and the use of a power number-Reynolds number correlation is a powerful tool.

However, if the mixing process depends on fluid shear stresses, then some additional relationships involving other parameters, as well as other chemical and fluid properties, are needed.

The fluid force parameters are well given by groups involving gravity forces, inertia forces, and viscous forces. However, for process scale-up there are many other kinds of ratios that are often investigated, are important, and are often used in correlations. Moreover, it is not possible, as a rule, to keep more than one parameter ratio constant if dimensional similarity is used.

It is best to consider that a particular parameter is to be "controlled" on scale-up, rather than being held constant, since in studying any one of the many hundreds of different mixing processes, there is usually no reason to assume that a given parameter should be constant on scale-up. It is normally more rewarding to look at the question from the viewpoint, "What is the relationship between tank size and the correlating parameter to give a particular process result?" (Table IV).

If it turns out that one of these parameters is constant,

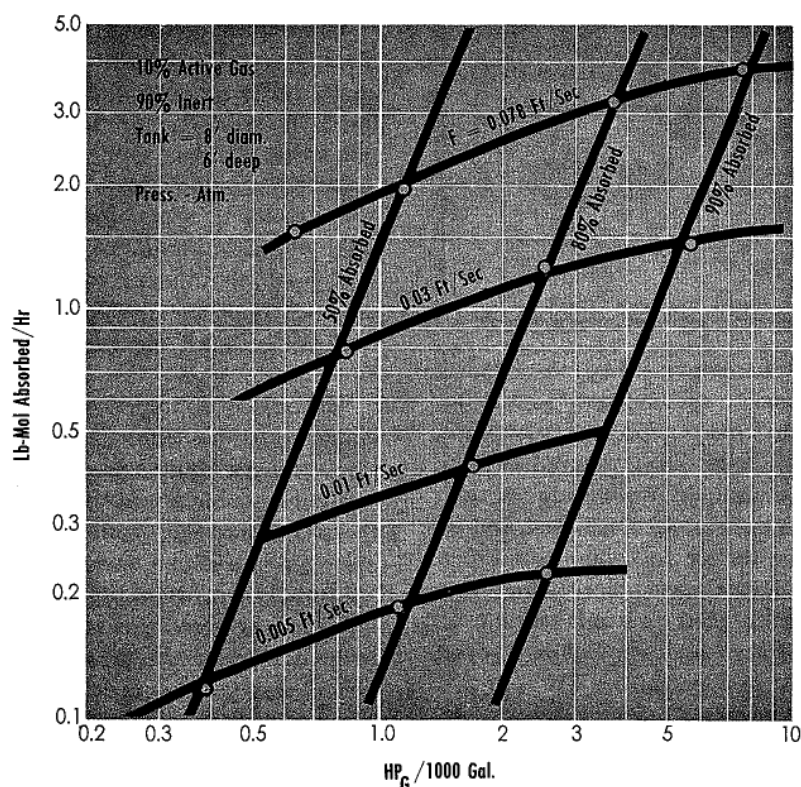


Figure 12. Plot of mass transfer as function of power and air flow for CO_2 absorption

TABLE IV. PROPERTIES OF FLUID MIXER ON SCALE-UP

Property	Pilot, 215 gal	Plant scale, 3200 gal, $T = 2.5$			
P	1.0	15.6	98	6.2	15.6
P/vol	1.0	1.0	6.2	0.4	1.0
N	1.0	0.54	1.0	0.4	0.26
D	1.0	2.5	2.5	2.5	3.9
Q	1.0	8.5	15.6	6.2	15.6
Q/vol	1.0	0.54	1.0	0.4	1.0
ND	1.0	1.4	2.5	1.0	1.0
$\frac{ND^2\rho}{\mu}$	1.0	3.4	6.2	2.5	1.57
D/T	0.33	0.33	0.33	0.33	0.52

so much the better. For most process scale-up considerations covering a wide spectrum of operations, the tendency is for the maximum shear rates to increase with scale-up, while the average shear rates decrease (3) (Figure 13). This means that the particle size and bubble size distribution would probably not be the same on full scale as on small scale if the bubbles are subject to deformation or degradation by either the average

or maximum fluid shear stresses. There are relatively few mixing processes that depend on knowing in detail this complex relation, but several examples will illustrate the importance:

(1) In non-Newtonian fluids, the average shear rate as represented by the impeller speed is sufficient to estimate the viscosity the impeller "sees."

(2) In emulsion polymerization, bubble degradation is controlled by maximum shear rates, while particle size distribution and the average particle size can be adjusted by other mixing parameters.

(3) Fermentation involves control of the maximum fluid shear rate to give optimum liquid-gas dispersion, yet prevents biological solids damage, while other mixing parameters are adjusted to get the proper overall mass transfer, blending, and suspension.

Circulating Capacity and Fluid Shear Rate

It is helpful to restate the relation that the fluid motion in the tank can be characterized into two components of circulating capacity and impeller head. The impeller head is related to the fluid shear rate. This fluid shear rate is a velocity gradient. Multiplying the fluid shear rate by the viscosity of the fluid at that shear rate, the fluid shear stress results:

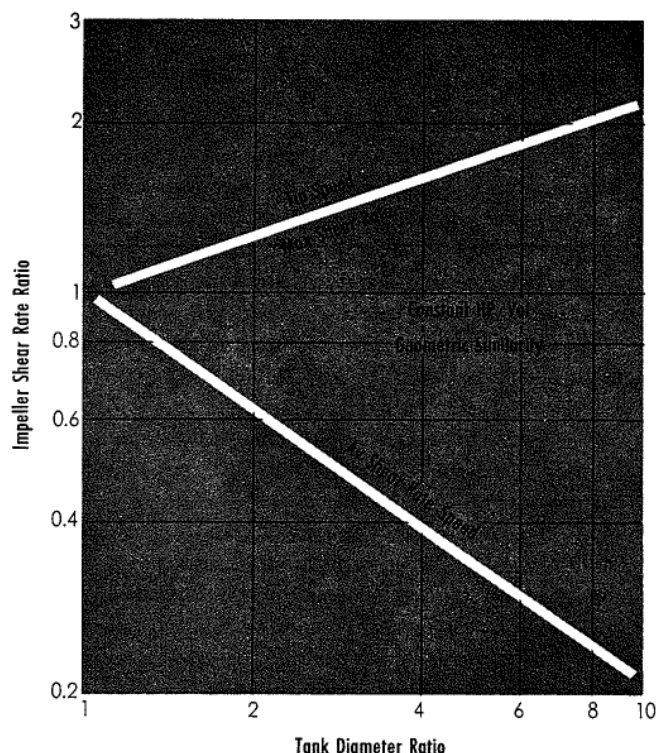


Figure 13. Effect of tank diameter scale-up on impeller shear-rate ratio

Fluid shear stress = viscosity (fluid shear rate)

If the fluid velocities are fluctuating, so are the fluid shear rates, and the fluid shear stresses are also going to be fluctuating. The next concern is the effect of the size of the particle being operated on by these shear stresses. The total velocity gradient across the particle is more meaningful than the individual shear stress existing at any point on the particle. One analogy is that we do not use an ultrasonic drill to drive a rivet, nor do we use a rivet hammer to drill a tooth.

There are also two other effects that must be considered. One is the average shear rate existing in the zone of the impeller and the other is the maximum shear rate existing in the zone. They must be of a size and frequency which the particles or bubbles can "see." It turns out that the shear rate, from the standpoint of defining the viscosity around the zone of the impeller, is the average shear rate and is related only to the impeller speed, regardless of impeller size, for a particular geometric series (3). It also turns out that the maximum shear rate which is experienced by most bubbles and particles in the range of 10 to 200 mesh, is a function of both the impeller diameter and impeller speed. This presents the characterization on scale-up with geometric similarity shown in Figure 13.

Table IV shows what can happen to several commonly used parameters on scale-up, if any one of these is held constant.

When there are particles or bubbles whose sizes and shapes are affected by the fluid shearing stresses in the tank, we have to consider the role of the fluid mixing in producing fluid shear rates and the mechanism for cascading of the fluid shear stresses generated.

Scale-down

To study a process that is operating full scale, two principles must be considered:

(1) If geometric similarity is used in scaling down to a small size pilot vessel, the Reynolds number of the pilot unit will normally be much lower than full scale, which means that the effect of viscosity on flow pattern can be different from what it was in the full-scale unit.

(2) Maximum shear rates will be much lower in the pilot plant than they were in the full size unit. If the action of any of the chemical ingredients in the full-size system depends on maximum shear rates, their effects would not show up in the same fashion in the small-size unit.

We then use a nongeometric scale-down, which means, in general, that the small-size unit should have a smaller blade width and/or a smaller D/T to achieve similar shear rates with other conditions constant. Table V illustrates an example of this scale-down.

Examples of Gas-Liquid-Solid Mixing

To illustrate the use of these principles in processes where gases, liquids, and solids are present, three examples are chosen. Practical industrial examples are carbonation of lime slurry, chlorination of paper stock, submerged aeration of fermentation broth, and hydrogenation of vegetable oil in the presence of a solid catalyst. When a new gas-liquid-solid process comes up, the principles previously discussed can help to determine the proper pilot plant program and scale-up.

TABLE V. PROPERTIES OF FLUID MIXER ON SCALE-DOWN

Property	Plant scale, 2500 gal	Pilot scale, 3.4 gal			
P	1.0	0.00137	0.0022	0.0022	0.0022
P/vol	1.0	1.0	1.6	1.6	1.6
N	1.0	4.33	5.07	6.4	10.1
D	1.0	0.11	0.11	0.097	0.097
Q	1.0	0.006	0.007	0.006	0.004
Q/vol	1.0	4.3	5.1	4.3	2.7
ND	1.0	0.48	0.56	0.62	1.0
$ND^2\rho$	1.0	0.07	0.08	0.06	0.09
μ					
D/T	0.35	0.35	0.35	0.30	0.30
D_w/D	1.0	1.0	1.0	1.0	0.25

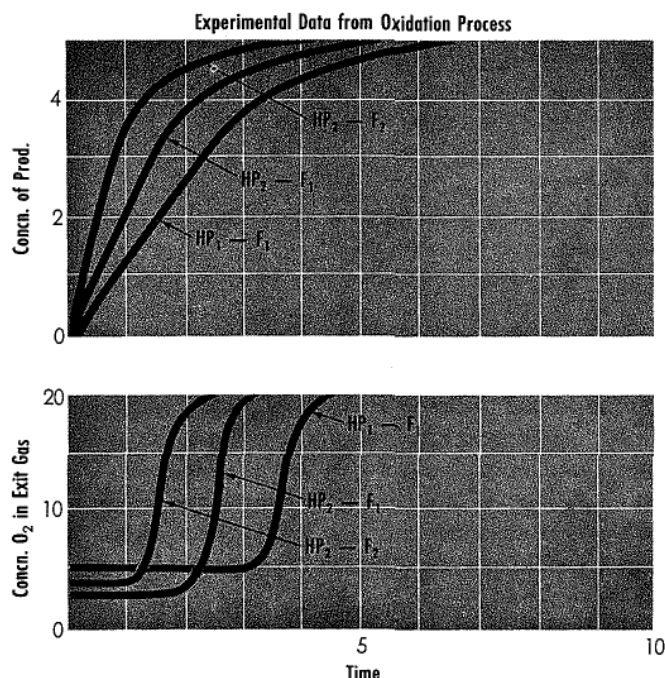


Figure 14. Experimental data on batch oxidation process at three different mixer horsepower and gas rate combinations

Two general principles for the pilot plant are:

(1) Several batch runs, regardless of whether the process is to be batch or continuous, can shed light on the relative effect of mixing variables. For continuous processes they can set the proper stage for the evaluation of continuous pilot planting and/or continuous full-scale operation.

(2) For gas-liquid absorption, with or without solids, three or four runs varying the mixer power level at a given gas rate and then a higher and lower gas rate at one of these power levels is normally a good basic series of runs.

Example 1: batch pilot plant. A slurry is contacted batchwise with a continuous flow of air. Three experimental runs out of a series of batch runs is shown in Figure 14.

In this study, varying the horsepower level and then the gas rate markedly affects the reaction rate in the initial stage of the reaction. This indicates that the chemical reaction is initially fast, and is being limited by the gas-liquid mass transfer rate.

The fact that gas rate markedly affects the first part of the reaction at constant mixer horsepower indicates that the gas-liquid mass transfer step is controlling rather than the solid-liquid. Changing gas rate doesn't normally affect liquid-solid mass transfer steps.

As the reaction nears completion, it slows down to a point where chemical reaction controls and mixer horsepower and gas rate do not have any further effect.

The overall conclusions are:

(a) Liquid-solid mass transfer rate does not have a marked effect. The effect of horsepower and gas velocity in the initial stage is too high for liquid-solid mass transfer rate to be controlling. The negligible effect of horsepower and gas rate in the final stage is too low for liquid-solid mass transfer.

(b) The first part of the process is controlled by gas-liquid mass transfer and mixer and gas variables have a very marked effect.

(c) The second part of the process is controlled by a chemical reaction. Horsepower and gas flow do not have any effect.

(d) For full-scale continuous flow operations, this would indicate the possibility of carrying out first the mass transfer controlled part of the operation in one tank at high mass transfer rate, and then carrying out the second portion of the reaction in one or more tanks with appropriate mixer design in each.

Example 2: single-stage reactor. A slurry is passed continuously through a single tank. A gas containing a reactant is passed through the slurry. Figure 15 indicates the type of data obtained with this single-stage continuous reactor when it is operated to give a constant effluent concentration of the product. The flow rate needed to produce this product allows the reaction rate to be calculated.

Figure 15 shows curves drawn through four or five data points at different horsepower levels at a constant gas rate. At a superficial gas velocity of 0.03 ft/min, the curve has a typical slope for gas-liquid mass transfer controlled processes, until the slope changes at high power levels to one more typical of liquid-solid mass transfer processes (7).

To confirm this speculation, another series of runs is made at 0.06 ft/min. This gives a slope typical of a liquid-

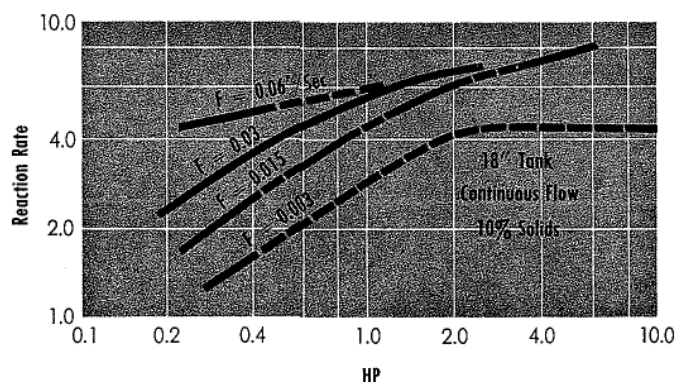


Figure 15. Typical reaction rate data from continuous process at various mixer horsepower levels and gas rates. Slurry flow rate to give a constant product composition

liquid mass transfer controlled process at all horsepower levels. The gas-liquid mass transfer step is higher than the solid-liquid mass transfer step at all power levels shown. One characteristic of liquid-solid mass transfer controlled processes is that gas rate has a negligible effect on the solid-liquid mass transfer rate.

At a lower gas rate, 0.015 ft/min, the interpretation is further confirmed by the resulting curve for a typical gas-liquid mass transfer controlled slope at low power levels and then merging into the liquid-solid mass transfer controlled rate at high power levels.

At extremely low gas rates, 0.003 ft/min, lower horsepower shows a typical gas-liquid mass transfer controlled slope, but then the curve flattens out when the stoichiometric amount of gas has been reacted, limiting the reaction rate.

From these data, the economics of different reactor volumes, horsepower levels, and gas rates can be examined, and proper scale-up techniques can be applied depending upon which range is selected for final process design. Scale-up techniques for gas-liquid mass transfer controlled processes are quite different from scale-up techniques for liquid-solid mass transfer controlled processes.

Example 3: batch scale-up. A batch slurry pilot plant study in a tank 18 in. in diameter and of 20-gal ungassed liquid was carried out with a head pressure of 5 psig. The full-size tank, 12 ft in diameter and 30 ft high, had a 15-psig head pressure. Table VI gives the details.

The first scale-up consideration used an equal volume of gas per volume of liquid per minute. The power level required to satisfy the gas-liquid absorption alone was too low to give good physical gas dispersion for the superficial gas velocity in the full-scale tank.

The exit concentration of gas showed that the volume of gas per volume of liquid could be reduced on scale-up and still stay within reasonable limits of per cent of active gas absorption. The scale-up to a second design at higher power level and lower gas velocity to give the same gas-liquid absorption rate was judged to be a practical operating unit since the impeller horsepower is sufficient physically to disperse the gas and have a smooth, nonpulsating gas dispersion flow pattern.

This scale-up involves geometric similarity, and it has been pointed out previously that the maximum shear rate in the impeller zone tends to go up. In this case, it was approximately 70% higher and was thought to have no effect on the solid particles present since they were not affected by fluid shear rates in this range. However, if maximum impeller fluid shear rates were to be reduced in the larger equipment, a change in either impeller size-to-tank size ratio or impeller blade proportions would be necessary.

TABLE VI. BATCH ABSORPTION SCALE-UP

Properties, tank	Pilot	Plant	
		A	B
Top head pressure, psig	5	15	15
T, tank diam	18"	12'	12'
P, impeller hp, gassed	0.020	52	90
D, impeller diam, nominal	6"	48"	48"
No. of turbines	1	2	2
Z, liq. level, gassed	18"	22'	22'
Volume, gal gassed	20	19,000	19,000
Volume, gal ungassed	18	15,000	16,000
V_G/V_L /min	1.6	1.6	0.7
Air flow, SCFM	4.2	3,500	1,600
F, ft/sec	0.03	0.22	0.1
Properties, process, relative values			
Mass transfer rate	1.0	1.0	1.0
K_{LA}	1.0	0.62	0.81
Δ Conc.	1.0	1.6	1.2
Max. fluid shear rate in turbine jet	1.0	1.4	1.7

Nomenclature

AFT	= axial flow turbine
c	= drag coefficient
C/Z	= impeller off-bottom to liquid depth ratio
d_A	= screen diameter size of particle
d_s	= diameter of sphere of equal volume as the particle
D	= impeller diameter
D/T	= impeller diameter to tank diameter ratio
D_w	= width of blade (projected vertical height)
D_w/D	= blade width to impeller diameter ratio
F	= superficial gas velocity at average temperature and pressure, ft/sec
g	= gravitational acceleration
hp	= horsepower
hp _G	= gassed horsepower
K_{LA}	= mass-transfer coefficient
N	= impeller speed
P	= power
Q	= volumetric fluid displacement of impeller
Re	= Reynolds No. for solid particles
T	= tank diameter
U	= overall heat transfer coefficient
Z	= liquid depth
Δ Conc.	= av concentration driving force, ppm
μ	= viscosity of liquid
ρ	= density of fluid
ρ_s	= density of solid particles
ψ	= sphericity of particle

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