Short Term Training Programme
On

"Process Engineering: Agitation & Mixing"
Conducted by
Anchor Institute
(Chemicals & Petrochemicals)
Promoted by
Industries Commissionerate
&

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GOVERNMENT OF GUJARAT

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Preface :

The Anchor Institute Chemicals & Petrochemicals, Dharmsinh Desai University, Nadiad has been actively involved in fulfilling its objective of designing and implementing Industry responsive courses considering the need of Industry in this sector right from the beginning. The courses are designed in consultation with Industry and Academia. We have also prepared teaching, learning and reference material for all these courses for the use of Faculty members. Anchor Institute DDU has conducted 42 training courses involving participants 720 from Industry, 447 faculty members and 1186 students till October 2012. This consists of 21 different subject Programmes as under.

1. CDM and Carbon Trading in India.
2. METLAB and its application.
3. Prevention and Management of Chemical Hazards & accidents.
4. Certificate course for Chemical Plant Operator.
5. Scheduling & Optimization of Process Plants.
6. Process Engineering (Debottlenecking & Time Cycle reduction).
7. Energy Conservation in Chemical Industry.
8. Advance Process Control Dynamics and Data analysis.
9. Process Engineering (Agitation & Mixing).
10. Use of animations in Chemical Engineering – Effective teaching learning process.
11. Process Engineering (Vacuum Technology).
12. Repair and Maintenance of Chemical Plant & Equipment.
13. Industrial Chemical Technology.
14. Energy conservation & audit.
15. Stoichiometry for Chemical process plants.
16. Chemical Process Simulations – Application of software in Chemical Industry.
17. Performance enhancement of ETP.
18. Chemical Engineering to non-chemical Engineers.
19. Environment Management System ISI 14001-2004.
20. Quality Management System ISO 9001-2008.
21. Piping systems in Chemical Industry.

It is our pleasure to present teaching, learning course material on **"Process Engineering- Agitation & Mixing"**

Dr. P. A. Joshi   Mr. S. J. Vasavada
Chairman,   Associate Coordinator,
Anchor Institute, DDU   Anchor Institute DDU

Anchor Institute- Chemicals & Petrochemicals Sector

Gujarat has witnessed rapid industrialization in the last two decades and has evolved as hub for Chemicals and Petrochemicals. In fact the State has become the *petro capital of the country*. Department of Chemicals & Petrochemicals under the Chemicals & Fertilizer Ministry of Government of India has signed memorandum of agreement with the Government of Gujarat to set up a *Petroleum, Chemicals and Petrochemicals Investment Region (PCPIR) in the state* at Dahej with estimated total proposed investment of Rs.50, 000 crore and expected to provide employment to 8 lakh people that include 1.9 lakh of direct employment over a period of time.

Hence, developing the man power on a massive scale for this sector is the prime issue as realized by the *Industries Commissionerate, Government of Gujarat (I.C., GOG).* The need for better quality and skilled technical manpower is increasing and will continue to increase in time to come. It therefore, has decided to tackle this issue and took proactive approach through *the industry responsive Training Courses and Skill Development Programmes.*

We are pleased to inform you that *I.C., GOG* entrusted *DDU* to take up the challenge to be an *Anchor Institute* for the fastest growing *Chemicals & Petrochemicals sector of the state*. Its Associates are L. D. College of Engineering, Ahmedabad as Co Anchor Institute, N. G. Patel Polytechnic, Afwa, Bardoli and ITI Ankleshwar as Nodal Institutes.

The objective of the Anchor Institute and its Associates is to take various initiatives in creating readily employable and industry responsive Man Power, at all level for Chemicals & Petrochemicals sector across the State.

To achieve the Objective **our major proposed activities ahead are as under**

- Identifying the training courses & skill development programs as per the need of the Chemical & Petrochemical Industries in Gujarat state for ITI, Diploma & Degree Level faculty members & students, SUCs, people in the industries, unemployed persons who are seeking jobs in this sector etc.
- Organizing faculty development programs (training for trainers)
- Mentoring and Assisting the Nodal Institutes.
- Benchmarking of the training courses
- Up-grading the Courses offered in Chemical & Petrochemical Engineering and make them Industry responsive.
- Identifying new and emerging area in this field.
  With above activities, we expect that following will be *the major beneficiaries*
- Unemployed technical manpower having completed the formal study
- Technical manpower already in job
- Faculty members and students of the technical institutions
- Chemicals & Petrochemicals industries

To accomplish this task, we involve the Experts from the Industries, consulting companies, , Engineering Companies & Trainers well known in this Sector.

Prof. (Dr.) Shirish L. Shah, University of Alberta, Canada Addressing at A.P.C. Course module II at DDU Nov 21-26, 2010.

Course Material Process Engineering: Agitation & Mixing

Preamble

Agitation is a means whereby mixing of phases can be accomplished and by which mass and heat transfer can be enhanced between phases or with external surfaces. In its most general sense, the process of mixing is concerned with all combinations of phases like Gas, Liquid, solid. It is the heart of the chemical industry.

McCabe rightly quoted "Many processing operations depend for their success on the effective agitation & mixing of fluids"

P I Industries ltd., Udaipur has wisely decided to give detailed exposure to this area to its Engineering and R & D officers.

I am sure that the identified Faculty members will deliver the lectures on the topics assigned to them to the best of their capacity and expertise and put their best efforts to satisfy the thirst of the participants. This course is the outcome of discussions on various topics among Experts from P I Industries Ltd and the expert faculty members for about 4 months. The entire course will cover the topics the following topics arranged in sequential order so that all of you are benefited to the best way.

- Introduction to Agitation and Mixing Process
- Agitator Design
- Mixing Time
- Mixing of Liquid System
- Design of Gas Dispersion process
- Design of Solid Liquid Mixing Process
- Design of Solid-Solid Mixing Process

These topics are discussed in detail by the respective faculty members.

At this Juncture, I also appeal all the participants to take full advantage of this learning and apply to the operations where ever needed to improve the efficiency leading to improvement in quality and quantity of the products.

I am thankful to Dr. H. M. Desai, Vice- Chancellor of Dharmsinh Desai University and my source of Inspiration, Prof. (Dr.) P. A. Joshi, the Chairman of the Anchor Institute and former Dean of Faculty of Technology and one of my best colleague since more than 35 years who has always supported me not only in this endeavor but also others. I extend my thanks to the management of P I Industries Ltd and Mr. Kamlesh Mehta, General Manager-Process development & Kapil Khanna, Manager – HR. I extend my gratitude to all the faculty members of Department of Chemical Engineering , Faculty of Technology, Dharmsinh Desai University and from the field to accept my invitation to join and spare their time , coming long a way and being with us to share their expertise. My sincere thanks are to Prof. Mihir P. Shah to help me in compiling this Course material to put before you on time.

Prof. H R Shah

Ex. Coordinator,

Anchor Institute- Chemicals & Petrochemicals.

DDU, Nadiad

CONTENTS.....

Chapter 1.
MIXING AND AGITATION

Agitation is a means whereby mixing of phases can be accomplished and by which mass and heat transfer can be enhanced between phases or with external surfaces. In its most general sense, the process of mixing is concerned with all combinations of phases of which the most frequently occurring ones are

1. gases with gases.
2. gases into liquids: dispersion.
3. gases with granular solids: fluidization, pneumatic conveying, drying.
4. liquids into gases: spraying and atomization.
5. liquids with liquids: dissolution, emulsification, dispersion
6. liquids with granular solids: suspension.
7. pastes with each other and with solids.
 8. solids with solids: mixing of powders.

Interactions of gases, liquids, and solids also may take place, as in hydrogenation of liquids in the presence of a slurred solid catalyst where the gas must be dispersed as bubbles and the solid particles must be kept in suspension.

Three of the processes involving liquids, numbers 2, 5, and 6, employ the same kind of equipment; namely, tanks in which the liquid is circulated and subjected to a certain amount of shear. This kind of equipment has been studied most extensively. Although some unusual cases of liquid mixing may require pilot plant testing, general rules have been developed with which mixing equipment can be designed somewhat satisfactorily

1. A BASIC STIRRED TANK DESIGN

The dimensions of the liquid content of a vessel and the dimensions and arrangement of impellers, baffles and other internals are factors that influence the amount of energy required for achieving a needed amount of agitation or quality of mixing. The internal arrangements depend on the objectives of the operation: whether it is to maintain homogeneity of a reacting mixture or to keep a solid suspended or a gas dispersed or to enhance heat or mass transfer. A basic range of design factors, however, can be defined to cover the majority of cases, for example as in Figure 1.

THE VESSEL

A dished bottom requires less power than a flat one. When a single impeller is to be used, a liquid level equal to the diameter is optimum, with the impeller located at the center for an all-liquid system. Economic and manufacturing considerations, however, often dictate higher ratios of depth to diameter.
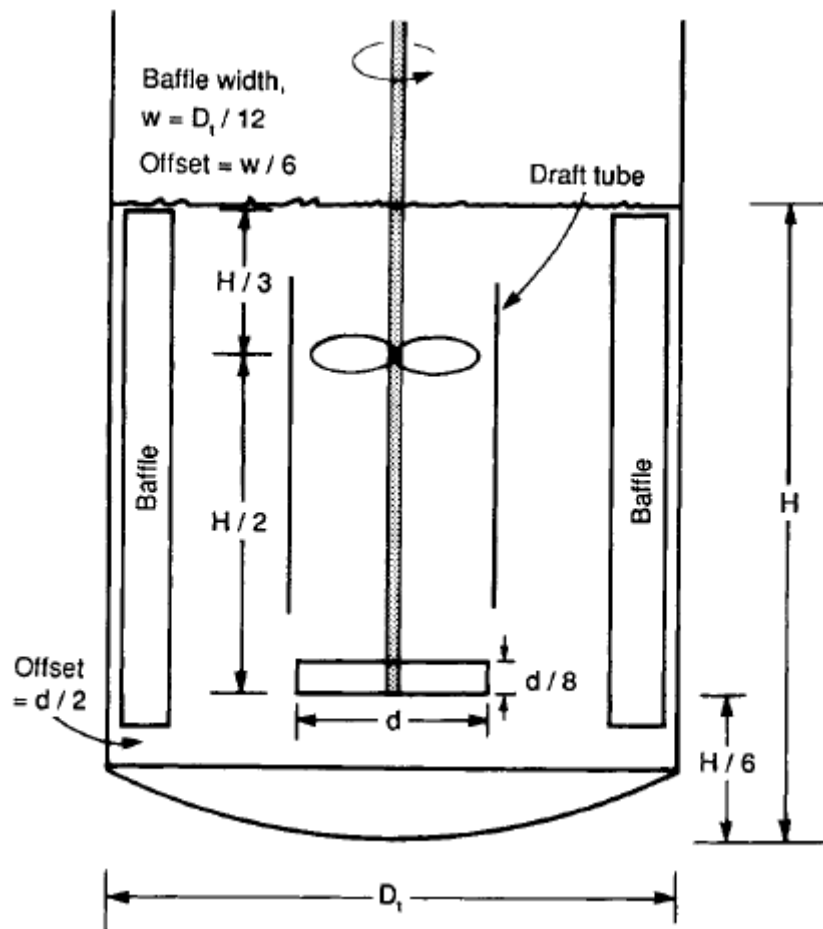
Figure 1. A basic stirred tank design, not to scale, showing a lower radial impeller and an upper axial impeller housed in a draft tube. Four equally spaced baffles are standard. H = height of liquid level, D,=tank diameter, d =impeller diameter. For radial impellers, 0.3 5d/D,50.6.

BAFFLES
Except at very high Reynolds numbers, baffles are needed to prevent vortexing and rotation of the liquid mass as a whole. A baffle width one-twelfth the tank diameter, w = $D_t/12$; a length extending from one half the impeller diameter, d/2, from the tangent line at the bottom to the liquid level, but sometimes terminated just above the level of the eye of the uppermost impeller. When solids are present or when a heat transfer jacket is used, the baffles are offset from the wall a distance equal to one- sixth the baffle width. Four radial baffles at equal spacing are standard; six are only slightly more effective, and three appreciably less so. When the mixer shaft is located off center (one-fourth to one-half the tank radius), the resulting flow pattern has less swirl, and baffles may not be needed, particularly at low viscosities.

DRAFT TUBES
A draft tube is a cylindrical housing around and slightly larger in diameter than the impeller. Its height may be little more than the diameter of the impeller or it may extend the full depth of the liquid, depending on the flow pattern that is required. Usually draft tubes are used with axial impellers to direct suction and discharge streams. An impeller-draft tube system behaves as an axial flow pump of somewhat low efficiency. Its top to bottom circulation behavior is of particular value in deep tanks for suspension of solids and for dispersion of gases.

IMPELLER TYPES
A basic classification is into those that circulate the liquid axially and those that achieve primarily radial circulation. Some of the many shapes that are being used will be described shortly.

IMPELLER SIZE
This depends on the kind of impeller and operating conditions described by the Reynolds, Froude, and Power numbers as well as individual characteristics whose effects have been correlated. For the popular turbine impeller, the ratio of diameters of impeller and vessel falls in the range, d/D_t=0.3-0.6, the lower values at high rpm, in gas dispersion, for example.

IMPELLER SPEED
With commercially available motors and speed reducers, standard speeds are 37, 45, 56, 68, 84, 100, 125, 155, 190, and 320 rpm. Power requirements usually are not great enough to justify the use of continuously adjustable steam turbine drives. Two-speed drives may be required when starting torques are high, as with a settled slurry.

IMPELLER LOCATION
Expert opinions differ somewhat on this factor. As a first approximation, the impeller can be placed at 1/6 the liquid level off the bottom. In some cases there is provision for changing the position of the impeller on the shaft. For off-bottom suspension of solids, an impeller location of 1/3 the impeller diameter off the bottom may be satisfactory. Criteria developed by Dickey (1984) are based on the viscosity of the liquid and the ratio of the liquid depth to the tank diameter, h / Q .
Whether one or two impellers are needed and their distances above the bottom of the tank are identified in this table:

Viscosity [cP (Pa sec)]	Maximum level h / D_t	Number of Impellers	Impeller Clearance	
			Lower	Upper
<25,000 (<25)	1.4	1	$h/3$	—
<25,000 (<25)	2.1	2	$D_t/3$	$(2/3)h$
>25,000 (>25)	0.8	1	$h/3$	—
>25,000 (>25)	1.6	2	$D_t/3$	$(2/3)h$

Side entering propellers are placed 18-24 in. above a flat tank floor with the shaft horizontal and at a 10" horizontal angle with the centerline of the tank; such mixers are used only for viscosities below 500 CP or so.
In dispersing gases, the gas should be fed directly below the impeller or at the periphery of the impeller. Such arrangements also are desirable for mixing liquids.

2. KINDS OF IMPELLERS
A rotating impeller in a fluid imparts flow and shear to it, the shear resulting from the flow of one portion of the fluid past another. Limiting cases of flow are in the axial or radial directions so that impellers are classified conveniently according to which of these flows is dominant. By reason of reflections from vessel surfaces and obstruction by baffles and other internals, however, flow patterns in most cases are mixed. When a close approach to axial

flow is particularly desirable, as for suspension of the solids as a slurry, the impeller may be housed in a draft tube; and when radial flow is needed, a shrouded turbine consisting of a rotor and a stator may be employed.

Because the performance of a particular shape of impeller usually cannot be predicted quantitatively, impeller design is largely an exercise of judgment so a considerable variety has been put forth by various manufacturers. A few common types are illustrated in Figure 2 and are described as follows:

a. The three-bladed mixing propeller is modeled on the marine propeller but has a pitch selected for maximum turbulence. They are used at relatively high speeds (up to 1800rpm) with low viscosity fluids, up to about 4000cP. Many versions are avail- able: with cutout or perforated blades for shredding and breaking up lumps, with saw tooth edges as in Figure 2(g) for cutting and tearing action, and with other than three blades. The stabilizing ring shown in the illustration sometimes is included to minimize shaft flutter and vibration particularly at low liquid levels.

b. The turbine with flat vertical blades extending to the shaft is **suited** to the vast majority of mixing duties **up to 100,000CP** or so at high pumping capacity. The simple geometry of this design and of the turbines of Figures 2(c) and (d) has inspired extensive testing so that prediction of their performance is on a more rational basis than that of any other kind of impeller.

c. The horizontal plate to which the impeller blades of this turbine are attached has a **stabilizing effect**. Backward curved blades may be used for the same reason as for type e.

d. Turbine with blades are inclined 45° (usually). Constructions with two to eight blades are used, six being most common. Combined axial and radial flow are achieved. Especially **effective for heat exchange** with vessel walls or internal coils.

e. Curved blade turbines effectively **disperse fibrous** materials without fouling. The swept back blades have a *lower starting torque* than straight ones, which is important when starting up settled slurries.

f. Shrouded turbines consisting of a rotor and a stator ensure a high degree of radial flow and shearing action, and are well adapted to **emulsification and dispersion**.

g. Flat plate impellers with saw tooth edges are suited to **emulsification and dispersion**. Since the shearing action is localized, *baffles are not required*. Propellers and turbines also are sometimes provided with saw tooth edges to improve shear.

h. Cage beaters impart a **cutting and beating** action. Usually they are mounted on the same shaft with a standard propeller. More violent action may be obtained with spinned blades.

i. Anchor paddles fit the contour of the container, prevent sticking of **pasty materials**, and promote *good heat transfer* with the wall.

j. Gate paddles are used in wide, shallow tanks and for **materials of high viscosity** when *low shear is adequate*. Shaft speeds are low. Some designs include hinged scrapers to clean the sides and bottom of the tank.

k. Hollow shaft and hollow impeller assemblies are operated at high tip speeds for **recirculating gases**. The gas enters the shaft above the liquid level and is expelled centrifugally at the impeller. Circulation rates are relatively low, but satisfactory for some hydrogenations for instance.

l. This arrangement of a shrouded screw impeller and heat exchange coil for **viscous liquids** is perhaps representative of the many designs that serve special applications in chemical processing.

3. CHARACTERIZATION OF MIXING QUALITY
Agitation and mixing may be performed with several objectives:
1. Blending of miscible liquids.
2. Dispersion of immiscible liquids.
3. Dispersion of gases in liquids.
4. Suspension of solid particles in a slurry.
5. Enhancement of heat exchange between the fluid and the boundary of a container.
6. Enhancement of mass transfer between dispersed phases.

When the ultimate objective of these operations is the carrying out of a chemical reaction, the achieved specific rate is a suitable measure of the quality of the mixing. Similarly the achieved heat transfer or mass transfer coefficients are measures of their respective operations. These aspects of the subject are not covered here. Here other criteria will be considered.
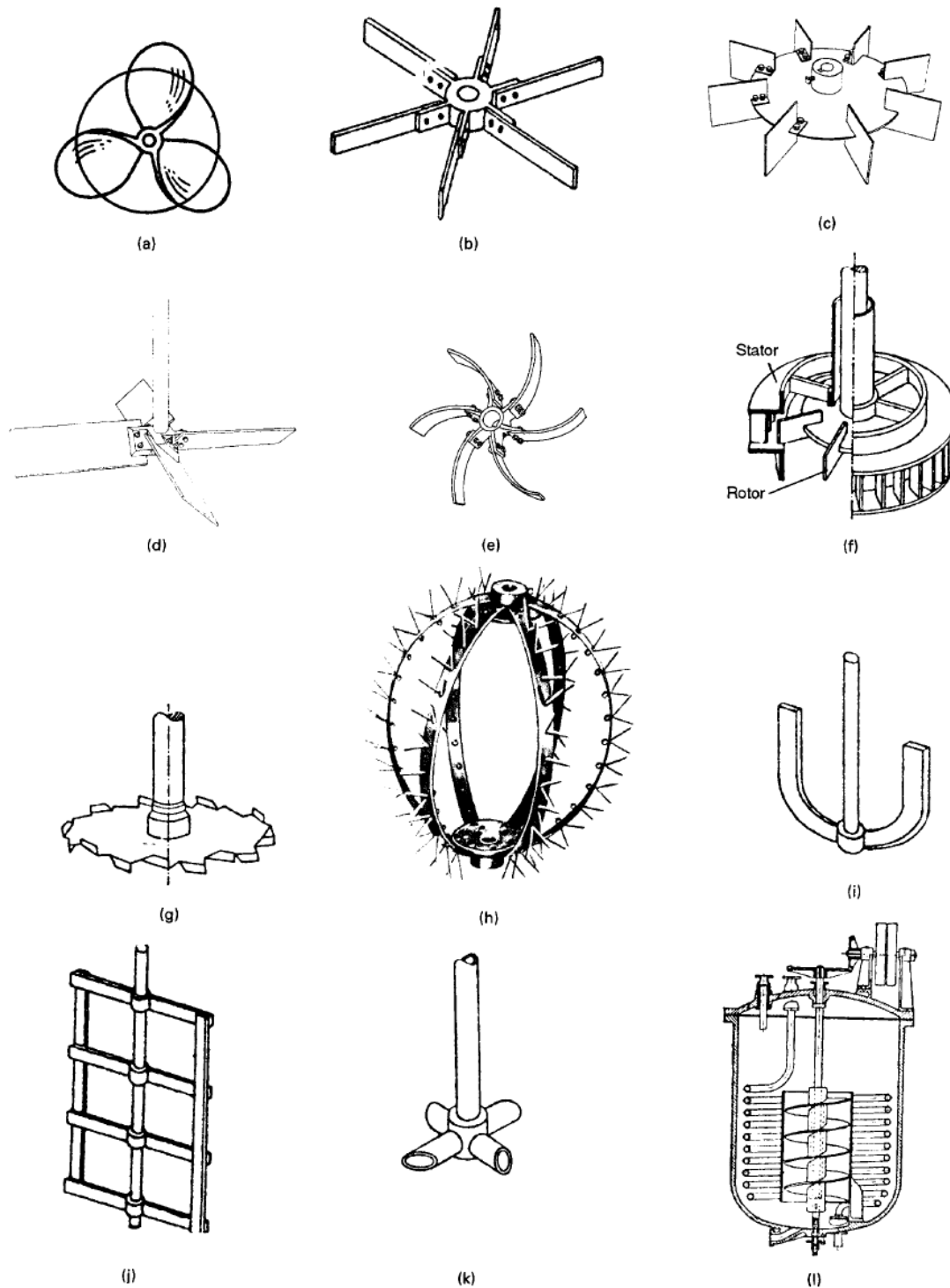
Figure 2. Types of Impellers

The uniformity of a multiphase mixture can be measured by sampling of several regions in the agitated mixture. The time to bring composition or some property within a specified range

(say within 95 or 99% of uniformity) or spread in values-which   is the blend time-may be taken as a measure of mixing performance.

Various kinds of tracer techniques may be employed, for example:
A dye is introduced and the time for attainment of uniform color is noted. A concentrated salt solution is added as tracer and the measured electrical conductivity tells when the composition is uniform. The color change of an indicator when neutralization is complete when injection of an acid or base tracer is employed.

The residence time distribution is measured by monitoring the outlet concentration of an inert tracer that can be analyzed for accuracy. The shape of response curve is compared with that of a thoroughly (ideally) mixed tank.
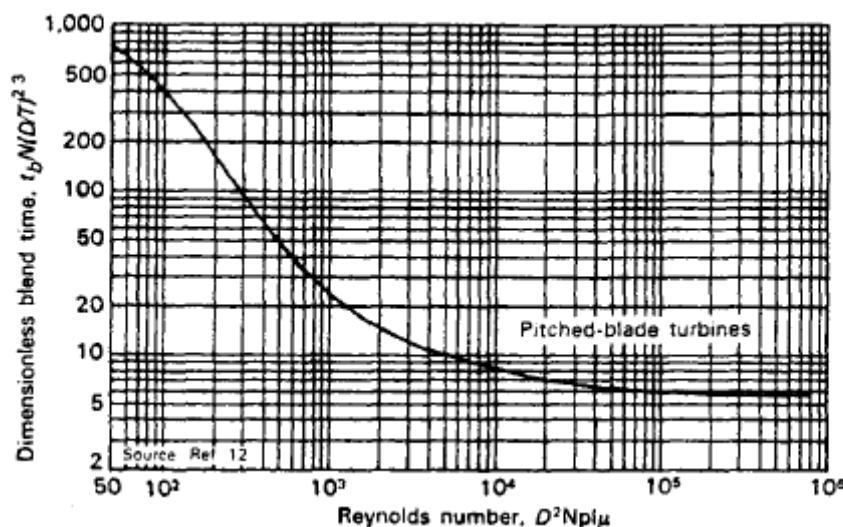

Figure 3. Dimensionless blend time as a function of Reynolds number for pitched turbine impellers with six blades whose WID= 1/5.66 [Dickey and Fenic, Chem. Eng. 145, (5Jan. 1976)l.

In most cases, however, the RTDs have not been correlated with impeller characteristics or other mixing parameters. Largely this also is true of most mixing investigations, but Figure 3 is an uncommon example of correlation of blend time in terms of Reynolds number for the popular pitched blade turbine impeller. As expected, the blend time levels off beyond a certain mixing intensity, in this case beyond Reynolds numbers of 30,000 or so. The acid-base indicator technique was used. Other details of the test work and the scatter of the data are not revealed in the published information.

An impeller in a tank functions as a pump that delivers a certain volumetric rate at each rotational speed and corresponding power input. The power input is influenced also by the geometry of the equipment and the properties of the fluid. The flow pattern and the degree of turbulence are key aspects of the quality of mixing. Basic impeller actions are either axial or radial, but, as Figure 4 shows, radial action results in some axial movement by reason of deflection from the vessel walls and baffles. Baffles contribute to turbulence by preventing swirl of the contents as a whole and elimination of vortexes; offset location of the impeller has similar effects but on a reduced scale.
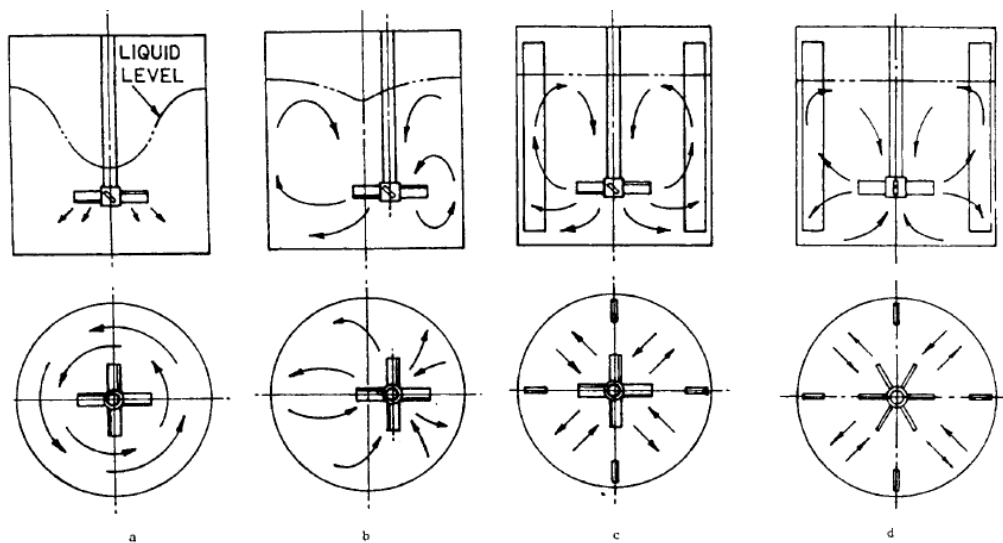
Figure 4. Agitator flow patterns. (a) Axial or radial impellers without baffles produce vortexes. (b) Off center location reduces the vortex. (c) Axial impeller with baffles. (d) Radial impeller with baffles.

Power input and other factors are interrelated in terms of certain dimensionless groups. The most pertinent ones are, in common units:

$N_{Re}=10.75Nd^2S/\mu$,   Reynolds number, (10.1)

$N_P= 1.523(10^{13})P/N^3d^5S$, Power number, (10.2)

$N_Q= 1.037(10^5)Q/Nd^3$,   Flow number, (10.3)

t_bN,   Dimensionless blend time, (10.4)

$NFr = 7.454(10^{-4})N^2d$,   Froude number, (10.5)

d = impeller diameter (in.),

D = vessel diameter (in.),

N = rpm of impeller shaft,

P = horsepower input,

Q =volumetric pumping rate (cuft/sec),

S = specific gravity,

t_b = blend time (min),

μ= viscosity (cP).

The Froude number is pertinent when gravitational effects are significant, as in vortex formation; in baffled tanks its influence is hardly detectable. The power, flow, and blend time numbers change with Reynolds numbers in the low range, but tend to level off above NRe= 10,000 or so at values characteristic of the kind of impeller. Sometimes impellers are characterized by their limiting N_p as an N_p =1.37 of a turbine, for instance. The dependencies on Reynolds number are shown on Figures 5 and 6 for power, in Figure 3 for flow and in Figure 7 for blend time.

Rough rules for mixing quality can be based on correlations of power input and pumping rate when the agitation system is otherwise properly designed with a suitable impeller (predominantly either axial or radial depending on the process) in a correct location, with appropriate baffling and the correct shape of vessel. The power input per unit volume or the

superficial linear velocity can be used as measures of mixing intensity. For continuous flow reactors, for instance, a rule of thumb is that the contents of the vessel should be turned over in 5 to 10 % of the residence time. Specifications of superficial linear velocities for different kinds of operations are stated later. For baffled turbine agitation of reactors, power inputs and impeller tip speeds such as the following may serve as guide:

Operation	HP/1000 gal	Tip Speed (ft/sec)
Blending	0.2–0.5	
Homogeneous reaction	0.5–1.5	7.5–10
Reaction with heat transfer	1.5–5.0	10–15
Liquid-liquid mixtures	5	15–20
Liquid-gas mixtures	5–10	15–20
Slurries	10	

4. POWER CONSUMPTION AND PUMPING RATE

These basic characteristics of agitation systems are of paramount importance and have been investigated extensively. The literature is reviewed, for example, by Oldshue (1983, pp. 155-191), Uhl and Gray (1966, Vol. l), and Nagata (1975). Among the effects studied are those of type and dimensions and locations of impellers, numbers and sizes of baffles, and dimensions of the vessel. A few of the data are summarized on Figures 5-7. Often it is convenient to characterize impeller performance by single numbers; suitable ones are the limiting values of the power and flow numbers at high Reynolds numbers, above 10,000-30,000 or so, for example

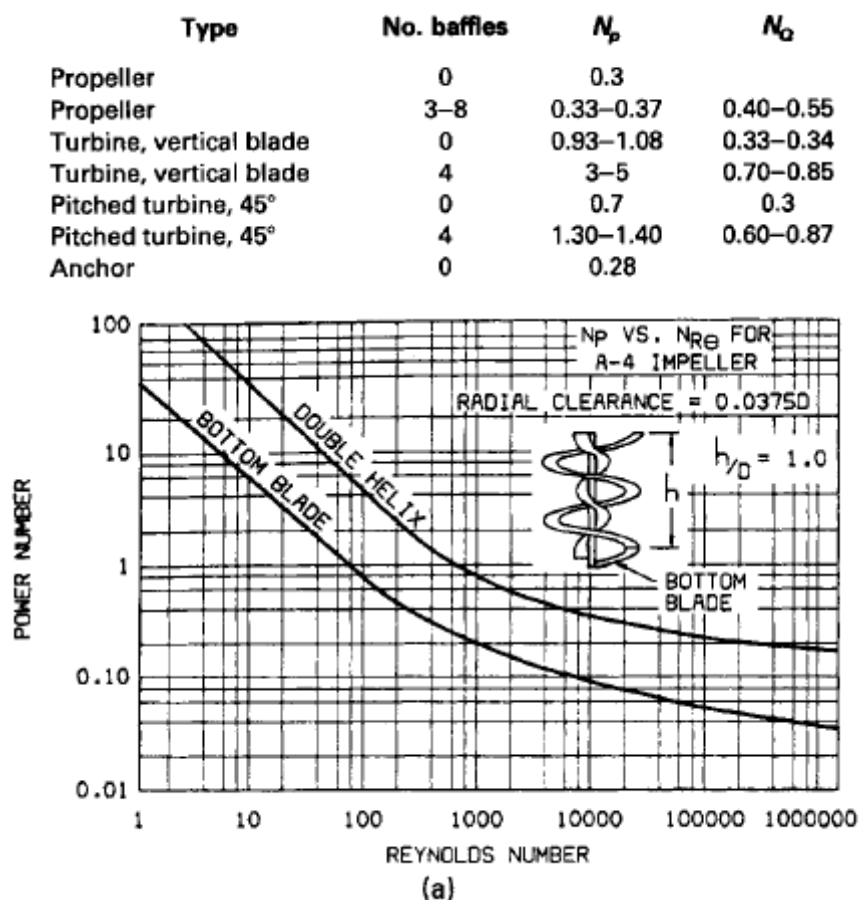| Type	No. baffles	N_P	N_Q
Propeller	0	0.3	
Propeller	3–8	0.33–0.37	0.40–0.55
Turbine, vertical blade	0	0.93–1.08	0.33–0.34
Turbine, vertical blade	4	3–5	0.70–0.85
Pitched turbine, 45°	0	0.7	0.3
Pitched turbine, 45°	4	1.30–1.40	0.60–0.87
Anchor	0	0.28	

Figure 5. Power number, N, = Pg,/N3D5p, against Reynolds number, NRe = NDzp/p, for several kinds of impellers: (a) helical shape (Oldshue, 1983); (b) anchor shape (Oldshue, 1983); (c) several shapes: (1) propeller, pitch equalling diameter, without baffles; (2) propeller, s =d, four baffles; (3) propeller, s =2d, without baffles; (4) propeller, s =2d, four

baffles; (5) turbine impeller, six straight blades, without baffles; (6) turbine impeller, six blades, four baffles; (7) turbine impeller, six curved blades, four baffles; (8) arrowhead turbine, four baffles; (9) turbine impeller, nclined curved blades, four baffles; (10) two-blade paddle, four baffles; (11) turbine impeller, six blades, four baffles; (12) turbine impeller with stator ring; (13) paddle without baffles (data of Miller and Mann); (14) paddle without baffles (data of White and Summerford). All baffles are of width 0.1D [afterRushton, Costich, and Everett, Chem. Eng. Prog. 46(9), 467 (1950)l
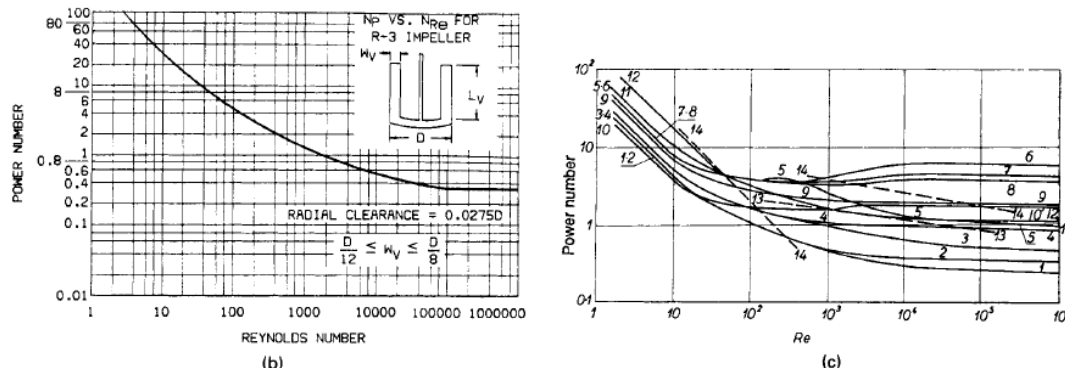
Figure 5 (continued)

A correlation of pumping rate of pitched turbines is shown as Figure 7.
Power input per unit volume as a measure of mixing intensity or quality was cited in Section 3. From the correlations cited in this section, it is clear that power input and Reynolds number together determine also the pumping rate of a given design of impeller. This fact has been made the basis of a method of agitator system design by the staff of Chemineer. The superficial linear velocity-the volumetric pumping rate per unit cross section of the tank-is adopted as a measure of quality of mixing. Table 2 relates the velocity to performance of three main categories of mixing: mixing of liquids, suspension of solids in slurries, and dispersion of gases. A specification of a superficial velocity will enable selection of appropriate impeller size, rotation speed, and power input with the aid of charts such as Figures 6 and 7.
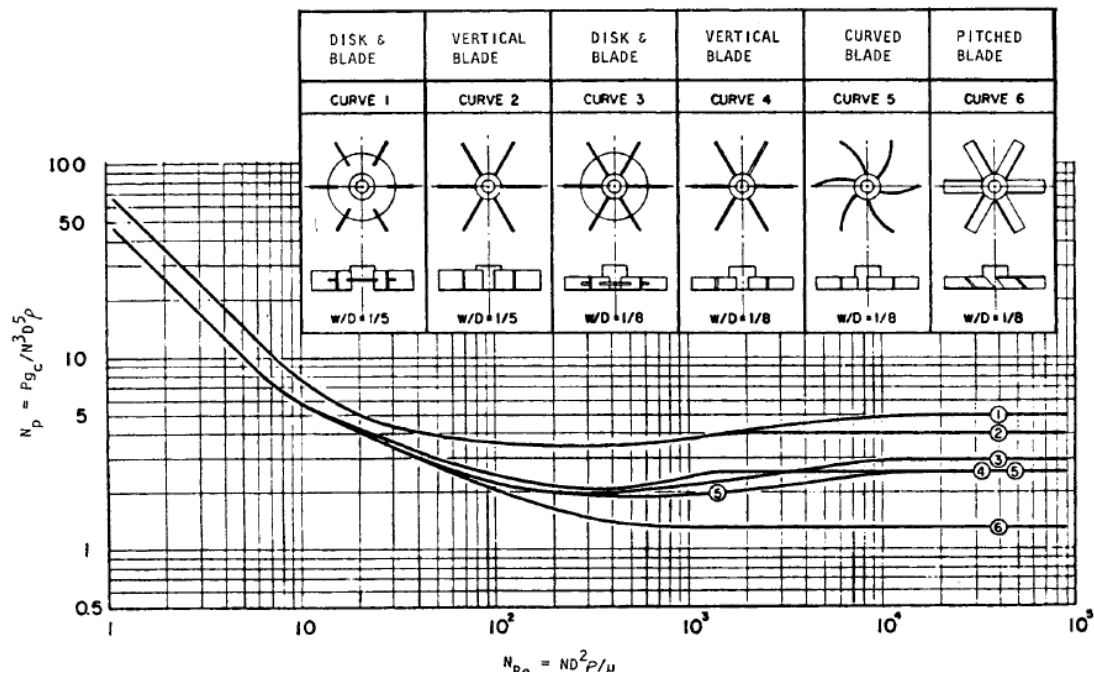
Figure 6. Power number against Reynolds number of some turbine impellers [Bates, Fondy, and Corpstein, Ind. Eng. Chem. Process. Des. Dev. 2(4) 311 (1963)l.

Figure 7. Flow number as a function of impeller Reynolds number for a pitched blade turbine with N, = 1.37. D I T is the ratio of impeller and tank diameters. [Dickey, 1984, 12, 7; Chem. Eng., 102-110 (26Apr. 1976)l.

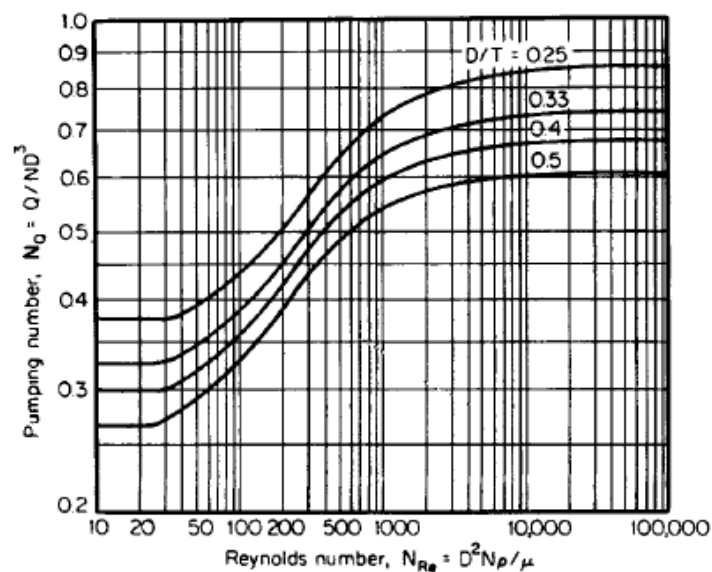TABLE 2. Agitation Results Corresponding to Specific Superficial Velocities

ft/sec	Description	ft/sec	Description
Liquid Systems			c. suspend all solids with the design settling velocity completely off the bottom of the vessel
0.1–0.2	low degree of agitation; a velocity of 0.2 ft/sec will		d. provide slurry uniformity to at least one-third of the liquid level
	a. blend miscible liquids to uniformity when specific gravity differences are less than 0.1		e. be suitable for slurry drawoff at low exit nozzle locations
	b. blend miscible liquids to uniformity if the ratio of viscosities is less than 100	0.6–0.8	when uniform solids distribution must be approached; a velocity of 0.6 ft/sec will
	c. establish liquid movement throughout the vessel		f. provide uniform distribution to within 95% of liquid level
	d. produce a flat but moving surface		g. be suitable for slurry drawoff up to 80% of liquid level
0.3–0.6	characteristic of most agitation used in chemical processing; a velocity of 0.6 ft/sec will	0.9–1.0	when the maximum feasible uniformity is needed. A velocity of 0.9 ft/sec will
	e. blend miscible liquids to uniformity if the specific gravity differences are less than 0.6		h. provide slurry uniformity to 98% of the liquid level
	f. blend miscible liquids to uniformity if the ratio of viscosities is less than 10,000		i. be suitable for slurry drawoff by means of overflow
	g. suspend trace solids (less than 2%) with settling rates of 2–4 ft/min	**Gas Dispersion**	
	h. produce surface rippling at low viscosities	0.1–0.2	used when degree of dispersion is not critical to the process; a velocity of 0.2 ft/sec will
0.7–1.0	high degree of agitation; a velocity of 1.0 ft/sec will		a. provide nonflooded impeller conditions for coarse dispersion
	i. blend miscible liquids to uniformity if the specific gravity differences are less than 1.0		b. be typical of situations that are not mass transfer limited
	j. blend miscible liquids to uniformity if the ratio of viscosities is less than 100,000	0.3–0.5	used where moderate degree of dispersion is needed; a velocity of 0.5 ft/sec will
	k. suspend trace solids (less than 2%) with settling rates of 4–6 ft/min		c. drive fine bubbles completely to the wall of the vessel
	l. produce surging surface at low viscosities		d. provide recirculation of dispersed bubbles back into the impeller
Solids Suspension		0.6–1.0	used where rapid mass transfer is needed; a velocity of 1.0 ft/sec will
0.1–0.2	minimal solids suspension; a velocity of 0.1 ft/sec will		e. maximize interfacial area and recirculation of dispersed bubbles through the impeller
	a. produce motion of all solids with the design settling velocity		
	b. move fillets of solids on the tank bottom and suspend them intermittently		
0.3–0.5	characteristic of most applications of solids suspension and dissolution; a velocity of 0.3 ft/sec will		

[Chemineer, Co. Staff, *Chem. Eng.*, 102–110 (26 April 1976); 144–150 (24 May 1976); 141–148 (19 July 1976)].

5. SUSPENSION OF SOLIDS

Besides the dimensions of the vessel, the impeller, and baffles, certain Physical data are needed for complete description of a slurry mixing problem, primarily:

1. Specific gravities of the solid and liquid.
2. Solids content of the slurry (wt %).
3. Settling velocity of the particles (ft/min).

The last of these may be obtained from correlations when the mesh size or particle size distribution is known, or preferably experimentally. Taking into account these factors in their effect on suspension quality is at present a highly empirical process.

Table 3: Mixing of Liquid; power and impeller speed (hp/rpm) for two viscosities, as a function of liquid superficial velocity, pitch blade turbine impeller

| | Volume (gal) | | | | | |
| | 5000 cP | | | 25,000 cP | | |
ft/sec	1000	2000	5000	1000	2000	5000
0.1	2/280 1/190	2/190 1/100	2/100	2/125 1.5/84	2/84 1.5/56	7.5/125 5/100 5/84 3/56
0.2	2/190 1/100	2/125 2/84 1.5/84	5/125 3/84 3/68 2/45	3/84 2/84 1.5/56	5/125 3/84 3/68 2/45	10/84 7.5/68 5/45 3/37
0.3	2/125 1.5/84	3/84 1.5/56	7.5/125 5/100 5/84 3/56	5/125 5/84 3/68 2/45	15/155 7.5/68 5/45 3/37	20/100 15/68 10/45 7.5/37
0.4	2/84 1.5/56	5/125 3/68 3/56 2/45	10/84 7.5/68 5/45 3/37	7.5/84 5/56	10/84 7.5/45	30/100 25/84 20/68 10/37
0.5	5/125 3/84	7.5/125 5/84	15/100 10/68 7.5/45	15/155 10/100 10/84 7.5/68	25/125 20/100 15/84 10/56	75/190 60/155 40/100 15/45
0.6	5/100 3/68 3/56 2/45	15/155 10/100 7.5/84 3/37	25/125 20/100 10/56 7.5/37	20/155 15/125	25/100 15/68 15/56 10/45	40/84 30/68 25/56 20/37
0.7	7.5/125 5/84	10/84 7.5/68 5/45	15/68 15/56 10/45 10/37	25/155 15/84	40/155 30/100 25/84 20/68	75/125 50/84 30/45
0.8	10/125 7.5/100	10/68 7.5/56	30/100 25/84 20/68 15/45	30/155 25/125 20/100	50/155 40/125	75/100 60/84 50/68 40/56
0.9	15/155 10/100 7.5/84	15/84 10/56 7.5/45	60/155 40/100	40/155 30/125 25/100	75/190 60/155 40/100 30/68	75/84 60/68 50/56
1.0	10/84 7.5/68	30/155 25/125 20/100 15/68	50/100 40/84 30/68 25/56	40/125 30/100	60/125 50/100 50/84 40/84	125/125 100/100 75/68 60/56

[Hicks, Morton, and Fenic, *Chem. Eng.*, 102–110 (26 April 1976)].

Table 4: Suspension of solids; power and impeller speed (hp/rpm) for two settling velocities, as a function of liquid superficial velocity, pitch blade turbine impeller

ft/sec	Volume (gal) 10 ft/min			25 ft/min		
	1000	2000	5000	1000	2000	5000
0.1	1/190	2/190	5/125	2/190	2/125	5/125
		1/100	3/84	1/190	2/84	3/84
			3/68	1/100	1.5/84	3/68
			2/45		1.5/56	2/45
0.2	1/100	2/125	7.5/125	2/125	3/84	15/155
		1.5/84	5/100			10/100
			5/84			7.5/68
			3/56			5/45
0.3	2/190	2/84	3/37	1.5/84	5/125	10/84
		1.5/56			3/68	
					2/45	
0.4	2/155	5/155	7.5/84	2/84	7.5/155	7.5/45
	1.5/100		5/56	1.5/56	5/100	
					3/56	
0.5	1.5/84	3/84	15/155	2/68	7.5/125	15/84
	2/125		10/100	2/56	5/84	10/56
			7.5/68			7.5/37
			5/45			
0.6	2/100	5/125	10/84	3/84	5/56	25/125
	1.5/68	3/68				20/100
		3/56				15/68
		2/45				10/45
0.7	2/84	7.5/155	15/84	7.5/155	15/155	30/100
	1.5/56	7.5/125	10/56	5/125	10/100	25/84
		5/84	7.5/45	5/100	7.5/84	20/68
			7.5/37	3/68	7.5/68	15/56
0.8	3/84	7.5/84	25/125	7.5/125	10/84	60/155
		5/56	20/100	5/84		40/100
			15/68			30/68
			10/45			25/56
0.9	7.5/155	15/155	40/155	10/125	15/84	75/190
	5/125	10/100	30/100	7.5/100		60/125
	5/100	7.5/68	25/84			50/100
	3/68		20/68			40/84
1.0	7.5/125	20/100	50/100	15/155	30/155	75/125
	5/84	15/84	40/84	10/100	25/125	75/100
		10/84	30/68		20/100	60/84
			25/56			50/84

[Gates, Morton, and Fondy, *Chem. Eng.*, 144–150 (24 May 1976)].

Suspension of solids is maintained by upward movement of the liquid. In principle, use of a draft tube and an axial flow impeller will accomplish this flow pattern most readily. It turns out, however, that such arrangements are suitable only for low solids contents and moderate power levels. In order to be effective, the cross section of the draft tube must be appreciably smaller than that of the vessel, so that the solids concentration in the draft tube may become impractically high. The usually practical arrangement for solids suspension employs a pitched blade turbine which gives both axial and radial flow.

For a given tank size, the ultimate design objective is the relation between power input and impeller size at a specified uniformity. The factors governing such information are the slurry volume, the slurry level, and the required uniformity. The method of Oldshue has corrections for these factors, as F,,F2, and F3. When multiplied together, they make up the factor 4 which is the ordinate of Figure 8(d) and which determines what combinations of horsepower and ratio of impeller and vessel diameters will do the required task.

6. GAS DISPERSION

Gases are dispersed in liquids usually to facilitate mass transfer between the phases or mass transfer to be followed by chemical reaction. In some situations gases are dispersed adequately with spargers or porous distributors, but the main concern here is with the more intense effects achievable with impeller driven agitators.

SPARGERS
Mixing of liquids and suspension of solids may be accomplished by bubbling with an inert gas introduced uniformly at the bottom of the tank. For mild agitation a superficial gas velocity of 1ft/min is used, and for severe, one of about 4 ftlmin.

Table 4: Dispersion of gas; power and impeller speed (hp/rpm) for two gas inlet superficial velocities, as a function of liquid superficial velocity, vertical blade turbine impeller

| | Volume (gal) | | | | | |
| | 0.07 ft/sec | | | 0.20 ft/sec | | |
ft/sec	1500	3000	5000	1500	3000	5000
0.1	2/56	5/84	7.5/68 15/155	3/56	7.5/68 10/100	10/45
0.2	2/45	7.5/125	10/84 7.5/45	3/45	15/155	15/68 20/100
				5/100	10/84	25/125
0.3	3/84 3/68 3/56	7.5/68 5/45 7.5/84 5/56	10/45 10/56		7.5/45	
0.4	5/125 5/84 5/100 5/45	10/84 10/100 10/45 10/56	15/68 20/100 15/84 20/68	5/84 7.5/155 5/56	10/45 10/56	30/155 20/68 15/45 15/56
0.5	7.5/125 7.5/155 7.5/68 7.5/84	15/155 15/68 15/84 15/45	25/125 25/84 25/100 25/56	7.5/125 7.5/68 7.5/84	15/68 15/84 15/45 15/56	25/84 25/100 25/56
0.6	10/84 10/100	20/100 20/68 20/45	30/155 30/100 30/125 30/68	10/84 10/100	20/100 20/68	30/100 30/125 30/68 30/45
0.7	10/56	25/125 25/84 25/100 25/56	40/155 40/84 40/100 40/56	10/56	25/125 25/84 25/100 25/56	40/155 40/84 40/100 40/56
0.8	15/155 15/84	30/155 30/100 30/125	50/100 50/68 50/84 50/45	15/155 15/84	30/155 30/100 30/125	50/100 50/68 50/84 50/56
0.9	15/68	30/68	60/125 60/155 60/84 60/56	15/68	30/68	60/125 60/155 60/84 60/56
1.0	20/100 20/68	40/155 40/84	75/190 75/100 75/125	25/125 25/84	40/155 40/84	75/190 75/100 75/125

[Hicks and Gates, *Chem. Eng.*, 141–148 (19 July 1976)].

6. GAS DISPERSION
power input as a factor is given by Treybal (Mass Transfer Operations, McGraw-Hill, New York, 1980, 156); presumably this is applicable only below the minimum power input here represented by Figure 11.

When mass transfer coefficients are not determinable, agitator design may be based on superficial liquid velocities with the criteria of Table 2.

SYSTEM DESIGN
The impeller commonly used for gas dispersion is a radial turbine with six vertical blades. For a liquid height to diameter ratio h/D51,asingleimpellerisadequate;intherange15h/D51.8 two are needed, and more than two are rarely used. The lower and upper impellers are located at distances of 1/6 and 2/3 of the liquid level above the bottom. Baffling is essential, commonly with four baffles of width 1/12 that of the tank diameter, offset from the wall at 1/6 the width of the baffle and extending from the tangent line of the wall to the liquid level. The best position for inlet of the gas is below and at the center of the lower impeller; an open pipe is commonly used, but a sparger often helps. Since un-gassed power is significantly larger than gassed, a two-speed motor is desirable to prevent overloading, the lower speed to cut in automatically when the gas supply is interrupted and rotation still is needed.

MINIMUM POWER
Below a critical power input the gas bubbles are not affected laterally but move upward with their natural buoyancy. This condition is called gas flooding of the impeller. At higher power inputs the gas is dispersed radially, bubbles impinge on the walls and are broken up, consequently with improvement of mass transfer. A correlation of the critical power input is shown as Figure 10.

POWER CONSUMPTION OF GASSED LIQUIDS
At least partly because of its lower density and viscosity, the power to drive a mixture of gas and liquid is less than that to drive a liquid. Figure 11(a) is a correlation of this effect, and other data at low values of the flow number Q / N d 3 are on Figure 11(b). The latter data for Newtonian fluids are correlated by the equation
$$P_g/P = 0.497(Q/Nd^3)^{-0.38}(N^2d^3\rho_L/\sigma)^{-0.018} \qquad (7)$$
where the last group of terms is the Weber number,
ρ_L, is the density of the liquid,
and is σ its surface tension.

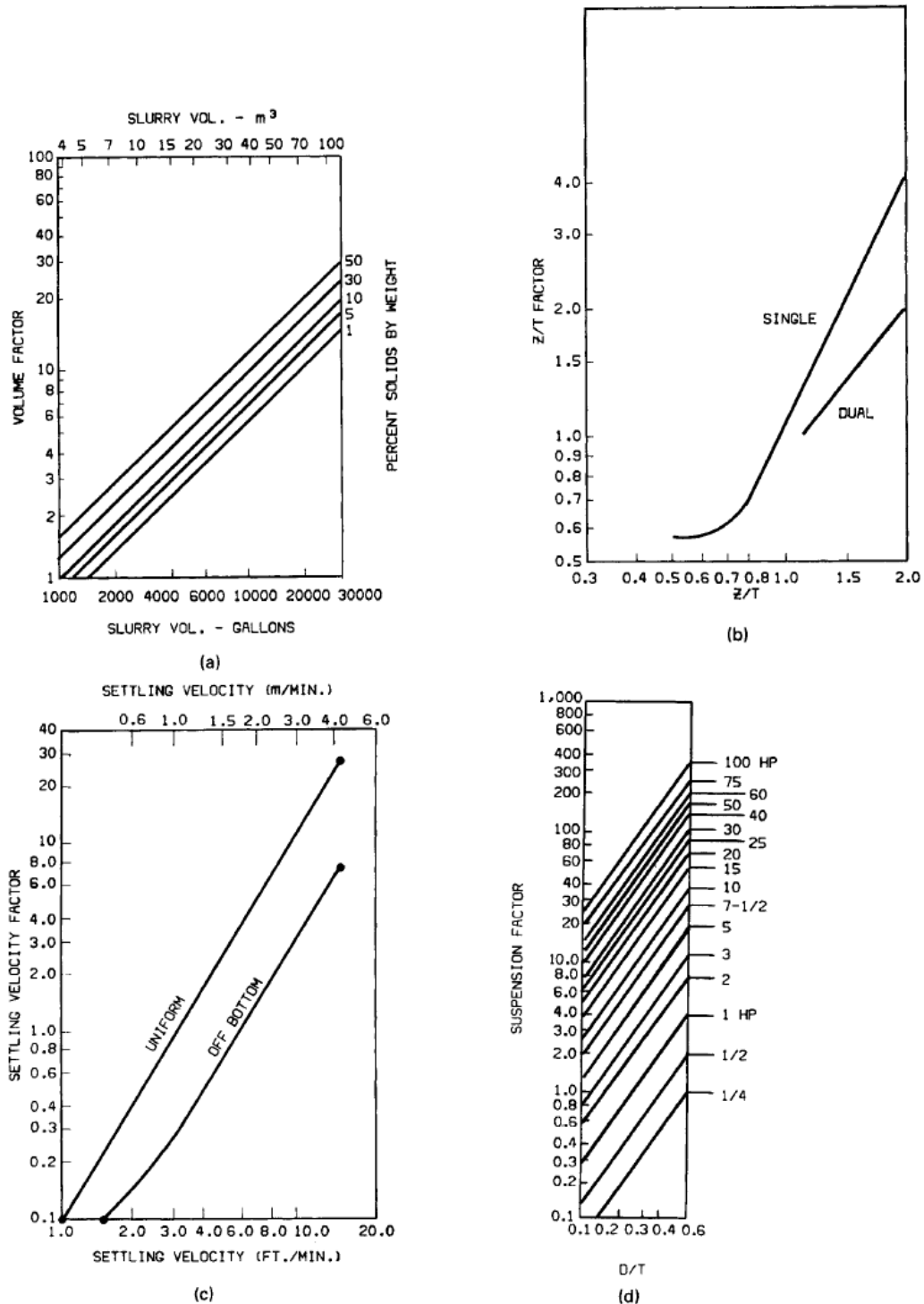Fire 8. Suspension of solids. Power and ratio of diameters of impeller and tank, with four-bladed 45" impeller, width/diameter =0.2. [method of Oldshue (1983)l.(a) The factor on power consumption for slurry volume, F1.(b) The factor on power requirement for single and dual impellers at various h/D ratios, F2. (c) The effect of settling velocity on power consumption, F3.(d) Suspension factor for various horsepowers: F4=F1,F2,F3
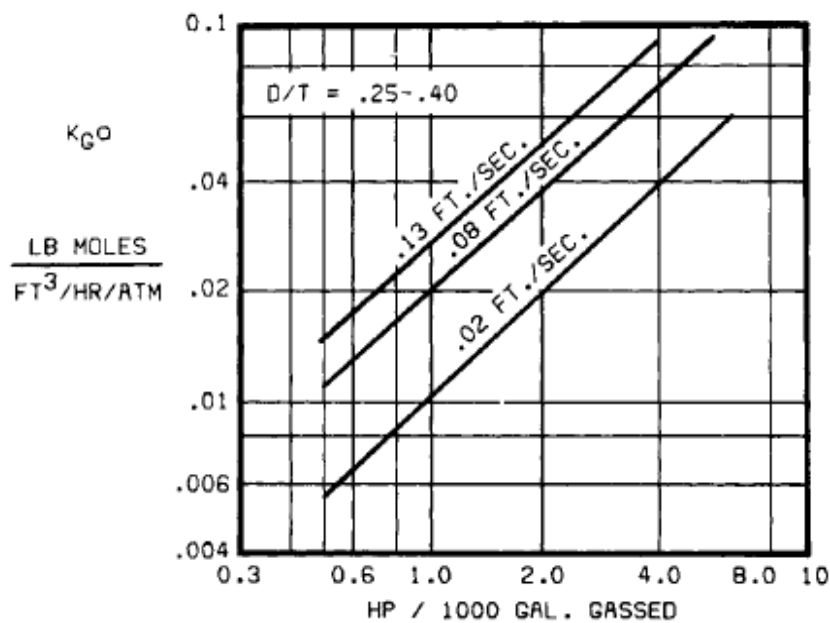
Figure 9: Typical data for mass transfer coefficient at various power levels and superficial gas rates for oxidation of sodium sulfite
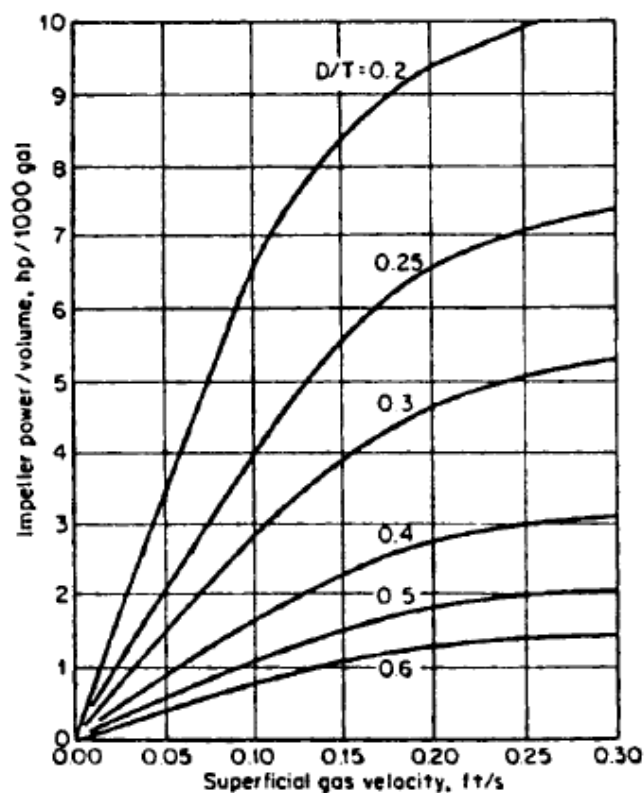

Figure10. Minimum power requirement to overcome flooding as a function of superficial gas velocity and ratio of impeller and tank diameters, d/D. [Hicksand Gates, Chem. Eng., 141-148 (19 July1976)].
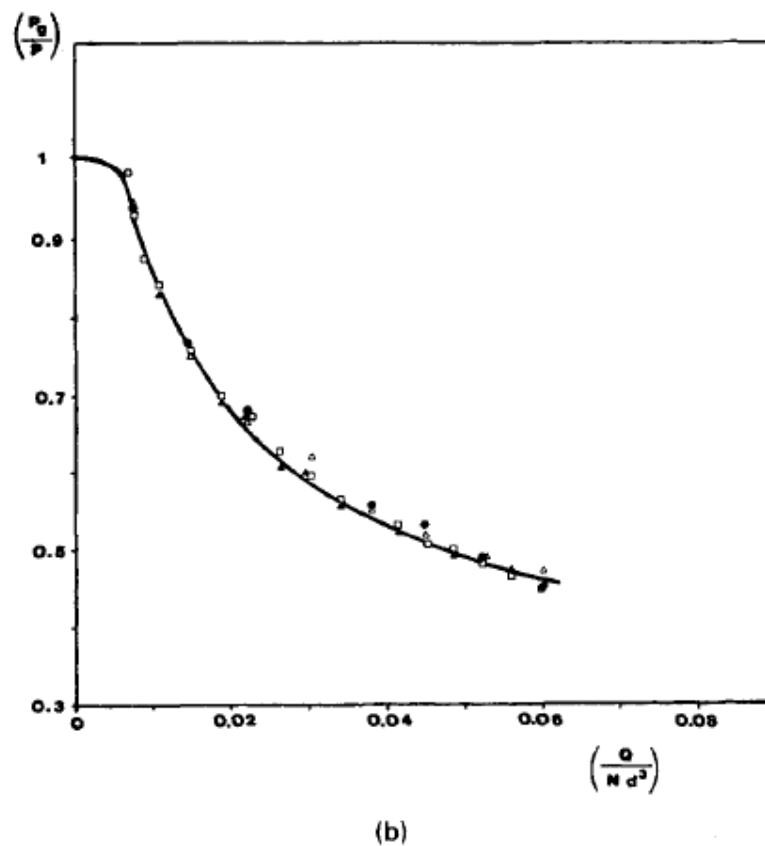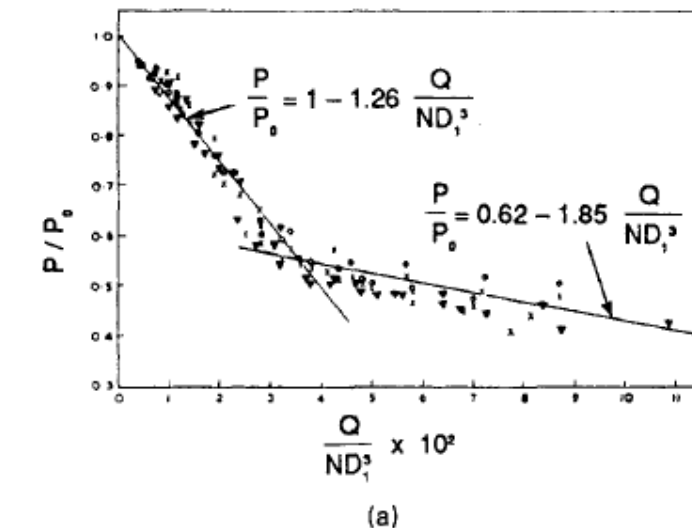
$$\frac{P}{P_0} = 1 - 1.26 \frac{Q}{ND_1^3}$$

$$\frac{P}{P_0} = 0.62 - 1.85 \frac{Q}{ND_1^3}$$

(a)

(b)

Figure 11.Power consumption. (a) Ratio of power consumptions of aerated and unaerated liquids. Q is the volumetric rate of the gas: (0)glycol; (x) ethanol; (V)water. [After Calderbank, Trans. Inst. Chem. Eng. 36, 443 (1958)l. (b) Ratio of power consumptions of aerated and unaerated liquids at low values of Q/Nd3.Six-bladed disk turbine: (0)water; (0)methanol (10%); (A)ethylene glycol (8%); (A) glycerol (40%); P, =gassed power input; P =ungassed power input; Q =gas flow rate; N =agitator speed; d = agitator-impeller diameter. [Luong and Volesky, AIChE J. 25, 893 (1979)j
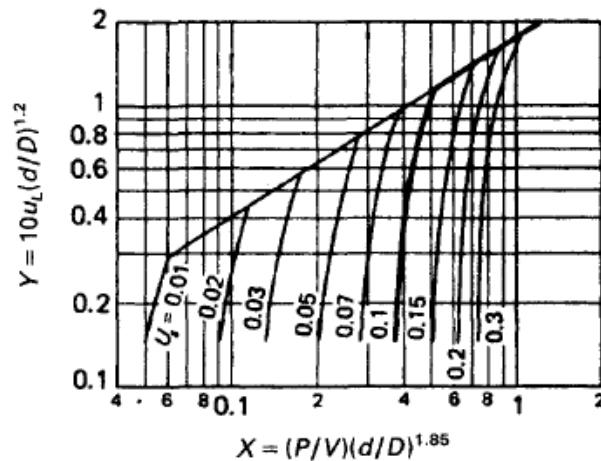
Figure 12. Relation between power input, P/V HP/1000 gal, superficial liquid velocity uL ft/sec, ratio of impeller and tank diameters, d/D, and superficial gas velocity u, ft/sec. [Hicksand Gates, Chem. Eng., 141-148 (19 July 1976)].

SUPERFICIAL LIQUID VELOCITY
When mass transfer data are not known or are not strictly pertinent, a quality of mixing may be selected by an exercise of judgment in terms of the superficial liquid velocity on the basis of the rules of Table 2. For gas dispersion, this quantity is related to the power input, HP/lOOOgal, the superficial gas velocity and the ratio d / D in Figure 12.

DESIGN PROCEDURES
On the basis of the information gathered here, three methods are possible for the design of agitated gas dispersion. In all cases the size of the tank, the ratio of impeller and tank diameters and the gas feed rate are specified. The data are for radial turbine impellers with six vertical blades.

The starting point of agitator design is properly a mass transfer coefficient known empirically or from some correlation in terms of parameters such as impeller size and rotation, power input, and gas flow rate. Few such correlations are in the open literature, but some have come from two of the industries that employ aerated stirred tanks on a large scale, namely liquid waste treating and fermentation processes. A favored method of studying the absorption of oxygen is to measure the rate of oxidation of aqueous sodium sulfite solutions. Figure 9 summarizes one such investigation of the effects of power input and gas rate on the mass transfer coefficients. A correlation for fermentation air is given by Dickey (1984, 12-17):
k_La = rate/(concentration driving force) =O.O64(Pg/V)$^{0.7}$ $u_g$$^{0.2}$, 1/ sec),~~~, (6)
with Pg/V in HP/lOOO gal and superficial gas velocity u_g in ft/sec. A general correlation of mass transfer coefficient that does not have power input as a factor is given by Treybal (Mass Transfer Operations, McGraw-Hill, New York, 1980, 156); presumably this is applicable only below the minimum power input here represented by Figure 11.

When mass transfer coefficients are not determinable, agitator design may be based on superficial liquid velocities with the criteria of Table 2.

SYSTEM DESIGN
The impeller commonly used for gas dispersion is a radial turbine with six vertical blades. For a liquid height to diameter ratio h/D 51, a single impeller is adequate; in the range 15h/D 51.8 two are needed, and more than two are rarely used. The lower and upper impellers are

located at distances of 1/6 and 2/3 of the liquid level above the bottom. Baffling is essential, commonly with four baffles of width 1/12 that of the tank diameter, offset from the wall at 1/6 the width of the baffle and extending from the tangent line of the wall to the liquid level. The best position for inlet of the gas is below and at the center of the lower impeller; an open pipe is commonly used, but a sparger often helps. Since ungassed power is significantly larger than gassed, a two-speed motor is desirable to prevent overloading, the lower speed to cut in automatically when the gas supply is interrupted and rotation still is needed.

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POWER CONSUMPTION OF GASSED LIQUIDS

At least partly because of its lower density and viscosity, the power to drive a mixture of gas and liquid is less than that to drive a liquid. Figure 11(a) is a correlation of this effect, and other data at low values of the flow number Q/Nd^3 are on Figure 11(b). The latter data for Newtonian fluids are correlated by the equation

$$Pg/P = 0.497(Q/Nd^3)^{-0.38}(N^2d^3\rho_L/\sigma)^{-0.018} \quad (7)$$

where the last group of terms is the Weber number, ρ, is the density of the liquid, and σ is its surface tension.

SUPERFICIAL LIQUID VELOCITY

When mass transfer data are not known or are not strictly pertinent, a quality of mixing may be selected by an exercise of judgment in terms of the superficial liquid velocity on the basis of the rules of Table 2. For gas dispersion, this quantity is related to the power input, HP/lOOOgal, the superficial gas velocity and the ratio d/D in Figure 12.
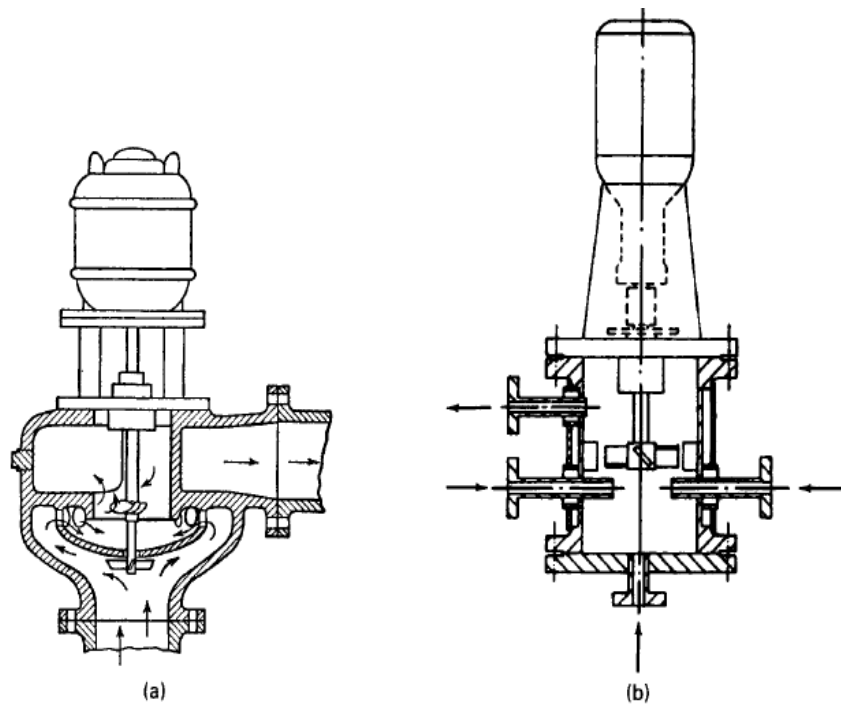
Figure 13. Motor-driven in-line blenders: (a) Double impeller made by Nettco Corp.; (b) three-inlet model made by Cleveland Mixer Co.
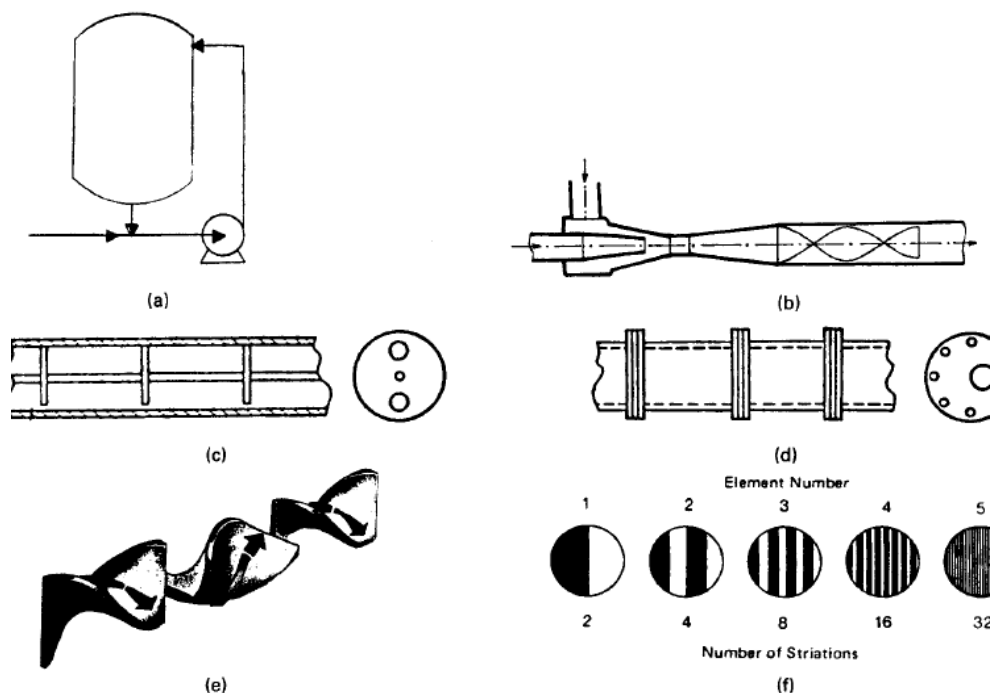

Figure 10.14. Some kinds of in-line mixers and blenders. (a) Mixing and blending with a recirculating pump. (b) Injector mixer with a helical baffle. (c) Several perforated plates (orifices) supported on a rod. (d) Several perforated plates flanged in. (e) Hellical mixing elements with alternating directions (Kenics Corp.). (f) Showing progressive striations of the flow channels with Kenics mixing elements.

7. IN-LINE BLENDERS AND MIXERS

When long residence time is not needed for chemical reaction or other purposes, small highly powered tank mixers may be suitable, with energy inputs measured in HP/gal rather than HP/1000gal. They bring together several streams continuously for a short contact time (at

most a second or two) and may be used whenever the effluent remains naturally blended for a sufficiently long time, that is, when a true solution is formed or a stable emulsion-like mixture.

The Kenics mixer, Figure 14(a), for example, consists of a succession of helical elements twisted alternately in opposite directions. In laminar flow for instance, the flow is split in two at each element so that after n elements the number of striations

becomes 2". The effect of this geometrical progression is illustrated in Figure 14(b) and points out how effective the mixing becomes after only a few elements. The Reynolds number in a corresponding empty pipe is the major discriminant for the size of mixer.

Other devices utilize the energy of the flowing fluid to do the mixing. They are inserts to the pipeline that force continual changes of direction and mixing. Loading a section of piping with tower packing is an example but special assemblies of greater convenience have been developed, some of which are shown in Figure 14. In each case manufacturer's literature recommends the sizes and pressure drops needed for particular services.

Besides liquid blending applications, static mixers have been used for mixing gases, pH control, dispersion of gases into liquids, and dispersion of dyes and solids in viscous liquids. They have the advantages of small size, ease of operation, and relatively low cost. The strong mixing effect enhances the rate of heat transfer from viscous streams. Complete heat exchangers are built with such 12-18 mixing inserts in the tubes and are then claimed to have 3-5 times normal capability in some cases.

8. MIXING OF POWDERS AND PASTES

Industries such as foods, cosmetics, pharmaceuticals, plastics, rubbers, and also some others have to do with mixing of high viscosity liquids or pastes, of powders together and of powders with pastes. Much of this kind of work is in batch mode. The processes are so diverse and the criteria for uniformity of the final product are so imprecise that the non-specialist can do little in the way of equipment design, or in checking on the recommendations of equipment manufacturers. Direct experience is the main guide to selection of the best kind of equipment, predicting how well and quickly it will perform, and what power consumption will be.

Where design by analogy may not suffice, testing in pilot plant equipment is a service provided by many equipment suppliers.
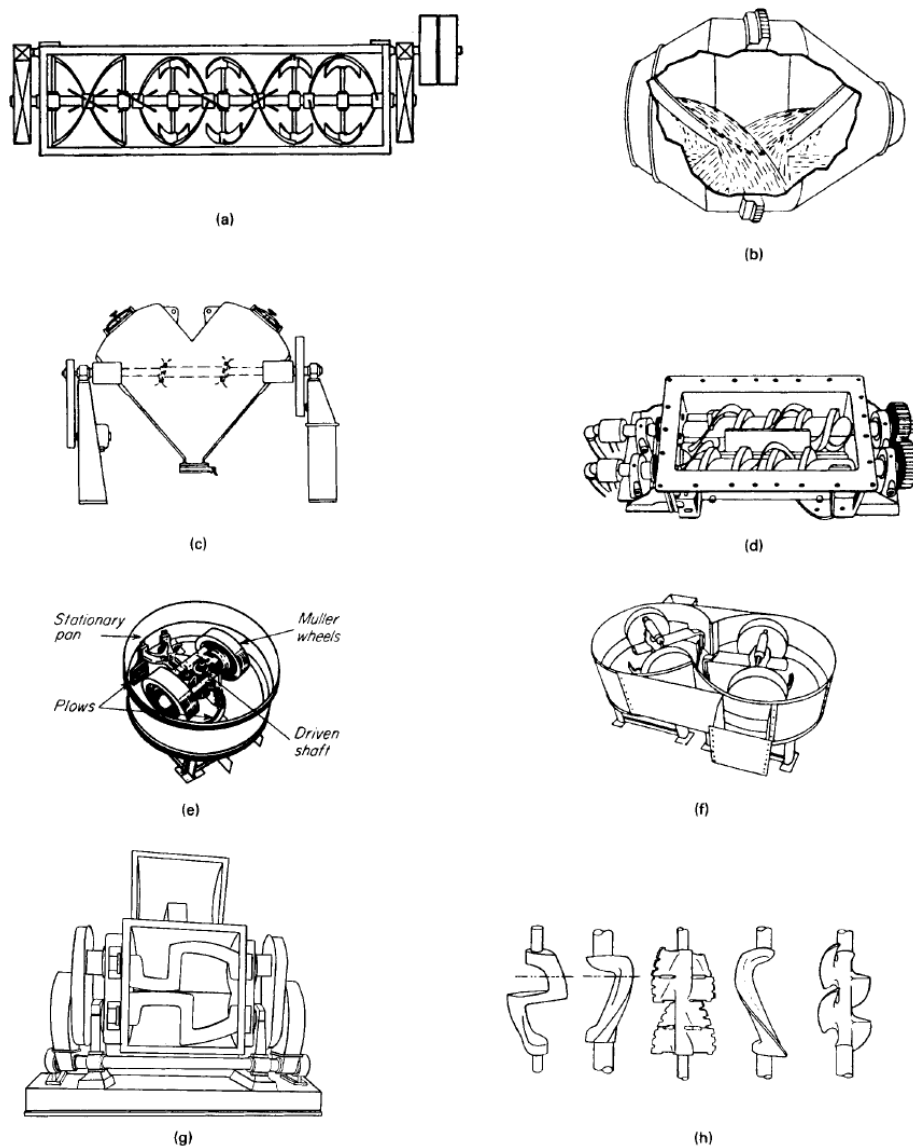
Figure 10.15. Some mixers and blenders for powders and pastes. (a) Ribbon blender for powders. (b) Flow pattern in a double cone blender rotating on a horizontal axis. (c) Twin shell (Vee-type); agglomerate breaking and liquid injection are shown on the broken line. (d) Twin rotor; available with jacket and hollow screws for heat transfer. (e) Batch muller. (f) Twin mullers operated continuously. (8) Double-arm
mixer and kneader (Baker-Perkins Znc.). (h) Some types of blades for the double-arm kneader (Baker-Perkins Znc.).

A few examples of mixers and blenders for powders and pastes are illustrated in Figure 15. For descriptions of available equipment-their construction, capacity, performance, power consumption, etc.-the primary sources are catalogs of manufac- turers and contact with their offices. Classified lists of manufacturers, and some of their catalog information, appear in the Chemical Engineering Catalog (Reinhold, New York, annually) and in the Chemical Engineering Equipment Buyers Guide (McGraw-Hill, New York, annually). Brief descriptions of some types of equipment are in Perry's Chemical Engineers Handbook (McGraw-Hill, New York, 1984 and earlier editions).

Chapter 2
AGITATOR DESIGN

"Many processing operations depend for their success on the effective agitation & mixing of fluids"

……McCabe

Mixing and agitation is the heart of the chemical industry. Almost all process equipments need some type of mixing or agitation. Uniformity of composition and desired flow pattern depends upon the type of agitator and the speed of agitation. It is also necessary to control the quality of the product, specifically where there is evaluation of heat and the temperature has to be maintained constant.

Agitation

- ➢ It is an induced motion of a material in a specified way.
- ➢ The pattern is normally circulatory.
- ➢ It is normally taken place inside a container.

Mixing

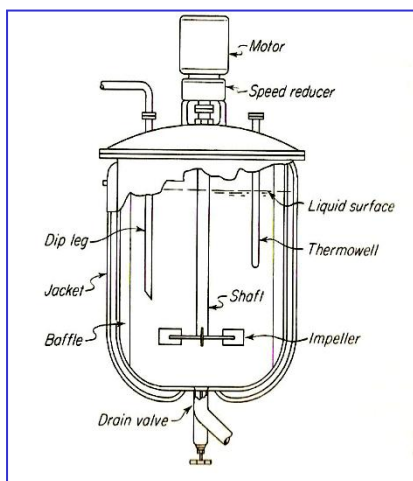- ➢ Random distribution, into & through one another of two or more initially separate phases

- ➢ Liquids are agitated in a tank
- ➢ Bottom of the tank is rounded
- ➢ Impeller creates a flow pattern.
- ➢ Small scale tank (less than 10 litres) is constructed using Pyrex glass.
- ➢ For larger reactors/tank, stainless steel is used.
- ➢ Speed reduction devices are used to control the agitation speed.
- ➢ Mixing Flow : 3 patterns (axial, radial, tangential flow)

Fig. 1 Agitated Vessel

Course Material Process Engineering: Agitation & Mixing

Different types of operations:
1. Blending of miscible liquids
2. Gas absorption
3. Gas dispersion
4. Dissolution
5. Crystallization
6. Heat transfer
7. Chemical Reaction
8. Extraction etc.

A good mixing should achieve the following:
1. Minimum power requirement.
2. Efficient mixing in optimum time.
3. Best possible economy.
4. Minimum maintenance, durable and trouble free operation.
5. Compactness.

Important consideration in the designing are:
➢ Determination of amount of energy required or power required for satisfactory performance of mixing operation.
➢ Process has to be well defined e.g. a mixing system is to be designed to make up and hold in uniform suspension a 15% slurry.
➢ Description of the components to be mixed. Their properties at initial stage, final stage, overall specific gravity, initial and final viscosity, concentration etc.
➢ Details of the tank geometry.
➢ Outline of the mixing cycle: It depends upon the nature of the operation. Decide whether the process is a batch, semi-continuous or continuous etc.

Factors affecting the designing of the agitator:
➢ Type of vessel
➢ Circulation pattern.
➢ Location of the agitator
➢ Shape and size of the vessel
➢ Diameter and width of the agitator
➢ Method of baffling
➢ Power required
➢ Shaft overhang
➢ Type of stuffing box or seal, bearing, drive system etc.

Mixing Flow patterns (3 types):
(i) **Axial flow.**
➢ Impeller makes an angle of less than 90° with the plane of rotation thus resultant flow pattern towards the base of the tank (i.e. marine impellers).
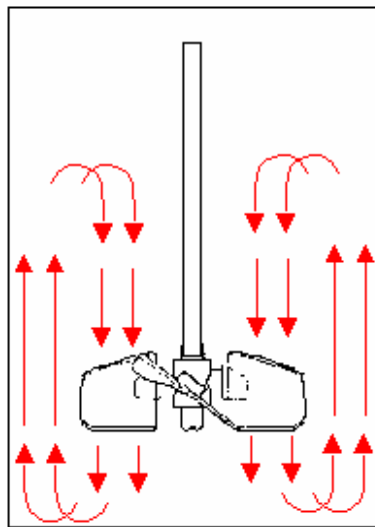
➢ More energy efficient than radial flow mixing.
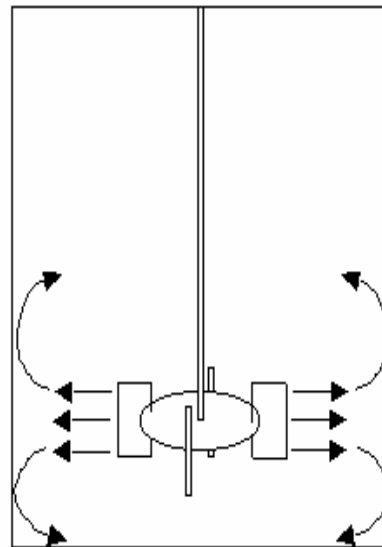➢ More effective at lifting solids from the base of the tank.

(ii) **Radial flow.**

➢ Impellers are parallel to the axis of the drive shaft.
➢ The currents travel outward to the vessel wall & then either upward or downward.
➢ Higher energy is required compared to axial flow impellers.

(ii) **Tangential flow.**

➢ The currents acts in the direction tangent to the circular path around the shaft.
➢ Usually, it produce vortex (disadvantageous) & swirling the liquid.

With axial impellers, the liquid is pushed in a downward direction; that is, along the axis of the reactor.

With radial flow impellers, the liquid is pushed towards the wall of the tank; that is, along the radius of the reactor
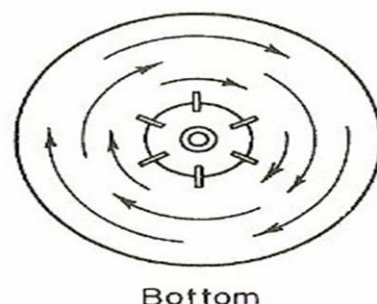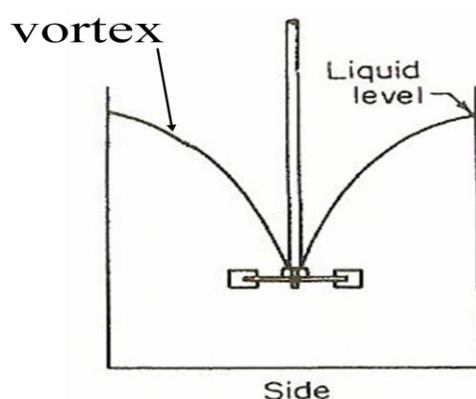
vortex

Liquid level

Side

Bottom

Fig. 2 (i) Axial Mixing (ii) Radial Mixing (iii) Tangential Mixing

Vortex

➢ If solid particles present within tank; it tends to throw the particles to the outside by centrifugal force.
➢ Power absorbed by liquid is limited.
➢ At high impeller speeds, the vortex may be so deep that it reaches the impeller.

➢ Method of preventing vortex
- baffles
- impeller in an angular off-center position

Preventing vortex

(i) Baffles on the tank walls

(ii) Impeller in an angular off-center position

(i) Baffles

Baffles are vertical plates (typically about 10% of the tank diameter) that stick out radially from the tank wall

➢ If simple swirling motion is required no baffling is necessary.

➢ Generally 4 baffles are used located 90° apart.

➢ Baffle width is 10-12% tower diameter

➢ Baffle height 2 times impeller height

➢ With coils in the tank, baffles are placed inside the coil.

Fig. 3 Flow Pattern in presence of baffles (i) Vertex (ii) Axial Flow turbine (iii) Radial Flow Turbine

➢ Without baffles, the tangential flow (swirling) occurred in a mixing tank causes the entire fluid mass to spin (more like a centrifuge than a mixer).

➢ With baffles, most impellers show their true flow characteristics.

➢ Most common baffles are straight flat plates of metal (standard baffles).

➢ Most vessels will have at least 3 baffles. 4 is most common and is often referred to as the "fully baffled" condition.

Fig 4. Baffled Vessel

Fig. 5. Agitation with and without baffles

(ii) Impeller in an angular off-center position

Mount the impeller away from the center of the vessel & tilted in the direction perpendicular to the direction of flow.

Fig. 6 Flow pattern for off mounted impeller

■ Figure 5. Off-center agitator mounting (left) reduces swirl in unbaffled vessels, while angled, off-center mounting (right) approximates the flow in fully-baffled vessels.

Fig. 7 (i) Side Mounted Impeller (ii) Angle Mounted Impeller

Types of impeller:

1. Paddle
2. Anchor

3. Propeller
4. Turbine
5. Beater
6. Gate Type
7. Helical
8. Ribbon
9. Toothed
10. Marine
11. Plate Type

Paddle type agitator

- Speed range 5-300rpm
- Used for large size vessels
- Agitator size almost touching vessel wall
- Normally used for reaction vessel having jacket by providing good heat transfer area
- Doesn't allow solid buildup at the wall

Fig. 14.1 Paddle Agitators: (a) Anchor, (b) Gate, (c) Gate with pitched cross arms, (d) Anchor with pitched cross arms, (e) Combined anchor and gate

Fig. 4.1 Types of Anchor Agitator

Fig. 8 Paddle and Anchor Type Agitators

Propeller type agitator

- Axial flow impellers
- Maximum flow is achieved at axis of agitator
- Maximum vessel size is 1m^3
- Maximum speed is 415 r/minute
- Diameter of propeller is 15-30% of vessel diameter

Fig. 14.7 Radial propeller agitator

Fig. 14.3 Propeller agitator

Fig. 9 Propeller Type Agitator

Turbine type agitator
- Motion is achieved due to rotary action of impeller
- Two types are available
    - Axial flow turbine
    - Radial flow turbine
    flat bladed
    pitched bladed
    curved bladed

(1) Pitched blade impeller   (2) Flat blade impeller   (3) Flat impeller with center disk   (4) Impeller hub

Fig. 4.8 Types of turbine impeller

Fig. 14.2 Turbine Agitators: (a) Straight flat blade, (b) Pitched blade,
(c) Curved back-sloped, (d) Straight blades attached to disc

3 blades   **4 blades**   **5 blades**

6 blades   **8 blades**   **12 blades**

(a)   (b)

Fig. 14.9 Shows the Flow Pattern Generated by a Turbine:
(a) Agitator without baffles, (b) With baffles

Fig. 10 Different arrangement in Turbine Type agitator and flow pattern in turbine type agitator

Helical or ribbon type agitator

- 4 types are available in market
  - Single helical
  - Double helical
  - Helical screw
  - Ribbon type
- Good for top to bottom liquid circulation
- Used for blending for pseudo plastic materials
- High power requirement

Fig. 4.2 Types of helical ribbon agitator

Fig. 11. Helical Ribbon Type Agitators

Specially designed Agitators

Fig. 14.5   Cone type agitator

➢ Produces high shear and helps disintegration of low density fibrous solids

Fig. 14.8   High speed disc agitator

➢ Cone type agitator is used for handling fibrous and dense slurries.
➢ Speeds are similar to turbine type
➢ Speed provides sufficient centrifugal force through surface friction to generate flow.

Selection of the agitator depends upon the viscosity of the fluid to be agitated.

When the blade area is small it can rotate at very high speed. For such cases the propeller and turbine type agitators are preferred. (μ = 1000 to 50000cps)

When the blade area is larger it will rotate at the slow speed. For such cases the anchor bolts and helical screw type agitators are used. (μ >> 50000cps)

Mounting of the agitators are done in mainly two ways:

➢ Top entering agitators ➔ Used in large units
➢ Side entering agitators ➔ Used in small units and economical

$$\text{Number of agitators} = \frac{\text{Maximum liquid height} * \text{average specific gravity}}{\text{tank diameter}}$$

Distance between two agitators is 1 to 15agitator diameter.

Agitator Drive system

- Electric motor supplies the power.

- IF rpm of motor shaft and agitator shaft is similar then gear box is not required.

- Gear box transmits power of electric motor shaft to agitator shaft directly or sometimes to the other shaft which is attached to agitator shaft.

- Coupling is used to connect two shafts.

Shaft seals

- During the process, liquid vapors or gases should not leak through agitator shaft nozzle.

- There should not be any exchange either from inside to outside or vise versa.

- Like in case of vacuum reaction

- Most common method for sealing shaft is with stuffing box and gland.

Stuffing box

Fig. 12 Stuffing Box

Mechanical Seal

- Stuffing box is having cylindrical shape, placed around the shaft.
- Inside diameter is greater than outside diameter of shaft, having packing material in between for sealing.
- Gland has ID nearly equal to OD of shaft.
- Gland is made of hard metals having sleevs attached to it made of soft metals.
- Glands is bolted to stuffing box using bolts.
- Used only when, $P \leq 10kg$, $T \leq 120C$ and $N \leq 300rpm$.
- If any of these conditions are not satisfied then better replace stuffing box with mechanical seals.

FIG. 5.4 (b) MECHANICAL SEAL

$$\frac{D_a}{D_t} = \frac{1}{3} \qquad \frac{H}{D_t} = 1 \qquad \frac{J}{D_t} = \frac{1}{12}$$

$$\frac{E}{D_t} = \frac{1}{3} \qquad \frac{W}{D_a} = \frac{1}{5} \qquad \frac{L}{D_a} = \frac{1}{4}$$

Fig. 13. Standard Turbine Design

Draft tubes

Fig. 14 Draft Tube arrangement

Circulation and Velocity in Agitated Vessels

- Volume of fluid circulated by impeller must be sufficient to sweep out entire vessel in reasonable time
- Velocity of stream leaving impeller must be sufficient to carry current to remotest parts of tank
- In mixing, also need turbulence
    - Results from properly directed currents and large velocity gradients in liquid
- Circulation and generation of turbulence both consume energy
- Large impeller + medium speed = flow
- Small impeller + high speed = turbulence

Mixing Types

- Laminar
- Turbulent

Mixing Mechanism

- Dispersion or Diffusion is the act of spreading out

- Molecular diffusion is diffusion caused by molecular motion and characterized by molecular diffusivity DAB.
- Eddy Diffusion or Turbulent Diffusion is dispersion in turbulent flows caused by motion of large groups of molecules called eddies; this motion is measured as the turbulent velocity fluctuations.
- Convection or bulk diffusion is dispersion caused by bulk motion.
- Taylor dispersion is a special case of convection, where dispersion is caused by mean velocity gradient. This is the case of Laminar flow conditions.

Scale of Mixing

- Macro mixing is mixing driven by largest scale of motion in the fluid. Characterized by mixing time or blend time in a batch system.
- Mesomixing is mixing on a scale smaller than the bulk circulation (tank diameter) but larger than the micromixing scales, where molecular and viscous diffusion become important. It is an evident of feed pipe scale of semibatch reactors.
- Micromixing is mixing on the smallest scales of motion and at the final scale of molecular diffusivity. It is usually limiting step in progress of fast reactions. It is more or less fixed value.

Measures of Mixedness

- Scale of segregation is a measure of the large scale break up process without the action of diffusion. (Fig.a)
- Intensity of segregation is a measure of difference in concentration between the purest concentration of B and the purest concentration of A in the surrounding fluid. (Fig. b)
- Molecular diffusion is needed to reduce the intensity of segregation.
- Even smallest eddy have a very large diameter relative to the size of molecule, turbulence may not be needed for reduction in intensity of segregation but it can definitely increase the rate of reduction.

Intensity and scale of segregation

Fig. 15. (a) Reduction in scale of segregation (b) Reduction in intensity of segregation (c) simultaneous reduction in scale and intensity of segregation

Turbulence Vs. Laminar

- The Concept was developed by Prandtl.
- Also called Prandtl mixing length.
- The Design and operation of mixing process in these two flow regime is quit different, the reason for this is
    - Reynolds number

- Low Reynolds number – Laminar Flow – Viscosity Dominant – Infinitesimal disturbances are damped out.
- Higher Reynolds number – Turbulent flow – Interfacial force Dominates - No viscosity effect on process

Fig. 16. Evaluating Mixing Performance

Heat and Mass Transfer

In figure 17 both materials are available in equal amount 50%. However, the internal contact area between the two constituent is higher in case (b). The parameter affecting mass transfer, the area available for transport and the diffusion distance, are both affected by the topology of mixture. The area available for transport is greater and diffusion distance is shorter in case (b). if this were a reactive mixture, these difference would result in a faster overall reaction rate in case (b). similar situation occurs in industrial mixture also.

Fig. 17 Interfacial area available

POWER REQUIREMENT

➢ For an effective mixing, the volume of fluid circulated in a vessel via an impeller must be sufficient to sweep out the entire vessel in a reasonable time.
➢ Stream velocity leaving the impeller must be sufficient to carry currents to the remotest part of the vessel.

FACTORS AFFECTING THE POWER REQUIREMENT:

č   Properties of fluid to be agitated

- č   Height of the liquid
- č   Tank size and dimensions
- č   Agitator type and size
- č   Speed of agitator

Terminology in Power Calculation

(i)   Flow Number

$$q \, \alpha \, n \, D_a^{\,3} \qquad \boxed{N_Q = \frac{q}{nD_a^3}}$$

➤ Where q is the volumetric flow rate, measured at the tip of the blades, n is the rotational speed (rpm), Da is the impeller diameter

➤ Total flow was shown to be→ $\boxed{q_T = 0.92nD_a^3\left(\frac{D_t}{D_a}\right)}$

➤ N_Q is constant for each type of impeller. For flat-blade turbine (FBT), in a baffled vessel, N_Q may be taken as 1.3; For marine propellers (Square pitch), $N_Q = 0.5$; For four blade 45° turbine, $N_Q = 0.87$;

➤ For HE impeller- N_Q=0.47

(ii)   The Reynolds number, N_{Re}

$$\boxed{N_{RE} = \frac{D_a^2 n\rho}{\mu}}$$

(iii)   The Froude number, N_{Fr}

$$\boxed{N_{Fr} = \frac{n^2 D_a}{g}}$$

■ Froude No. is a measure of the ratio of the inertial stress to the gravitational force per unit area acting on the fluid. It appears in the dynamic situations where there is significant wave motion on a liquid surface. Important in ship design. Unimportant when baffles are not used or Re< 300

Why Dimensionless Numbers?

- Empirical correlations to estimate the power required to rotate a given impeller at a give speed, with respect to other variables in system
    - Measurements of tank and impeller
    - Distance of impeller from tank floor
    - Liquid depth
    - Dimensions of baffles
    - Viscosity, density, speed

Dimensionless Correlations

Power versus Reynolds

N_P | laminar | transition | turbulent

N_{Re}

Velocity versus Reynolds

D-less vel. | laminar | transition | turbulent

N_{Re}

Pumping number versus Reynolds

$\dfrac{Q}{N D^3}$ | laminar | transition | turbulent

N_{Re}

Blend time versus Reynolds

$t_B N$ | laminar | transition | turbulent

N_{Re}

Not a predefined empirical equation to find out the power requirement it will depend upon all factors described above some unexpected problems.

In the power curve:

Region A   : Viscous range

Region B   : Transition range

Region C   : Turbulent region

The curve is drawn for 6flat bladed turbine with height of the liquid is equal to diameter of the vessel and vessel is having 4 baffles.

Power number N_P versus N_{Re} for six-blade turbines. With the dashed portion of curve D, the value of N_P read from the figure must be multiplied by N_{Fr}^m.

Fig. 18. Power Curve for Turbine Type agitator

Curve A = vertical blades, $W/D_a = 0.2$

Curve B = vertical blades, $W/D_a = 0.125$

Curve C = pitched blade

Curve D = unbaffled tank

The power number is calculated as:

$$\phi = N_p = \frac{Pg_c}{8N^3 D_a^5} \qquad \text{and} \qquad N_{Re} = \frac{\rho N D_a^2}{\mu}$$

where N_p = power number

   P = power requirement, kg.m

   g_c = gravitational acceleration, m/sec^2

   ρ = density of the fluid, kg/m^3

   μ = viscosity of the fluid, kg/m sec

   D_a = Diameter of the vessel

For unbaffled vessel:

$$\phi = N_p = \frac{Pg_c}{\rho N^3 D_a} \qquad \text{for } N_{Re} \leq 300$$

$$\phi = N_p = \frac{\alpha - \log_{10} N_{Re}}{N_{Fr}\beta} \qquad \text{For } N_{Re} > 300$$

$$N_{Fr} = \frac{N^2 D_a}{g} \qquad \text{Fraud Number}$$

Here the values of α & β are given as the function of Diameter of the agitator:

Diameter Da	Da/D	α	β
10	0.3	1.0	4.0
15	0.33	1.0	4.0

If the configuration changes the graph will be changed and the also the values.

After calculating the power requirement from the above calculations the losses of power we have to consider because equation gives the actual values not the losses.

Losses due to the fittings and other attachments.

Losses due to transmission and glands (0.5 – 5.0hP or 10 – 20% of maximum requirement)

    Basic quantities
    Characteristic length: Impeller diameter D (m)
    Characteristic time: Inverse impeller speed: $1/N$ (s)
    Characteristic mass: Liquid density and cube
        of impeller diameter: ρD^3 (kg)

    Derived quantities
    Characteristic velocity: Impeller diameter and speed: DN (m/s)
    Characteristic pressure: Density and velocity
        square: $\rho D^2 N^2$ (Pa)
    Characteristic flow rate: Velocity and area $ND^3 \left(m^3/s \right)$

Reynolds number

$$N_{RE} = \frac{D_a^2 n \rho}{\mu}$$

baffled → **Power number** N_P (Fig. 1) ⟹ **Power**

$$P = \frac{N_P n^3 D_a^5 \rho}{g_c}$$

Unbaffled (curve D) → **Froude number**

$$N_{Fr} = \frac{n^2 D_a}{g}$$

$$N_{P(Corr)} = N_P \times N_{Fr}^m$$

$$m = \frac{a - \log_{10} N_{Re}}{b}$$

Constants a & b (Table 9.1)

Power Number Curves for Various Type of Impeller

(i) Power number N_P vs. Reynolds number Re for turbines and impellers

(ii)   Power number N_P vs. Reynolds number Re for marine propellers and helical ribbons

(iii)   Power correlation for a 6-blade turbine in pseudo plastic liquids

(iv)   Power required for complete suspension of solids in agitated tanks using pitched-blade turbines

(v)   Power correlation for a 3-blade Propeller Type agitator

Fig. 4.6 Power correlations for single three bladed propellers

(vi)   for turbine type agitator with 6 flat blades liquid height equal to vessel height and 4 baffles are installed

Fig. 4.9 Power correlations for baffled turbine impellers

impeller width : D - impeller diameter.

Power Consumption

- Power required to drive impeller $q = nD_a^3 N_Q$, $E_k = \dfrac{\rho(V_2')^2}{2g_c}$

- V'$_2$ slightly less than tip speed, u$_2$ $\alpha = V_2'/u_2$ $V_2' = \alpha \pi n D_a$

- Power Requirement $P = \dfrac{\rho n^3 D_a^5}{g_c}\left(\dfrac{\alpha^2 \pi^2}{2} N_Q\right)$ $P = \dfrac{N_P n^3 D_a^5 \rho}{g_c}$

- At low N_{Re} (<10), density is no longer a factor $N_P = \dfrac{K_L}{N_{Re}}$

$$P = \dfrac{K_L n^2 D_a^3 \mu}{g_c}$$

- At N_{Re}>10,000 in baffled tanks, P is independent of N_{Re} and viscosity is not a factor

$$N_P = K_T$$

$$P = \dfrac{K_T n^3 D_a^5 \rho}{g_c}$$

- K_L and K_T are constants for various types of impellers and tanks

Type of Impeller	K_L	K_T
Propeller, 3 blades Pitch 1.0 Pitch 1.5	41 55	0.32 0.87
Turbine 6-blade disk (S_3=0.25 S_4=0.2) 6 curved blades (S_4=0.2) 6 pitched blades (45°, S_4=0.2) 4 pitched blades (45°, S_4=0.2)	65 70 - 44.5	5.75 4.80 1.63 1.27
Flat paddle, 2 blades (45°, S_4=0.2)	36.5	1.70
Anchor	300	0.35

Example

A flat-blade turbine with six blades is installed centrally in a vertical tank. The tank is **1.83 m** in diameter, the turbine is **0.61 m** in diameter & is positioned **0.61 m** from the bottom of the tank. The turbine blades are **127mm** wide. The tank is filled to a depth of **1.83m** with a solution of 50% caustic soda at 65.6°C, which has a viscosity of **12cP** and a density of **1498 kg/m³**. The turbine is operated at **90 rpm**. What power will be required to operate the mixer if the tank was baffled?

**Solution:**

Solution (a) baffled

n = 90rpm / 60 s = 1.5 r/s

D_a = 0.61m

μ = 12cP = 12x10^{-3} kg/ms

$$N_{RE} = \frac{D_a^2 n \rho}{\mu} = \frac{(0.61)^2 (1.5)(1498)}{12 \times 10^{-3}} = 69600$$

For Re > 10000, Np = K_T = 5.8 from curve A for baffle (NRe = 69600),
NP = 5.8 (or from table 2 given before)

$$P = \frac{N_P n^3 D_a^5 \rho}{g_c}$$

$$= (5.8)(1.5)^3 (0.61)^5 (1498) = 2476.6 \; mN/s =$$

$$2476.6 W$$

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Solution (b) unbaffled

From Fig 1, curve D (N_{Re} = 69600), N_P = 1.07

Froude number,

$$n = 90\text{rpm} / 60 \text{ s} = 1.5 \text{ r/s}$$
$$D_a = 0.61\text{m}$$
$$\mu = 12\text{cP} = 12 \times 10^{-3} \text{ kg/ms}$$
$$N_{RE} = 69600$$

$$N_{Fr} = \frac{n^2 D_a}{g} = \frac{(1.5)^2 (0.61)}{9.81} = 0.14$$

$$m = \frac{a - \log 10 \, N_{Re}}{b} = \frac{1.0 - \log_{10} 69600}{40} = -0.096$$

$$N_{P(Corrected)} = N_P \times N_{Fr}^m = 1.07 \times 0.14^{-0.096} = 1.29$$

Thus power,

$$P = \frac{N_P n^3 D_a^5 \rho}{g_c} = (1.29)(1.5)^3 (0.61)^5 (1498)$$
$$= 550 mN / s = 550 \, W$$

Newtonian and non-Newtonian Fluids
➢ Newtonian fluids:
   Fluids which obey the Newton's law of viscosity are called as Newtonian fluids. Newton's law of viscosity is given by
➢ Non-Newtonian fluids:
   Fluids which do not obey the Newton's law of viscosity are called as non-Newtonian fluids. Generally non-Newtonian fluids are complex mixtures: slurries, pastes, gels, polymer solutions etc.

$$\tau_{xy} = \mu \frac{\partial v_x}{\partial y}$$

There is also one more - which is not real, it does not exist - known as the ideal fluid. This is a fluid which is assumed to have no viscosity. This is a useful concept when theoretical solutions are being considered - it does help achieve some practically useful solutions.

Fig. 19. Various Non-Newtonian Fluids

Fig. 2.8 (a) Shear stress, τ, and (b) apparent viscosity, η, as a function of deformation rate for one-dimensional flow of various non-Newtonian fluids.

Power Consumption in Non-Newtonian Liquids

Newtonian fluid:

$$\tau_{xy} = \mu \frac{dv_x}{dy} = \mu \dot{\gamma} \qquad : \qquad \dot{\gamma} = \frac{dv_x}{dy}$$

Power Law Fluid:

$$\tau_{xy} = K \dot{\gamma}^n \qquad : \qquad \mu_a = \frac{\tau_{xy}}{\dot{\gamma}} = K \dot{\gamma}^{n-1}$$

When n<1, viscosity decreases with shear

When n>1, viscosity increases with shear

- Non-Newtonian liquids – viscosity varies with shear rate
- Use apparent viscosity, μ_a

$$N_{Re,n} = \frac{n D_a^2 \rho}{\mu_a}$$

$$\mu_a = K' \left(\frac{du}{dy} \right)_{av}^{n'-1}$$

- For a straight-blade turbine in pseudoplastic liquids

$$\left(\frac{du}{dy} \right)_{av} = 11n$$

SHAFT DESIGN

→ Shaft can be attached to the vessel in vertical, horizontal or angular positions.

→ It is preferable to use the bearing either at top of the vessel or at bottom. It can be placed externally or internally to the vessel.

DESIGNING OF THE SHAFT CAN BE DONE BY THREE WAYS:

1. Based on torque
2. Based on Bending moment calculations
3. Based on the critical speed of the agitator.

Based on torque:

Continuous average rated torque is given by:

$$T_c = \frac{hp * 75 * 60}{2\pi N}$$
where N is speed in rpm

Maximum Torque possible in agitation system is at start up condition and the value is given by

$T_m = 1.5$ to $2.5 T_c$

Once equipment is in the running mode it will have the stresses because of torque, viscous force of the fluid, turbulence will create the centrifugal force.

Maximum stresses are given by the equation

$$f_s = \frac{T_m}{Z_p}$$

where for   Shear stress $Z_p = \frac{\pi}{32} d^3$

Bending stress $Z_p = \frac{\pi}{16} d^3$

Based on Moment Calculation

Maximum bending moment is given by $M_e = \frac{1}{2}\left[M + \sqrt{M^2 + T_M^2}\right]$

Torque T_m is resisted by a force for acting at a radius of $0.75R_b$ from the axis of agitator shaft;

$$F_m = \frac{T_m}{0.75R_b}$$

where R_b = radius of blade

$M_{max} = F_m * \ell$ where ℓ = shaft length

And the stresses are given by

$$f = \frac{M_{max}}{Z}$$ $< f\text{J}$

Based on Critical Speed:

It is difficult to calculate the unbalanced forces due to asymmetric construction of agitator. Fixing certain counter balance weight in the opposite direction to it can easily eliminate this.

It is necessary to control the deflection of shaft by adequate support.

The speed at which the shaft vibrates violently is called as the critical speed of the shaft.

Range of 70% to 130% of critical speed should be avoided.

Diameter should be so chosen that the normal working speed should not fall in this range.

The deflection due to concentrated load

Fig. 14.14   Shaft supported in bearings

$$\delta = \frac{W\ell^3}{3EI}$$

Deflection due to weight of shaft

$$\delta_s = \frac{w\ell^4}{8EI}$$

$$N_c = \frac{60*4.987}{\left(\delta_1 + \delta_2 + \delta_3 + \dfrac{\delta_s}{1.27}\right)^2} \quad \text{rpm}$$

Main parts of agitator are HUB & BLADES

Hub is attached to shaft by Keys & Bolts.

The load on the blade is assumed to act as 75% of the agitator radius. This will create a bending moment, which will be maximum at the point where the blade is attached to the hub.

Maximum bending moment $M_{max} = F(0.75R_b - R_h)$

If the blades are flat $f = \dfrac{\text{max B.M}}{Z} = \dfrac{F\left(0.75R_b - R_h\right)}{b_t * b_w^2 / 6}$

Where b_w = Blade width & b_t = blade thickness

The hub is fixed to the shaft by key, which transmits the shaft torque to the impeller. The hub is subjected to the bending moment due to force on the blade and to shear force due to the torque. It is assumed the outside hub diameter as twice the shaft diameter and check the shear stress due to torque.

Couplings:

It is used to join the agitator shaft to the drive shaft. There are three types of coupling in the industry: (1) Flange coupling (2) split muff coupling (3) simple coupling.

Stabilizers:

There are some unbalanced forces which causes the shaft to vibrate vigorously. To damp such vibrations a stabilizer is used. This is essentially in form of the rings.

Chapter 3
MIXING TIME

1. INTRODUCTION

The *'mixing time'* is the time measured from the instant of addition until the vessel contents have reached a specified degree of uniformity when the system is said to be 'mixed'.

Time taken for a volume of fluid added to a fluid in mixing vessel to blend throughout the rest of the mixing vessel to a *pre-chosen degree of uniformity*.

Mixing provides confidence that:

a. Any sample of product drawn from the vessel will have known chemical composition.
b. An added species will become well distributed with certain length of time.
c. Any thermal gradients due to chemical reaction or difference in bulk and feed temperature will be eliminated.
d. All regions of the mixing vessel are moving and mixing with all other regions.
e. Measurement of the bulk temperature or the concentration of a chemical species is representative of the entire vessel contents.

2. FLOW VISUALIZATION

This is a necessary step before carrying out mixing time calculation.

Objectives of flow visualization
⦿ Suitable point of addition of tracer material to study mixing system.
⦿ Possible choice of feed location.
⦿ Probe location for mixing time calculation.

Provides information on
• Flow of fluids in vessel
• Highlights region of poor fluid motion
• Flow compartmentalization within mixing system

Flow Visualization Technique
• Light sheet visualization
• Hot wire anemometry
• PIV method

Fig. 1 Flow Visualization

3. EVALUATION OF MIXING PERFORMANCE

Methods to evaluate mixing performance:

a. Characterization of homogeneity.
b. Blending time.

General methods to characterize homogeneity:

a. Visual uniformity.
b. Quantitative change in local concentration as a function of time.
c. Review instantaneous statistics about the spatial distribution of the species.
    i. Average concentration
    ii. Minimum and maximum
    iii. Standard deviation in the concentration.
    iv. Coefficient of variation CoV = standard deviation/average.

4. PROBE LOCATION

- As mixing time techniques work on the principle of addition of material to the vessel which has different properties from the bulk.
- The decay of material property fluctuations is used to measure the mixing time for the system.
- From flow visualization study one can identify position of probe so with minimum number of probe one can extract maximum data.
- Identified locations are:
  - First probe is installed at less intensely agitated area where mixing rates are lower.
    - Last point to be mixed in vessel.
    - Will control and limit the mixing rate for entire mixing process.
  - Second probe should be chosen to indicate well mixed bulk of fluid.
  - Third probe should be used to provide axial velocity for flow compartmentalize.
- Measuring the tracer concentration as a function of time c(t) in one or more points in the vessel, is a common experimental method.
- The mixing time is then the time it takes for the measured concentration c(t) to stay within a certain range of the final concentration c*.
- Advantage: easy to use in experiments.
- Disadvantage: uses only one or a few points in the vessel.
- Does not use all information present in a CFD simulation.

Fig.2 Probe Location

Fig.3 Concentration History

MEASUREMENT OF MIXING TIME

Measure degree of mixing using residence time distribution (RTD) of a tracer dye e.g.
Residence time distributions are important when calculating scale-up factors

Non-Ideality in Chemical process Vessels

Measure conc of dye

Up-flow reactor with substrate
added at the base of the reactor,
product etc exits at the top

Add dye as
a pulse

**Examine pattern of appearance of the pulse of dye in exit stream I.e plot
concentration versus time**

| Plug flow | Normal | Channelling | Back mixing |

RESIDENCE TIME DISTRIBUTION

⦿ The concept of residence time distribution (RTD) and its importance in flow processes first
developed by Danckwerts (1953) was a seminal contribution to the emergence of chemical
engineering science.

- ◉ The residence time distribution measures features of ideal or non-ideal flows associated with the bulk flow patterns or *macro-mixing in a reactor or other* process vessel.
- ◉ The term *micro-mixing, applies to spatial* mixing at the molecular scale that is bounded but not determined uniquely by the residence time distribution.
- ◉ The bounds are extreme conditions known as *complete segregation and maximum mixedness.*
- ◉ Suppose that a sample of fluid is collected and analyzed. One may ask: Is it homogeneous?
- ◉ Standard measures of homogeneity such as the striation thickness in laminar flow or the coefficient of variation in turbulent flow can be used to answer this question quantitatively.

RTD EXPERIMENTS

- ◉ Transient experiments with inert tracers are used to determine residence time distributions.
- ◉ Different RTD functions are given in [Literature](#).

Name	Symbol	Input Signal	Output Signal	Physical Interpretation	Properties
Washout function	$W(t)$	Negative step change in tracer, concentration from an initial value of C_0 to a final value of 0	$W(t) = C_{out}(t)/C_0$	$W(t)$ is the fraction of particles that remained in the system for a time greater than t.	$W(0) = 1$ $W(\infty) = 0$ $dW/dt \leq 0$
Cumulative distribution function	$F(t)$	Positive step change in tracer concentration from an initial value of 0 to a final value of C_∞	$F(t) = C_{out}(t)/C_\infty$	$F(t)$ is the fraction of particles that remained in the system for a time less than t.	$F(0) = 0$ $F(\infty) = 1$ $dF/dt \geq 0$
Differential distribution function	$f(t)$ or $E(t)$	Sharp impulse of tracer	$f(t) = \dfrac{C_{out}(t)}{\int_0^\infty C_{out}(t)\,dt}$	$f(t)\,dt$ is the fraction of particles that remained in the system for a time between t and $t + dt$.	$f(t) \geq 0$ $\int_0^\infty f(t)\,dt = 1$
Convolution integral	$C_{out}(t)$	Any time-varying tracer concentration	$C_{out}(t) = \int_{-\infty}^{t} C_{in}(\theta)f(t-\theta)\,d\theta$ $= \int_0^\infty C_{in}(t-\theta)f(\theta)\,d\theta$	The output signal is a damped response that reflects the entire history of inputs.	$[C_{out}]_{max} \leq [C_{in}]_{max}$

- ◉ What is Tracer?
  Any component that has difference in color than actual system can be used as tracer.
- ◉ How to select Tracer?
  Selected based on hydrodynamic and chemical properties.
  Density and viscosity should be very near to the fluid under consideration.
  Chemically inert to the system.
- ◉ What should be quantity of Tracer?
  Depends upon the volume of the reactor and also on the flow rates of reactants and products.
- ◉ How to do analysis
  If possible go for gravimetric analysis.
  GC-MS or spectrophotometer can also be used.

EXPERIMENTAL METHOD FOR MIXING TIME

- Acid/base/indicator reactions
- Electrical conductivity variations
- Temperature variations
- Refractive index variations
- Light-absorption techniques

Factors to be considered

- The manner of tracer addition
- The position and number of recording points
- The sample volume of the detection system
- The criterion adopted for deciding the cut-off point of the end of the experiment.

Batch Experiments
- Mixing time depends upon the way in which tracer is added and the location of point A.
- End point of the experiment is difficult to detect with precision.

Fig. 3. Batch Experimental Setup

Fig. 4 Results of Batch Experiments

Colorimetric Method

Fig. 5 Step by Step Colorization of tank

- Add dye and see how dye moves through the fluid.
- If colored tracer is added to an agitated vessel, it is not possible to identify the last point of mixing because any dye in front of or behind the poorly mixed region will mask the last mixed point.
- So better to go for Decolorization reaction.

Fig. 6 Schematic of Dye addition for mixing visualization in 3D.
(a) Poorly mixed pocket is uncolored and cannot be seen in 2D
(b) A poorly mixed pocket is colored and can clearly be seen in 2D

Above figure suggest that why method Decolorization is better than colorization for mixing time study. In colorization it is not possible to identify the last point of mixing because any dye in front of or behind the poorly mixed region will mask the last mixed point to be mixed remains colored or marked while the rest of the tank is stripped of its color.

To avoid this problem, technique commonly used is known as dye Decolorization. The entire contents of the vessel are colored using one chemical, and then a second chemical is added that removes the color. A poorly mixed region stands out as a pocket or color after the rest of vessel has cleared as shown in fig. 6(b).

Most common method is pH change with an appropriate indicator.
- Eg. Iodine change color in presence of starch
- Eg. NaOH change color in presence of phenolphthalein.

e.g. Iodine-Thiosulfate Method for Mixing Time
The reaction between sodium Thiosulfate and iodine in the presence of starch is a most satisfactory Decolorization reaction, with a strong color change from deep blue to clear.
Preparation of chemicals required:
- 2M Iodine: Dissolve 400g of iodate-free potassium iodide in 0.5L of distilled water in a 1L flask. Add 254g iodine to the flask and agitate until all the iodine has dissolved. Cool to room temperature, and make up to 1L. Store in a dark place to prevent deterioration.
- 1 M sodium Thiosulfate: weigh 248g of sodium Thiosulfate pentahydrate into 1 L volumetric flask and make upto 1L with distilled water. Add a small quantity of sodium carbonate (about 0.2g) to aid preservation.
- Starch indicator: manufacture a paste containing 10g of soluble starch in a little water. Add this drop-wise to a beaker containing 1L of very hot (not boiling) water while stirring constantly until a clear liquid is obtained. Cool the liquid to ambient temperature and dissolve it in 20b of potassium iodide. This solution should be protected from the air.

Procedure for use
- Add about 50ml of starch indicator to the vessel.
- Add sufficient iodine solution to the vessel to produce an intense black color. Note the quantity used.
- While agitator running at the desired speed, quickly incorporate sodium Thiosulfate into the vessel at a rate approximately 2.5 times quantity of iodine solution used. As mixing progresses, the liquid will clarify.
- After each test, slowly add iodine solution until the neutral point is again reached, demonstrated by a hint of light blue appearing. The test can be repeated.

Measurement of Concentration
- **Off-Line Sampling**: If an off-line analysis technique is used, a chemical marker such as a particular salt, dye, or acid is added to the mixing vessel, and samples are removed regularly. The concentration of the marker in each sample is measured, and the degree of uniformity is inferred from these measurements. Installation of a suitable sampling system can be difficult, and this technique is not suitable if the mixing time is very short, since there will generally be a finite sampling time.
- **Schlieren Effect**: The Schlieren based technique relies on the light scattering that occurs when two liquids with different refractive indices are mixed. Light shone through the mixing vessel is scattered by the layers of different liquids and the tank appears cloudy. When the liquid is fully blended, the tank becomes clear once more, giving a mixing time (Van de Vusse, 1955). This technique does, however, require transparent bulk and added liquids, and the liquids must have different refractive indices. The liquids are also, therefore, likely to have different physical properties.
- **Thermocouple based mixing time measurement**: A thermocouple-based mixing time test can be performed by adding a liquid that has a different temperature from the bulk. The temperature at different points in the mixing vessel is monitored over time, and the probe outputs are used to calculate the mixing time. This technique can be used with opaque and/or non-conducting liquids. A disadvantage of this technique is that it may not be suitable if the bulk liquid physical properties are very sensitive to changes in temperature, since the viscosity would then be a function of the temperature (and concentration) of the added liquid.
- **Conductivity probe technique**: The conductivity probe mixing time technique uses an electrolyte in the added liquid as the marker. Conductivity probes monitor the local conductivity as a function of time. If the electrolyte concentration is low, concentration is directly proportional to conductivity. The probe outputs are processed to calculate the mixing time for the system under consideration. This technique is not suitable for measurements in non-conducting systems and cannot be used in systems where the rheological properties of the bulk are sensitive to changes in salt concentration: for example, with certain gums and carrageenans. The technique is, however, cheap and easy to use. Conductivity probes can give very rapid response times, allowing measurements in mixing systems with short mixing times. Further details on probe designs suitable for conductivity-based mixing time measurements are available in Khang and Fitzgerald (1975). The probe itself is made by embedding pieces of platinum or stainless steel wire, soldered onto the

screened signal lead, in a "bullet" of epoxy resin. The diameter of the bullet is approximately 6 mm

Fig. 7 Conductometer Probe

COEFFICIENT OF VARIATION (COV)

When Two fluids are mixed, the quality of radial mixing can be described by the Coefficient of Variation (CoV), which is defined as

$$CoV = \left[\frac{\sum_{i=1}^{n}(C_i - C_{avg})^2}{n-1} \right]^2 \frac{1}{C_{mean}}$$

Where,

　　Ci = sample concentration
　　Cavg = average concentration
　　Cmean = mean concentration
　　n = number of samples

⦿ It describes the deviation of local concentration from the mean within a cross section of vessel.

⦿ Assuming normal distribution, the following can be applied approximately,

- About 2/3 of the values well lie within C(1±CoV)
- About 95% of the values will lie within C(1±2CoV)
- About 99% of the values will lie within C(1±3CoV)

This can be shown graphically as

Fig. 8 CoV Measurement (67% Confidence Interval)

Fig. 9 CoV Measurement (95% Confidence Interval)

Fig. 10 CoV Measurement (99% Confidence Interval)

Fig 11. Experimental Verification of CoV

Processing Mixing Data

Fig. 12 Experimental Data plotted with 95% confidence interval

Log variance

Fig. 13 Log Variance of Output data

Change in Variance with Time

Fig. 14 Decrease in CoV with time showing better mixing phenomena

Mixing Time Calculation

⊙ Normalized Output

$$C_i' = \frac{C_i - C_0}{C_\infty - C_0}$$

⊙ Mixing time is defined as the time required for the normalized probe output to reach and remained between 95% and 105% of final equilibrium value.

⊙ This value is called 95% mixing time.

Degree of Mixing is given by

$$\frac{\theta_n}{\theta_{95}} = \frac{\ln[1.0 - (n/100)]}{\ln(1.0 - 0.95)}$$

Mixing Time

- $t_M = fn(\rho, \mu, N, g, D, T, \text{ other geometrical parameters})$
- $Nt_M = fn(pND^2/\mu, N^2D/g, T/D, \text{ other geometrical ratios})$
- $Nt_M = fn(Re, Fr, \text{ geometrical ratios})$
- For *geometrically similar systems, neglecting the effect of the Froude number,* the dimensionless mixing time becomes a function of Reynolds number only"
- $Nt_M = fn(Re)$

Variation in Mixing time with Reynolds number

Fig. 15 Mixing time variation in Transition Region

Mixing Time in Turbine Type Agitator

Fig. 16 Mixing time calculation for Turbine type agitator

Mixing Time for Turbine (6-bladed)

- ◉ Based on Circulation Loop Model
- ◉ t_T is much greater when N_{Re} is 10-1000, even though power consumption is comparable to the turbulent range

$$q = 0.92 n D_a^3 \left(\frac{D_t}{D_a} \right)$$

$$t_\tau \approx \frac{5V}{q} = 5 \frac{\pi D_t^2 H}{4} \frac{1}{0.92 n D_a^3 \left(\dfrac{D_t}{D_a} \right)}$$

or

Validation of Model with experiments

$$nt_\tau \left(\frac{D_a}{D_t} \right)^2 \left(\frac{D_t}{H} \right) = const = 4.3 \approx 5$$

Fig. 17 Mixing time factor with Reynolds number

Norwood – Metzner Model

$$f_t = \frac{t_\tau \left(nD_a^2\right)^{2/3} g^{1/6} D_a^{1/2}}{H^{1/2} D_t^{3/2}} = nt_\tau \left(\frac{D_a}{D_t}\right)^2 \left(\frac{D_t}{H}\right)^{1/2} \left(\frac{g}{n^2 D_a}\right)^{1/6}$$

Fig. 18 Empirical Correlation for Mixing time

Fox and Gex Model

$$f_t = \frac{t_\tau \left(nD_a^2\right)^{2/3} g^{1/6}}{H^{1/2} D_t} = nt_\tau \left(\frac{D_a}{D_t}\right)^{3/2} \left(\frac{D_t}{H}\right)^{1/2} \left(\frac{g}{n^2 D_a}\right)^{1/6}$$

for a high efficiency impeller (Turbulent regime)

$$nT_T = 16.9 \left(\frac{Dt}{Da}\right)^{1.67} \left(\frac{H}{Dt}\right)^{0.5}$$

Problem:

An agitated vessel 1.83 m in dia contains a six –blade straight-blade turbine 0.6 m in diameter, set one impeller diameter above the vessel floor, and rotating at 80 rpm. It is proposed to use this vessel for neutralizing a dilute aqueous solution of NaOH at 70 ^{0}F with a stoichiometrically equivalent quantity of concentrated HNO3. The final depth of the liquid in the vessel is 1.83 m. assuming that all the acid is added to the vessel at one time, how long will it take for the neutralization to be complete?

Solution:

- ◉ Dt = 1.83 m Da = 0.61 m E = 0.61 m
- ◉ n = 80 /60 = 1.333 /s, density of liquid (given) = 1000 kg/m3, viscosity of liquid (given)
- ◉ Find Re = n Da^2.density/viscosity = 503000
- ◉ Find nt$_T$ from figure and then t$_T$.

$$nt_T\left(\frac{D_a}{D_t}\right)^2\left(\frac{D_t}{H}\right)=const=4.3$$

Mixer Selection
- Choice of impeller can also affect mixing time
- Propellers typically require longer mixing times compared to turbines
  - Propellers have lower power consumption
- Gas bubbles, liquid drops, or solid particles also increase blending time
- No direct relation between power consumed and amount or degree of mixing
- When mixing time is critical, best mixer is one that mixes in required time with least amount of power
  - Mixing time is a compromise arrived at by considering energy cost for mixing and capital cost of equipment

CFD Analysis of Mixing Time

Fig. 19. Mass Fraction variation in different parts of Vessel

Comparison of Various Injection of Tracer using Fluent

Fig. 20 Difference in Velocity profile due to different injection of tracer in tank

Mixing Time for different System

- In Pseudo plastic Liquid, blending time at Reynolds number at about 1000 is much longer than Newtonian Fluid under same impeller conditions.
- At higher Reynolds number there is not much difference between mixing time calculation.
- When gas bubbles, Liquids or solids particles are dispersed in liquid, the mixing time for continuous phase is increases.
- For viscous liquids, the mixing time can be increased up to twice when gas hold up is at 10%.

- As different fluids show a different balance of elasticity and pseudo plasticity it is potentially misleading to make judgments on the basis of low behavior index only.
- Measurements of the mixing rate of pseudo plastic fluids with a turbine impeller 26 gave rates up to 50 times lower than would be expected from similar Newtonian correlations.
- Viscoelasticity leads to flow pattern dampening and reports of both decreased and increased mixing rates have been made. It seems therefore that Viscoelastic properties affect the performance of different mixer designs in different ways and that there may be some opportunities to exploit Viscoelastic properties in mixer design.

Effect of Agitation Speed on Mixing Time

Figure 8.11   Progress of mixing in terms of intensity of segregation ($I_s = \sigma^2/\sigma_0^2$).
(a) Influence of impeller speed (N) on mixing rate; (b) approximate first-order relationship for all speeds in terms of impeller revolutions (Nt)

Chapter 4
Mixing of liquids

Introduction

Mixing may be defined as the 'intermingling of two or more dissimilar portions of a material, resulting in the attainment of a desired level of uniformity, either physical or chemical, in the final product'. Since natural diffusion in liquids is relatively slow, liquid mixing is most commonly accomplished by rotating an agitator in the liquid confined in a tank. It is possible to waste much of this input of mechanical energy if the wrong kind of agitator is used. Parker (1964) defined agitation as 'the creation of a state of activity such as flow or turbulence, apart from any mixing accomplished'. A rotating agitator generates high speed streams of liquid which in turn entrain stagnant or slower moving regions of liquid resulting in uniform mixing by momentum transfer. As the viscosity of the liquid is increased, the mixing process becomes more difficult since frictional drag retards the high speed streams and confines them to the immediate vicinity of the rotating agitator.

In general, agitators can be classified into the following two groups.
1 Agitators with a small blade area which rotate at high speeds. These include turbines and marine type propellers.
2 Agitators with a large blade area which rotate at low speeds. include anchors, paddles and helical screws.
The second group is more effective than the first in the mixing of high viscosity liquids.
The mean shear rate produced by an agitator in a mixing tank +, is proportional to the rotational speed of the agitator

Thus

$$ym = kN$$

where **k** is a dimensionless proportionality constant for a particular system. For a liquid mixed in a tank with a rotating agitator, the shear rate is greatest in the immediate vicinity of the agitator. In fact the shear rate decreases exponentially with distance from the agitator Thus the shear stresses and strain rates vary greatly throughout an agitated liquid in a tank. Since the dynamic viscosity of a Newtonian liquid is independent of shear at a given temperature, its viscosity will be the same at all points in the tank. In contrast the apparent viscosity of a non-Newtonian liquid varies throughout the tank. This in turn significantly influences the mixing process. For shear thinning liquids, the apparent viscosity is at a minimum in the immediate vicinity of the agitator. The progressive increase in the apparent viscosity of a shear thinning liquid with distance away from the agitator tends to dampen eddy currents in the mixing tank. In contrast, for shear thickening liquids, the apparent viscosity is at a maximum in the immediate
vicinity of the agitator. In general shear thinning and shear thickening liquids should be mixed using high and low speed agitators respectively. It is desirable to produce a particular mixing result in the minimum time t and with the minimum input of power per unit volume **P_A/V.**
Thus an efficiency function E can be defined as

$$E = \left(\frac{1}{P_A/V}\right)\left(\frac{1}{t}\right)$$

Small blade high speed agitators

Small blade high speed agitators are used to mix low to medium viscosity liquids. Two of the most common types are the six-blade flat blade turbine and the marine type propeller shown in Figures 1 and 2 respectively. Flat blade turbines used to mix liquids in baffled tanks produce radial flow patterns primarily perpendicular to the vessel wall as shown in Figure **3** In contrast marine type propellers used to mix liquids in baffled tanks produce axial flow patterns primarily parallel to the vessel wall as shown in Figure **4.** Marine type propellers and flat blade turbines are suitable to mix liquids with dynamic viscosities up to 10 and 50 Pa **s,** respectively. Figure 5 shows a turbine agitator of diameter D_A in a cylindical tank of diameter D_T filled with liquid to a height H_L. The agitator is located at a

height H_A from the bottom of the tank and the baffles which are located

I

     Fig. 1 Six-blade flat blade turbine   **Fig. 2** Marine propeller

Fig. 3 Fluid flow for chemical engineers

Figure 4 Standard tank configuration

Immediately adjacent to the wall have a width b. The agitator has a blade width **a** and blade length **r** and the blades are mounted on a central disc of diameter **s. A** typical turbine mixing system is the standard configuration defined by the following geometrical relationships: a six-blade flat blade turbine agitator

1 a six-blade flat blade turbine agitator
2 $D_A = D_T/3$
3 $H_A = D_T/3$
4 $a = D_T/5$
5 $r = D_T/4$
6 $H_L = D_T$
7 4 symmetrical baffles
8 $b = D_T/10$

Processing considerations sometimes necessitate deviations from the Agitator tip speeds **ilT** given by equation **3** are commonly **used** as a standard configuration.measure of the degree of agitation in a liquid mixing system.

$$U_T = \pi D_A N$$

Tip speed ranges for turbine agitators are recommended as follows:
2.5 to **3.3 m/s** for **low** agitation
3.3 to **4.1** *m/s* for medium agitation
4.1 to 5.6 m/s for high agitation

If turbine or marine propeller agitators are used to mix relatively low viscosity liquids in un baffled tanks, vortexing develops. In this case the liquid level falls in the immediate vicinity of the agitator shaft. Vortexing increases with rotational speed N until eventually the vortex passes through the agitator. As the liquid viscosity increases, the need for baffles to reduce vortexing decreases. A marine propeller can be considered as a caseless pump. In this case its volumetric circulating capacity Q_A is related to volumetric displacement per revolution V_D by the equation
where

$$Q_A = \eta V_D N$$

where n is a dimensionless efficiency factor which is approximately **0.6.** V_D is related to the propeller pitch p and the propeller diameter D_A by equation

$$V_D = \frac{\pi D_A^2 p}{4}$$

Most propellers are square pitch propellers where $p = D_A$ so that equation

$$V_D = \frac{\pi D_A^3}{4} \qquad Q_A = \frac{\eta \pi N D_A^3}{4}$$

defined a tank turnover rate I_T by the equation

$$I_T = \frac{Q_A}{V}$$

where V is the tank volume and I_T is the number of turnovers per unit time. To get the best mixing, I_T should be at a maximum. For a given tank volume V, this means that the **-11%** lilating capacity QA should have the highest possible value for the minim n consumption of power. The head developed by the rotatii,, agitator hA can be written as $hA = ClN2Di$
where C1 is a constant.
Combining equations gives the ratio

$$\frac{Q_A}{h_A} = \frac{CD_A}{N}$$

where C is a constant.
Since the mean shear rate in a mixing tank ym is given by equation

$$ym = kN$$

The ratio of circulating capacity to head Q_A/h_A is low for high shear agitators. For mixing shear thinning liquids a high circulating capacity Q_A and a high shear rate ym or head hA are both desirable. In this case a compromise has to be made.

Large blade low speed agitators
Large blade low speed agitators include anchors, gates, paddles, helical ribbons and helical screws. They are used to mix relatively high viscosity liquids and depend on a large blade area to produce liquid movement throughout a tank. Since they are low shear agitators they are useful for mixing shear thickening liquids. A gate type anchor agitator is shown in Figure Anchor agitato

Fig. 5 Gate type anchor agitator

operate within close proximity to the tank wall. The shearing action of the anchor blades past the tank wall produces a continual interchange of liquid between the bulk liquid and the liquid film between the blades and the wall [Holland and Chapman .Anchors have successfully been used to mix liquids with dynamic viscosities up to 100Pas,

Helical screws normally function by pumping liquid from the bottom of a tank to the liquid surface. The liquid then returns to the bottom of the tank to fill the void created when fresh liquid is pumped to the surface. A rotating helical screw positioned vertically in the centre of an unbaffled cylindrical tank produces a mild swirling motion in the liquid. Since **the** liquid velocity decreases towards the tank wall, the liquid at the wall of an

Fig. 6 Flow pattern in a baffled helical screw system

unbaffled tank is nearly motionless. Baffles set away from the tank wall create turbulence and facilitate the entrainment of liquid in contact with the tank wall. The flow pattern in a baffled helical screw system is shown in Figure 5.7. Baffles are not required if the helical screw is placed in **an** off-centred position since in **this** case the system becomes self-baffling. However, off-centred helical screws require more power to produce a comparable mixing result. Gray investigated the mixing times of helical ribbon agitators and found the following equation to hold:

$$Nt = 30$$

where N is the rotational **speed** of the helical ribbon agitator and *t* is the batch mixing time.

Designing of Agitator for Liquid-Liquid Mixing

Dimensionless groups for mixing :
In the design of liquid **mixing** systems the following dimensionless groups are of importance.

The power number

$$Po = \frac{P_A}{\rho N^3 D_A^5}$$

The Reynolds number for mixing Re_M represents the ratio of the applied to the opposing viscous drag forces.

$$Re_M = \frac{\rho N D_A^2}{\mu}$$

The Froude number for mixing FYM represents the ratio of the applied to the opposing gravitational forces.

$$Fr_M = \frac{N^2 D_A}{g}$$

The Weber number for **mixing** WeM represents the ratio of the applied to the opposing surface tension forces.

$$We_M = \frac{\rho N^2 D_A^3}{\sigma}$$

The power number Po can be related to the Reynolds number for mixing Re_M, and the Froude number for mixing Fr_M , by the equation

$$Po = CRe_M^x Fr_M^y$$

 where *C* is an overall dimensionless shape factor which represents the
geometry of the system.
Equation can also be written in the form

Example
Calculate the theoretical power for a six-blade flat blade turbine agitator with diameter *DA* = **3** m
running at a speed of *N* = 0.2 reds in a tank system conforming to the standard tank configuration

illustrated in Figure The liquid in the tank has a dynamic viscosity μ = 1 Pa s and a density of p = 1000 kg/m3.

$$Re_M = \frac{\rho N D_A^2}{\mu}$$

Substituting the given values

$$Re_M = \frac{(1000 \text{ kg/m}^3)(0.2 \text{ rev/s})(9.0 \text{ m}^2)}{1.0 \text{ Pa s}}$$

$$Re_M = 1800$$

From the graph of ϕ against Re_M in Figure 5.8

$$\phi = Po = 4.5$$

The theoretical power for mixing is

$$P_A = Po\rho N^3 D_A^5$$

$$= (4.5)(1000 \text{ kg/m}^3)(0.008 \text{ rev}^3/\text{s}^3)(243 \text{ m}^5)$$

$$= \underline{8748 \text{ W}}$$

Example *2*
Calculate the theoretical power for a six-blade flat blade turbine agitator with diameter DA = 0.1 m running at N = 16 reds in a tank system without baffles but otherwise conforming to the standard tank configuration illustrated in Figure The liquid in the tank has a dynamic viscosity of μ = 0.08Pas and a density of p = 900kg/m3. For this configuration α = 1.0 and β = 40.0.

Calculations:
The Reynolds number for mixing is given by

$$Re_M = \frac{\rho N D_A^2}{\mu}$$

$$= \frac{(900 \text{ kg/m}^3)(16 \text{ rev/s})(0.01 \text{ m}^2)}{0.08 \text{ Pa s}}$$

$$= 1800$$

$$P_A = \phi \rho N^3 D_A^5 \left(\frac{N^2 D_A}{g}\right)^y$$

Now

$$y = \frac{\alpha - \log Re_M}{\beta}$$

with

$$\alpha = 1.0 \text{ and } \beta = 40.0$$

and

$$\log 1800 = 3.2553$$

Therefore

$$y = \frac{-2.2553}{40} = -0.05638$$

Substituting known values

$$N^2 D_A \underline{\quad} (256 \text{ rev}^2/\text{s}^2)(0.1 \text{ m})$$

Substituting known values

$$\frac{N^2 D_A}{g} = \frac{(256 \text{ rev}^2/\text{s}^2)(0.1 \text{ m})}{9.81 \text{ m/s}^2}$$

$$= 2.610$$

So

$$\left(\frac{N^2 D_A}{g}\right)^y = 2.610^{(-0.05638)} = 0.9479$$

Therefore

$$P_A = (2.2)(900 \text{ kg/m}^3)(4096 \text{ rev}^3/\text{s}^3)(0.00001 \text{ m}^5)\,(0.9479)$$

$$= \underline{76.88 \text{ W}}$$

Chapter 5
GAS-LIQUID DISPERSION DESIGN

1. **What is Dispersion?**
    - As in solid suspension, the size and surface area of the solid particles exposed to the liquid are fixed, in Gas-Liquid or Liquid-Liquid dispersion operations, the size of bubbles and drops and total interfacial area between two phases vary with conditions and degree of mixing.
    - New area will be created against the force of interfacial tension.
    - Drops or bubbles are continuously coalescing and being redispersed.
    - In gas dispersion process, bubbles rise through the liquid pool and escape from the surface and replaced by new ones.

2. **What an effecting dispersion should achieve?**
    - Gas must be *effectively* and *efficiently* contacted with liquid to provide better *Mass Transfer* and *Chemical Reaction*.
    - Sometimes gas *provides energy* for Mixing the liquid.
    - Mass Transfer is important in processes like *fermentation* and *effluent treatment processes*.
    - Chemical Reaction is important in processes like *chlorination, sulfonation* etc.

3. **Scope of Gas Liquid Dispersion Process**
    - Required residence time for either phase
    - Allowable pressure drop
    - Relative flow rates of gas and liquid
    - Need for countercurrent contact
    - Local mass transfer performance (dispersion size and turbulent mass transfer)
    - Need to supply or remove heat
    - Corrosion considerations
    - Presence of solid particles
    - Foaming behavior and phase separation
    - Relative importance of micro mixing
    - Flow pattern requirements of reaction scheme
    - Interaction of reaction with mass transfer
    - Rheological behavior in laminar and transitional flow regimes

4. **The roles of the agitator in a gas-liquid system**
    - Break the gas into small bubbles for high interracial area;
    - Disperse the bubbles throughout the liquid;
    - Keep the bubbles in the liquid (i.e. recirculation) for sufficient time;
    - Mix the liquid throughout the vessel;
    - Provide turbulent eddies to feed liquid to and from the interfaces;
    - and possibly
    - Move the liquid past heat exchange surfaces and maximize heat transfer coefficients;
    - Maintien particules in suspension

5. **Classification of Gas-Liquid Equipments**
    Contactors in which the liquid flows as a thin film

     a. Packed columns

     b. Trickle bed reactors

     c. Thin-film reactors

     d. Rotating disk reactors

Contactors in which gas is dispersed into the liquid phase

     a. Plate columns (including control cycle reactors)

     b. Mechanically agitated reactors (principally stirred tanks)

     c. Bubble columns

     d. Packed bubble columns

     e. Sectionalized bubble columns

     f. Two-phase horizontal contactors

     g. Co-current pipeline reactors

     h. Coiled reactors

     i. Plunging jet reactors, ejectors

     j. Vortex reactors

Contactors in which liquid is dispersed in the gas phase

     a. Spray columns

     b. Venturi scrubbers

6. **Selection of Equipments**

   Emphasis is based on the basic mechanisms involved.

   Turbulent process – high speed agitators

        a. Higher contact time

        b. When reaction is important

   Laminar regions – Static Mixers

        a. Lower contact time

        b. When mass transfer is important

Table 1. General Guideline for Selection of Gas-Liquid Contacting Equipments

Type of Absorber	Typical Gas Velocity $\times 10^2$ (m/s)	Residence Time Distribution		Residence Time of Liquid	Fractional Liquid Holdup	$k_L \times 10^4$ (m/s)	a (m²/m³)	$k_L a \times 10^2$ (s⁻¹)
		Gas	Liquid					
Film type: packed column and trickle bed reactors	10–100	Plug	Plug	Very low	0.05–0.1	0.3–2	2–35	0.06–7
With gas dispersed as bubbles in liquids								
Bubble columns	1–30	Plug	Mixed	Unlimited	0.6–0.8	1–4	2.5–100	0.25–40
Packed bubble columns	1–20	Plug	Mixed	Unlimited	0.5–0.7	1–4	10–30	1–12
Bubble cap plate columns	50–200	Plug	Mixed	Unlimited	0.7–0/7	1–4	10–40	1–16
Plate columns without downcomers	50–300	Plug	Mixed	Limited variation	0.5–0.5	1–4	10–20	1–8
Mechanically agitated contactors	0.1–2	Mixed	Mixed	Unlimited	0.5–0.8	1–5	20–100	2–50
Horizontal pipeline contactors	5–300	Plug	Plug	Low	0.1–0.8	2–6	10–40	2–24
Static mixers	0.05–20	Plug	Plug	Low	0.01–0.99	1–20	10–100	10–200
With liquid dispersed in gas								
Spray columns	5–300	Mixed	Plug	Very low	—	0.5–1.5	2–15	0.1–2.25
Sieve plate in spray regime	100–300	Plug	Mixed	Unlimited	—	1–3	5–20	0.5–6

This data is based on air-water system but can be used for system having same hydrodynamics properties of fluids.

7. **Process Objectives**

- *Turbulent Mechanisms*. The processes of liquid mixing, generation of interface area, and gas–liquid mass transfer in turbulent systems are controlled primarily by the *power* dissipated in the fluids and the *gas volume fraction* φ. The power (together with the fluid properties) influences the *bubble size*. The gas is broken up into a dispersion of bubbles in a high-shear zone such as at

the discharge from the sparger holes in a bubble column, the impeller tips in an agitated vessel, or the gas inlet and wall-shear zones in a static mixer. It is the power dissipated in that zone which controls the bubble breakup process. However, with agitated vessels the design correlations are commonly based on the average energy dissipation per unit mass in the vessel, $P/\rho V$. The power in this expression is the sum of the shaft power and the (principally potential) energy introduced as a result of injecting the gas at depth (Middleton et al., 1994). It may be noted that the ratio of local to average energy dissipation rates can be large and will differ between impeller types.

The bubbles may or may not subsequently recoalesce to some extent, depending on the local fluid dynamics and the interfacial behavior. The unpredictability of this rules out a priori prediction of bubble size and interface area in general, so design via scale-up from experiments is preferred. The *gas fraction* in an agitated vessel is determined by the bubble size and the degree of bubble recirculation [itself a function of agitation, bubble size, and scale (Middleton, 1997)]. For a static mixer, φ is largely set by the ratio of the average gas flow to liquid flow, but with corrections for bubble "slip," which depend on flow orientation and the bubble size.

- *Factors Influencing the Power*. In a given baffled agitated vessel, with given fluid properties, the independent variables controlling P and φ are the impeller type, impeller diameter, impeller speed, and gas rate. However, the gas rate for a process is usually set by the process flow sheet, that is, by the stoichiometry and the required inlet and outlet gas compositions (or absorption efficiency), so the contribution of gas buoyancy to the total energy dissipation rate is fixed. Calculation of the other (usually main) contribution, being the impeller power input per unit mass, $P/\rho V$, is well established for single phase systems. For some (unfortunately, still common) impeller types such as the Rushton disk turbine and the down flow pitched blade turbine, the impeller power draw is greatly reduced when gas is introduced. The power draw is affected by the degree of gas recirculation and to some extent by the detailed geometry of the equipment. Modern gas–liquid impellers, such as the concave-blade disk turbines (Scaba SRGT, Chemineer BT6 and CD6 impellers, and the Lightnin R130; sample shown in Figure 11-5) and the up-pumping wide-blade hydrofoils (Lightnin A345, Prochem MaxfloW, APV B6; sample shown in Figure 11-5), maintain more than 70% of their ungassed power draw on gassing.

  For an inline mixer, a value for the specific power $(P/\rho V)$ can easily be estimated from the manufacturer's (or measured) friction factor, adjusted for the gas–liquid ratio using the correction of Lockhart and Martinelli (1944). Although this may not be rigorously applicable, some success has been achieved by applying the approach to static mixers using the laminar gas–turbulent liquid regime factors.

- *Liquid Mixing*. The bulk circulation is the rate-determining step for *liquid mixing* (*blending)* in stirred vessels. The turbulence ensures that mixing on smaller scales (meso mixing and micromixing) is comparatively fast. (Note, however, that extremely fast reactions can be even faster than the micromixing.) Again, the gas affects this. At modest gas rates, the gas affects the intensity of liquid mixing because of its effect on the impeller power, and its location because of changes to the flow field. At high gas fractions, presumably, the gas buoyancy must contribute.

- *Gas–Liquid Mass Transfer*. Good *mass transfer* performance requires large interface area between gas and liquid (resulting directly from small bubble size and high gas fraction, given the fixed gas rate) and a high mass transfer coefficient (associated with local levels of turbulence). A high gas fraction is not always desirable since the profitability of a reactor is largely controlled by the quantity of liquid it contains. Excessive gas retention may also lead to overreaction. It is only necessary to allow enough time for the required mass transfer.

- *Heat Transfer*. *Heat transfer* in the turbulent regime is essentially a macro mixing process. Heat transfer coefficients are controlled by the turbulence levels (hence boundary layer

thickness) near the heat transfer surfaces. In many cases the process demands of suspension or dispersion and mass transfer are more than sufficient to ensure adequate heat transfer.

- *Solid Particles*. *Particle suspension* from the base and *drawdown* from the surface are often required in gas–liquid agitated vessels and are influenced in a complex manner by gassing. There are no well-established correlations for the influence of gas. Particle suspension is probably controlled by the energy and frequency of turbulent bursts, and drawdown by details of local flow patterns and vortices at the surface, both of which could be expected to be affected by the presence of gas bubbles.

- *Flow Patterns*. Flow patterns can be important. A "slow" reaction scheme (occurring in the bulk liquid) with competing steps may exhibit selectivity dependent on the local concentration of a liquid or dissolved gas reactant. In this case the liquid flow pattern (i.e., whether the liquid undergoes back mixing or plug flow or, as is almost always the case, somewhere in between) is important. A "fast" reaction scheme (occurring mainly near the gas–liquid interface) with dependence of selectivity on local dissolved gas concentration, will be sensitive to the history of gas concentration in the bubbles as they travel through the reactor. In other words, the selectivity will be sensitive to the degree of back mixing of the gas phase, and therefore to the bubble flow pattern. Even for simple gas–liquid mass transfer, the gas flow pattern is critical unless a very small proportion of the dissolvable gas is absorbed per pass. For example, if 95% of the inlet dissolvable gas is absorbed, its mean concentration in the gas phase (and hence its mean transfer rate to the liquid), if in plug flow, is 5.17 times that for an perfectly back mixed gas phase.

 Here a conflict can arise in an agitated vessel. High power input per unit mass is required to enhance mass transfer area and heat transfer coefficient, but this will result in a high degree of gas recirculation, reducing the mean gas phase concentration "driving force" for mass transfer. Local shear rates will also increase with power input. The balance will vary with scale.

8. **Configuration of GL equipments**

 If it is not clear whether it is practical to achieve turbulent flow, an outline design will be useful. For example, if an agitated vessel is to be used (see below), take a typical power number (e.g., 0.8 if a Lightnin A345 up flow hydrofoil is to be used, or 5.0 for a six-blade Rushton turbine) and an impeller diameter of 0.4T (or 0.33T), where T is the intended vessel diameter, and calculate the speed N required to provide a specific power input of, say, 2 kWm−3 (see Section 11-1.4.2). The Reynolds number can then be calculated and compared with that required to give turbulent mixing. If the fluid is non-Newtonian, an appropriate viscosity will be that at a shear rate of about 10 times N, the agitator speed (Metzner and Otto, 1957; see Section 9-3). Skelland (1967) gives a table of the constant for a number of impeller types.

Tank diameter	T
Liquid depth	H
Agitator speed	N
Impeller diameter	D
Off-bottom clearance	C
Blade width	W
Submergence	S

Four baffles, T/10 or T/12
C in range T/4 to T/2
D in range T/4 to T/2
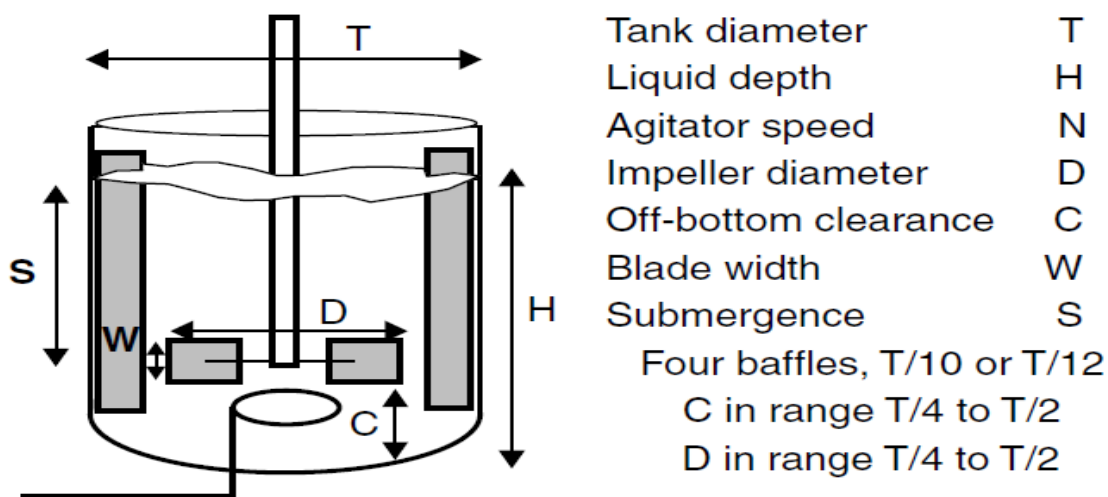
Fig. 1 Standard Agitated Vessel with H = T

Multiple Impeller with Hydrofoil
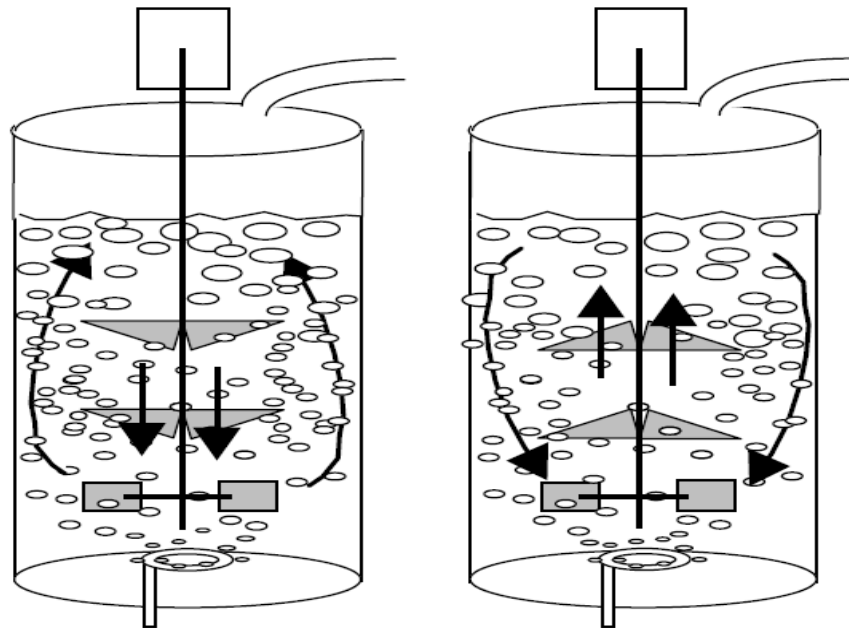
Fig.2 Multi Impeller System

For in-line mixing, gas will generally be fed to the inlet of a static mixer (see Figure 3), preferably via an axially positioned feed pipe or, with larger mixer diameters, via a multipoint distributor.

- *Simple bubble columns: I*n which gas is sparged into the liquid at the bottom of the contactor
- *Plate columns: G*as is redispersed and disengaged at each plate up the column
- *Mechanically agitated vessels: I*n which an impeller rotates in a tank (usually baffled) to give enhanced rates of mass transfer
- *In-line static mixers: M*ixing energy is derived from the flow itself
- *Jet ejectors: G*as is sucked into a liquid flow in a tube and dispersed by intense turbulence
- *Plunging jets : G*as is entrained into the liquid surface by a liquid jet
- *Surface aerators: A*gitators at the liquid surface entrain gas from the head space into the liquid.
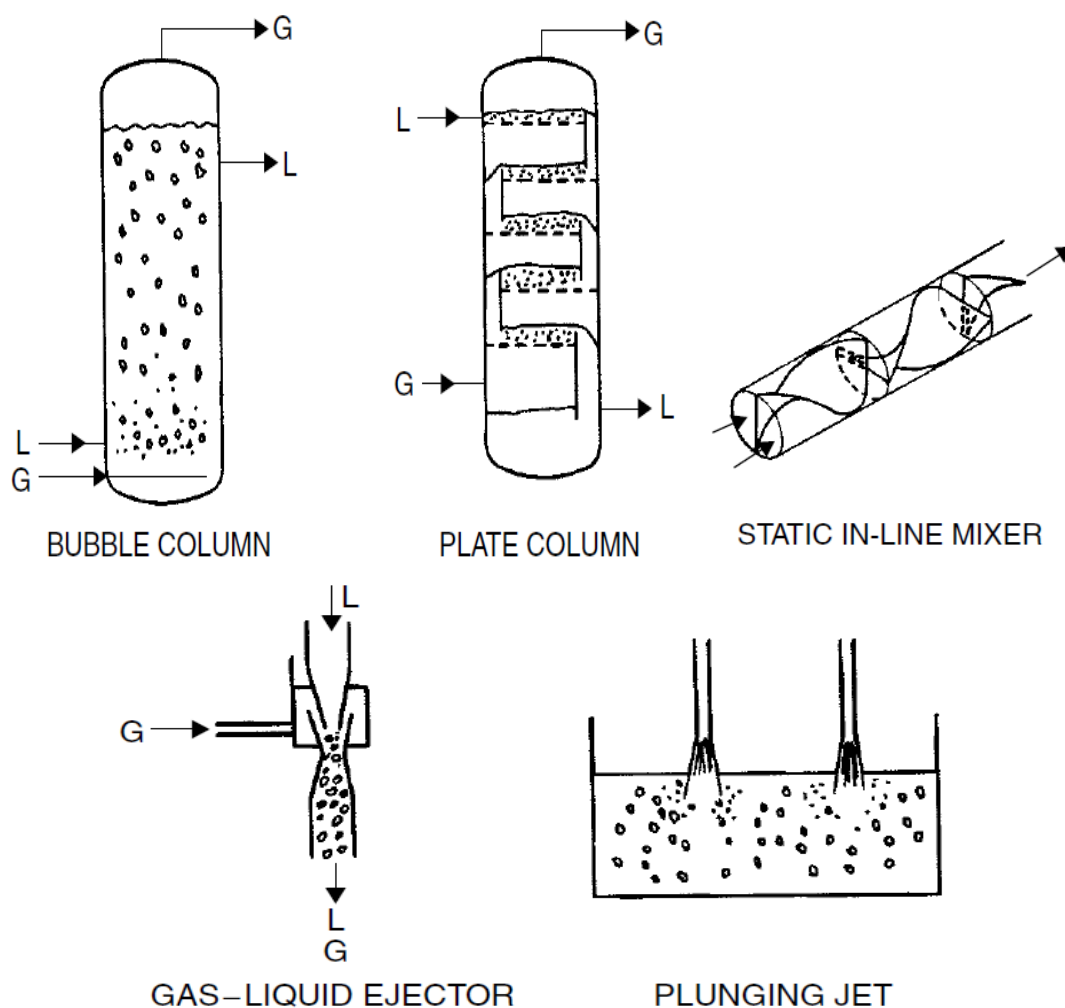
Fig 3. Gas–liquid contacting equipment for low viscosity liquids. (From Middleton, 1997; reproduced by permission of Butterworth–Heinemann.)

In some cases, sufficient gas pressure may not be available (e.g., if avoiding the dangers of compressing hydrogen) and the gas can be drawn in by means of the energy in the liquid flow. In a vessel, gas is drawn down from the headspace using, preferably, a proprietary *self-inducing agitator*, which draws gas down a hollow shaft to the impeller (see Figure 4).
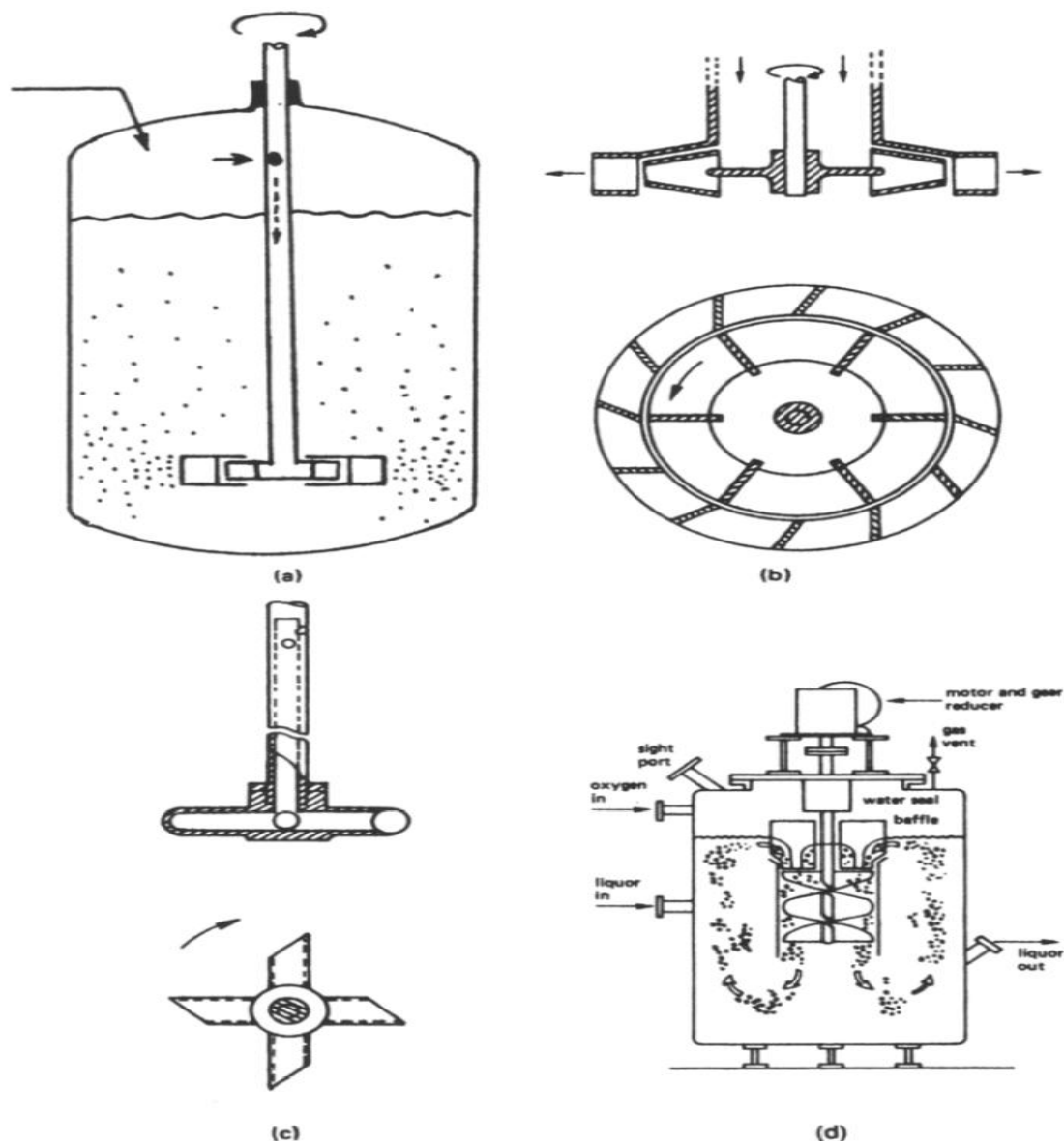
Fig. 4 Self Inducing Agitators

- These communicate the low pressure region behind the blades with the head space *(Figure a) or a gas supply source via a hollow shaft and* therefore use some of the input shaft energy to draw in the gas.
- In many cases multiple blades, sometimes curved like a centrifugal pump impeller, and a close-clearance stator around the agitator.
- Some self-inducing agitators of this type are shown in *Figure (b) and (c). They eliminate the* need for recycle compressors, but are not capable of drawing in very large gas flow rates and are, of course, inflexible in that gassing rate depends upon agitator speed.
- Another type based on a screw injector principle which offers better bulk mixing and enhanced solid suspension for three-phase systems has been developed by Praxair 63 *(Figure d).*

Facts About self Inducers
- Applied to effluent treatment areas.
- Simplest is agitated vessel.

- Not applied in case of drawing gas depth greater than 2.5m.

Mixing of non-Newtonian Fluids

(a)   dough mixer

(b)   dynamic in-line mixer

(c)   scraped-film contactor

(d)   rotating disc contactor
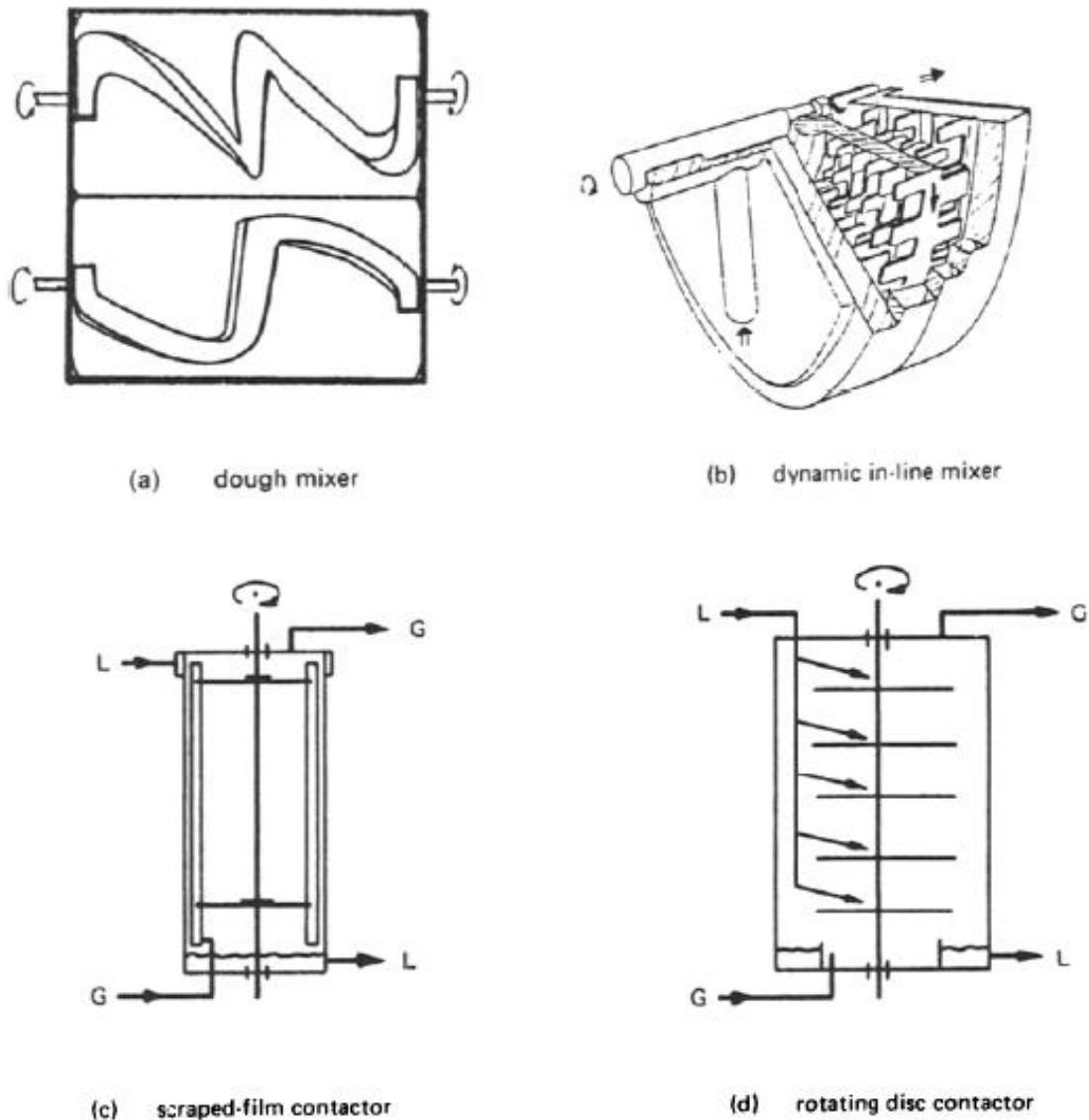
Fig. 5 Gas Liquid Contacting Devices for High Viscosity (Non-Newtonian) Fluids

- For high viscosities, turbulence cannot be achieved in practice and a mechanism which is a combination of distributive and laminar shear or elongation mixing has to be used for incorporation of gas.
- *Dough mixers*. Bubbles are incorporated into the liquid surface;
- *Dynamic in-line mixers*. These are capable of generating high shear rates suitable for producing a high gas content;
- *Scraped-film devices*. A thin film is generated by a blade moving near a surface.

Factors To be Considered for Selection of Equipment
- The rheological properties of the liquid
- The presence of solid particles
- The location of the main mass transfer resistance (in the gas or the liquid)
- The optimization of heat transfer requirements;
- The desired flow pattern for each phase (especially for reactors)
- The importance of gas disengagement and foaming.

9. **Types and configurations of turbulent gas-liquid**
   Single Impeller
   An impeller that approximately maintains the ungassed power level when gas is introduced will give more stable operation and minimal scale-up difficulties. Recommended types (Figure 11-5) include, for radial flow, hollow-blade designs such as the Scaba SRGT, Chemineer CD6 or BT6, Lightnin R130, or for axial flow, an upward-pumping wide-blade hydrofoil such as the Lightnin A345 or A340 or the Prochem-Chemineer MaxfloW. Down flow hydrofoils or pitched blade turbines may be unstable during gas–liquid operation. The liquid flow induced by a down pumping impeller is opposed to the natural tendency of buoyant gas to rise. With a single impeller this is evidenced in the transition between indirect and direct loading that occurs as the gas flow is increased. At certain impeller speeds there may be an accumulation of gas below the impeller plane which can become hydrodynamically unstable. These physical phenomena, which are independent of scale, have been found, within the authors' experience, to lead to an unpredictable loading of the impeller and a source of mechanical problems.
   A single up flow hydrofoil may not be optimum in a vessel with H = T, if the D/T ratio is larger than say 0.5 (which may occur if high $P/\rho V$ is required), since recirculation will be localized and zones of high local gas fraction will be formed.
   Multiple Impeller
   In vessels taller than H/T = 1.2, or when Reynolds numbers are below about 5000, additional impellers may be required. These would improve the liquid mixing, but also, especially in the heterogeneous regime or at high gas velocities, will help to redisperse and redistribute gas from the large bubbles which otherwise tend to bypass the impellers. Generally, spacing between impellers should be larger than their diameter D; otherwise, the flow patterns will interact and the power dissipated by the combined impellers will be less than the sum of the individuals. Multiple radial impellers tend to generate zoned or compartmentalized flow fields, in contrast with the better top-to-bottom circulation generated by multiple axial flow configurations. A combination of a radial flow impeller to produce dispersion together with one or more axial flow impellers is often recommended. Many operators use upward-pumping wide blade hydrofoils (D/T approximately 0.6) even though there is a tendency for these to develop regions of very high gas fraction in the upper part of the vessel.
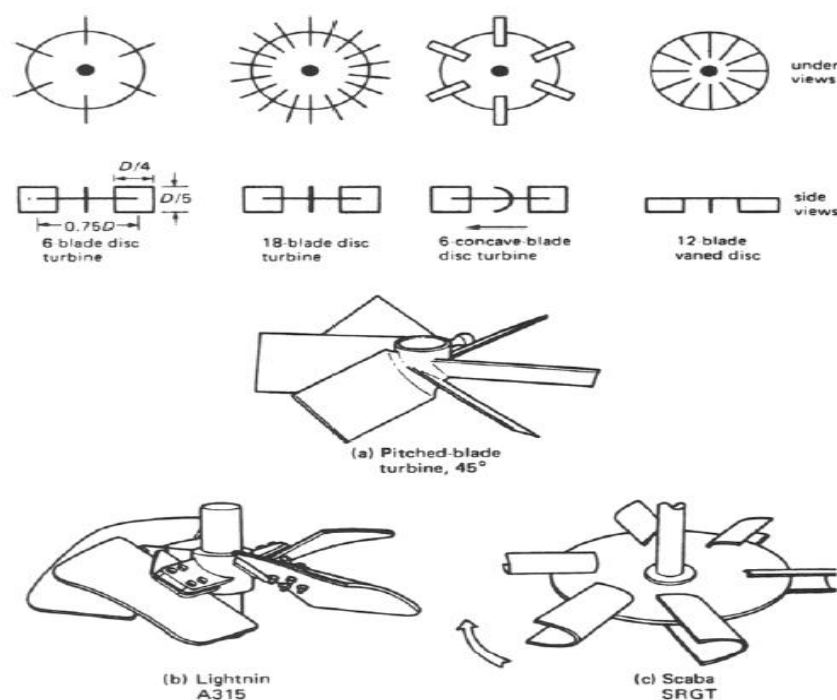
Fig. 6 Different Impellers used for Gas-Liquid Dispersion

10. Intensity of mass transfer and turbulence required

- *Low intensity:* kLa values (air–water equivalent) of order 0.005 s−1; for slow reactions, without a severe particle suspension or heat transfer duty. Large liquid volume is required since the reaction occurs throughout the liquid phase. Here a *bubble column* should be considered: possibly with packing to enhance the plug flow characteristics of the gas. Where it is appropriate to enhance the driving force for mass transfer by using countercurrent flow, or if the liquid needs to be nearer plug flow, a *plate column* may be selected. To meet low cost and intensity requirements when liquid flow pattern is not an issue, *plunging jets* could be considered. See Figure 3 and Table 1.
- *Moderate intensity:* kLa of order 0.05 s−1; for fast reactions with other slower steps; where particle suspension and/or heat transfer require enhancement. *Agitated vessels* are useful here, and indeed are often selected where the intensity needs are uncertain, or may vary widely (as in general purpose reactors). The larger top surface area per unit volume than can be achieved with bubble columns allows higher exit gas flow rates without liquid entrainment and carryover.
- *High intensity:* kLa of order 0.5 s−1; for very fast reactions and short residence times: *Static mixers in turbulent flow* offer plug flow in both phases. *Thin-film contactors* such as wiped-film columns or spinning disks offer large surface per unit volume, giving very rapid mass transfer and evaporative flux.

11. Flow Pattern and Operating regime

- For a mass transfer rate controlling process gas flow patterns are important
- For a reaction rate controlling process liquid flow patterns are important
- Flow patterns are either Perfectly Plug Flow or Perfectly Back Mixed.
- When Plug flow is required a stirred vessel or loop reactor with inline mixer is preferred.
  - For long residence time a cascade of stirred vessel or loop reactor is commonly used.
  - For shorter residence time static mixer or ejector is used.
- Continuous flow systems required perfect back mixed flow.

In the *homogeneous regime* in an agitated vessel, the superficial gas velocity, vS < 0.02 to 0.03 m/s (lower value for lower N), and the bubbles have a mono modal size distribution with a small mean size, generally between 0.5 and 4 mm. Here, the impeller controls the flow pattern and bubble size. At higher gas superficial velocities, the *heterogeneous regime* occurs, in which the bubble size distribution is bimodal, with some large bubbles (say 10 mm or greater), and is controlled more by the gas velocity (possibly void fraction) than by the agitator. In this regime the influences of impeller speed and gas rate are different from those in the homogeneous regime.

Gas flow pattern is important. It controls the degree of recirculation and back mixing of the gas phase, which in turn determines the mean concentration driving force for mass transfer. It can also profoundly affect the liquid-phase macro circulation and homogenization. One way to quantify the gas back mixing is to use the *recirculation ratio*, α, defined as the ratio of the gas flow recirculated to the impeller to that sparged. Since in the homogeneous regime gas is mixed with other gas only at the impeller, α represents the degree of back mixing of the gas. This implies that there is little coalescence in the bulk of the two-phase mixture in the reactor. In large scale equipment (larger than about 1 m3) liquid velocities are usually less than in small scale vessels, so even when the gas distribution is described as homogeneous (e.g., mon omodal in size distribution), it is unusual for much gas to be recirculated below the level of the (bottom) impeller.

- Degree of Recirculation and Back Mixing will decide mean concentration driving force (ΔC).
- Mean Concentration driving force depends upon Recirculation Ratio (α) = ratio of gas flow recirculated to the impeller to the gas flow sparged.

$$\alpha = c\left(\frac{PD}{V}\right)^{1.42}$$

α is used in mass transfer calculations to estimate the overall mean concentration driving force, as follows: If ΔC is the mean mass transfer driving force $(C^* - C_L)$, where C^* is the equilibrium dissolved gas concentration at the gas–liquid interface and C_L is the bulk dissolved gas concentration, the mean driving force for the vessel is given approximately by

$$\Delta C_{Mean} = \left(C^* - C_L\right) = \frac{\Delta C_{IN} - \Delta C_{OUT}}{(\alpha+1)\ln\left[(\Delta C_{IN} + \alpha\Delta C_{OUT})/(\alpha+1)\Delta C_{OUT}\right]}$$

The flow pattern of the gas depends on the regime of gas–impeller interaction. For six-blade disk-turbine impellers, three regimes of flow in the vessel can be defined, as shown in Figure 7.

Overall dispersion	Partial recirculation	Impeller flooded	Churn-turbulent
Low gas rate, high impeller speed	Moderate gas rate at normal speeds	Moderate void fractions, buoyancy controlled	Extreme void fractions, coalescence controlled

(a) (b)
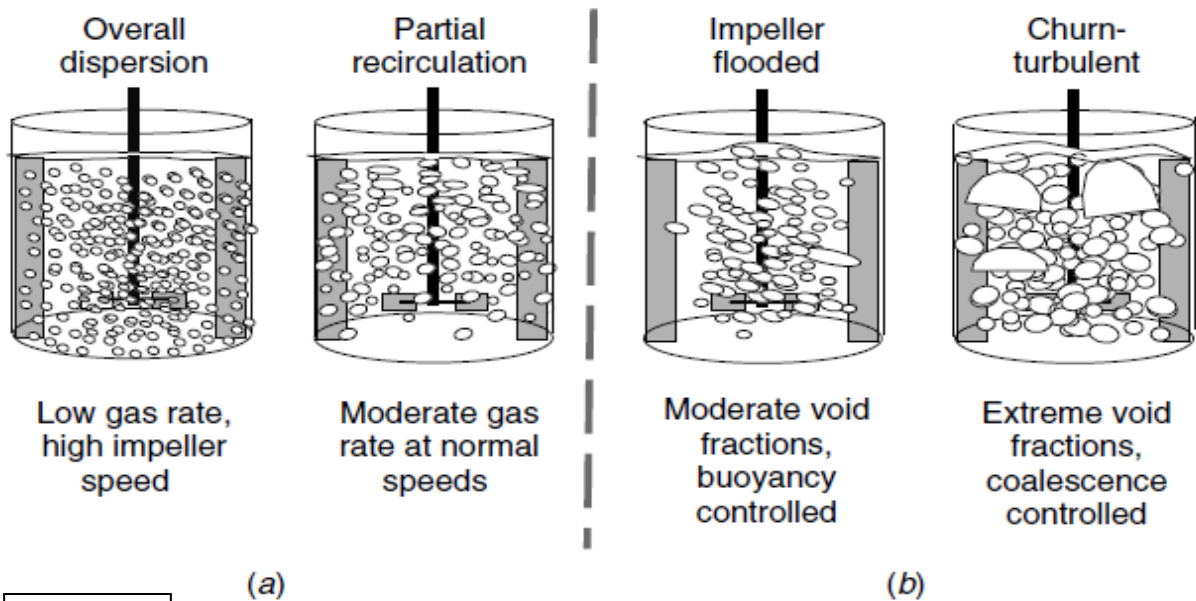
Fig. 7 Typical void fraction distributions in vessels with a single impeller: (a) impeller-controlled regimes; (b) void fraction-controlled regime.

Nienow et al. (5th Europ Conf on Mixing) has given idea about change in mixing pattern with change in gas velocity and impeller speed.
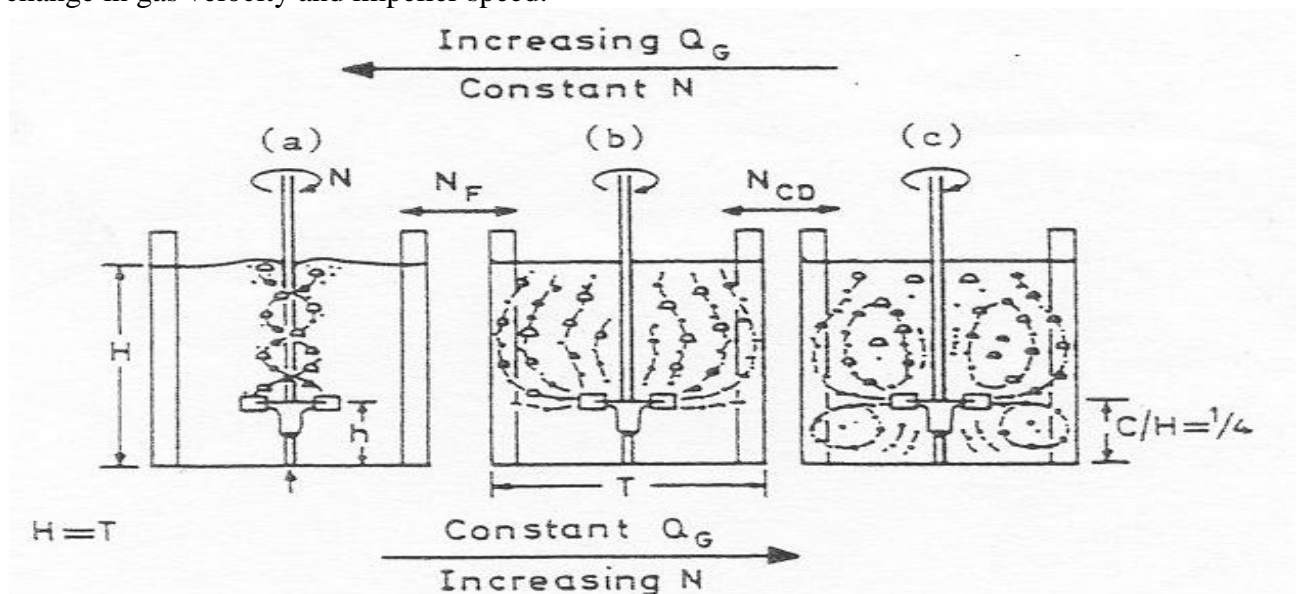
Fig. 8 Mixing Pattern with Gas velocity and impeller speed

Gas Impeller Interaction
- *Flooding* in which the impeller is overwhelmed by gas and gas–liquid contact; mixing, and so on, are very poor
- *Loading* in which the impeller disperses the gas through the upper part of the vessel
- *Complete dispersion* in which gas bubbles are distributed throughout the vessel and significant gas is recirculated back to the impeller
- Cavitation -→ Important in designing
  - At very high gas velocities creating a low pressure region behind baldes.
  - The measure of various interaction is RPD (Relative Power demand)
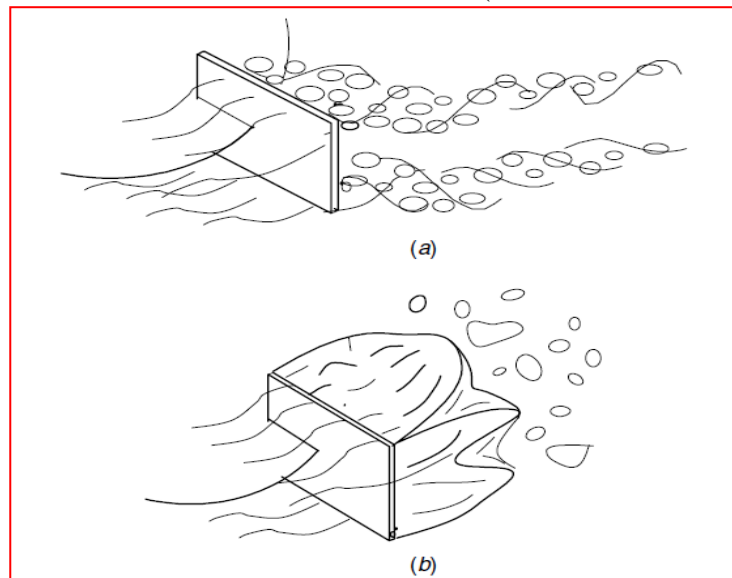
Fig. 9. Fluid at Impeller Blade (a) Vertex Cavity (b) Large Cavity
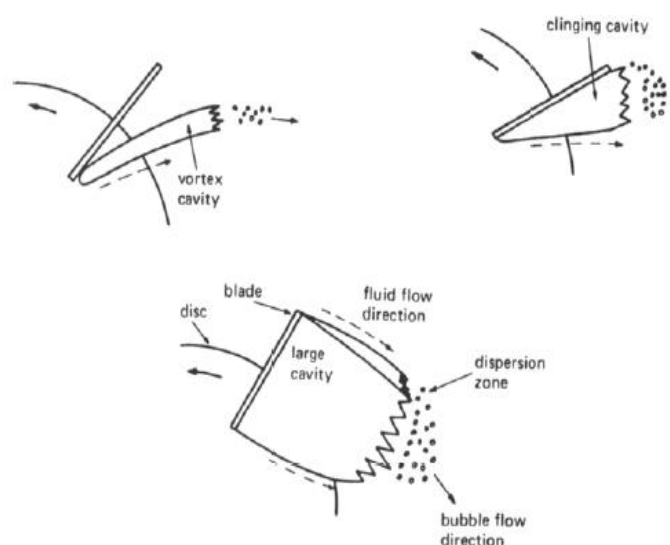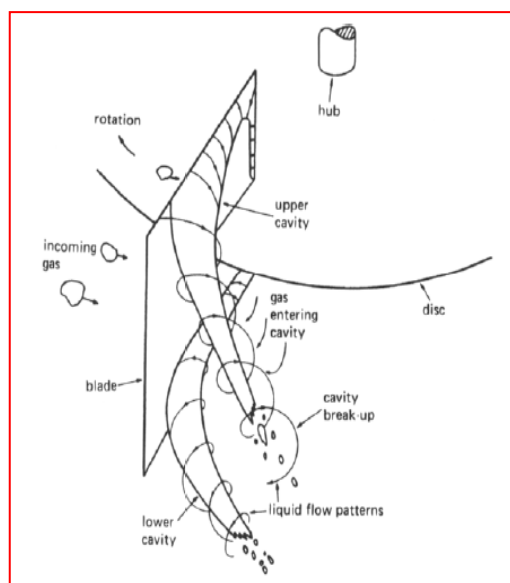
Fig. 10 Vertex Formation behind Impeller Blade

Position of impeller and Interaction
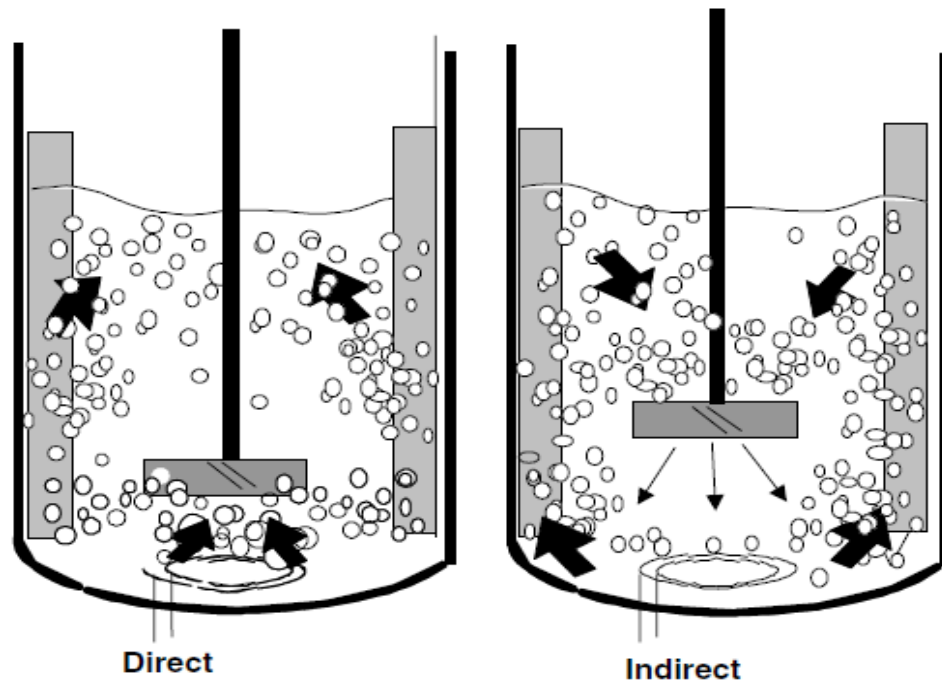
Fig. 11. Effect of Impeller Position on fluid Motion (Curtsy to Warmoeskerken *et.al.* 1984)

The transitions between the various regimes generated by a gassed Rushton turbine can be characterized with the main dimensionless numbers, the gas flow number *(FlG = QG/ND3)*, the impeller Froude number *(Fr = N2D/g)*, and the geometry *(D/T)*

1. Below a certain minimum speed, the impeller has no discernible action. This is approximately when

$$Fr < 0.04 \qquad (1$$

2. The gas flow will swamp the impeller (flooding) if

$$Fl_G > 30 Fr \left(\frac{D}{T}\right)^{3.5} \qquad (11\text{-}4)$$

3. Large cavities are developed by a Rushton turbine when

$$Fl_G > {\sim}0.025 \left(\frac{D}{T}\right)^{-0.5} \qquad (11\text{-}5)$$

4. Nienow et al. (1977) developed a relationship for the speed of a Rushton turbine that would recirculate a given gas rate which can be reformulated and expressed as

$$Fl_G < 13 Fr^2 \left(\frac{D}{T}\right)^{5.0} \qquad (11\text{-}6)$$

Flow Maps for various Impellers

Regimes: (1) below minimum dispersion speed (2) vortex cavities, no recirculation (3) vortex cavities with recirculation, (4) flooded, (5) loaded with large cavities (6) large cavities with recirculation.
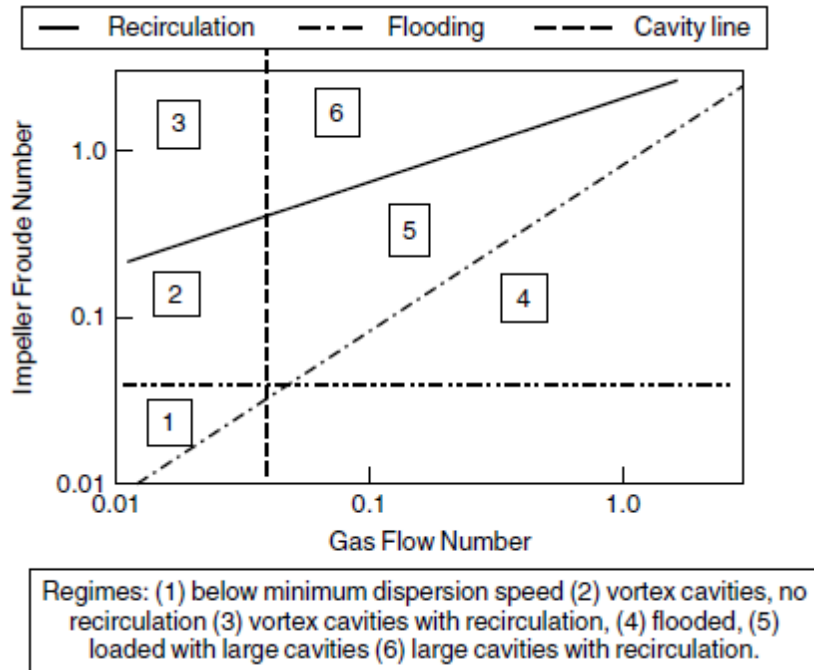
Fig. 12 Flow Map for Single Ruston Type Turbine

$$Fl_G = 3.8 \times 10^{-3} Re^2 (Fr)^{1/15} (T/D)^{0.5}$$
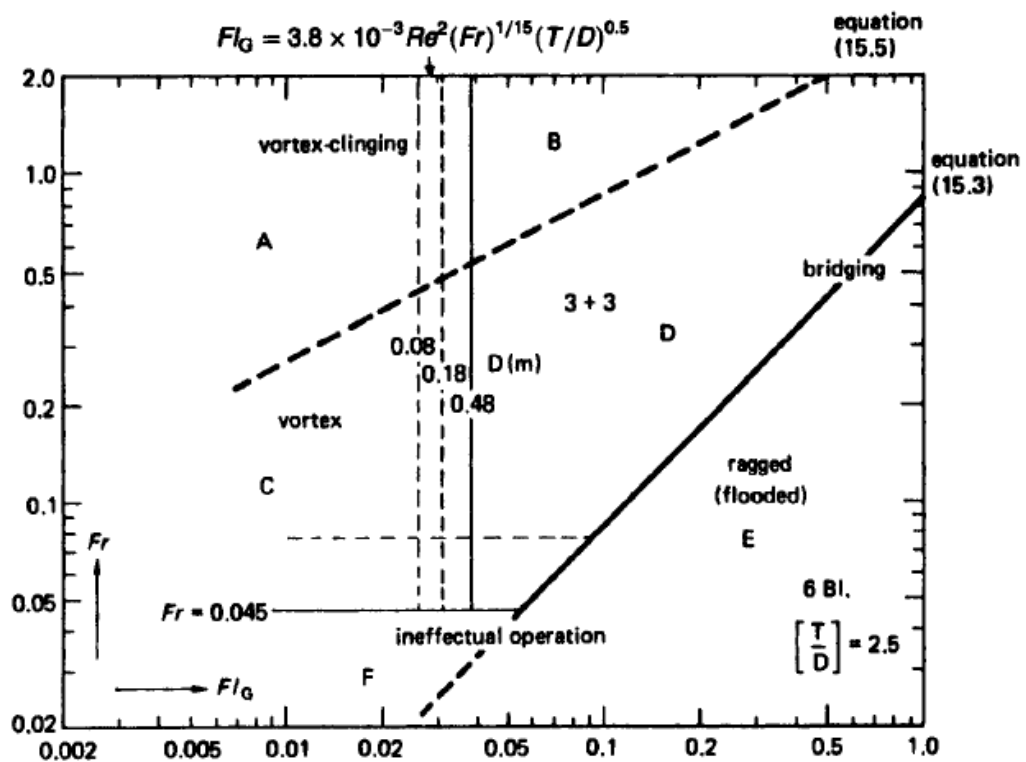
Fig. 13 Flow Map for Turbine
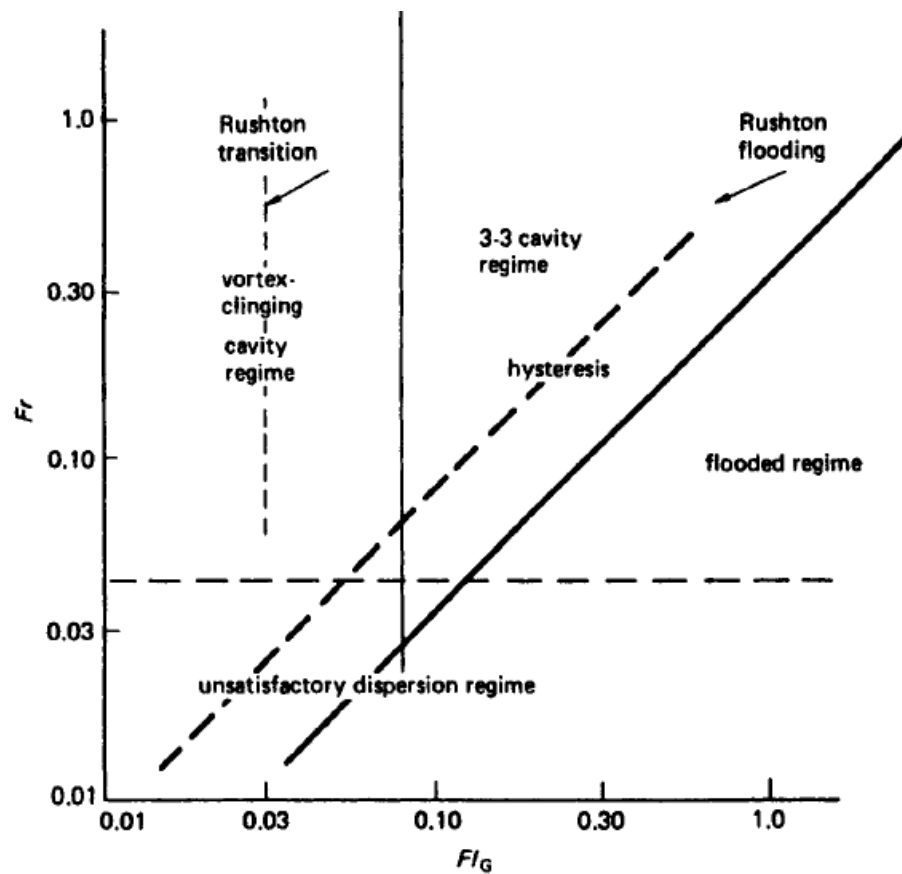
Fig. 14 Concave Blade Type Turbine Impeller
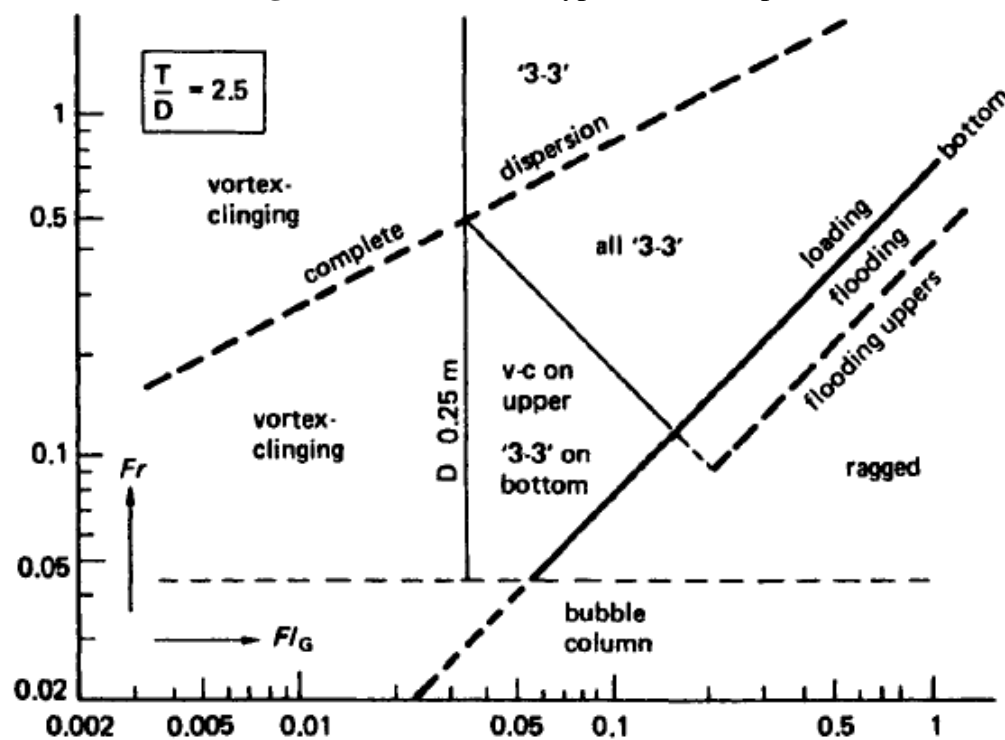
Fig. 15 Two Disc Turbine Type Impeller
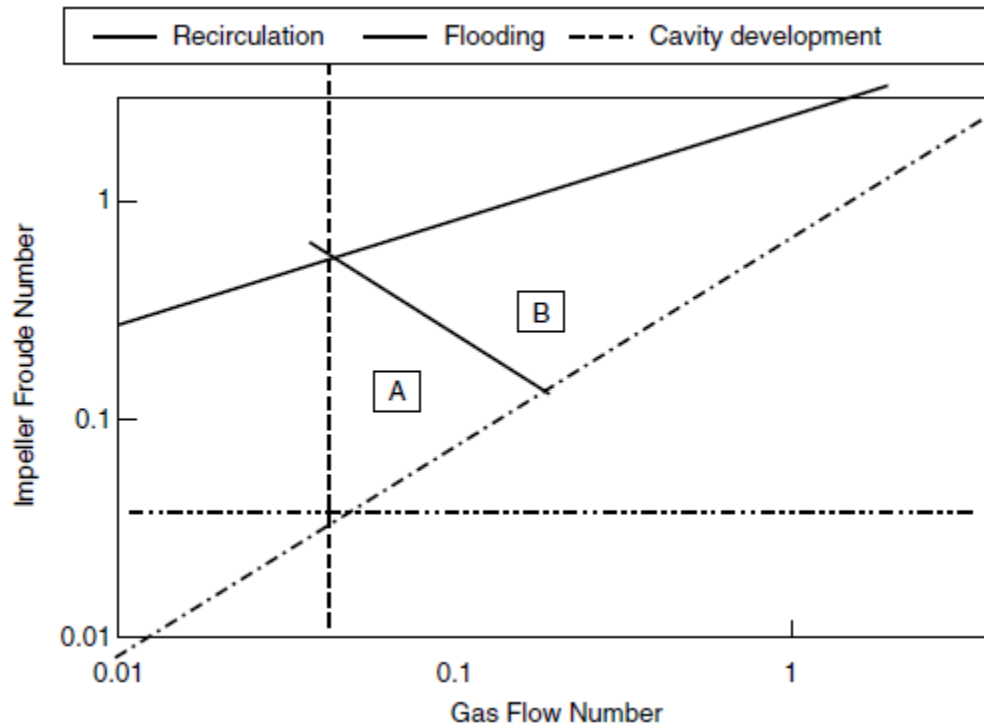
Fig. 16 Flow Map for Triple Ruston Type Impeller

Design Of Gas-Liquid System

*   A design will typically be required to meet specified rates of gas-liquid mass transfer and heat transfer and achieve suitable mixing of the liquid and gas phase at a certain throughput (or batch size).
*   The designer will need to know
    *   The power consumption,
    *   Gas holdup (voidage) fraction,
    *   Foam height
    *   Mixing Time
*   associated with such duty.

Factors to be considered for Design

*   *System Variables*: Viscosity, density and thermal conductivity of the liquid, interfacial tension, diffusion coefficients, chemical reaction rate constants;
*   *Operating variables*: Impeller speed, gas flow rate, liquid volume, pressure;
*   *Equipment Variables*: Impeller type and diameter, geometry of the equipment.

Mass Transfer Rate

$$J = k_L a V \left(C_L^* - C_L \right)$$

Where

J = Mass Flux

k_L = the interphase mass transfer coefficient

a = the interphase surface area,

C* - C = the appropriate average mass transfer driving force

V = the dispersion volume, which is related to the gas hold-up in the vessel,

Main Limitation of above equation is: Coalescing Effect is not considered. No direct Application

Power Number

- Knowledge of the power, P_g, absorbed by the gas-liquid dispersion from the agitator is required for
  - The determination of mass transfer rates, $k_L.a$,
  - Gas holdup
  - Interfacial area on the large scale from small scale tests and,
  - Mechanical design.

Power Input to Turbine Dispersers

- Power input required for gas-liquid system will be less than that of liquid system because of difference in viscosities & Densities of gas and liquid.
- The ratio of powers (RPD) when gas is present to that for the liquid alone depends mainly on the superficial gas velocity and to stirrer speed, tank size, impeller diameter and properties of liquid.

$$P = P_O(RPD)\rho N^3 D^5$$

Table 2. Recommended Power Number for Newtonian Fluid with Re>10^4

Geometry	Po
4 flat blade disc turbine	4.0
6 flat blade disc turbine	5.5
12 flat blade disc turbine	8.7
18 flat blade disc turbine	9.5
6-concave blade disc turbine	4.0
16 blade vaned disc ($W/D=0.1$)	4.0
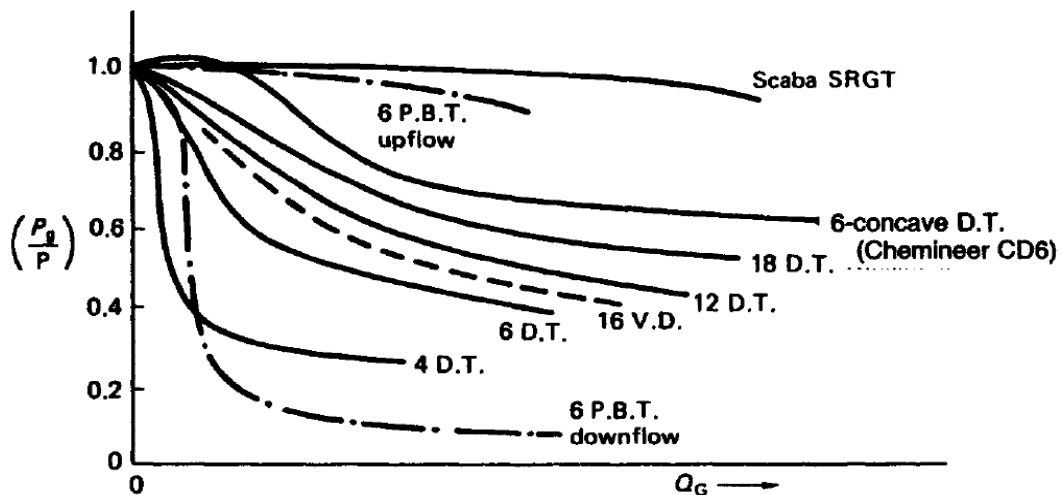
Fig. 17 Degassed Power Number for various Impeller (*D. T. = disc turbine;V.D. = vaned disc; P.B.T. = pitched blade turbine. All curves for one N and D.*)
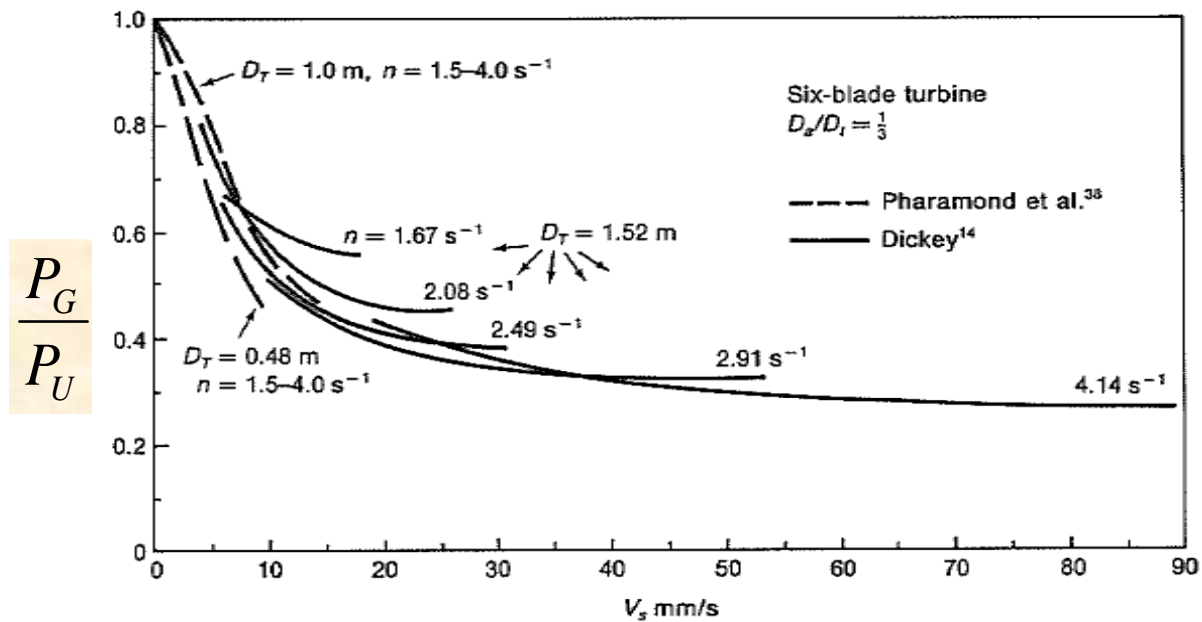
Fig. 18 Power consumption in Air-Water agitated vessel

Power Number

- Function of Type of Impeller used:
    - Perabolic concave blades RPD ≈ 0.4
    - Semicircular Blades ≈ 0.7
    - Flat Bladed turbine with more than 6 blades ≈ 0.4
    - Wide Bladed Hydrofoils ≈ 1.0
    - Axial Flow impellers upflow directions ≈ 0.75
    - Axial Flow impellers downflow directions > 1 (Unfavorable conditions)

Table 3. Comparison of Po for various Impellers

Impeller Type	Po	$(RPD)_{Fl=0.1}$
Radial flow		
6 blade disk turbine[a] D = T/3	5	0.4
12 blade disk turbine, D = T/3	10	0.6
18 blade disk turbine, D = T/3	12	0.7
Chemineer CD6	2.3	0.8
Chemineer BT6	2.0	0.9
Scaba 6SRGT	1.5	0.9
Axial upflow		
4 pitched blade turbine, D = T/3, C = T/3	1.3	0.75
6 pitched blade turbine, D = T/3, C = T/3	1.7	0.75
Lightnin A345, D = 0.4T	0.8	0.75
Axial downflow		
4 pitched blade turbine, D = T/3, C = T/3	1.3	0.3
6 pitched blade turbine, D = T/3, C = T/3	1.7	0.4
Prochem MaxfloW 5, D = 0.45T	1.3	0.7
Lightnin A315, D = 0.4T	0.8	0.7

Bujalski et.al. 1987
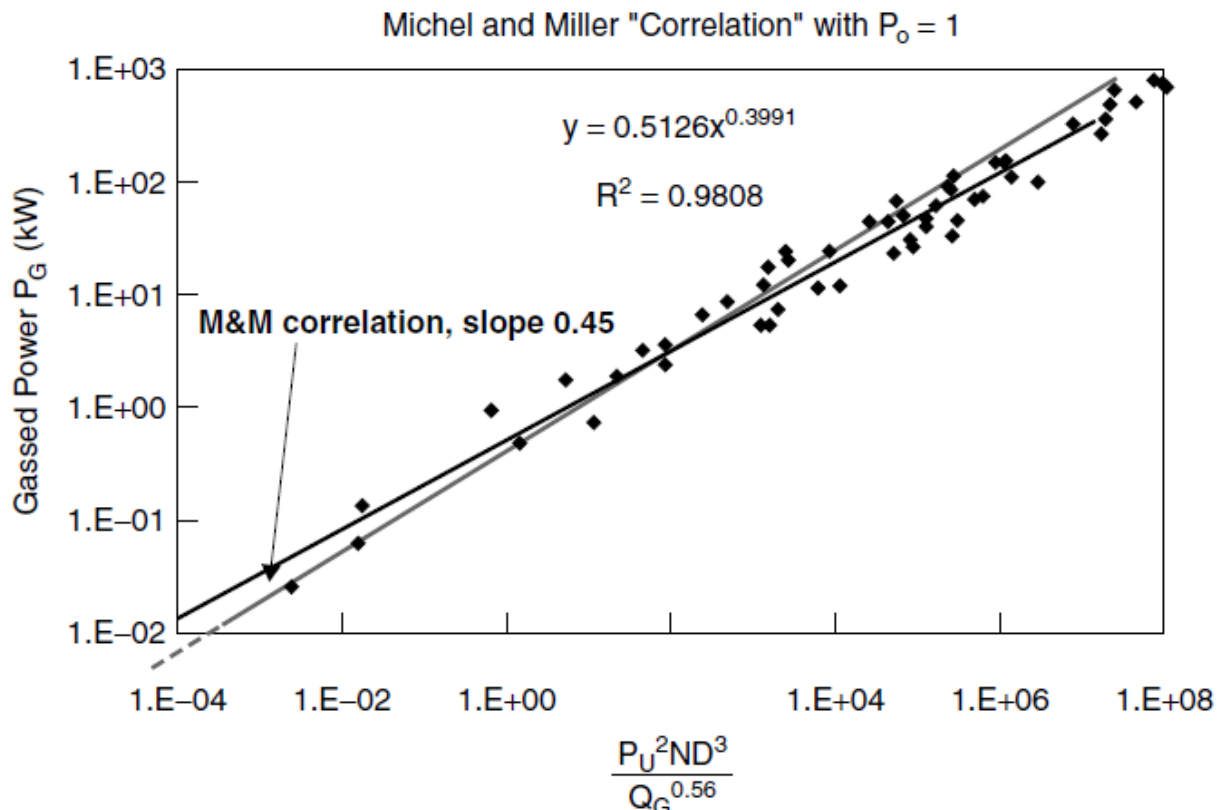
Fig. 19 Michel and Miller Equation 1962

Power Requirement

- Pharamond *et.al.* showed that relative power is almost independent of stirrer speed in low velocity region where Pg/Po > 0.5.
- *The relative power is about 10-15% higher for Dt = 1.0m than for Dt = 0.48 or 0.28m.*
- *Dickey covered higher range of gas velocities and showed that power decreases with higher speed and also as velocity increases.*

Prediction of gassed power

- **For Rushton turbines:**

Calderbank (1958) fitted two straight lines to curve:

$$\frac{P_G}{P_U} = 1 - 12.6 Ae \qquad \text{for}: Ae < 0.035$$

$$\frac{P_G}{P_U} = 0.62 - 1.85 Ae \qquad \text{for}: Ae > 0.035$$

Volesky (1979):

$$\frac{P_G}{P_U} = 0.10 \left(\frac{Q_G}{NV} \right)^{-1/4} \left(\frac{N^2 D^4}{gwV^{2/3}} \right)^{-1/5}$$

- **Bakker, Smith & Myers (*Chemical Engineering*, 199x):**

$$\frac{P_G}{P_U} = 1 - (b - a\mu)Fr^d \tanh(cAe)$$

	a	b	c	d
Rushton	0.72	0.72	24	0.25
Smith	0.12	0.44	12	0.37

Liquid Mixing Time
- Homogeneous aerated system

$$N \times t_{M(90\%)} = Cons\,tant \qquad \textbf{Cook et.al. 1988}$$

- Zhao et.al. 2001, Compared liquid mixing time in ungasses, cold and hot sparged and boiling Conditions.
    - They concluded that mixing time depends upon specific power input and superficial gas velocity, but they could not give specific correlation.
    - More over they also concluded that mixing time was least in boiling liquid system.
- Gezork et.al. 2001 showed that in heterogeneous mixing system because of larger bubble size mixing time was even less.
- Cook et.al. 1988 showed that for one to three radial flow impellers for Re > 4400

$$t_{M(90\%)}[P_o(RPD)]^{0.33} N\left(\frac{D_a}{D}\right)^{2.4}\left(\frac{D}{H}\right)^{2.4} = 3.3$$

Characteristics of dispersed phase
- Mean diameter of drop or bubble D_p
- Interfacial area a
- Dispersed phase hold up ϕ
- Total number of droplet in the volume N

Volume of spherical drop/bubbl es for unit volum e

$$\phi = \frac{\pi N D_p^3}{6}$$

Total surface area of the drops/bubbles in this volume

$$a = \pi N D_p^2$$

Ratio of these two equation will give

$$D_p = \frac{6\phi}{a}$$

The diameter of drop/bubble will differ from time to time, so this average diameter is called as Sauter Mean Diameter.

Flooding Velocity
- **For Rushton and Smith turbines (0.22 < D / T < 0.55):**

$$Ae_F = K_F Fr_F \left(\frac{D}{T}\right)^{3.5}$$

$$\frac{Q_G}{N_F D^3} = K_F \frac{N_F^2 D}{g} \left(\frac{D}{T}\right)^{3.5}$$

$$N_F = \left(\frac{Q_G g}{K_F D^4} \left(\frac{T}{D}\right)^{3.5} \right)^{1/3}$$

- **For a Rushton turbine:** $K_F = 30.$
- **For a Smith turbine:** $K_F = 70.$

Complete Dispersion
- **For Rushton and Smith turbines (0.22 < D / T < 0.55):**

$$Ae_{CD} = K_{CD} Fr_{CD}^{0.5} \left(\frac{D}{T}\right)^{1/2}$$

$$\frac{Q_G}{N_{CD} D^3} = K_{CD} \left(\frac{N_{CD}^2 D}{g} \frac{D}{T} \right)^{1/2}$$

$$N_{CD} = \left(\frac{Q_G (Tg)^{1/2}}{K_{CD} D^4} \right)^{1/2}$$

- **For a Rushton turbine:** $K_{CD} = 0.2.$
- **For a Smith turbine:** $K_{CD} = 0.4.$

Bubble diameter under Laminar Mixing Conditions
- By equating the net buoyant force on the bubble to the opposing drag force at the edge of orifice,

Where
D_o = orifice diameter
σ= interfacial tension
ρ_L = liquid density
ρ_v = Vapor density

$$F_b - F_d = \frac{g}{g_c} \frac{\pi D_p^3}{6}(\rho_L - \rho_v)$$

$$F_d = \pi D_o \sigma$$

$$\therefore D_p = \left[\frac{6 D_o \sigma g_c}{g(\rho_L - \rho_v)}\right]^{1/3}$$

Gas Dispersion in agitated vessel
- Gas is transferred to the vessel by
    - Open end of submerged pipe
    - Sparger
    - Porous ceramic plate
    - Porous metal plate

Generally motor driven turbine type impeller is used for gas dispersion.

Bubble Diameter
- For hold up $\phi < 0.15$, SMD is given by *Calderbank et.al. 1958*
- Average bubble diameter does not increase 2 to 5mm.
- The diameter will be even small in higher speed where shear effect is large but they will coalesce rapidly, and average size will be balanced.

$$D_p = 4.15 \frac{(\sigma g_c)^{0.6}}{(P g_c / V)^{0.4} \rho_L^{0.2}} \phi^{1/2} + 0.9$$

Interfacial Area

$$a = 1.44 \frac{(Pg_c/V)^{0.4} \rho_L^{0.2}}{(\sigma g_c)^{0.6}} \left(\frac{\bar{V}_s}{u_t}\right)^{\frac{1}{2}}$$

Where
V= volume of vessel
V_s = superficial velocity of gas
u_t = bubble rise velocity in stagnant liquid

Gas-Hold Up in Liquid
- Most published correlations for gas hold-up are derived from experiments with either pure liquids ('coalescing' systems) or aqueous solutions of electrolytes ('non-coalescing') and are of the form

$$\phi \propto \left(\frac{P_g}{V_L}\right)^A (\upsilon_s)^B$$

- Where A = 0.2 to 0.7 and B = 0.2 to 0.7
- *Smith et.al*. A = 0.48 and B = 0.4 for υs = 0.005 – 0.05m/s

$$\phi = \left(\frac{\bar{V}_s \phi}{u_t}\right)^{\frac{1}{2}} + 0.216 \frac{\left(Pg_c/V\right)^{0.4} \rho_L^{0.2}}{(\sigma g_c)^{0.6}} \left(\frac{\bar{V}_s}{u_t}\right)^{\frac{1}{2}}$$

Gas Handling Capacity
- If gas through put to a turbine agitated vessel is progressively increased, the impeller eventually floods and can no longer disperse the gas effectively.
- In agitated vessel transition is not as distinct as packed bed vessel.
- One possible definition is based on visual inspection when most of bubbles rise vertically between the turbine blades rather than being dispersed radially from tip of the blade.
- The critical gas velocity for this transition was given by **Gao et.al. 2001**
- $\bar{V}_{s,c} \propto P_g D_t / V$

$$\bar{V}_{s,c} = 0.114 \left(\frac{P_g}{V}\right)\left(\frac{D_t}{1.5}\right)^{0.17}$$

- Where
  - P_g/V is in W/m3
  - D_t is in m
  - $V_{s,c}$ is in mm/s

Liquid Mixing
- It has been shown by Middleton (1929) that for gas hold-up fractions up to *at least 0.1*, liquid circulation time (related to mixing time) is *changed only slightly* by the presence of the gas, so correlations for liquid only may be used.
- However, above about *20% hold-up*, circulation time is increased, i.e. *mixing is worse* than without gas.
- At much higher gas *hold-ups (> 0.7)* as encountered in some foamy boiling systems, the *liquid is mostly in the form of films between bubbles*, with very restricted mobility and the agitated tank cannot provide good liquid mixing in such cases.
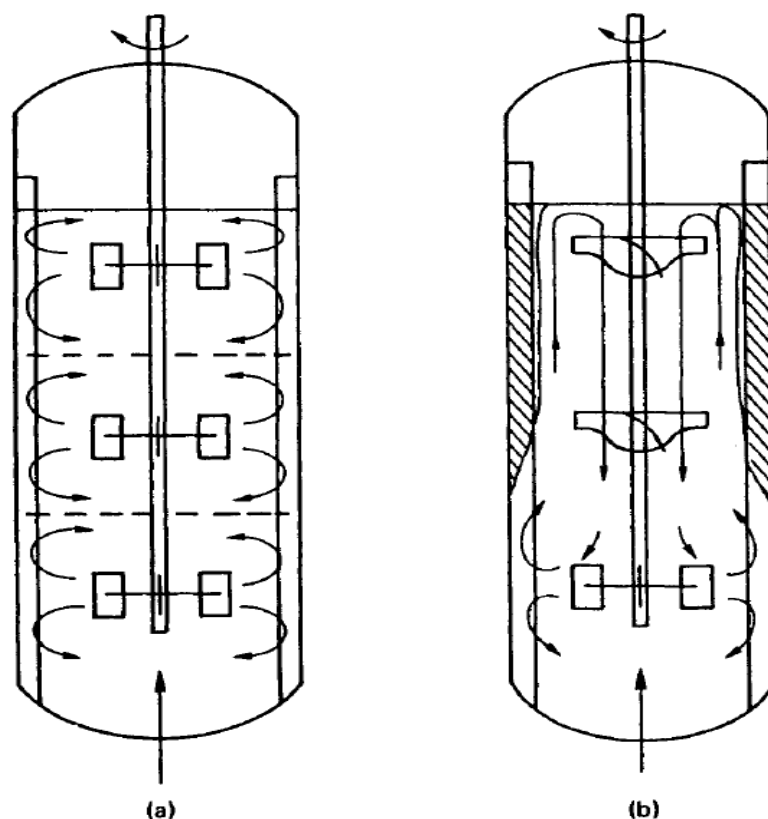
Fig. 20 Mean flow patterns with multiple agitators

With tall vessels and multiple impellers, care must be taken with selection of impeller type and power distribution. Multiple disc turbines give high velocities at the walls (for heat transfer etc.) but compartmentalized mixing patterns with poor transfer between zones, hence poor overall axial mixing *(Figure 20a).* A common alternative is to use a radial flow gas dispersing lower agitator as before but with down flow axial hydrofoil agitator(s) *(Figure 20b).* These improve axial mixing, but have lower velocities near the upper walls, especially at high gas rates and/or with shear-thinning liquids. Up flow impellers are becoming popular for their better gassed performance and good blending.

Gas Mixing

Gas is mainly mixed with other gas only at the agitator in the gas cavities behind the blades and minimal mixing occurs elsewhere in the vessel. If bubbles do coalesce after formation, they do so very rapidly and very near the agitator, where the gas has just been mixed anyway so bubble-bubble coalescence has a negligible influence on gas back mixing except that it controls average bubble size which influences gas recirculation. Thus, there is an apparent anomaly in that 'non-coalescing' systems have a greater amount of gas back mixing than 'coalescing'.

Gas Recirculation Rate
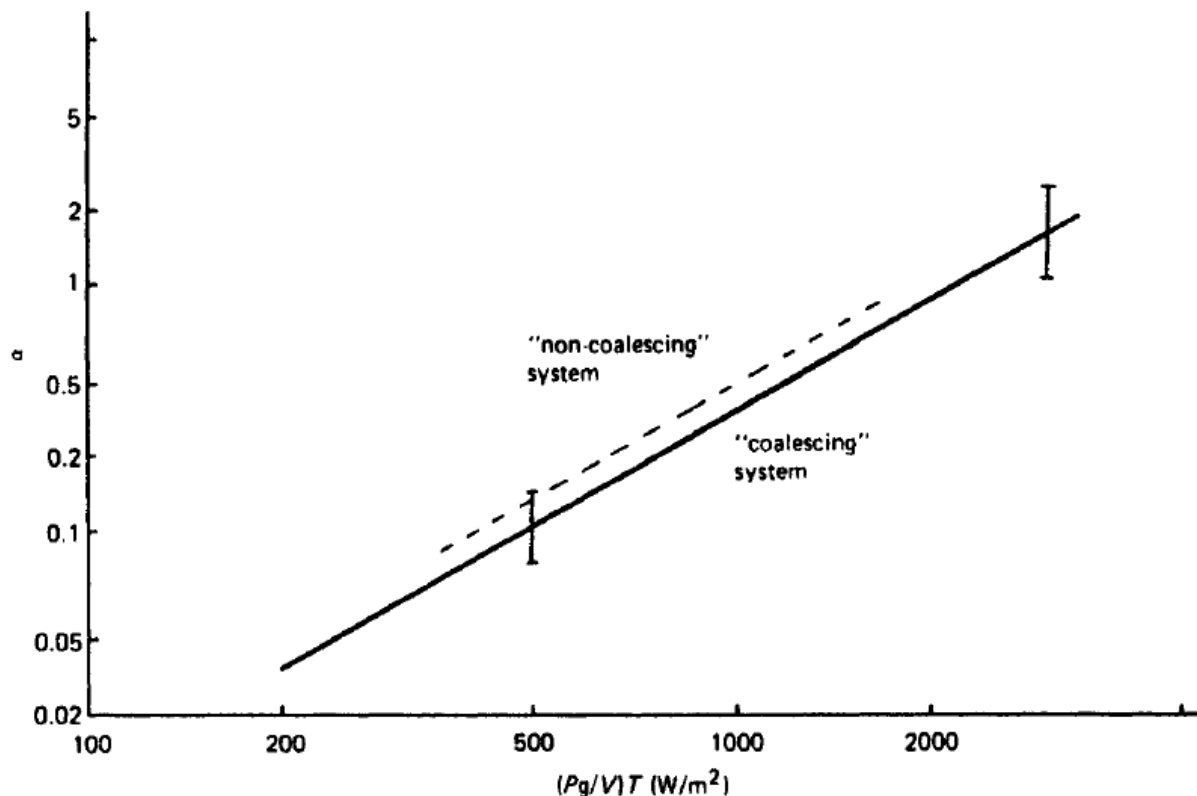
Fig. 21 *Gas recirculation ratio (α) correlation. Data covers: vessels 0.6–1.8 m diameter, H = T; C = T/3 and T/2; tip speed 1.6–4.7 m/s; $v_s = 5.40 \times 10^{-3}$ m/s*

Gas-Liquid Mass Transfer
- Controlling Factor : Film Diffusion on Liquid Side given by MTC k_L.
- From Mass Balance Equation:

$$\text{Overall Transfer Rate} = K_L a.V.\Delta C_{mean}$$

- Basic problem with above equation:
  - Doesn't required Bubble size in calculation

$$\frac{1}{K_L} = \frac{1}{Ek_G} + \frac{1}{k_L}$$

Gas Mass Transfer Coefficient
- For small bubbles (<0.2mm),

$$k_G = \frac{2\pi^2 D_{AG}}{2d_B} = 6.6\frac{D_{AG}}{d_B}$$

- For small contact time

$$k_G = 2\sqrt{D_{AG}/\pi t}$$

- With larger bubbles (de i> 2 mm), k_G is increased by internal convection up to 2.25 times the above values
  - kG controls the rate if $k_G E << k_L$.
  - Usually $k_L << k_G E$, so that the liquid side resistance predominates and therefore K = k_L.

Liquid Mass Transfer Coefficient

$$k_L a = \alpha'' \left(\frac{P}{\rho V} \right)^{\beta''} (v_s)^{\gamma''}$$

- P includes shaft power and buoyancy but not kinetic energy of gas.
- For air – water $\alpha = 1.2$, $\beta = 0.7$ & $\gamma = 0.6$, P = w, V = m3, vs = m/s
- For air-Electrolyte $\alpha = 2.3$ all other parameters will be same.
- Smith 1977 found that β & γ are not function of type of fluid, impeller or scale but α is strong function of type of fluid and impeller.
- After long runs Cook et.al. 1988 found that $\quad k_L a \propto \left(\frac{P}{\rho V} \right)^{0.5} (v_s)^{0.3} \mu^{-1}$

- Calderbank Equation $\qquad$ For $d_B \leq 2.5mm$

$$k_L = 0.31 \left[\frac{(\rho - \rho_g)\mu g}{\rho^2} \right]^{1/3} \left[\frac{\mu}{\rho D_{AL}} \right]^{-2/3}$$

For $d_B \geq 2.5mm$

$$k_L = 0.42 \left[\frac{(\rho - \rho_g)\mu g}{\rho^2} \right]^{1/2} \left[\frac{\mu}{\rho D_{AL}} \right]^{-1/2}$$

- K_L correlation for 6-Bladed Turbine
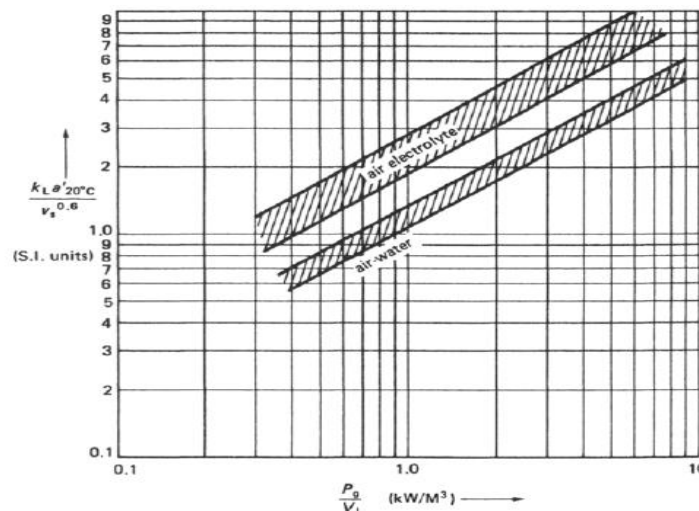
Fig. 22 Liquid Side Mass Transfer Coefficient

Heat Transfer
- In many instance when gas is dispersed into a liquid it is necessary to add or remove heat at some stage during the process.
- In mechanically agitated vessels this can be achieved using either jacketed vessels or helical or baffle coils immersed in the vessel.
- Edney and Edwards (1978) indicate that, for holdups < 15%, the rates of heat transfer with gas addition are very close to the values obtained without gas addition, i.e. with the single phase liquid only being agitated.

Gas-Liquid Mixers as Reactors
- This section shows how mixing is applied to the selection and design of gas-liquid reactors in which the absorbed gas reacts with a liquid component. It concentrates on mixing vessels as gas-liquid reactors, although the degree and mode of mixing can have an important effect

on the conversion and yield of a reaction and in some cases other devices, e.g. packed columns, bubble columns, liquid film contactors, with one or both phases in plug flow are more appropriate.

- The treatment concentrates around the relative rates of gas-liquid mass transfer and reaction. The reaction rate controls the overall rate if it is very slow; but often it is fast, so that the overall rate is mass transfer controlled. Very fast reactions can influence the diffusion process, causing enhancement of mass transfer above the purely physical rate. This enhancement is itself a function of the reaction rate. In extreme cases ('instantaneous' reaction), mass transfer again controls the overall rate of the process.
- Ideally the reactor design requires quantitative data for both the chemical kinetics and the mixing and mass transfer. The difficulties of predicting the physical effects have been outlined in the earlier sections. Often the chemistry is also difficult to analyse, and simplification must be made. In any case, small scale experiments must be carried out to determine relevant rates and other necessary reaction characteristics such as heat of reaction, degree of foaming and bubble coalescence etc. These experiments then allow scale-up to the final plant design. If time and techniques permit, this experimentation also guides the choice of reactor type and indicates whether and how complex mathematical modeling should be carried out.
- Generally the difficulties of chemical analysis together with the uncertainties in gas-liquid fluid dynamics render scale-up of gas-liquid reactors a hazardous procedure. Experiments therefore must be done under the same conditions of temperature, pressure, catalyst type etc. as it is intended to use on the final plant. Allowances for uncertainty must be made in scaling-up to the final plant because of the difficulty and expense of experiments. In the past, designs have been rather arbitrary and therefore often sub-optimal and the large stirred pot has been very popular, mainly because it resembles a chemist's flask whilst also providing reasonable heat transfer, particle suspension and flexibility. This section aims to assist in deciding whether these expensive vessels are really the most efficient and economic for a given process.
- It covers briefly the theory of interacting chemical rate processes with physical ones and the possible influence of mixing on reaction yields. Some suggestions are given on experimentation to elucidate the relative importance of the two rate processes and finally a summary of reactor modeling techniques suitable for gas-liquid systems.

Reaction theory

- Let $nA + mB \rightarrow$ products $q = m/n$

- Rate of reaction $\quad -r_A = k_{nm} C_{LA}^n C_{LB}^m$

- Reaction Time $t_R = \left(\dfrac{n+1}{2 k_{nm} C_{LA}^{n-1} C_{LB}^m} \right)$

- Diffusion Time $\quad t_D = D_{AL} \Big/ k_L^2$

Regimes of Gas-Liquid Mass Transfer with Reaction

When the reaction rate is comparable to that of the mass transfer through the diffusion film, interactions must be taken into account. The interactions can be delineated as five regimes, as shown in Figure 23. These are identified by the value of the *Hatta number*, Ha, which is defined as the square root of the ratio of the diffusion time, t_D, to the reaction time, t_R.

The regime dictates the choice of reactor. From Figure 23, the following choice of equipment for each regime can be inferred:
- *Regime I:* reaction in bulk, modest kLa: bubble column
- *Regimes II, IV, and V:* high a and kLa: stirred vessel
- *Regime III:* all reaction in film, high a: thin-film reactor (packed column or spinning disk)
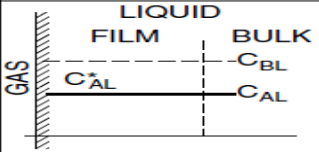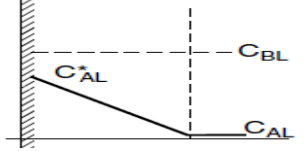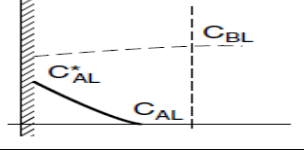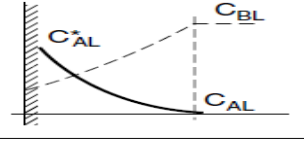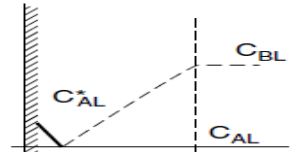
	REGIME	CONDITIONS	IMPORTANT VARIABLES	CONCENTRATION PROFILES
I	Kinetic control Slow reaction	$\sqrt{\dfrac{t_D}{t_R}} < 0.02$	Rate α ϵ_L α k_{nm} α $(C^*_{AL})^n$ α $(C_{BL})^m$ Independent of a (if a adequate) Independent of k_L	
II	Diffusion control Moderately fast reaction in bulk of liquid $C_{AL} \approx 0$	$0.02 < \sqrt{\dfrac{t_D}{t_R}} < 2$ Design so that $\dfrac{\epsilon_L}{a} > 100 \dfrac{D_{AL}}{k_L}$	Rate α a α k_L α C^*_{AL} Independent of k_{nm} Independent of ϵ_L (if ϵ_L adequate)	
III	Fast reaction Reaction in film $C_{AL} \approx 0$ (pseudo first order in A')	$2 < \sqrt{\dfrac{t_D}{t_R}} < \dfrac{C_{BL}}{qC^*_{AL}}$ $C_{BL} \gg C^*_{AL}$	Rate α a α $\sqrt{k_{nm}}$ α $(C^*_{AL})^{(n+1)/2}$ Independent of k_L Independent of ϵ_L	
IV	Very fast reaction General case of III	$2 < \sqrt{\dfrac{t_D}{t_R}}$ $C_{BL} \sim C^*_{AL}$	Rate α a depends on k_L, k_{nm}, C^*_{AL}, C_{BL} Independent of ϵ_L	
V	Instantaneous reaction Reaction 'at interface'. Controlled by transfer of B to interface from bulk. $J \alpha k_L a$	$\sqrt{\dfrac{t_D}{t_R}} \gg \dfrac{C_{BL}}{qC^*_{AL}}$	Rate α a α k_L Independent of C^*_{AL} Independent of k_{nm} Independent of ϵ_L	

Fig. 23 Regimes of Gas-Liquid Reactions

Chapter 6
Solid–Liquid Mixing

INTRODUCTION

In this chapter the focus is on mixing operations involving, primarily, solid and liquid phases carried out in agitated or stirred vessels. Fundamental aspects of the hydrodynamics and mass transfer as well as practical design issues for solid–liquid mixing of both settling and floating solids in ungassed or gassed suspensions are discussed. Settling solid particles have a higher density than the liquid and will settle without agitation. Solids that float without agitation include solids that are less dense than the liquid, dense solids with trapped gas, and solids
that are difficult to wet. Often, solid–liquid mixing operations are carried out in the presence of gas bubbles. These are known as *gassed suspensions*, in contrast to un-gassed suspensions in the absence of gas bubbles. The gas bubbles may be introduced, directly as in solid-catalyzed hydrogenation reactions, entrained inadvertently or deliberately from the headspace, or evolved as in an evaporative crystallization or as a gaseous reaction product.

Solid suspensions are typically carried out in mechanically agitated or stirred vessels. Pumped liquid jets have also been used to suspend low concentrations of relatively slow settling solids. Although static mixers have been used to disperse fine solids into polymers, application of the technology is limited and beyond the scope of the present discussion.

Not included in this chapter are several solid–liquid contacting operations, such as:
1. Dispersion of very fine particles in liquids where interfacial phenomena dominate both the dispersion process and the rheology of the suspension. An application of this technology is in the preparation of a stable solid suspension such as an agricultural "flowable" formulation by the addition of suspending aids, stabilizers, and so on.
 2. Liquid or gas fluidized beds.
3. Liquid–solid contacting in fixed bed systems.

Scope of Solid–Liquid Mixing

The primary objectives of solid–liquid mixing are to create and maintain a slurry and/or to promote and enhance the rate of mass transfer between the solid and liquid phases. The mixing operation promotes the

• Suspension of solids

• Resuspension of settled solids

• Incorporation of floating solids

• Dispersion of solid aggregates or control of particle size from the action of fluid shear as well as

any abrasion due to particle–particle and impeller–particle impacts • Mass transfer across the solid–liquid interface

Unit Operations Involving Solid–Liquid Mixing

Solid–liquid mixing is a key aspect of common unit operations in the chemical industry, including:
1. Dispersion of solids
2. Dissolution and leaching
3. Crystallization and precipitation
4. Adsorption, desorption, and ion exchange
5. Solid-catalyzed reaction
6. Suspension polymerization

These unit operations, with the exception of dispersion, involve mass transfer between the solid and liquid phases.

Dispersion of solids is a physical process where solid particles or aggregates are suspended and dispersed by the action of an agitator in a fluid to achieve a uniform suspension or slurry. Applications include the preparation of a slurry of solid reactants or catalyst to feed a reactor as well as dispersion of solid pigments and other materials into a liquid.

Dissolution is a mass transfer unit operation during which the solid particle decreases in size and ultimately disappears as it is incorporated as solute in the liquid. In *leaching*, a soluble component of the solid dissolves, usually leaving a particle of different size, density, and/or porosity. For some rubber or plastic materials, the particles may actually swell initially. The density and viscosity of the resulting liquid may differ considerably from the original liquid for some systems. The process goal here is to achieve the desired rate of dissolution or leaching by agitation.

Crystallization and *precipitation* start with a solid-free liquid phase if unseeded. The solid particles form during the crystallization or precipitation operation. The solids grow in size as well as in population. The viscosity and density of the slurry thus formed usually increase. The process goals include control of the rate of nucleation and growth of the particles as well as the minimization of particle breakage or attrition. Both the average size and the particle size distribution are important properties. Liquid-phase mixing to achieve uniformity of super saturation or to avoid local high concentration regions is important in achieving particle size control.

In *adsorption*, *desorption*, and *ion exchange*, there is mass transfer between the solid and the solution. Mass transfer is from the liquid phase into the solid in adsorption and from the solid into the liquid phase for desorption. In ion-exchange operations there is an exchange of ions between the solid and the liquid.

Solid-catalyzed reactions usually involve adsorption of reactants onto the surfaces of the catalyst particles where the reactions take place, followed by the desorption of the reaction products from the surface. A uniform suspension of catalyst particles ensures a uniform concentration of reactants and reaction products throughout the vessel. In addition, agitation reduces the diffusional mass transfer boundary layer, thus enhancing the solid–liquid mass transfer.

Suspension polymerization starts with the creation of a stabilized dispersion of monomer droplets. As polymerization proceeds, the monomer droplets polymerize, usually passing through a sticky phase. The protective coating of suspending agents (surfactants, etc.) and agitation conditions keep the droplets from coalescing. They also control particle size and size distribution. The mixing objective here is to produce and maintain, by agitation, a dispersion of uniform size drops and suspension of both monomer drops and eventually, polymer particles.

Process Considerations for Solid–Liquid Mixing Operations
The desired process results for solid–liquid mixing vary from process to process as indicated above in the brief discussion of several unit operations. It is the responsibility of the process researcher and/or process engineer to determine the pertinent and specific process needs. Sometimes, results

associated with other mixing operations—blending, gas–liquid, liquid–liquid, heat transfer, and so on—may be more important. Therefore, it is essential to consider and understand, early in the process development stage, all the physical and chemical phenomena necessary to achieve the desired process results. In particular, how these phenomena are influenced by the process chemistry, the properties of the solid and liquid phases, and the operational variables of mixing must be understood. The key considerations include the:

1. *Mode of process operations:* batch, semibatch [continuous addition to batch (con-add)], or continuous
2. Phases—solid, liquid, and/or gas phases—that are present or occur from the beginning to the end of the process
3. Properties of the solid and liquid phases, including stickiness and tendency to agglomerate
4. Unit operations involved from the beginning to the end of the process
5. Vessel geometry and internals
6. *Mixing parameters:* local or average fluid velocity or flow, local or average shear rates, blend time, power input, and so on.

Key Process Questions for Solid–Liquid Mixing.

For each mixing operation, several key process-related issues must be addressed before scale up and design. For solid–liquid mixing operations, key process questions include the following.

• *What is the process mode of operation: batch, semi batch, or continuous?*

Whether a process is best run as a batch, semi batch, or continuous operation depends on the unit operation, upstream and/or downstream operations, and the volume of materials processed. For example, in a single stirred tank, a solid–liquid mixing operation requiring complete solid dissolution or complete reaction of the solid must, of necessity, be batch or semi batch. The solid–liquid mixing operations where a slurry is the end product can be batch, semi batch or continuous. For batch operations, the mixing requirements often change during the batch as a result of changes in physical and chemical properties and/or changes in the mixing volume for semi batch operations. It is therefore important to determine all the physical and chemical phenomena taking place during the entire duration of the batch. For continuous operations, the physical and chemical phenomena occurring during startup and shutdown must also be determined.

• *What phases are present or occur during the process?*

 The type of mixing operation to study, and the degree of difficulty in achieving the desired
process result, depend on the phases present. The presence of solid and liquid phases only suggests that the mixing problem of interest is one of solid–liquid mixing operation. For example, the mixing problem is blending rather than solid–liquid mixing if the settling velocity is less than about 0.5 ft/min or 0.0025 m/s. This condition occurs if the viscosity of the suspending liquid is very high, the solid particles are so small, and/or the density difference between the solid particle and the liquid is small. The presence of gas bubbles and/or immiscible liquids can significantly influence the ability to suspend the solids.

• *Is there a chemical reaction of the solid with the liquid?*

 Solid–liquid mixing operations involving chemical reactions often require a high relative
velocity between the solid particle and the liquid—high local shear rate or agitation intensity—to minimize the thickness of the boundary layer for mass transfer. This is also true for the dissolution of a sparingly soluble solid.

• *What are the physical properties of the solid and liquid phases present?*

The degree of difficulty in solid suspension depends on several properties of the fluid and solid particles discussed in Section 10-2. The properties of interest include the relative density of the

solid and liquid phases, the viscosity of the liquid, the wetting characteristics of the solid, the shape of the solid particles, and the mass or volume ratio of solids to liquid.

Large and dense solids are more difficult to suspend than small light ones; spherical particles are also more difficult to suspend than thin flat disks. The impact of these properties on solid–liquid mixing must be studied and understood early in process research and development.

• *What degree or level of suspension is required?*

The required degree or level of suspension depends on the desired process result and the unit operations involved. For example, a higher degree of suspension is required in a crystallizer or slurry feed vessel than in a vessel for the dissolution of a highly soluble solid.

• *What is the minimum agitator speed to suspend the solids?*

 In stirred tanks, there is always an impeller speed below which settling solids will tend to accumulate on the bottom of the vessel. This speed is different for different types of impellers and for identical impellers located at different clearances from the bottom of the vessel. It also depends on the properties of the solid and liquid phases. The minimum speed may be estimated for certain impeller and tank geometries using the Zwietering correlation. It is advisable, however, to determine this value experimentally for processes where
solid–liquid mixing is deemed critical.

• *What happens to the suspension when agitation is decreased or interrupted?*

Obviously, solids will settle or float depending on the properties of the solid relative to the liquid phase. The more important issues are whether the solids agglomerate and/or cake as they settle or how easy it is to resuspend them when agitation is increased or restored. This information is crucial for the proper mechanical design as well as instrumentation and control of the agitation.

• *What happens to the suspension when agitation is increased?*

Most solid–liquid mixing operations operate above the minimum speed for suspension. A higher agitation speed improves the degree of suspension and enhances mass transfer rates. The higher speed also translates into higher turbulence as well as local and average shear rates, which for some processes may cause undesirable particle attrition. Obviously, there is also a practical economic limit on the maximum speed of agitation.

• *What effect does vessel geometry have on the process?*

The geometry of the vessel, in particular the shape of the vessel base, affects the location of
dead zones or regions where solids tend to congregate. It also influences the minimum agitation speed required to suspend all particles from the bottom of the vessel. In flat-bottomed vessels, dead zones and thus "fillet formation" tend to occur in the corner between the tank base and the tank wall, whereas in dished heads the solids tend to settle beneath the impeller or midway between the center and the periphery of the base. The minimum agitation speed is typically 10 to 20% higher in a flat-bottomed vessel than in one with a dished head. Both the minimum agitation speed and the extent of fillet formation are also a function of impeller type, ratio of impeller diameter to tank diameter, and location of the impeller from the vessel bottom. In general, a dished-head vessel is preferred to a flat-bottomed vessel for solid–liquid mixing operations. There is little or no difference between ASME dished, elliptical, or even hemispherical dished heads as far as solid–liquid mixing is concerned. However, elliptical heads are preferred for higher-pressure applications.

• *What is the appropriate material of construction for the process vessel?*

The main issue here is that, for steel or alloy vessels, the standard four wall mounted baffles provide a better environment for solid–liquid mixing. The standard glass-lined vessels are usually under baffled because of a deficiency of nozzles from which to mount baffles.

HYDRODYNAMICS OF SOLID SUSPENSION AND DISTRIBUTION

Solid suspension requires the input of mechanical energy into the fluid–solid system by some mode of agitation. The input energy creates a turbulent flow field in which solid particles are lifted from the vessel base and subsequently dispersed and distributed throughout the liquid.   Recent measurements of the 3D velocity of both the fluid and the suspension confirm the complexity. Solids pickup from the vessel base is achieved by a combination of the drag and lift forces of the moving fluid on the solid particles and the bursts of turbulent eddies originating from the bulk flow in the vessel. This is clearly evident in visual observations of agitated solid suspensions as in the video clip included on the accompanying CD ROM. Solids settled at the vessel base mostly swirl and roll around there, but occasionally, particles are suddenly and intermittently lifted up as a tornado might lift an object from the ground. The distribution and magnitude of the mean fluid velocities and large anisotropic turbulent eddies generated by a given agitator determine to what degree solid suspension may be achieved. Thus, different agitator designs achieve different degrees of suspensions at similar energy input. Also for any given impeller the degree of suspension will vary with D/T as well as C/T at constant power input. One of the video clips on the accompanying CD ROM shows the effect of D/T on solid suspension for a pitched blade impeller at constant power input.

For small solid particles whose density is approximately equal to that of the liquid, once suspended they continue to move with the liquid. The suspension behaves like a single-phase liquid at low solid concentrations; the mixing operation is more like blending than solid suspension. For heavier solid particles, their velocities will be different from that of the liquid. The drag force on the particles caused by the liquid motion must be sufficient and directed upward to counteract the tendency of the particles to settle by the action of gravity.

The properties of both the liquid and the solid particles influence the fluid–particle hydrodynamics and thus the suspension. Also important are vessel geometry and agitation parameters. The important fluid and solid properties and operational parameters include:

1. Physical properties of the liquid, such as:
a. Liquid density, ρ_l (lb/ft3 or kg/m3)

b. Density difference, $\rho_s - \rho_l$ (lb/ft3 or kg/m3)

c. Liquid viscosity, μ_l (cP or Pa · s)

2. Physical properties of the solid, such as:
a. Solid density, ρ_s (lb/ft3 or kg/m3)
b. Particle size, d_p (ft or m)
c. Particle shape or sphericity, ψ (dimensionless factor defined by the ratio of surface area of a spherical particle of the same volume to that of a nonspherical particle)
d. Wetting characteristics of the solid
e. Tendency to entrap air or headspace gas
f. Agglomerating tendencies of the solid
g. Hardness and friability characteristics of the solid

3. Process operating conditions, such as:
a. Liquid depth in vessel, Z (ft or m)

b. Solids concentration, X (lb solid/lb liquid or kg solid/kg liquid)

c. Volume fraction of solid, φ

d. Presence or absence of gas bubbles

4. Geometric parameters, such as:

a. Vessel diameter, T (ft or m)

b. Bottom head geometry: flat, dished, or cone-shaped

c. Impeller type and geometry

d. Impeller diameter, D (ft or m)

e. Impeller clearance from the bottom of the vessel, C (ft or m)

f. Liquid coverage above the impeller, CV (ft or m)

g. Baffle type and geometry and number of baffles

5 Agitation conditions, such as:

a. Impeller speed, N (rps)

b. Impeller power, P (hp or W)

c. Impeller tip speed (ft/s or m/s)

d. Level of suspension achieved

e. Liquid flow pattern

f. Distribution of turbulence intensity in the vessel

States of Solid Suspension and Distribution

In agitated vessels, the degree of solids suspension is generally classified into three levels: on-bottom motion, complete off-bottom suspension, and uniform suspension. These are illustrated in Figure

On-Bottom Motion or Partial Suspension.

This state is characterized by the visual observation of the complete motion of all particles around the bottom of the vessel. It excludes the formation of *fillets*, a loose aggregation of particles in corners or other parts of the tank bottom. Since particles are in constant contact with the base of the vessel, not all the surface area of particles is available for chemical reaction or mass or heat transfer. On-bottom motion conditions are sufficient for the dissolution of highly soluble solids.

Off-Bottom or Complete Suspension.

The state of suspension known as off-bottom or complete suspension is characterized by the complete motion of all particles, with no particle remaining on the base of the vessel for more than 1 to 2 s. This condition is known as the *Zwietering criterion*. Under this condition, the maximum surface area of the particles is exposed to the fluid for chemical reaction or mass or heat transfer. The "just suspended" condition refers to the minimum agitation conditions at which all particles attain complete suspension. In mechanically agitated vessels, the minimum agitation speed for the just suspended state, Njs, has been the subject of many experimental and theoretical analyses.

Uniform Suspension.

Uniform suspension corresponds to the state of suspension at which particle concentration and particle size distribution are practically uniform throughout the vessel; any further increase in agitation speed or power does not appreciably enhance the solids distribution in the fluid.
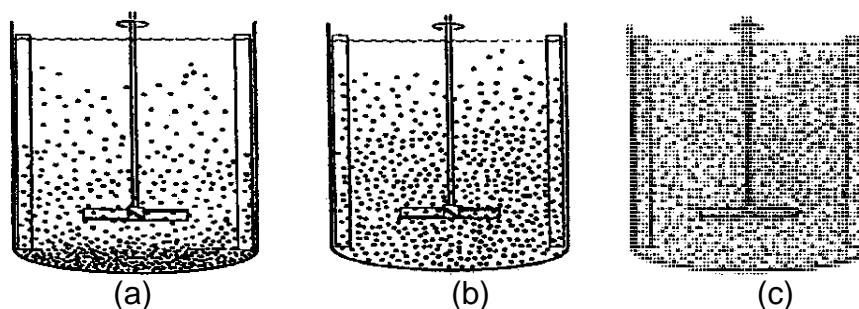
Figure Degrees of suspension. (*a*) Partial suspension: some solids rest on the bottom of the tank for short periods; useful condition only for dissolution of very soluble solids. (*b*) Complete suspension: all solids are off the bottom of the vessel; minimum desired condition for most solid–liquid systems. (*c*) Uniform suspension: solids suspended uniformly throughout the vessel; required condition for crystallization, solid catalyzed reaction.

MASS TRANSFER IN AGITATED SOLID–LIQUID SYSTEMS

As noted earlier, with the exception of the purely physical process of producing a slurry, unit operations involving solid–liquid mixing are mass transfer processes. These include:

• Leaching

• Dissolution of solids with or without chemical reaction

• Precipitation

• Crystallization–nucleation and crystal growth

• Adsorption

• Desorption

• Ion exchange

• Solid-catalyzed reactions

• Suspension polymerization
.

SELECTION, SCALE-UP, AND DESIGN ISSUES FOR SOLID–LIQUID MIXING EQUIPMENT

The selection, scale-up, and design of the components that make up the mixing system are based on the fundamental and experimental descriptions of the hydrodynamics and mass transfer aspects of solids suspension discussed earlier. The following issues must be addressed:

1. Process needs assessment, including:
a. Phases—solid, liquid, and gas—present or occurring during the process
b. Mixing operations and the desired process results
c. Unit operations of interest
d. Quantities and properties of solid and liquid phases

2. Vessel design and internals, including:
a. Bottom head design
b. Size and dimensions
c. Baffles and other internals

3. Selection and design of the agitator or mixer components, including:
a. Impeller type, number, and dimensions
b. Impeller location in the vessel
c. Impeller speed and power
d. Shaft diameter and length
e. Drive and seal system

Process Definition

The first task in analyzing a mixing problem, determining experiments to perform for mixer scale-up, or designing a mixing system is to define the process needs. It is important to consider carefully the potential impact of mixing on all the physical and chemical phenomena necessary to achieve the desired process result. Invariably, one of these phenomena will be the critical operation on which to base the selection, scale-up, or design of the mixing system.
The definition should include:

• A list of all the phases of matter (gas, liquid, solid) involved or that can occur, even by accident, from start to end of the process; in particular, instances where two or more phases coexist must be noted.

• A list of all the mixing operations (blending, solids suspension, gas dispersion, immiscible liquids dispersion, etc.) involved in the process or carried out in the same vessel.

• A statement of the purpose and duty of the mixing operations, including the desired process result.
For solids suspension, one must choose from among the applicable process objectives as well as the desired degree of suspension. The selection must be based on knowledge of the process determined experimentally or by comparison with a similar process.

• The quantities of solid and liquid phases involved as well as the properties of the solid and liquid to assess how difficult it might be to achieve the aforementioned desired results.

Process Scale-up.

Scale-up is an effort to understand the fundamental phenomena occurring in a process in order to predict the performance in larger scale equipment. It begins with process research at the bench scale, often in small glassware, through pilot scale studies to full production. The value of scale-up is captured in the following comment attributed to L. H. Baekland, the father of plastics: "Commit your blunders on a small scale and make your profits on a large scale." In solid–liquid mixing applications, the purpose of scale-up is to determine the operating conditions at different scales at which mixing yields equivalent process results. The tasks involve:
1. Definition of the appropriate desired process result, such as level of uniformity of the solid distribution in a vessel, the time to achieve complete dissolution, the rate of reaction between a solid and a liquid reactant, and
2. Developing reliable correlations that describe the effects of key process properties, mixer design, and operating variables on the desired process result by either experimentation or mathematical analysis of the physicochemical phenomena
3. Determining and confirming the key controlling physicochemical phenomena and the associated correlating parameters, preferably in dimensionless form
4. Applying the key correlations to predict the process performance at different scales Occasionally, heuristics based on extensive experience with similar processes are sufficient. Often, especially for processes involving multiple phases or fast reactions, it is necessary to perform several experiments

at two or more different scales, where the vessel size based on diameter is varied by at least a factor of 2.

Laboratory or Pilot Plant Experiments

Simple laboratory or pilot plant experiments carried out in transparent vessels, such as glassware, where one can observe the behavior of the various phases during agitation often provides great insight and understanding of the mixing challenges and opportunities. Often, these are augmented with pilot scale tests to determine or evaluate pertinent scale-up requirements. The lab experiments should be designed to answer specific process-related questions such as those discussed. Ultimately, the tests should provide information including:
1. The desired level of suspension required by the process
2. The properties of the solids and liquids required to estimate the necessary solid–liquid mixing parameters, including:
a. Settling velocity, V_t
b. Minimum speed for suspension, N_{js}
c. Solid–liquid mass transfer coefficient, k_{SL}
d. Materials of construction
In the various correlations presented earlier, the magnitude and sign of the exponents on the variables establish their parametric effects and may be used as a guide for selection of the more sensitive parameters to explore in a laboratory or pilot plant.

Typical lab experiments must include evaluation of the following effects:

1. Impeller speed to establish the effect, if any, on the process result as well as the speed beyond which there is no further significant gain in or deterioration of the desired process results
2. Particle size to determine the effect on reaction rates for solid-catalyzed reactions: in particular, the particle size at which mass transfer effects are negligible
3. Addition rate of solids and/or liquid, as well as the ratio of solids to liquid to determine their effects on rheology, suspension level, reaction, or other mass transfer rate
4. Impeller design and geometry to explore the relative effects of flow and shear distribution in the vessel for particle size control, micromixing for fast kinetics, and so on. Geometric ratios of importance include:
  a. Ratio of the impeller to tank diameter, D/T, to determine the effect of the ratio of overall pumping capacity to fluid shear
  b. Blade width to impeller diameter, W/D, to evaluate the relative effects of microscale and macroscale mixing processes and also fluid shear rates
5. Number and location of the impeller to explore the effect of liquid coverage on headspace gas entrainment, uniformity of solids distribution, and so on. Parameters of interest include:
  a. Ratio of the impeller clearance from vessel bottom to tank diameter, C/T
  b. Ratio of liquid coverage above impeller to tank diameter, CV/T
6. Baffle design and location to explore effects of vortex formation for entrainment of floating solids, and so on.

Tips for Laboratory or Pilot Plant Experimentation
In any laboratory or pilot plant tests, the first thing to vary is the impeller speed. This changes pumping capacity, blend time, and shear rates.

• On-bottom motion or partial suspension is rarely a useful desired mixing result except, perhaps, for the dissolution of very soluble solids.

• Complete suspension is the minimum desired mixing goal for most solid–liquid mixing operations involving settling solids. The equivalent condition for floating solids is complete incorporation and dispersion of the floating solids.

• Uniform suspension is required for crystallization, solid-catalyzed reactions, and suspension polymerization where high local concentrations may lead to poor yields of the desired product. Also, as practical as possible, crystallization slurries must be fed to a centrifuge at a uniform solids concentration for the proper cake buildup required for effective filtration and washing of the solid cake.

• Specified mass transfer rate such as dissolution rate, reaction rate, and so on, may be the desired process result to achieve a given production capacity.

• Particle size control may be the desired result in certain formulation operations.

• The measurement of power on a full or pilot scale vessel is best accomplished with a wattmeter. Ammeter readings, at best, must be ratioed to the full-load nameplate amperage, which varies with voltage, power factor, and motor type.

• For the fractional-horsepower motor used in the laboratory or pilot plant, power draw is best determined by calculation using the defining equation for the power number. This requires power number versus Reynolds number data or correlation.

• To estimate the viscosity of complex non-Newtonian slurries, it is recommend the use of a mixing viscometer that mimics the hydrodynamic environment likely to be encountered in an agitated vessel.

Recommendations for Solid–Liquid Mixing Equipment

Solids suspension is usually carried out in mechanically agitated vessels with or without draft tubes. In the following sections we provide several design guidelines and examples of the selection, design, and operation of equipment for solid–liquid mixing.

Vessel Geometry and Vessel Nozzles.

The vessel design, in particular, the bottom head design, can have a profound effect on the agitation requirements for a given desired result. The bottom head geometry influences
the flow patterns responsible for lifting solids up from the vessel bottom.

Design Tip. Dished heads (ASME dished, elliptical, or torispherical heads) are the preferred design. To achieve complete suspensions, flat-bottomed heads require 10 to 20% higher impeller speeds than for dished heads. Conical bottoms must be avoided.

The aspect ratio of the vessel—actually, the ratio of liquid depth, H, to vessel diameter, T is an important determinant of the number of impellers to be used. The fluid velocities decrease with increasing distance from the impeller region and may not be sufficient to counteract the
tendency of the solid to settle. Also, impellers mounted far above the vessel base may not generate enough turbulent velocity at the base of the vessel to lift any settled solids.

Design Tips

• A single impeller is usually sufficient for off-bottom suspension in vessels with dished heads, H/T < 1.3.

• Dual impellers are recommended for vessels with 1.3 < H/T < 2.5, used for uniform suspension of fast-settling solids.

• Three impellers may be required if 2.5 < H/T. A vessel with such a high aspect ratio is a poor choice for solid suspension.

• Vessel nozzles should be located and oriented to avoid or minimize any interference with the mixing system's performance.

Nozzle Design Tips

• Nozzles and dip pipes for liquid addition should not allow the liquid jet to impinge directly on the impeller. At too high a liquid jet velocity, the jet force will contribute to higher shaft deflections.

• Dip pipes and other probes must be supported—usually by attaching to wall-mounted baffles—or stiff enough to withstand the bending moments imposed by the fluid forces.

• Install grating or screen on nozzles for solid addition to keep very large solid chunks or foreign matter from the liquid.

• Bottom nozzles should be as short as practical and be installed with flush bottom valves to prevent solids from collecting.

Baffles

Baffles are highly recommended for solids suspension operations involving solids that are heavier than the liquid. They convert the swirling motion into top-down or axial fluid motion that helps to lift and suspend the solids. For floating solids, consider the use of submerged or partial baffles to achieve a controlled vortex to draw down the floating solids.

Baffle Design and Installation Tips

• In steel or alloy vessels, the recommended baffle design for solid suspension of settling solids is four flat blade baffles, each with width, B, equal to T/12 at a wall clearance of at least T/72. The baffles should extend to the lower edge of the lower impeller or to the lower tangent line.

• In glass-lined equipment, the recommended baffles are either fin or beavertail type A minimum of two baffles is recommended.
These baffles are generally less effective than the standard four flat blade baffles.

• Fin baffles must be installed with the edge of the fin pointing toward the vessel wall; the flat face must be perpendicular to the tangential flow.

Selection and Design of Impeller

Solids suspension and solids distribution is governed primarily by the bulk or convective flows in a vessel. *High efficiency impellers* whose discharge is flow dominated as well as axially directed, are more efficient than others in achieving solids suspension. However, high efficiency impellers may be a poor choice when the solid suspension is accompanied by other mixing duties, such as liquid–liquid dispersion or gas dispersion. For these cases a multiple-impeller system consisting of a high efficiency impeller in combination with a $45°$ pitched blade impeller should be evaluated in pilot plant studies. Small pitched blade impellers with diameter $D < T/2.5$, located nearer the vessel base $(C < T/4)$, are good for solid suspension). They also aid in the discharge of the solids during slurry transfer. Typical values for impeller clearance are T/4 for hydrofoils and T/3 for pitched blade turbines. For glass-lined vessels, one is no longer limited to the Pfaudler "crowfoot," also known as the *retreat blade* or *retreat curve impeller* (RCI). Most impeller designs can now be obtained with a glass lining. Removable glassed impeller designs are preferred over the integral glassed shaft-impeller design

Impeller Speed and Power

The impeller speed recommended will in general be higher than Njs, the speed required for the just suspended state estimated by the Zwietering correlation. The speed required should be based on experimental data.

Design Tip. For multiprocess batch reactors, mixers equipped with variable speed drives permit the mixer to be operated at different impeller speeds to accommodate the different mixing needs of the various steps in the process.

Shaft, Hub, and Drive

In the design of the shaft and drive system, careful consideration should be given to issues, including the need for:

• Startup of the mixer in settled solids.

• Filling and emptying while the mixer is running—the fluid forces on the impeller and shaft are amplified significantly when the liquid surface runs through the impeller, causing severe shaft deflections and vibrations.

• Ensuring that the suspension is maintained during emptying of the vessel to very low levels—for top-mounted agitators, a longer shaft fitted with a smaller-diameter impeller; a tickler, located at the lowest possible clearance from the base of the vessel, is required.

• Employing the same mixer for multiple mixing operations in the same process or for different processes.

Design Tip. The need for startup of a mixer in settled solids will require a larger shaft. This should be stated clearly in any mixer specification or request for quotation. The American Gear Manufacturing Association (AGMA) service rating for the gearbox will be higher. The shaft and gearbox design should be based on a minimum service rating factor of 2. An experienced mechanical engineer should be consulted for help in specifying the mixer or in reviewing any vendor proposals or quotations. Mixing equipment suppliers have calculation tools to size the shaft to minimize shaft deflections.

Design Tip. Sizing mixers to handle startup in settled solids requires measuring torque under test conditions with actual settled solids. In the absence of such a measurement, any design for such conditions can only be a "wild" guess. Use other means, such as air sparging, lancing with high-pressure liquid, heating to melt or dissolve the solid, and so on, to loosen the settled solid first. Before attempting to start the agitator drive, check and confirm by hand-turning the shaft that the impeller is indeed free.

Chapter 7.
Solids-Solid Mixing

INTRODUCTION

Solids mixing is essential to many industries, including ceramics, metallurgy, chemicals, food, cosmetics, coal, plastics, and pharmaceuticals. To give an idea of the magnitude of applications involving granular processes, worldwide production annually accounts for over a trillion kilograms of granular and powdered products, much of which must be uniformly blended to meet quality and performance goals. In this chapter we present an example-oriented overview of current understanding of mixing and de-mixing mechanisms of importance to powder blending operations. We focus on blending in tumblers, which simultaneously comprises the bulk of solids blending operations and represents the greatest opportunity for future predictive modeling.

Numerous distinct mechanisms for both mixing and de-mixing of granular materials have been cataloged, including convection, diffusion, shear, and percolation, and in most applications several mechanisms act concurrently and interact in complex ways. For example, details of loading of powders into blenders of common design can alter the time needed to homogenize them by two orders of magnitude, and by the same token, given that a certain blender can be designed to deliver acceptable performance in the laboratory, we have no consistent a priori mechanism to scale the process up and achieve the same performance in blenders of industrial size. The opposite problem, lack of dynamical similarity during process scale-down, is also quite common, haunting practitioners who attempt to undertake benchtop product design or wish to reproduce manufacturing problems in the lab. Nevertheless, although comprehensive predictive understanding of practical blending problems remains a distant goal, it has recently become possible to define models that generate respectable agreement with observations in practical granular devices (e.g., 3D tumblers). Progress has been made to develop systematic techniques to analyze new products and equipment. Some of these advances are reviewed in this chapter, following a description of the current level of understanding of blending and segregation mechanisms in commonly used industrial devices.

CHARACTERIZATION OF POWDER MIXTURES

A prerequisite to meaningful evaluation and interpretation of mixing is the development of a reliable measure of mixing. Straightforward though this concept may seem, some care needs to be exercised in its implementation. Any mixing measure is obtained by first evaluating a relevant quantity, typically concentration, in specified sample regions. Ideally, for the samples to be representative, they should be taken uniformly from a flowing stream that is itself uniform in both space and time. In tumbling blenders, this is not practical, and sampling usually consists of extracting small samples from a static bed. We discuss techniques for extracting such samples shortly, but first it is worthwhile to review the description of ideal mixtures, for which particle distributions are known throughout the mixture.

Ideal Mixtures versus Real Mixtures

Mixing is so common an everyday experience to both specialist and layperson that it is often taken for granted. Throughout the undergraduate curriculum in engineering, processes that are clearly mixing-dependent (such as chemical reaction, crystallization, die filling) are assumed to be homogeneous. This widespread preconception is also reflected in the common attitude toward powder mixtures, especially for relatively small particles that, due to their ability to scatter visible light, tend to look more uniform to the naked eye than is often warranted.

Early conceptions of homogeneous particle assemblies assumed the particles to be distributed in a state of perfect homogeneity, such that any sample containing a large number of particles would have essentially the same composition. Three conceptual approaches to such blissful state—perfect, random, and ordered mixtures— are discussed below. Real mixtures, unfortunately, tend to show at least some degree of heterogeneity, obeying to one of three main causes: incomplete mixing, agglomeration, and segregation, resulting in different types of textures, also discussed below.
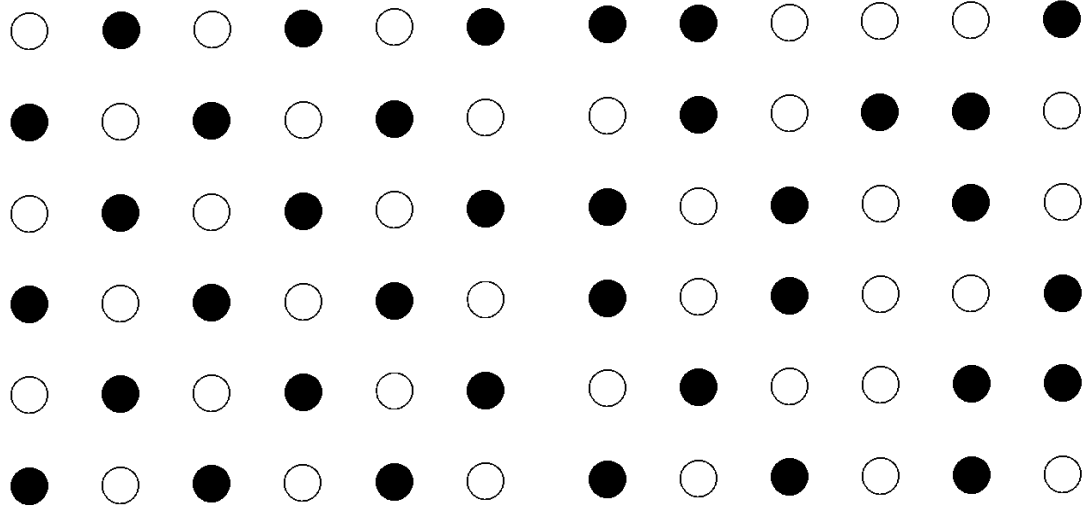
Figure 1 Simulated mixtures: (a) perfect mixture; (b) random mixture.

Perfect Mixtures versus Random Mixtures

 The first and simplest conception of a homogeneous system is the perfectly uniform mixture, where particles alternate themselves along a lattice (Figure 1a), very much resembling the position of atoms of different species inside a perfect crystal. Samples taken from such a mixture are necessarily identical. This highly ordered state is never achieved unless painstakingly created by po sitioning particles one at a time. If the particles are freely moving and differing from one another by a property that does not affect their movement in any way (such as, perhaps, color for identically sized glass beads), the best achievable state is that of a random mixture (Figure 1b), rigorously defined as a mixture where the probability of a particle belonging to a certain moiety is statistically independent of the nature of its neighbors. Sample extracted from such a mixture follow a binomial (or multinomial) distribution.

Ordered Mixtures.

For cohesive systems where the particles apply surface forces to one another, it is common to observe the formation of agglomerates. Depending on the relative magnitude of forces between like-particles and unlike-particles, it is possible to see agglomerates of a single species (the "guest"), as well as agglomerates where a small-size moiety essentially coats another, larger moiety (the "host"). This latter situation motivated the concept of an "ordered mixture" (which the reader should distinguish from the situation depicted in Figure 1a). In the ideal case, the same exact number of identical guest particles covers every identical host (Figure-2a). Samples taken from such a system would be, once again, identical, thus resulting in a higher degree of sample homogeneity than the random mixtures depicted in Figure 1b. In reality,

(a)   (b)
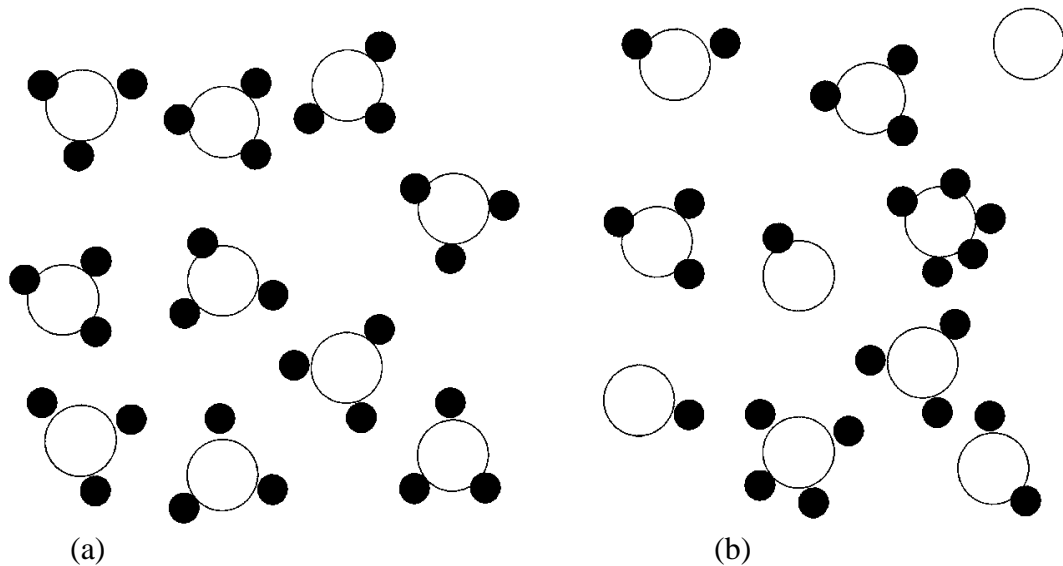
Figure-2 Distributions of individual particles that form an (a) ideal random mixture and a (b) random mixture. Part (b) illustrates a less structured blend that is still well mixed but does not exhibit long-range order in the spatial distribution of particles.

This distribution has been called the ideal random mixture, one for which the location of any particle has no influence on the particle (or particles) that are adjacent to it. In other words, a particle that is removed from any location in the mixture has an equivalent chance of being of either species type. In practical terms, this distribution is often the best attainable for a real system of interacting particles. one observes a distribution in the number of guests on each host, as well as free (unassociated) guests, leading to a less homogeneous outcome (Figure 2b).

Textured (Segregated) Mixtures.

The most troublesome mixtures are those that exhibit long scale texture (i.e., segregation), complicating description of mixture distributions and characterization. Textured mixtures form when a characteristic of one or more particle species causes that component to separate
into specific regions of the mixture, depending on the type of agitation applied to the bulk mixture. Also, dead zones or incomplete agitation of the powder can lead to segregated regions in blenders. In general, more free-flowing mixtures exhibit more extreme segregated states. Cohesivity acts to inhibit mixture segregation, as individual particles have trouble moving independently of the bulk mixture. Determining mixture quality of textured mixtures depends on accurately determining the size, location, and severity of the segregated regions.

This drawing depicts segregation of ordered units with different-sized carrier particles, but segregation of ordered units with leftover adherent particles is also possible. In any real mixture there will be areas that correlate closely to many of the ideal distributions discussed previously. Unfortunately, the characterization of mixture quality cannot currently be done by viewing particle distributions throughout the mixture. For real systems, samples are extracted from specific regions of the mixture and it is important to ensure that the sample size is representative.

Powder Sampling

Real systems do not yield complete and pristine data on the distribution of particulate
species within the bed. Instead, it is necessary to extract a finite, typically small number of samples from the mixture. These samples often have important limitations and biases, as discussed here.

The most common means for sampling powder constructs is through the use of sampling thieves. These devices are inserted into the bed and extract samples from the interior. When devising a sampling scheme, it is important to adequately sample all regions of the powder bed. As mentioned, granular materials can segregate spontaneously, and can

mix very slowly (especially when dispersion is the major mixing mechanism). Hence, sampling at only a few locations can lead to significant under sampling as regions of poor mixing are completely missed or underrepresented. Furthermore, post processing of a powder mixture can cause a previously well-mixed sample to de-mix and adversely affect further applications.

Physical Sampling Methods

The behaviors of two popular types of thief samplers are shown in Figures 4 and 5. In Figure 4a we illustrate the bed disturbances that occur when using a side sampling thief. This device consists of a tube with a slot in its side that can be opened to allow particles to flow into a cavity, and closed to extract the sample.
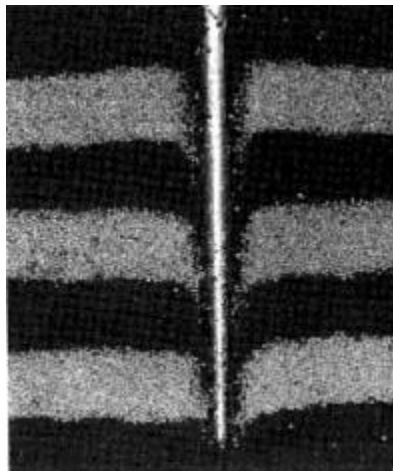
Figure 4 : Systematic sampling errors introduced by a side-sampling thief. (a) Initially layered configuration of large (light) and small (dark) particles are noticeably disturbed as the thief entrains particles during insertion.

An initially layered system of light gray 200 and dark 60 μm particles is visibly disturbed by inserting the probe. Particles are entrained along the insertion route, causing local particle rearrangements that typically result in the bed appearing to be anomalously well mixed. It is also significant that side-sampling thieves rely on particle flow into the sampling cavity to obtain particles; consequently, free-flowing or smaller particles can flow into the sampling cavity more readily than more cohesive or larger particles. These observations are quantified in Figure 4 that shows the fraction of smaller beads in samples obtained using a side-sampling thief in separate experiments in which 60 μm particles are initially arranged in a single thick layer over a bed of 200 μm particles. The thief obtains samples almost entirely consisting of the smaller species, irrespective of the actual concentration at the sampling location.

Sampling problems that arise from differences in particle flow into the sampling cavity can be mitigated through the use of end-sampling thieves, such as the one shown in Figure 5. For these thieves, the sampling tube is inserted to a desired depth in the bed, an aperture at the distal end of the probe is opened, and then the probe is pushed deeper into the bed to capture the sample; closing the aperture allows extraction of the sample. Particles are actively forced into the cavity rather than passively flowing into it, as in side-sampling thieves. Thus, this device is relatively free of differential sampling problems caused by differences in particle flow ability. However, Figure 5

demonstrates that these devices are typically bulky and consequently entrain and disturb considerable material during their insertion. For the case
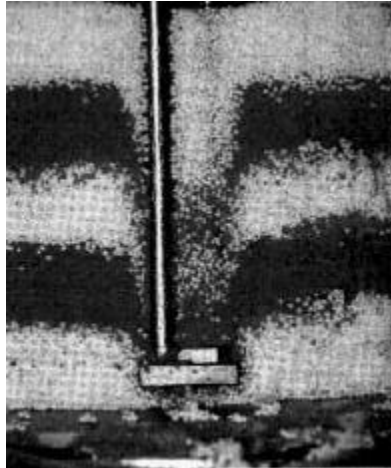
Figure 5 Sampling errors introduced by an end-sampling thief differ from those introduced by a side-sampling thief, but persist nonetheless. In this type of thief, a window is opened at the bottom of the sampling tube, and particles are forced into a cavity by further insertion of the thief. This eliminates the bias toward particles that passively fill a cavity more easily than others, but on the other hand, (*a*) these thieves entrain more particles during insertion.

discussed here, the resulting sample concentration measurements are improved over those of the side-sampling thief but remain very inaccurate, as data consistently overestimates mixture quality. An alternative that is nearly free of either entrainment anomalies is the core sampler. This sampler extracts an entire contiguous core of particles throughout the depth of insertion. At its simplest, the probe consists of a thin-walled tube that is inserted into a granular bed, together with a mechanized extrusion apparatus to permit samples to be extracted in a last-in, first-out manner after the tube has been removed from the bed. For capturing free-flowing particles, which can flow out of the tube, an end cap that can be opened during insertion and then closed during extraction is added to the device. Unlike the end-sampling thief, the end-cap mechanism here is internal to the sampling tube, and an entire core is extruded from the bed. The behavior of this device is demonstrated in Figure 6.
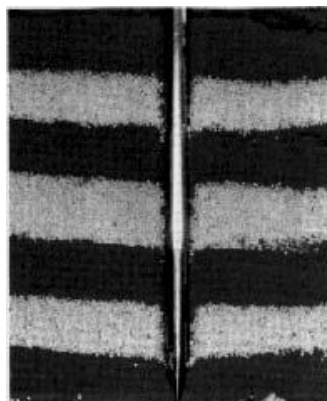
Figure 6 Core sampler with end cap can be used for freely-flowing (e.g., granulated) materials that would escape from the sampling tube during removal from the bed without the end cap. (a) Very little entrainment is visible after insertion.

Importantly, in the core sampler the core extends through the depth of the sampling tube, allowing for precise determination of concentrations between different layers of the bed. Furthermore, sample size is completely variable and can easily be adjusted for different mixtures, core sampler diameters, or changes in process parameters. By foregoing use of the end cap, core sampler performance is improved further.

Once samples have been obtained, one can use a variety of available chemical, optical, spectroscopic, chromatographic, or other assays to determine concentration. For example, data in Figure 15-7 were obtained using a calibrated densitometric technique in which one of the two species was colored in advance. Similar results have been obtained using other assay techniques, such as reflection near infrared spectroscopy to evaluate concentrations of magnesium stearate (a common pharmaceutical lubricant) or conductivity assays to evaluate the mixing of salt (NaCl, KCl) in anionic excipients (Avicel).

Noninvasive Methods

Other, more technologically complex techniques have also been developed for visualizing the interior of granular beds. These include:
• Diffusing wave spectroscopy, where statistics of fluctuations in relatively thin, Hele–Shaw configurations are measured
• Positron emission tomography, where a single radioactive particle is tracked during flow within a granular bed using an array of external photomultipliers
• Magnetic resonance imaging, where magnetic moments of hydrogenated particles are aligned in structured configurations (e.g., stripes) and these structures are tracked for short periods of time
• X-ray tomography, where a population of radiopaque particles are tracked in a flow of interest
These techniques are typically expensive and cumbersome to implement; nevertheless, they reveal flows within an optically opaque bed and provide valuable information not available otherwise. For example, in Figure 8, we display results of x-ray tomography experiments that show the evolution of the interior mixing structure within a double-cone blender using molybdenum-doped tracer particles.
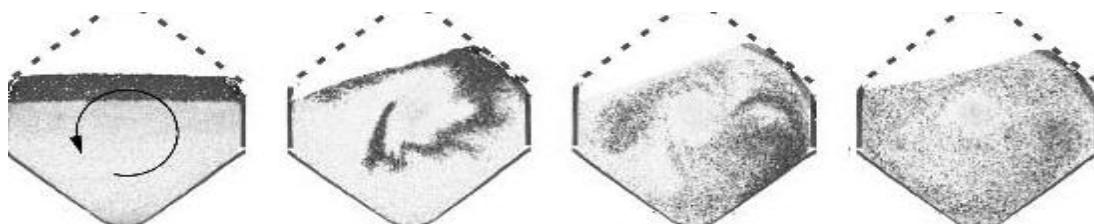
Figure 8 X-ray tomographic time series of blending of radiopaque grains in a double- cone blender is representative of several new techniques available for on-line and in situ assays of blending mechanisms.

Data of this kind reveal a complexity in flow and mixing evolution that simultaneously represents the cause of historical difficulty in understanding the subject and the opportunity for future developments. As these methods are improved, they will yield more quantitative information about mixture quality, leading to more robust methods for characterization of powder mixtures.

Mechanisms of Mixing: Freely-Flowing Materials

In tumbling applications, dilation and flow principally play out near the unconstrained upper surface of a granular bed, and except for solid-body rotation, the bulk of grains beneath are thought to remain nearly motionless during rotation of the blender. This simplified picture changes for some blenders (notably the V-blender, in which flow is strongly intermittent; see Moakher et al., 2000), but predictive models for blending in most common blending geometries can be derived by disregarding all transport beneath the free surface. In the sections following, we summarize the best existing models and methods and describe their
application to common tumbler designs. A useful design choice for the purposes of illustration is the horizontal drum tumbler. The horizontal drum is used in many chemical, metallurgical, and pharmaceutical industries in the form of ball mills, dryers, rotary kilns, coating pans, and mixers.

Flow in rotating drums with increasing tumbling speed has been described qualitatively in terms of regimes termed: slipping (or slumping), avalanching, rolling, cascading, cataracting, and centrifuging. These are defined as follows.

Slipping.

The slipping regime occurs when the granular bed undergoes solid body rotation and then slides, usually intermittently, against the rotating tumbler walls. This occurs most frequently in simple drums that are only partially filled and is typically counteracted by including baffles of various designs along the inner walls of the tumbler. While the slipping regime is not important for blending purposes per se, it is encountered even in effective blending systems, and an evaluation of the number of times a bed turns over per tumbler revolution will often reveal the presence of some slipping.

Avalanching

A second regime seen at slow tumbling speeds is avalanching flow, also referred to as slumping. In this regime, flow consists of discrete avalanches that occur as a grouping of grains travel down the free s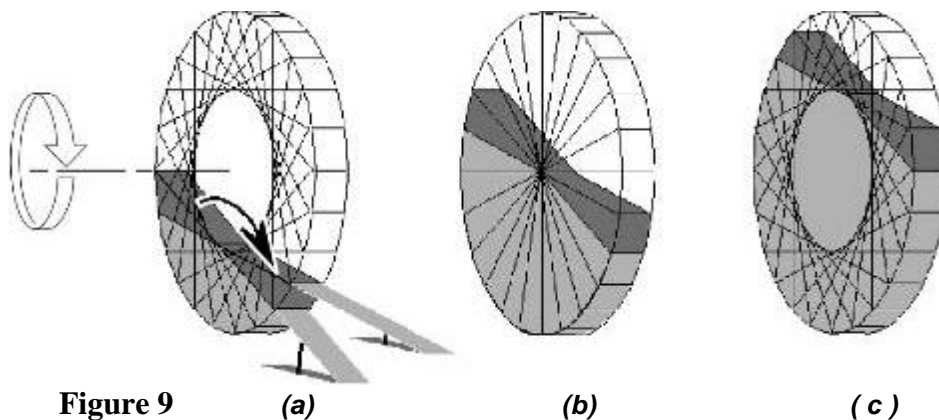urface and come to rest before a new grouping is released from above. The avalanching regime is not seen in tumblers larger than a few tens of centimeters
in diameter, but it is an instructive case because a flow and mixing model can be derived in

Figure 9   *(a)*   *(b)*   *(c)*

closed form for simplified drum geometries. To analyze this problem, one needs only bserve that if the angle of repose at the free surface immediately before an avalanche is θi, and after an avalanche is θf, the effect of the avalanche is to carry a wedge of material in the angle θf − θi, downhill, as sketched in Figure 12a for an idealized two dimensional disk blender. The same behavior occurs for all fill levels, and one can readily use this model to make several concrete predictions. First, mixing occurs during avalanches through two distinct echanisms: (1) particles within a wedge rearrange during a single avalanche, and (2) particles rearrange globally between wedges during successive avalanches. Second, at 50% fill (Figure 12b) no two avalanching wedges intersect, so no global mixing between separated regions can exist, and mixing must slow. Third, since flow occurs only near the avalanching surface, at high fill levels a nonmixing core necessarily develops (Figure 12c). Although this model is oversimplified and neglects material variations, boundary effects, and other important phenomena, these conclusions carry over to more realistic tumbling systems.

Rolling

At higher tumbling speeds, discrete avalanches give way to continuous flow at the surface of a blend . Grains beneath this surface flowing layer rotate nearly as a solid body with the lender until they reach the surface. One can solve for flow and transport subject to certain simplifying

assumptions in this regime as well. For this solution, one assumes that the grains are so small as to be regarded as a continuum and one takes the free surface to be nearly flat, as sketched in Figure 13a. The interface between the flowing layer and the bed beneath has been determined experimentally and computationally to be roughly parabolic in shape, and by demanding mass conservation at this interface, one can construct continuum flow equations for this system. If one simulates the mixing in an idealized disk blender of mechanically identical grains initially separated by color to left and right of avertical central plane, one obtains the results displayed in Figure 13 b (for a particular fill level and flowing layer depth). Corresponding experimental results are shown in Figure 13c.

Figure 10   *(a)*   *(b)*   *(c)*
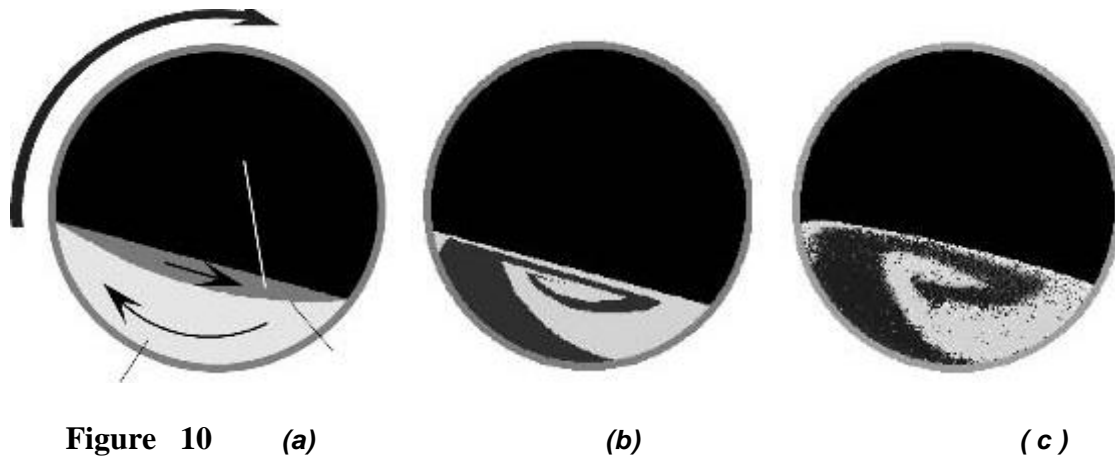
Cascading, Cataracting, and Centrifuging.

For larger tumblers, or for tumblers rotated at higher speeds, the surface is manifestly not flat, as shown in Figure 11 in a 1 m diameter disk tumbler. This flow, termed *cascading* differs qualitatively from the rolling flow solution; here the flowing layer is thin, is nearly uniform in speed and thickness, and has been modeled as depth-averaged pluglike flow. As the rotation speed of the tumbler is increased, the surface becomes increasingly sigmoidal until grains become airborne, and at higher speeds yet, the grains centrifuge against the tumbler wall. These regimes are termed *cataracting* and *centrifuging*, respectively, and have not been well analyzed.
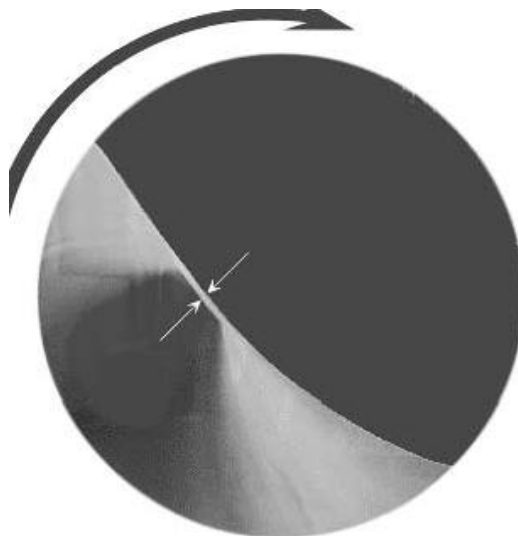
Figure 11   Cascading flow occurs in large tumblers or during tumbling of fine but freely flowing grains. This snapshot shows a 1 m diameter 1 cm wide transparent disk tumbler partially filled with colored ~500 μm irregular grains. Initially, light and dark grains were placed to the left and right of a central plane; this napshot shows the mixing pattern at one-half revolution of the disk. This tumbler is thin, so grains are under the strong influence of wall effects; nevertheless, this example serves to illustrate that the free surface is manifestly not flat, and the cascading layer is thin and nearly uniform along the flowing surface.

Mechanisms of Mixing: Weakly Cohesive Material

Another mechanism of granular and powder mixing is associated with blending of weakly cohesive materials. Weakly cohesive materials (e.g., powders and fine grains in the size range 50 to 300 μm) exhibit stick-slip motion so that flow becomes intermittent rather than continuous. This is a situation of practical importance since most industrial applications use particles across a broad range of sizes and materials. As the size of grains diminishes or as interparticle cohesion grows, stick-slip flow transforms mixing interfaces from a smooth, regular patterns as shown in Figure 12 (500 or 700 μm cases) to a complex, irregular pattern, shown in Figure 12 (300 or 100 μm cases). In simple geometries this response to shear can be modeled accurately: If we assume that the flowing surface of a bed sticks and slips periodically, the mechanism displayed in Figure 12a can be embellished by allowing the shear band between flowing layer and bed to deform periodically. This produces mixing 100 μm Model
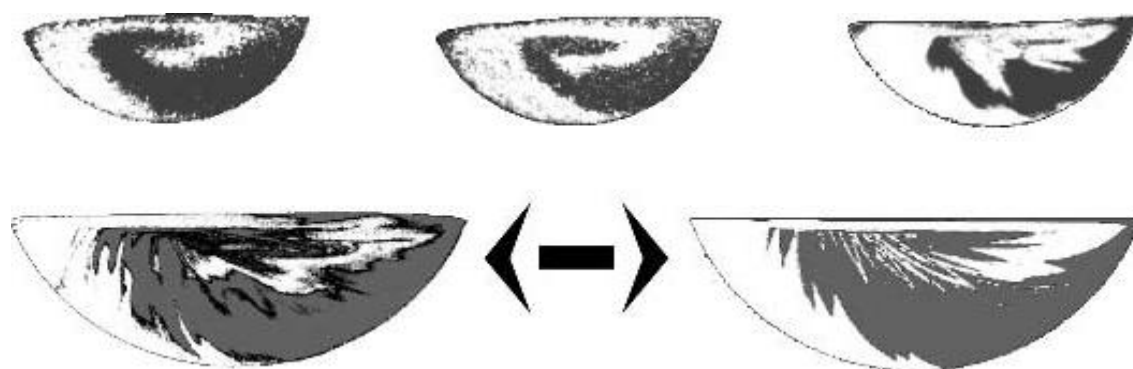
Figure 12   Mixing patterns after one revolution in identical drum tumblers loaded with identical (except for color) grains in four experiments using successively finer grains as well as in a model simulation of idealized stick-slip flow. At 700 and 500 μm, the mixing interface remains smooth and regular; below about 300 μm, it becomes variegated due to intermittent slipping of the cascade. Each experimental snapshot shows a view from the interior of a blend using the solidification technique and all cases began with light grains to the left of center and dark grains to the right.

identical grains that are substantially similar to experimentally observed ones, as shown at the bottom of Figure 12 This is important for blending because in smooth regular flow, adjacent particles remain nearby for long periods of time, while in intermittent stick-slip flow, particles can rapidly relocate across the blender, resulting in an exponentially rapid growth of interfaces between separated regions of grains.

For particles smaller than about 100 μm, cohesive forces (believed to be due to van der Waals interactions for intimate contacts, and to surface tension of adsorbed water layers for lubricated contacts) between particles become comparable to particle weights, and small particles can stick to one another in relatively rigid aggregates. Unless such aggregates are destroyed, the system will behave as if it had an effective particle size much larger than the primary particle size. For strongly cohesive materials, it is typically necessary to fragment agglomerates through the introduction of high-shear, intensification devices, such as impellers or mills that energetically deform grains on

the finest scale. Many forms of intensification are used in industrial practice. Some common approaches include passing the blend through shaker sieves or through hammer or pin mills between blending stages, as well as using high-speed devices within blenders, such as intensifier bars in tumbling or choppers in high-shear granulator-style mixers. Essentially no detailed systematic quantitative information is available concerning the effects of intensification on blend quality. We are aware of no studies investigating the micromixing quality as a function of intensity and duration of applied shear. Scale-up and design information provided by equipment vendors is largely limited to advising the user to keep intensifier tip speed and time of operation constant during scale-up. Although this guideline is reasonable in lieu of rigorous information, it is clear that in situ intensifiers apply shear only locally, and nonuniformly, to the mixture; the end result is almost guaranteed to be affected by the interplay of the intensity of the shear field, the residence time of particles in the shear field, and the global homogenization capabilities of the blender. At the present time, laboratory devices for applying shear uniformly and at a known rate are unavailable, making study of the problem even harder. Given a tendency across industry to deal with ever smaller, ever more cohesive materials, understanding the role of shear on blend quality is undoubtedly one of the areas in greatest need of attention by the scientific community.

De-mixing

Processing blends of dissimilar grains almost invariably promotes de-mixing, also referred to as segregation, characterized by the spontaneous emergence of regions of nonuniform composition. Segregation due to differences in particle size in a blend has drawn the greatest attention in the literature, including studies of fluidized beds, chutes, hoppers, vibrated beds, and tumbling blenders, but segregation due to differences in particle density, shape, and triboelectric order have also been recorded. As a practical matter, segregation manifests itself in granular mixing that characteristically improves over a brief initial period, while convection generates large scale mixing, and then degrades, often dramatically as slower segregation fluxes take over. De-mixing should not be confused with the phenomenon of overblending, which is also frequently encountered in blending applications. Overblending is associated with physical degradation of material properties, as occurs, for example, when a waxy lubricant is excessively deformed, causing it to coat pharmaceutical grains and reduce their bioavailability or when coated granules are damaged through abrasion or fracture. At the present time, mechanisms for segregation, even in the simple tumbling drum, remain obscure, and work on more complex and industrially common blender geometries is extremely limited. Three distinct types of de-mixing are moderately well characterized in tumblers: radial de-mixing, axial de-mixing, and competitive patterned de-mixing. We describe each of these in turn.

Segregation typically proceeds in two stages. First, large grains rapidly segregate radially, producing a central core of fine grains surrounded by larger grains, identified in Figure 13 or a simple drum tumbler. Unlike the core seen in overfilled tumblers, this core appears at fill levels under 50% and is associated exclusively with migration of fine grains toward the center of an overturning blend. Radial segregation is seen in both quasi-2D and fully 3D blenders of various geometries. In simpler 3D geometries, such as the drum, double-cone, or tote, the core is nearly always apparent when blending significantly dissimilar grains, while in more complicated geometries such as the V-blender or slant cone, the core becomes significantly distorted and may only be conspicuous for higher fill levels or in certain (e.g., upright) orientations of the blender. Even in the simplest case of the drum tumbler, however, the location and dynamics of the core remain somewhat enigmatic—for example, as shown in Figure 13, the core is actually located upstream of the geometric center of the granular cascade. The core appears to form as a result of two cooperative influences. First, smaller grains percolate through the flowing layer to occupy successively lower strata each time the bed overturns. Second, once a sufficient volume of smaller

grains has accumulated, the larger grains tend to roll increasingly freely over the (comparatively smooth) substrate of smaller grains. This higher-speed surface flow reinforces the segregated state by expelling remaining slower small grains. These mechanisms are very robust, and cores are almost invariably found in tumbling of freely flowing grains with diameter ratios between about 1 : 1.5 and 1 : 7. As the diameter ratio approaches unity, the core becomes more diffuse, while as the
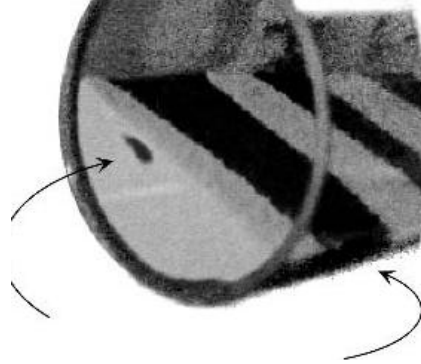

Figure 13   Typical segregation pattern seen between fine (dark) and coarse (light) grains in a small transparent drum tumbler. A core of fines travels along the entire length of the tumbler, connecting the bands that emerge at the surface in a single bulging tube. The coarse grains are constrained to flow within the confines defined by this tube. This constraint is important for understanding mechanisms of de-mixing in more complex geometries.

diameter ratio grows sufficiently large, fine grains can percolate increasingly freely through a matrix of larger grains or, if sufficiently fine, can coat the larger species.

Axial De-mixing
 A second stage of segregation occurs in drum tumblers as grains in the core migrate along the tumbling axis. Numerical and experimental investigations have attributed this migration to conflicting causes (e.g., a secondary flow within the core leading to a bulging of the core toward the surface versus different angles of repose of fine, mixed, and coarse grains). Whatever the ultimate cause, the result of this axial migration is the formation of a series of bands as shown in Figure 15-16. In this final state, two pure phases of material are formed, divided by sharp boundaries with very little intermixing.

Competitive Patterned De-mixing
In more complex, and more common tumbler geometries, several distinct segregation patterns have been observed. These patterns are believed to arise from a competition between surface segregation of coarse grains flowing over a radially segregated core of fine grains and interactions with the boundaries of the tumbler. Despite significant differences between common blender geometries, there is substantial commonality in the ultimate patterns seen. For example, mixing of large, light-gray, and small, dark-gray grains in a double cone and a V-blender generate similar patterns in both experiments and particle-dynamic simulations as shown in Figure 14. As parameters such as fill level, tumbler speed, and concentrations of the different particle species are varied, the patterns observed change significantly. Importantly, there appear to be few dominant and recurring patterns that are seen in both experiments and simulations in all blender geometries. Notably at high fill levels and tumbling speeds, the left–right state shown in Figure appears to dominate. This pattern and two other common variants are shown at the top of Figure 15 in top views of the surface of a double-cone blender. Each of these patterns appears reproducibly and spontaneously whenever different-sized grains are tumbled in any of several blender geometries. Simulations shown beneath the experimental figures in Figure 15-18 use a continuum model in which large
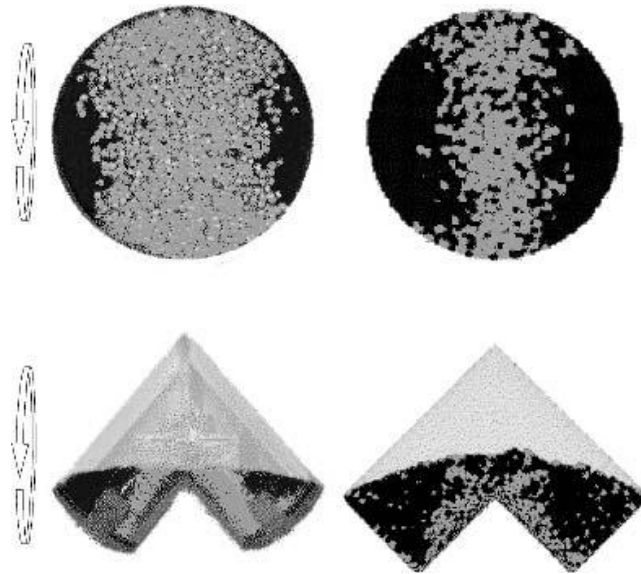
Figure 14 Axial segregation in top views of double-cone blender from (*a*) experiment and (*b*) particle-dynamic simulation using large, light and small, dark spherical grains. Similar patterns are seen in other tumbler designs: for example, in the V-blender in (*c*) experiment and (*d*) simulation.
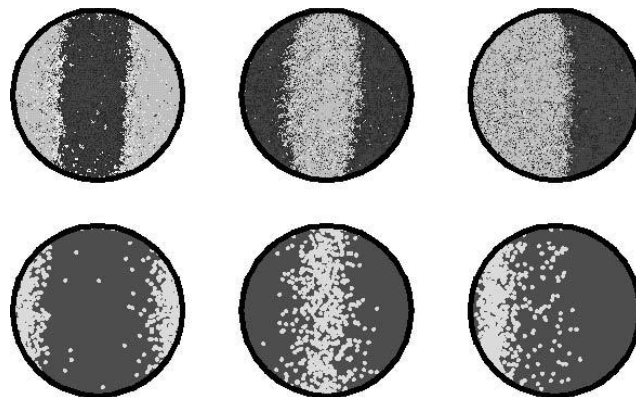
Figure 15   Three common segregation patterns between large (light) and small (dark) grains seen in top views of a double-cone blender. Top: experimental snapshots; bottom: simplified continuum simulations.

BATCH MIXERS AND MECHANISMS

Tumbling Mixers

Although drum blenders represent a convenient paradigm for the purpose of categorizing granular behaviors, most blending operations occur in more complex tumbler geometries. Three of the most common geometries used in pharmaceutical operations are the double cone, the V-blender, and the bin blender, sketched in Figure 16. Each of these geometries possesses many variants; for example, symmetry can be broken to introduce cross-flow by slanting the double cone, by elongating one of the arms of the V-blender, or by inserting baffles in a bin. To model flow and blending in complicated geometries, particle-dynamic simulations have been applied. In these simulations, particles are treated as individual entities with physical properties (e.g., size, static and dynamic friction coefficients, coefficient of restitution, etc.) appropriate to the problem of interest, and Newton's laws of motion are integrated for each particle. Particle-dynamic simulations are similar in concept to molecular-dynamic simulations but include features of importance to the flow of macroscopic particles (e.g., static and dynamic friction models)
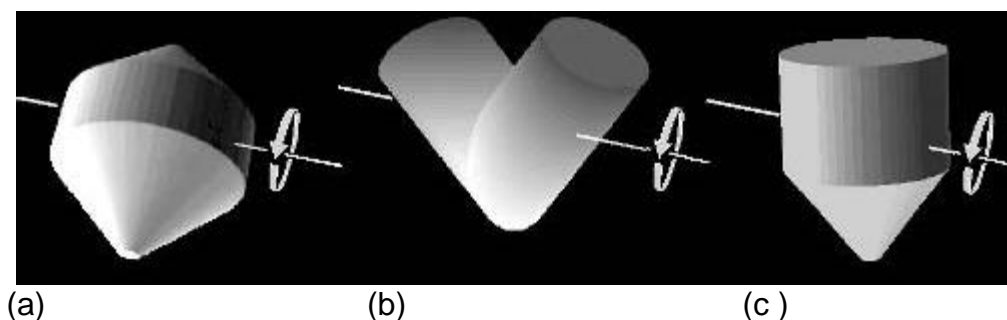
Figure 16   Three common tumbler designs: (*a*) double cone, (*b*) V, and (*c*) tote or bin blenders.

in place of microscopic properties (e.g., bond strengths and chemical potentials). Particle-dynamic simulations come in many different types, depending on how they treat physical parameters, such as rolling friction and particle shape, or numerical issues, such as search algorithms and routines to maintain computational stability. As such, results of distinct computational simulations can differ, sometimes significantly, and the importance of experimental validation of numerical results cannot be overemphasized. Two of the most common classes of particle-dynamic simulations are termed hard-particle and soft-particle methods. Hard-particle methods calculate particle trajectories in response to instantaneous, binary collisions between particles, and allow particles to follow ballistic trajectories between collisions. This class of simulation permits only instantaneous contacts and is consequently often used in rapid flow situations such as are found in chutes, fluidized beds, and energetically agitated systems. Soft-particle methods, on the other hand, allow each particle to deform elastoplastically and compute responses using standard models from elasticity and tribology theory. This approach permits enduring particle contacts and is therefore the method of choice for tumbler applications. The simulations described in this chapter use soft-particle methods and have been validated and found to agree in detail with experiments.

V-Blender

Mixing in all tumbling blenders consists of a fast convective stage, driven by the mean velocity of many particles, followed by a much slower dispersive stage, caused by velocity fluctuations leading to rearrangements of individual particles. Convection in grains (as in fluids) is by far the faster and more efficient mixing mechanism, yet at the same time it suffers from the same mixing limitations known for fluids: convective flows can—and very often do—possess barriers to mixing (e.g., islands) that do not interact with surrounding material. Two pathologies are readily observed: overfilled mixers develop elliptic, nonchaotic islands that rotate as a unit in the center of the granular bed and symmetric blenders (seen in most standard designs) exhibit separatrices that divide the flow into noninteracting sectors. Beyond this, little is currently known of details of particle flow patterns and mixing barriers in practical, three dimensional blender geometries, although there is strong evidence indicating that flow bifurcations analogous to those seen in fluids may be present in granular tumblers. Convection in the context of granular blenders refers to transport associated with flow driven by gravity (in tumbling blenders) or impellers (in intensified, ribbon, or other blenders). Convection is observed in all functioning blender geometries and can be visualized using particle-dynamic simulations. In Figure 17 we display successive front and side views taken a quarter revolution apart of 20 000 identical but colored spheres tumbled in a V-blender in the cascading regime. These snapshots illustrate the qualitative motion produced in this blender, which causes the bed to overturn from top to bottom. Mixing due to convective flow grows linearly with time insofar as the area of an interface grows characteristically linearly with time. Similar qualitative behaviors are seen in all tumbler geometries, although the quantitative mixing seen can differ considerably between geometries.
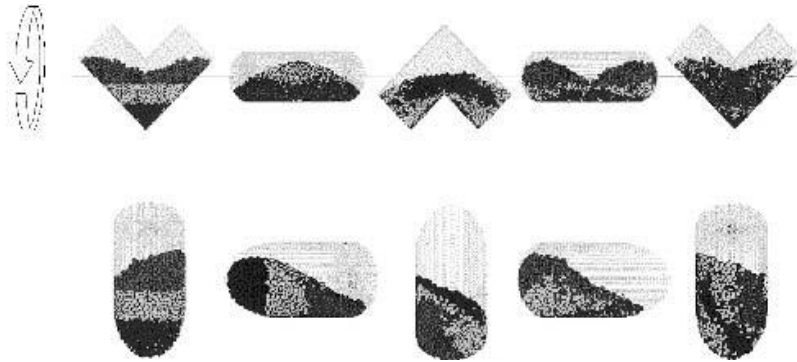
Front

Side

Figure 17   Rapid convective flow seen in particle-dynamic simulation of identical but colored spheres in a V- blender. Top: Front view reveals that unlike in some designs, convection in this blender drives grains axially alternately outward toward the tumbler arms and inward toward its center. This axial flow strongly influences mixing. Bottom: side view indicates that transport is dominated by a spiraling flow, seen also in drums and other blenders.

Bin Blender.

In contrast to convection, which can effectively intersperse grains in a tumbler within tens to hundreds of revolutions, is dispersion, or diffusion. Dispersion refers to the random relocation of individual grains due to collisions between adjacent particles and can take hundreds to thousands of revolutions to act. Thus, particles can only cross a plane separating the two arms of the V-blender (or an equivalent symmetry plane in many other blender geometries) as a result of occasional collisional happenstances and not
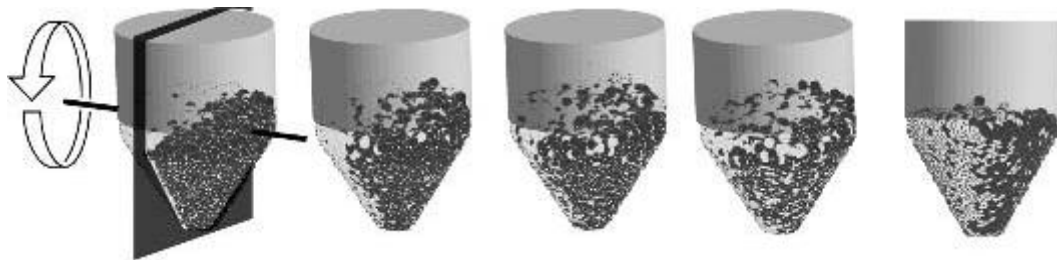
Figure 18   Dispersive mixing is slow across the symmetry plane of a blender, here a bin design. After 10 revolutions, a front view reveals clear evidence of the initial left–right distribution of identical but colored spheres in this particle-dynamic simulation

as a result of an overall mean flow. Various stratagems, including the use of baffles, asymmetric cross-flow designs (referred to earlier), irregular rotation protocols, and axial rocking, have been introduced to mitigate this limitation. Notwithstanding these improvements, dispersion is the rate-limiting mechanism for mixing, and there is much potential for improvement of dispersive mixing.

Although convection is typically orders of magnitude more rapid than dispersion, the relative contribution of each mechanism to blending is strongly influenced by the initial distribution of species in the mixer. Thus, ingredients loaded in horizontal layers (as in Figure 18 can be mixed relatively rapidly, while ingredients layered side by side, either intentionally (as in Figure 18) or inadvertently (as a result of careless loading of a tumbler), will typically mix enormously more slowly.

To visualize this effect, in Figure 18 we display dispersive mixing of 8000 identical but colored grains loaded side by side, in a bin blender. With each successive revolution, only a few particles cross the interface separating the two symmetric halves of the tumbler, and as a result, after 10 revolutions the original particle ordering is still unmistakable. Systematic assays obtained from

experiments of blending of realistic pharmaceutical excipients and actives confirm that imperfectly loaded blends retain any initial asymmetry for many hundreds of tumbler revolutions.

Convective Mixer

The second class of blenders commonly used in industrial applications is convective blenders. In contrast to tumbling blenders, convective blenders primarily mix by transporting material throughout a mixing vessel by the motion of a stirring device. A typical convective blender consists of a stationary chamber swept out by stirring mechanisms, such as rotating impellers. Convective blenders have a broad range of applications and can be used to blend components that cannot be adequately combined in tumbling blenders, such as materials that are prone to segregate or agglomerate. Many convective blenders can be designed to accommodate continuous rather than batch processing, further adding to their utility.

Figure 19   Well-mixed powder bed in a five-spoke ribbon blender, with both sampled cores and regions blocked by mixing blades visible.

However, despite this versatility, we have only a very limited understanding of the dynamics and performance of convective blenders. This is due to both the inherent difficulty involved in characterizing powder and granular mixing and the fact that few researchers have investigated these devices (see, e.g., Masiuk, 1987). In fact, much of the limited quantitative analysis was performed over 30 years ago (Adams and Baker, 1956; Greathead and Simmons, 1957; Poole et al., 1964; Ashton and Valentin, 1966; Harnby, 1967; Williams and Khan, 1973) and suffers from many of the limitations and difficulties of powder bed and granular mixing characterization discussed elsewhere in this chapter. Convective blenders also pose special challenges to powder bed analysis, as the impellers may hinder or block acquisition of powder samples, rendering thorough and uniform sampling difficult. Figure 19 shows a blended powder bed in a ribbon blender, a common convective blender. Several cores have been extracted from this bed, but it is evident that certain locations were impossible to sample, due to blades obstructing that location. Process design and optimization utilizing convective blenders are therefore performed predominately on a case-by-case basis.

Blending Mechanisms
Powder blending in a convective mixer is accomplished primarily by convection and shear effects. The motion of the impeller initiates blending by transporting material from region to region within the mixing chamber. This motion also generates slip planes, which often results in nonnegligible shear mixing as well. Diffusive mixing typically has a minimal role in mixing in these blenders.

This combination of mixing mechanisms produces superior results in many applications. Segregation of particles of different properties can be a major problem in tumbling mixers, potentially resulting in de-mixing or agglomeration (accretion of smaller particles into larger clumps). The motion of a convective mixing impeller both prevents the formation of and destroys any preexisting agglomerates. Convection is the mixing process least likely to result in pronounced segregation caused by difference in size or other physical properties. Convective mixers typically transfer relatively large amounts of material throughout the mixing vessel, allowing only limited opportunities for particles to segregate. This is in contrast to tumbling blenders, where the material has many opportunities to segregate during the tumbling/avalanching phase of the mixing cycle. In addition, mixing occurs throughout the chamber in a convective blender, whereas in a tumbling blender, mixing is typically confined to the surface of the powder bed, leaving large regions undisturbed during long periods of the mixing cycle.

While providing resistance to segregation and agglomeration, the mixing mechanisms employed by a convective blender can produce potentially adverse consequences. The motion of the impeller induce high levels or rates of shear, which may damage the material being mixed. Two other effects that may occur in a convective blender are attrition (grinding the powder into finer particles) or intense heating of the powder. Attrition is a common phenomenon in these blenders, as the shearing action of the impeller blades can cause rapid reduction in the individual particle size. This effect is sometimes desirable, for example, when designing a process to equalize the particle size of the material being processed. The motion of the impellers may aerate the mixture and cause the regions near the surface to fluidize. Convective blenders are typically operated at rotational speeds (15 to 60 rpm) and fill levels (>50%) similar to those of tumbling blenders. However, no systematic study of mixing performance in terms of mixing time, rotation rate, fill level, or loading patterns has been published to date for any convective blender.

Applications and Types.

The design of convective blenders allows for the efficient processing of a wide range of material states, including powders,granular solids, slurries, liquids, pastes, and combinations of these. Consequently, these blenders are utilized in a wide range of industries, including construction, agriculture, chemicals, pharmaceuticals, and foods. Convective blenders come in a wide variety of designs, all following the basic design of a stationary vessel swept by a rotating impeller. The stationary mixing chamber may be conical, cylindrical, or trough-shaped, and impeller designs range from ribbons to paddles
to screws. The impeller may sweep though the entire mixing chamber each rotation, or it may stir small regions progressively until the entire chamber is stirred. Common convective blenders include the ribbon blender (a cylindrical vessel with a helical ribbon impeller mounted to a horizontal shaft), the paddle blender (a modified ribbon blender with paddles instead of a helical ribbon), and the Nauta blender (a vertically oriented conical tank swept out by a rotating and processing screw impeller). Other blenders used in industrial applications include the Forberg mixer (two paddle blender drives sweeping two connected troughs), the Z-blade blender (a cylindrical vessel swept out by a Z-shaped blade), and the Lodige (similar to a kitchen mixer, where plough-shaped shovels rotate within a cylindrical drum).

Ribbon Blender.

The ribbon blender is one of the most common general purpose mixers, as it is capable of effectively performing a wide range of mixing processes including liquid, solid, and liquid–solid blending. Common industrial applications of these blenders include mixing the powder components

of pharmaceutical tablets, blending oils and shortenings into dry ingredients to form a cake batter, and combining gravel and asphalt. A batch ribbon blender is depicted in Figure 20. The motion of the ribbon blades near the vessel walls can result in pinch points, regions of high shear and compression which may damage fragile materials or cause attrition. The capacity of these blenders is set by the span of the ribbon, which must clear the top of the powder bed in order to mix the entire bed. As is true for many convective blenders, the intensity of shear can result in heating that can adversely affect the quality of the product. During operation of a standard ribbon blender, two sets of helical ribbon blades transport material in opposite directions; the outer ribbons will transport material toward the center of the mixing vessel while the inner ribbons transport material toward the ends of the vessel (Figure 21a). Turbulent convective currents caused by these counter rotating elements act to blend the different components. Unlike many tumbling blenders, a ribbon blender is often not completely discharged by gravity, requiring additional blade rotation to complete this process. This can result in additional mixing, segregation, and attrition, which must be taken into account during process design.
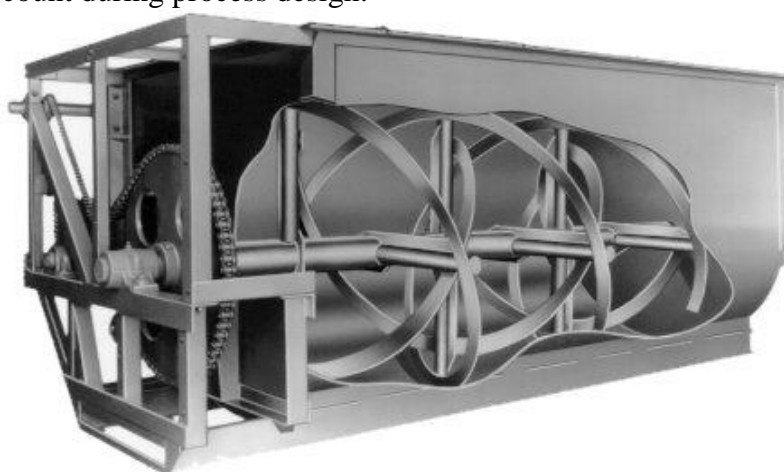
Figure 20   Schematic of a ribbon blender, consisting of a cylindrical vessel swept out by a rotating helical impeller blade. (Courtesy of H.C. Davis Sons Manufacturing Co., Inc., Bonner Springs, KS, printed with permission.)

Ribbon blenders are simple to modify for specific processes, and many refinements have been commercialized. The most common type of modification is to change the ribbon blade design; for example, Figure 21 shows some variations of the standard ribbon blade (Figure 21a). Figure 21b shows a common variation: the center-discharge ribbon. Here the two outer ribbons bring material to the center of the blender, while the inner ribbons force the ingredients outward to each end of the vessel. The paddle agitator (Figure 21c) contains both forward and reversing paddles in place of the smooth ribbon, constantly moving ingredients from one end to the other. To limit pinch points between the paddles and the blender surface, notches, or saw teeth, can be cut into the paddles (Figure 21d). Other types of modifications include creating a hybrid ribbon-paddle agitator or adding components to or removing sections of a ribbon to vary shear effects. Two agitators of the latter type are the cut-it-in ribbon agitator (a standard ribbon supplemented with cutting wires mounted on ribbon arms) used to cut thick materials (fats, oils, shortening) into powders (flour) and the cutout agitator (alternating sections of a standard ribbon are removed) used for heavier materials. Although there are many variations on the simple ribbon, these are all ad hoc, and there are few rigorous and usable scientific studies of these devices.
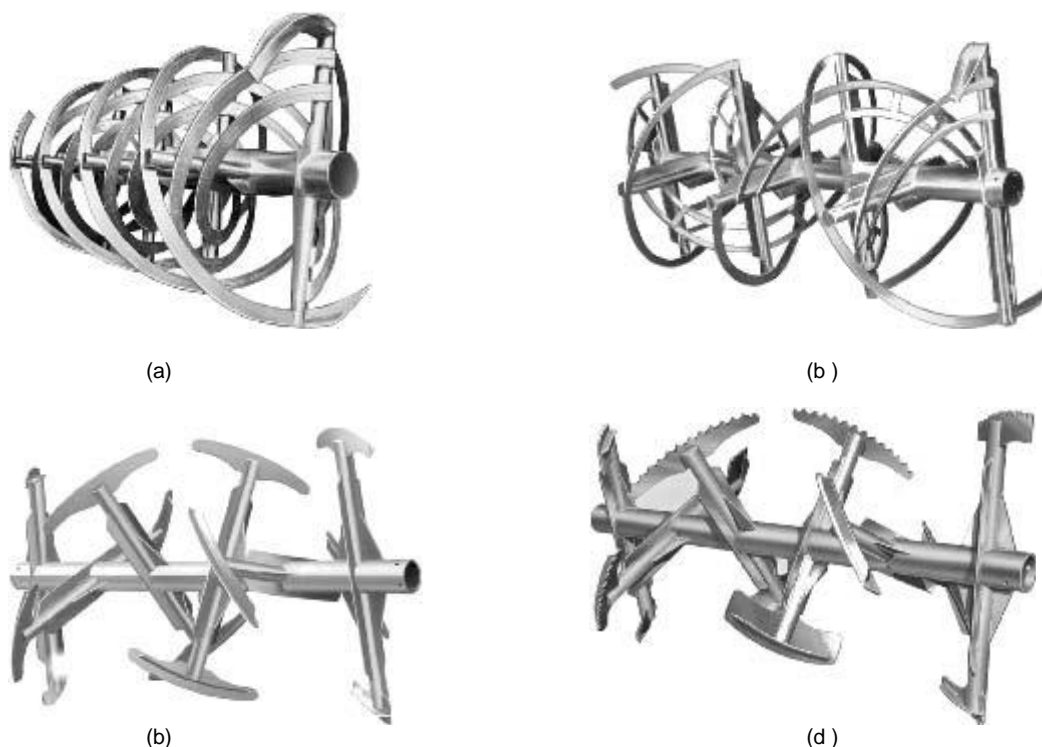
Figure 21   Four convective blender impellers: (*a*) double ribbon agitator; (*b*) center- discharge ribbon agitator; (*c*) paddle agitator; (*d*) sawtooth paddle agitator. (Courtesy of H.C. Davis Sons Manufacturing Co., Inc., Bonner Springs, KS, printed with permission.)

SELECTION AND SCALE-UP OF SOLIDS BATCH MIXING EQUIPMENT

As mentioned above, tumbling blenders can be grouped into two categories: convective blenders and tumbling blenders. Convective blenders rely on the action of impellers or paddles within a bowl, trough, cone, and so on, to move powders around and to generate a well-mixed product. Tumbling blenders consist of a hollow vessel attached to a rotating shaft; different blender types are identified by the geometry of the vessel. In either case, powders are mixed by rotating the blades or the vessel itself at a fixed rotation rate.

From a design and scale-up view point, the major difference between the two types of blenders is the amount of shear imparted to the mixture during the blending process. In the absence of intensifier bars, tumbling blenders provide low-shear environments and are used when materials are shear sensitive or non-agglomerating.

Convective blenders impart much more shear into the mixture and tend to be utilized for cohesive materials. Some tumbling blenders are equipped with a high-speed impeller, which can greatly increase the shear environment and allow for blending of some cohesive mixtures. From a manufacturing standpoint, tumbling blenders are often preferred because they come in a wide range of capacities and have shorter cleaning times. The choice of mixer often comes down to the properties of the mixture in question. Unfortunately, without well-established methods for measuring cohesion or agglomerating tendencies for different mixtures, it is impossible to develop a priori rules for blender choice based on the characteristics of the mixture materials.

As mentioned previously, the description of mixing mechanisms in convective blenders has not been the subject of considerable experimental investigation work, relegating scale-up considerations to trial and error.

Final Scale-up and Scale-down Considerations

Generally, mixture characteristics (rather than blender characteristics) play the determining role in scale-up of tumbling blenders. For free-flowing mixtures, changing vessel size has an enormous impact on axial mixing rates, making the loading conditions the most important variable affecting the change in mixing rates with changes in scale. For cohesive mixtures, the amount of shear in the mixing process has the greatest effect on the mixing rate because the breakup and dispersion of agglomerates is necessary for creating well-mixed blends. Thus, one must be cautious of *increases* in vessel size for free-flowing mixtures, but for cohesive mixtures it is *decreases* in vessel size that pose the most problems. It is clear that the processes that control mixing in these devices will not be fully understood until quantitative methods for measuring cohesion of powder mixtures are developed.

CONCLUSIONS

Despite sustained efforts during the past decade both at Rutgers and elsewhere, powder blending remains largely an "art," governed by empiricism and subject to frequent failure. In the opinion of the authors, the chief limitations in advancing the scientific understanding of powder mixing is a lack of effective experimental methods for measuring powder flow and powder constitutive behavior. Lacking hard data, constitutive modeling remains in its infancy, severely limiting our ability to achieve effective models for predicting powder flow and mixing from first principles. This situation is in stark contrast with the state of the art in fluid flow and mixing, which is discussed abundantly in the remainder of this book. For fluids, well-established methods for measuring constitutive behavior have greatly advanced our understanding of fluid rheology, which is a major building block of CFD models. Excellent EFD technology facilitates validation of CFD efforts. The net result is that fluid mixing systems can be designed with increasing reliability, and in many cases, entirely by computer. Although limitations exist, the fluids community is taking them by storm, one after another, at a rapid pace.

Thus, to advance beyond the current state of the art, we need to improve experimental techniques for measuring powder flow and powder constitutive behavior. Important efforts, mentioned earlier in this chapter, are under way using a variety of noninvasive technologies, but much remains to be done before powder blending processes can be designed and scaled-up reliably. The reader is advised to stay tuned, as the field is likely to evolve rapidly in coming years.

Chapter 8
Scale Up of Mixing Process

The main objective of scale-up is to design a large scale mixing system that will achieve the same mixing quality as in a laboratory tank. Since the distributions of shear rate and energy dissipation widen as the volume is increased, the mixer design must be adjusted to obtain the same process result. Therefore, it is important to understand the impact of these differences on the process. The scale-up criteria depend strongly on the process type and requirements.

Some scale-up methods emphasize geometric similarity. This refers to holding constant the impeller geometry, the impeller dimensional ratios (such as D/T, W/D, C/T), the liquid height/tank diameter ratio, and baffling. There are many situations when complete geometric similarity is not feasible: for example, when the aspect ratio of commercial scale tanks needs to be larger than the laboratory tank.

There are two commonly used scale-up criteria based on holding power per unit volume (P/V) or torque per unit volume (TQ/V) constant on scale-up.
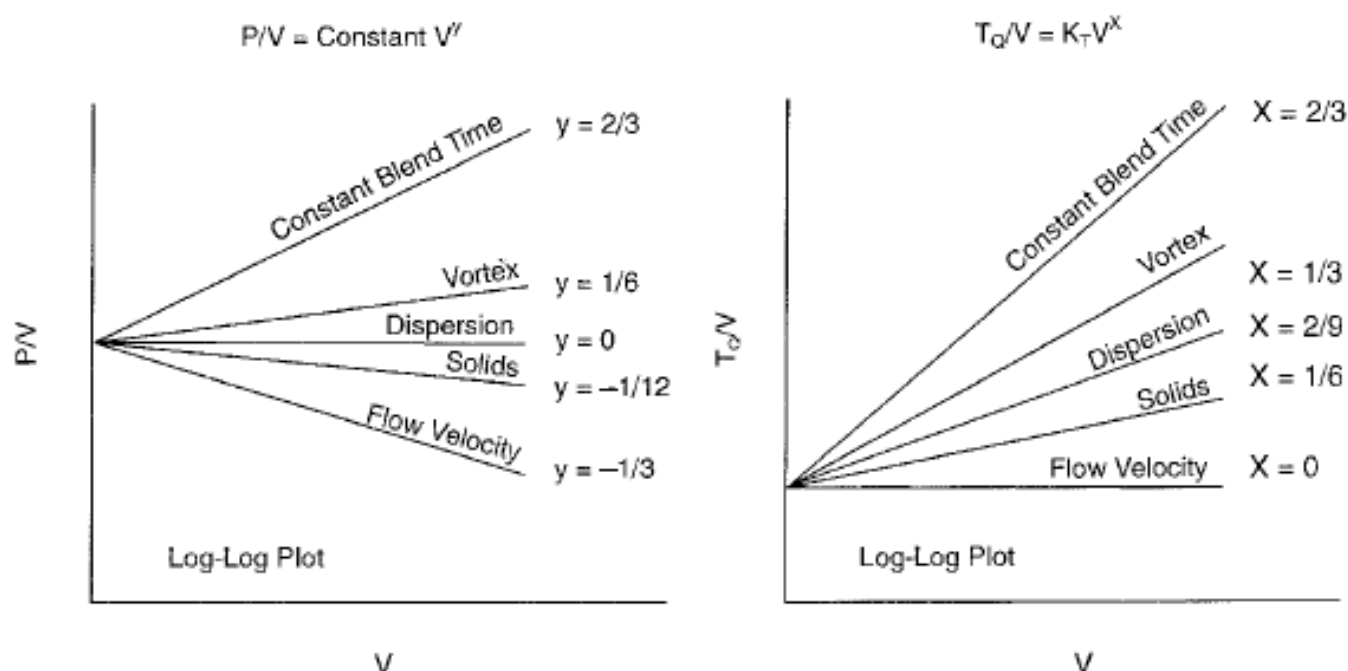
Fig. 1 Scale up method for different types and requirement

Figure 6-28 shows changes in these parameters as the vessel volume is increased for several processes. The exponents x and y in Figure 6-28 should be determined experimentally or verified even for the processes listed in these plots. Some mixing equipment vendors prefer to use the TQ/V criterion because it has a direct impact on the overall size and cost of the mixer, including the gearbox.

When choosing a scale-up method, one must consider changes in other flow and power parameters and their impact on the process result. Table 6-8 shows how these important parameters change on scale-up to 10 times the diameter and 1000 times the volume of laboratory mixing tank. Scale-up methods based on constant blend time require the mixer speed in the commercial vessel to be the same as in the laboratory vessel. This, however, results in a very large increase in the motor power. Such a demanding criterion is necessary for very fast to instantaneous reactions where the reaction lifetime may be a few seconds. Commercial reactors for such systems are, therefore, relatively

small in size. Using constant P/V, the mixer speed decreases by 78%, but the blend time increases by a factor of 4.6. If constant P/V is used in scaling up a reacting system, the reactors may need to be sized for longer residence time than the laboratory reactor because of the increase in blend time.
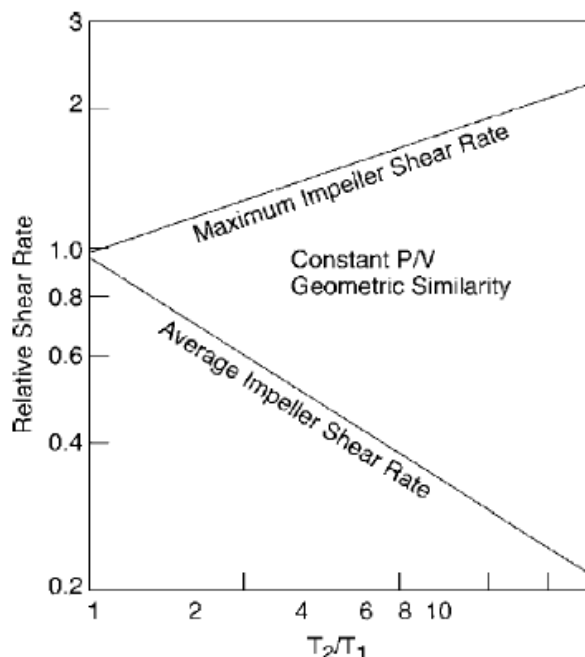
Fig. 2. Changes in maximum and average shear rate on scale-up.

It should be noted that the Reynolds number increases by a factor of 21.5, and therefore, the flow regime may significantly change and affect the mixing quality. Also, the Weber (We) number increases by a factor of 48.4, which may decrease the dispersed phase drop size on scale-up of an immiscible liquid system. Constant tip speed and equal TQ/V are some other scale-up criteria and are used only when flow velocities in the impeller region need to be the same as in the laboratory tank.

Table 1 Most Important Changes in Mixing Parameters on Scale-up by a Factor of 10 in Diameter and 1000 in Volume for Geometrically Similar Systems

Quantity	N	Q/V	Tip Speed	Re	T_Q/V	We	P/V	P
Changes	1	1	10	100	100	1000	100	10^5
in	0.1	0.1	1	10	1	10	0.1	100
parameters	0.22	0.22	2.2	21.5	4.8	48.4	1	1000

It must be recognized that rotational speed and shear rate change significantly on scale-up at constant P/V.

SCALE UP OF SOLID-LIQUID MIXING

Scale-up is an effort to understand the fundamental phenomena occurring in a process in order to predict the performance in larger scale equipment. It begins with process research at the bench scale, often in small glassware, through pilot scale studies to full production. The value of scale-up is captured in the following comment attributed to L. H. Baekland, the father of plastics: "Commit your blunders on a small scale and make your profits on a large scale."

In solid–liquid mixing applications, the purpose of scale-up is to determine the operating conditions at different scales at which mixing yields equivalent process results. The tasks involve:

a. Definition of the appropriate desired process result, such as level of uniformity of the solid distribution in a vessel, the time to achieve complete dissolution, the rate of reaction between a solid and a liquid reactant, and so on.

b. Developing reliable correlations that describe the effects of key process properties, mixer design, and operating variables on the desired process result by either experimentation or mathematical analysis of the physicochemical phenomena

c. Determining and confirming the key controlling physicochemical phenomena and the associated correlating parameters, preferably in dimensionless form

d. Applying the key correlations to predict the process performance at different scales

SCALE UP OF GAS-LIQUID MIXING

In the example a gas–liquid reaction with particulate solids (e.g., a catalyst) operating in regime II in a stirred reactor with a Rushton turbine is to be scaled up. The primary process requirement is for the same degree of reaction conversion at each scale, which means the same number of moles of gas transferred per mole of liquid fed:

$$\frac{k_L a V (C_A^* - C_A)}{C_{LBfeed} Q_L} = \text{constant}$$

Assume for simplicity that $CA = 0$ (a good approximation for regime II) and that the degree of gas backmixing is the same at all scales (this should be checked at the end of the calculation and reiterations performed if necessary). Given a constant feed concentration at all scales,

$$k_L a V C_L \propto Q_L$$

Sometimes it is necessary for the outlet gas concentration to be constant (e.g., with hazardous gases); then from the mass balance this becomes

$$k_L a V \propto Q_G$$

Substituting a suitable correlation for $k_L a$, for example

$$k_L a \propto \left(\frac{P}{V} \right)^{0.7} (v_S)^{0.6}$$

and a curve fit for the gassed power curve, such as

$$P = N^{3.3} D^{6.3} Q_G^{-0.4}$$

not necessarily reliable! an expression such as

$$N^{3.4} T^{6.0} \propto Q_L$$

Will Result.

Another constraint will then fix the design. In this example maintaining $N > N_{JS}$ for the suspension of the catalyst particles is important, so $NT^{0.76} = \text{constant}$ could be added (although not strictly applicable to gassed systems), giving

$$Q_G \propto Q_L \propto T^{3.4}$$

This scale-up method has the effects, on increasing the scale, of:
• Increasing vS, so foaming and entrainment become more likely
• Decreasing P/V
• Decreasing the heat transfer flux per unit throughput
• Nearer approach to poor gas dispersion

• Longer liquid mixing time

SCALE UP OF LIQUID-LIQUID MIXING

Scale-up of agitated immiscible liquid–liquid systems can be a challenge that should not be taken lightly. The problems arise from incomplete or inaccurate process information and few quantitative tools to deal with complex technology. In this section we describe some proven practices for scale-up and caution that liquid–liquid dispersion technology is highly system specific.

Most problems are not observed in glass bench scale equipment because unrealistically high rates of circulation mask coalescence and suspension problems. These problems usually surface at the time of scale-up. Throughout this chapter it has been emphasized that production scale vessels are dominated by coalescence, whereas small vessels are dominated by dispersion. As discussed previously, Sprow (1967b) worked with a coalescing system in a small bench scale vessel and found that different regions of the vessel responded differently to agitation. The technology to cope with these complex issues lags other mixing operations, such as blending and solids suspension. Often, all three flow dependent phenomena—dispersion, coalescence, and drop suspension—must be dealt with simultaneously.

A successful scale-up does not mean that identical results are obtained at two different scales, but rather, that the scale-up results are predictable and acceptable. Problem correction at large scale is costly, time consuming, and sometimes not possible (see Section 12-9.2.2). Scale-up errors can lead to losses in capacity, quality, safety, and profits. For example, an explosion resulted from increasing agitation for an inadequately suspended mixed acid nitration. Faster agitation created a large increase in interfacial area at reaction temperatures and led to an uncontrolled exothermic reaction and property loss.

The scale-up of certain liquid–liquid processes can be straightforward. Dilute dispersions are the easiest processes to scale up. The most difficult ones involve simultaneous coalescence, dispersion, suspension, mass transfer, and chemical reaction. If multiple complex reactions are involved, inadequate mixing often leads to yield losses.

The first step is to understand the goals of the process and to acquire accurate data for all components, including physical, chemical, and interfacial properties as well as reaction kinetics. This also includes the influence of minor impurities. Differences in the quality of raw materials need to be considered.

It is important to undertake bench scale studies that simulate the poorer mixing conditions in the larger vessel. For example, simulate the large scale vessel circulation time. Although dispersion is apt to be unrealistic, coalescence and settling problems can be observed. Examination of the flow patterns in the proposed full scale vessel using CFD can help visualize potential problems related to design.

Once the CFD model has been developed and validated, design and operating parameters can be compared to determine design sensitivities. One observation seems to hold universally—better results are always obtained in small equipment.

Identify applications by types likely to cause problems, and separate these from more trivial applications.

For example, mixing is critical in the following applications:
• Chemical reactors/polymerizers in which reaction rates are equal to, or faster than, mixing rates
• Competing chemical reactions when yields depend on good mixing
• Mass transfer dependent reactions involving coalescence and dispersion

Less demanding tasks include:
• Heat transfer
• Reactors involved with slow chemical reactions

Scale-up Rules for Dilute Systems

Many processes have been scaled successfully using ND^X = constant. This simple rule is based on years of industrial experience. To apply it, the tank Reynolds number must be greater than 10^4 and vessels must be geometrically similar. Table 2 lists the rule and the application best suited to the rule. Other operations, such as blending and solids suspension, are included to provide the reader

Table 2   Table 12-7   "Rules" for Scale-up of Geometrically Similar Vessels at Turbulent Based on ND^X = Constant

Value of X	Rule	Process Application
1.0	Constant tip speed, constant torque/volume	Same maximum shear; simple blending; shear-controlled drop size.
0.85	Off-bottom solids suspension	Used in Zwietering equation for N_{js}, for easily suspended solids; also applies to drop suspension (see Section 12-6.2).
0.75	Conditions for average suspension	Used for applications of average suspension difficulty.
0.67	Constant P/V	Used for turbulent drop dispersion; fast settling solids; reactions requiring micromixing; gas–liquid applications at constant mass transfer rate.
0.5	Constant Reynolds number	Similar heat transfer from jacket walls; equal viscous/inertial forces.
0.0	Constant speed	Equal mixing time; fast/competing reactions.

with an overview of how the exponent on impeller diameter varies from operation to operation. One can see from the table that different scale-up rules apply for suspension, dispersion, heat transfer, and reaction, making it necessary to focus on the most important or limiting task. As mentioned earlier, the indiscriminate use of rules can lead to problems.

Scale-up of Concentrated, Non-coalescing Dispersions

Dilute, low viscosity dispersions are nearly always controlled by turbulence. At high dispersed phase concentrations, small scale turbulent eddies are damped out by the drops and bulk viscosity increases. As a result, laminar shear forces can control drop dispersion in concentrated systems. Turbulence theories developed for dilute dispersions can sometimes apply to concentrated, noncoalescing systems. However, in other cases, they may not. This is illustrated, by example, below for the scale-up of a suspension polymerization application, described by Leng and Quarderer (1982).

The system consisted of free radical initiated styrene–divinylbenzene monomers dispersed in water containing 0.2% dissolved polyvinyl alcohol. The dispersed phase was 50 vol %. The process was to be carried out in a vessel containing a loop impeller operating at low-shear conditions. Bench scale studies showed important variables to be speed, impeller diameter, baffling, selection of the suspending agent, and continuous phase viscosity. Polymerization reactions were completed and bead size distributions were determined by sieve analysis.

Theories based on laminar and turbulent dispersion conditions were developed, and tested by comparing bead size against each specific variable. Results showed that beads of size greater than 300 μm were formed under laminar shear-controlled conditions, and smaller beads were formed under turbulence controlled conditions.

Leng and Quarderer (1982) reasoned that dispersion occurred in the boundary layer adjacent to the loop impeller surfaces and that the impeller vertical elements could be approximated by cylinders moving through the suspension at the relative impeller tip speed. When laminar shear forces predominated, it was shown that

$$d_{max} = C_{28}\sigma \left(\frac{D_C}{\mu_c \rho_c}\right)^{1/2} \frac{1}{[ND(1-k_v)]^{3/2}} \left(\frac{(\mu_d/\mu_c)+1}{1.19(\mu_d/\mu_c)+1}\right) f(\mu_d/\mu_c)$$

where DC is the diameter of the cylinder and kv is the ratio of the tangential velocity at the impeller tip to the tip speed. All other variables follow earlier use.

The equation for turbulent dispersion was based on the classical development of Chen and Middleman (1967) (see Section 12-2), with the energy dissipation term calculated for drag on a cylinder. Two cases were assumed for the dissipation volume in the wake region behind the cylindrical impeller blade. The first was that an eddy length proportional to the cylinder diameter determined the dissipation volume. The second was that this volume was proportional to the velocity of the cylinder (tip speed) and a characteristic eddy decay time. Equation (12-74) results from the second case. It showed reasonable agreement with data taken at higher speeds.

Design Calculation

The conditions in the large vessel are close as possible to the pilot scale/lab scale unit

Criterion (i) Constant Mixing Time

When the volume of the vessel is increased the length of the flow path for bulk also increases. To keep mixing time constant, the velocity of the fluid in the larger tank should be increased in proportion to the size. Power input per unit volume is proportional to the square of the velocity. So large power is needed to maintain constant mixing time and so this is not feasible. So this criterion for scale up cannot be used.

Criterion (ii): Constant power input per Unit Volume

$$P = \frac{N_P n^3 D_a^5 \rho}{g_c} \qquad V = \frac{\pi}{4} D_t^2 H \qquad \frac{P}{V} = \frac{N_p \rho n^3 D_a^5}{\frac{\pi}{4} D_t^2 H}$$

$$D_t = \phi_1 D_a \qquad H = \phi_2 D_t \qquad H = \phi_2 \phi_1 D_a$$

$$\frac{D_t}{D_a} = \phi_1 \qquad \frac{H}{D_t} = \phi_2 \qquad \frac{P}{V} = \frac{N_p \rho n^3 D_a^2}{\frac{\pi}{4} \phi_1^3 \phi_2}$$

$$\frac{P}{V} = \frac{N_p \rho n^3 D_a^5}{\frac{\pi}{4} \phi_1^2 D_a^2 \phi_2 \phi_1 D_a} \qquad \frac{P}{V} = \frac{N_p \rho n^3 D_a^5}{\frac{\pi}{4} \phi_1^3 \phi_2 D_a^3}$$

$$\frac{P}{V} = Kn^3 D_a{}^2 \qquad \left(\frac{P}{V}\right)_{plant} = Kn_{plant}{}^3 D_{plant}{}^2 \qquad \left(\frac{P}{V}\right)_{lab} = Kn_{lab}{}^3 D_{lab}{}^2$$

If we will say $\left(\dfrac{P}{V}\right)_{lab} = \left(\dfrac{P}{V}\right)_{plant}$ then

$$Kn_{lab}{}^3 D_{lab}{}^2 = Kn_{plant}{}^3 D_{plant}{}^2$$

$$\frac{n_{lab}}{n_{plant}} = \left(\frac{D_{aplant}}{Da_{lab}}\right)^{2/3}$$

Criterion (iii) Same impeller tip speed
- nt_m- dimensionless no. represents the number of stirrer rotations required to homogenize the liquid. At high Re, $n_i t_m$ is independent of Re.

$$nt_m = \frac{1.54V}{D_a{}^3}$$

Problems:

1. A fermentation liquid of viscosity 0.1 poise and density of volume 2.7 m³ using Rushton turbine impeller with 1000 kg/m³ is agitated in a baffled than Estimate the mixing time for a dia of 0.5 m and stirring at a sped of 600 rpm. Estimate the mixing time.

$$n_i t_m = \frac{1.54V}{D_a{}^3} \qquad t_m = \frac{1.54V}{n_i D_a{}^3}$$

$$t_m = \frac{1.54 \times 2.7}{600 \times (0.5)^3} = 0.055\,\text{min} = 3.32s$$

2. A pilot plant vessel 1 ft (305 mm) in dia is agitated by a six blade turbine impeller 4 in (102 mm) in dia. When the impeller Re no. is 10000, the blending time of two immiscible liquids is found to be 15 s. The power required is 2 Hp/1000 gal (0.4 kW/m³. (a) What power input is

required to give the same blending time in a vessel 6 ft (1830 mm) in dia (b) What would be the blending time in the 6 ft (1830 mm) vessel if the power input per unit volume were the same as in the pilot plant vessel?

$$N_P = K_T$$

$$P = \frac{K_T n^3 D_a^5 \rho}{g_c}$$

At high Re no's

For a given density

$$\frac{P}{D_a^3} = K' n^3 Da^2 \qquad \frac{P}{V} \propto n^3 Da^2$$

$n_a = n_b$ (given)

The ratio of the power inputs per unit volume in the two vessels are

$$\frac{P_b / V_b}{P_a / V_a} = \frac{n_b^3 D_{ab}^2}{n_a^3 D_{aa}^2} = \left(\frac{D_{ab}}{D_{aa}}\right)^2 = \left(\frac{6}{1}\right)^2 = 36$$

$$\frac{P_b}{V_b} = 36 \frac{P_a}{V_a} = 36 \times 0.4 = 14.4 kW/m^3 \qquad \text{(For constant blending time)}$$

(b) If the power input per unit volume is same then

$$\frac{P_a}{V_a} \propto n_a^3 D_a^2 \qquad \frac{P_b}{V_b} \propto n_b^3 D_b^2$$

$$\frac{P_a}{V_a} = \frac{P_b}{V_b}, then\ n_a^3 D_a^2 = n_b^3 D_a^2$$

$$\frac{n_b}{n_a} = \left(\frac{D_a}{D_b}\right)^{2/3} = \left(\frac{6}{1}\right)^{2/3} = 3.3$$

The blending time in 6 ft vessel $= 3.3 \times 15s = 49.5s$

$$\mathrm{Re} = \frac{D_a{}^2 n\rho}{\mu} = \frac{0.82 \times 90 \times 1120}{60 \times 120} = 8.96 \angle 10$$

$$(So \quad la\min ar)$$

$$N_p \text{ x Re} = 65; \; N_p = 65 \, /8.96 = 7.25$$

3.
$$P = N_{P.} \, N^3 \, Da^5 \, \rho = 8980 \; W = 12hP$$

A vertical tank 2.4 m dia is provided with a flat blade turbine impeller (6 blades) mounted centrally in the tank at a height of 0.8 m from bottom. The turbine is 0.8 m in dia and the blades are 167 mm wide. The tank is filled to a depth of 2.4 m With rubber latex compound having density 1120 kg/m^3 and viscosity 120 kg/m.s. If the tank is baffled and turbine is Rotated at 90 rpm, what is the power consumption in hP? Take N_p. Re = 65 for laminar flow and 5.75 for turbulent flow.

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Faculty Profile:

Dr. P.A. Joshi

Dr.P.A.Joshi is Chairman Anchor Institute Chemical & Petrochemical sector DDU and Professor of Chemical Engineering and Dean, Faculty of Technology, Dharmsinh Desai University, Nadiad (GUJARAT). He received a bachelor's degree in chemical engineering from Gujarat University, M.Chem.Engg. from UDCT, Mumbai, Ph.D. from IIT Bombay, FDPM from IIM Ahmedabad and a course on Cleaner Production from University of Adelide, Australia. He had worked as PG Coordinator in Institute of Technology, Nirma University of Science and Technology, Ahmedabad, Professor and Head, Chemical Engineering Department, Government Engineering College, Gandhinagar and Research Engineer with M/s. Standard Alkali, Thane, a Mafatlal Group Company. He has 30 years of UG and PG teaching, research, administrative and consultancy experience and has interest in energy and environment, fluidization engineering and mathematical modeling. He has visited several universities in Australia and USA.

Dr Joshi has about 30 papers to his credit and delivered several invited lectures in various institutes. He has organized more than 20 national level seminars/ workshops and conferences and two continuing education programs sponsored by ISTE. He has been technical advisor/ consultant to three chemical industries and has designed effluent treatment plants for dyestuff industries and an organic chemical manufacturing unit. He was involved in conducting environmental audit (EA) for about 15 industries in Gujarat and currently is associated with EA Cell of DD University as Technical Advisor. The cell has generated revenue of nearly 15 million rupees in last five years and conducts audit of IPCL, KRIBHCO, GSFC, GNFC, Lupin and many other renowned companies. He has supervised 12 PG dissertations and several UG projects.

Prof H R Shah

Coordinator, Anchor Institute-Chemicals & Petrochemicals

Prof Harshad R shah(H R Shah) after graduating in Chemical Engineering from Dharmsinh Desai Institute of Technology(TO day's faculty of Technlogy, Dharmsinh Desai University) Nadiad in 1974 joined L D College of Engineering Ahmedabad. He completed his M. Tech in Process Engineering from Indian Institute of Technology Delhi.

He remained with Chemical Engineering Department of L D College of Engineering for about 22 years   He served Chemical Engineering Department of Nirma University at Degree and Diploma level for nearly 10 years. During his academic career Prof Shah actively participated in Department Development and Institute building activities.   He was also member of board of Studies of Gujarat University, Nirma University. He was member of several Institute , State and Central government committees . He supervised   around 10 M.E./ M. Tech dissertations and conducted examinations of various universities in the state OF GUJARAT. He was invited to give expert lectures by various industries like IFFCo Kalol Unit, Indian Rayon Veraval and leading Chemical engineering Departments of the Gujarat state. At present Prof Shah is coordinator of Anchor Institute- Chemicals and Petrochemicals promoted by Industries Commissionerate in Partnership with Dharmsinh Desai University and over responsible for its functioning.

Prof Shah is life member of Indian Institute of chemical engineers and Indian society for Technical Education and involved in their various activities

Prof. D. J. Vyas

Prof. D. J Vyas is a Senior Faculty Member in the Department of Chemical Engineering at Dharmsinh Desai University Nadiad since last 25 years. He has obtained his B.E. (Chem.Engg.) & M.E. (Chem.Engg.) from Gujarat University & pursuing his Ph.D. from D D University. He is a member of Board of Studies, examiner and paper setter of various Universities like Nirma, Gujarat, Sardar Patel, Saurashtra, M.S., and DDU etc.

Recently he has been awarded as the *BEST TEACHER AWARD* from Indian Society for Technical Education - New Delhi. He has organized many National & Inter-national workshop & conferences like on Disaster Management, Pollution Control & Safety in Chemical Process Industries, Nanotechnogy from Chemical Engineering Perspectives etc. in associations with Institution of Engineers (India), Vatva Industries Association-Ahmedabad & ISTE under the banner of D. D. University. Many Technical Proceedings of conferences, presentations & publications of technical papers are credited to him. He has served as a resource person for Vidhya Dairy – baby Amul. He was also invited as a speaker for many seminars on different topics by various Technical Institutions, Engg. Colleges, Polytechnics & NGO like Lions Club, Rotary Club etc.

He is associated with many professional organizations as Life Member like ISTE, IIChE, DDUAA, IE (India) etc. He is also working as Secretary of ISTE chapter Nadiad & credited Three ISTE Best Chapter Awards in the last five years. He is also working as a Nodal Officer for Admission Committee for Professional Courses - Gujarat State. He has coordinated many institutional activities and represented at various level.

Porf. Anand P.Dhanwani

Porf. Dhanwani is working as an associate professor in Department of chemical Engineerign, FOT, DDU, Nadiad since last 15years.

He did his graduation in Chemical Engineering from L.D.Engineering College, Ahmedabad in 1993 and his M.E.Chemical from D.D.University in 2002.

He also has 3years of industrial experience.

He has 2 publications and 10 national and international presentations to his credit.

Details of Research Interest

OPTIMIZATION:
Scheduling: In multi product and multipurpose batch, semi continuous and continuous plants, different products are manufactured via the same or different sequence of operations by sharing available pieces of equipment, intermediate materials and other production resources. Low volume, high value products with changing demand pattern requires inherent operational flexibility. We are trying to optimize the recourses and infrastructure through optimal scheduling.
HEAT TRANSFER:
Batch heat transfer cooling of Newtonian and Non-Newtonian fluids using different geometry like vertical tube coil and helical coil in agitated vessel to study heat transfer and developed the correlation for evaluating overall heat transfer coefficient.

Prof. Mihir P. Shah

Mihir Shah is working as an Assistant Professor in Department of Chemical Engineering since last 10years.

He did his B.E. Chemical in 1996 and M.E. Chemical in 2006 from Department of Chemical Engineering, D.D.University, Nadiad.

He has worked on Separation of Pollutants and Metal Ions from aqueous phase using emulsion liquid membranes. He has also completed a GUJCOST (Gujarat Council on Science and Technology, A State Govt. Agency) sponsored project in the same field.

Currently he is working on simulation of Chemical Processes using ASPEN Plus and MATLAB.

He has 4 national presentations and 1 international presentation to his credit.

Prof. (Dr.) Shirish L. Shah, University of Alberta, Canada, Prof. D. G. Panchal, Dean, FOT, & Prof. H. R. Shah AI Lighting the Lamp at the inaugural function of APC Module II at DDU, Nov 21-26,2010

Dr. Kannan Moudgalya, Head CEP IITB taking class on SCILAB At DDU Training Course Nov 09-13, 2009

The Participant and Faculty at Anchor Institute DDU course at PI Industries, Panoli in June, 2010.

Dr. S. Ganeshan G.M. Project In charge (IRUP), Toyo Eng. India Ltd. Taking class at CEP-IITB Course June 08-12, 2010

Dr. P. A. Joshi, Dean, FOT, DDU & Chairman, AI addressing the Participants at Ahmedabad Training Course Feb 12-13, 2009

Faculty and Participant at Lead Auditors course on ISO 14001-2004 Environment Management System June 25-30, 2012

Dharmasinh Desai University (DDU):

Dharmsinh Desai Foundation was established at Nadiad in Gujarat, by an eminent Parliamentarian and a social worker, *Late Shri Dharmsinh Desai*. From an affiliated College, started in 1968 as *Dharmsinh Desai Institute of Technology (DDIT),* offering Degree and Diploma in Chemical Engineering, *DDU* has now become a trusted name amongst a variety of stake holders, namely, students, their parents, researchers, academicians, employers, other academic institutions offering higher level education, National level Institutions and State & Central Government Agencies. Some of the salient features are as under.

- ISO 9001:2008 certification since past eight years.
- Member of the Association of Indian Universities(AIU) and Association of Commonwealth Universities(ACU)
- Mission: to undertake programs and projects for development of human resources, both through formal and non-formal delivery systems, in areas of professional pursuits in all walks of human endeavors, with accent on relevance, value addition, societal needs and futuristic pilot projects.
- It's Faculty of Technology, Faculty of Pharmacy, Faculty of Dental Science, Faculty of Management and Information system and Faculty of Commerce Caters the need of Doctoral, Post Graduate under Graduate and Diploma, Level Professional Education.   The university will add Faculty of Medical Science in near future.

Faculty of Technology (FOT):

Faculty of Technology offers various Doctoral, Graduate Under Graduate and Diploma, level Programmes in Engineering & Technology. Some of the salient features and achievements of *FOT* are as under

- All Engineering Programmes are accredited by National Board of Accreditation (NBA) for five years.
- ISO 9001-2000 certification
- Well equipped Environmental Audit and Consultancy Service Cell. Some of the major clients are RELIANCE, IPCL, GNFC, GEB, L&T, Narmada Cement, NCPL, BASF, Bayer, Rallis India Ltd, Sun Pharma, Zydus Cadila, Birla Cellulosic, Cadila Pharma, GAIL India Ltd. Kanoria Chemicals Wockhardt, Tata Chemicals etc.
- MOU with University of IOWA, USA for 5 years BE +MS Programme and KHS Germany for industrial training.
- The total grant   Rs. 15.6757 crores have been received /approved / sanctioned by various Departments of Government of Gujarat, Government of India and sponsored projects from Industries in 2009 -10.

Chemical Engineering Department:

- Established in 1968, Department of Chemical Engineering, Faculty of Technology, D.D. University, is second oldest in Gujarat offering Chemical Engineering courses at Diploma, Degree, Postgraduate and Ph.D. level. Its Salient Features are:
- Highly qualified and experienced faculty members.
- Accredited by National Board of Accreditation (NBA) for a period of five years.
- More than 3000 students have passed out and many of them have occupied highest positions in the Industries/Academia in India and in Abroad.
- D.D. University is recognized as State level Anchor institute inn Chemical & Petrochemical Sector due to its strength in Chemical Engineering faculty.
- Well-equipped laboratories with latest analytical Instruments.
- Offers four Graduate level programs Viz General Chemical Engineering, CAD & Control, Environment Engineering, Bio Technology and Nano Technology.
- The Shah-Schulman Centre for Surface Science and Nanotechnology headed by a world known scientist, Dr. Dinesh Shah of University of Florida.