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MIXING AND BLENDING

Mixing, an important operation in the chemical process industries, can be divided into five areas: liquid–solid dispersion, gas–liquid dispersion, liquid–liquid dispersion, the blending of miscible liquids, and the production of fluid motion. Mixing performance is evaluated by two criteria. The first is physical uniformity, ie, a physical relationship is required in terms of samples of uniformity in various parts of the mixing vessel. Specifications describe this requirement. The other criterion is based on mass transfer or chemical reaction. The elements of mixer design are: (*1*) process design—fluid mechanics of impellers, fluid regime required by process, scale-up, and hydraulic similarity; (*2*) impeller power characteristics—relate impeller power, speed, and diameter; and (*3*) mechanical design—impellers, shafts, and drive assembly. The curves in Figure 1 give data for a wide variety of impeller types and systems (1). The power consumption curves shown in Figure 1 are completely independent of process performance.

Homogeneous chemical reactions require a knowledge of the overall concentration in the tank, primarily on a microscale mixing level. A study showed that the root-mean-square (RMS) velocity fluctuation value correlated well with the direction in competitive consecutive second-order reactions. This indicates that this type of microscale mixing can affect certain kinds of chemical reaction (2). The speed of a chemical reaction is related to the blend time on both macro- and microscale levels (3–4). Thus, mixing may affect the progress of a homogeneous chemical reaction.

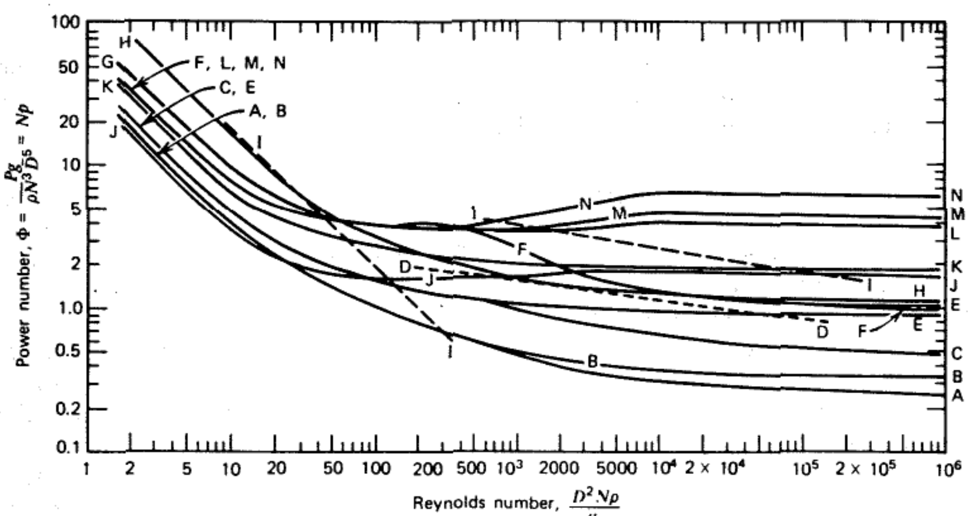

Figure 1. Power characteristics of mixing impellers. When no baffles (for N_{Re} over 300), $\Phi = (P_g/\rho N^3 D^6)(g/N^2 D)(a - \log N_{Re}/b)$. A, propeller, pitch equal to diameter, no baffles; B, propeller, pitch equal to diameter, four baffles each 0.1 T; C, propeller, pitch twice the diameter, no baffles; D, paddle, no baffles; E, propeller, pitch twice the diameter, four baffles each 0.1 T; F, flat six-blade turbine, no baffles; G, shrouded six-blade turbine, four baffles each 0.1 T; H, shrouded six-blade turbine, stator ring with 20 blades; I, paddle, no baffles; J, flat paddle, two-blade, four baffles each 0.1 T; K, fan turbine, eight-blade, four baffles each 0.1 T; L, arrowhead six-blade turbine, four baffles each 0.1 T; M, curved six-blade turbine, four baffles each 0.1 T; and N, flat six-blade turbine, four baffles each 0.1 T.

Fluid Mechanics

Mixer power P produces a pumping capacity Q expressed in kg/s, and a specific velocity work term H expressed in J/kg (5).

$$P = QH$$

The term H is related to the square of the velocity and, therefore, to fluid shear rates. In low and medium viscosity, the pumping capacity is related to speed and diameter of the impeller (6):

$$Q \propto ND^3$$

The power drawn by the impeller is proportional to N^3D^5:

$$(P \propto \rho N^3 D^5)$$

These relationships can be combined to show that at constant power input, the flow-to-velocity work ratio is related to the impeller diameter:

$$(Q/H)_P \propto D^{8/3}$$

This is sometimes expressed as the flow-to-fluid shear ratio which is correct conceptually but is not in terms of the mathematical equation above. This equation also does not hold as a constant on scale-up, and is, therefore, used primarily to evaluate the effect of geometric variables. It does not have a ready evaluation in terms of the actual numbers when comparing large to small tanks.

Most mixing applications are sensitive primarily to fluid-pumping capacity. Thus, if the pumping capacity of different impellers is compared (7) or the overall flow pattern in the tank considered, process results are proportional in the same fashion to the actual circulating capacity of the impeller and in the tank. The equations relate to pumping capacity of the impeller itself and do not include the entrainment provided as the jet from the impeller circulates through the tank.

Most open impellers, propellers, turbines, and axial-flow turbines are normally limited to an upper range of about 0.6–0.7 D/T, because at that point, no further entrainment is possible to provide additional flow in the mixing tank.

If a process is dependent primarily upon pumping capacity, the fluid velocities and the individual shearing rates, both on a macro- and a microscale, are above a certain minimum level to allow other process requirements to proceed unhindered. If the pumping capacity is increased and some of the other velocity and shear rate values are decreased below some minimum, then fluid shear stress enters into the overall design.

Other process applications sensitive to fluid shear rates include fermentation (qv), crystallization (qv), solids dispersion, polymerization, and many others. Shear rates should be considered if they are part of the overall process requirement or mechanism. However, oversimplification of the complex mixing phenomenon involved can result in serious errors in analysis and interpretation. Thus, high pumping capacity and low impeller velocity work can be obtained by large impeller diameters running at slow speeds for a given power level. Low pumping capacities and high impeller velocity work are obtained by running small impellers at high speeds.

In a general way, a minimum circulation rate in the tank sets the entire volume of fluid in motion. This is usually a minimum value to make the processing vessel a full participant in the process. Furthermore, the level of specific impeller velocity work or fluid shear rate which is needed to carry out the blending, dispersion, or diffusion required by the process objective has to be determined.

Most manufacturers supply portable mixers with either direct drive or gear drive. Direct-drive portable mixers have small impellers at high speeds whereas the gear-drive unit has large impellers and low speeds. The former are used for applications that require high shear rates, whereas the latter are used for applications requiring lower shear rates and higher pumping capacity.

Figure 2 shows the velocity profile of the blade of a flat-blade turbine radial-flow impeller. It was determined by measuring the average velocity at a point in a baffled tank (Fig. 3). By taking the slope at any point on this profile, the velocity gradient dv/dy is obtained which is the definition of fluid shear rate. Thus, the maximum shear rate around the impeller zone and the average shear rate can be evaluated (8–9).

The shear rate multiplied by the viscosity gives the shear stress, ie, shear stress = μ (shear rate). The shear stress carries out the dispersion and diffusion required in a process. Even at low viscosities, eg, 1–5 mPa·s (= cP), the shear stress increases five times from the same shear rate around the impeller.

Increasing the viscosity increases shear stress at a given shear rate. Thus, an increase in viscosity might improve the mixing performance. This is correct in terms of the actual shear stress produced but the material has to circulate through the shear zone and throughout the entire tank. Increasing the viscosity may introduce additional requirements for pumping capacity, blending, and other variables in the overall tank, hence the net result of the total energy required in the system depends on the viscosity. However, it is easier to disperse materials into a high viscosity liquid since the impeller has the ability to generate high shear stresses, assuming that the material is being circulated through the impeller shear zone.

A high frequency-response velocity probe that could pick the instantaneous velocity fluctuations in low viscosity turbulent-flow systems (Fig. 4) would give a plot as shown in Figure 5. This plot permits the calculation of the RMS velocity fluctuation which is a measure of microscale-mixing shear rates. Thus, in a mixing tank large shear stresses are generated by the average velocities at a point and the resulting shear stress (10) whereas macroscale mixing is generated by the existing high frequency turbulent fluctuations.

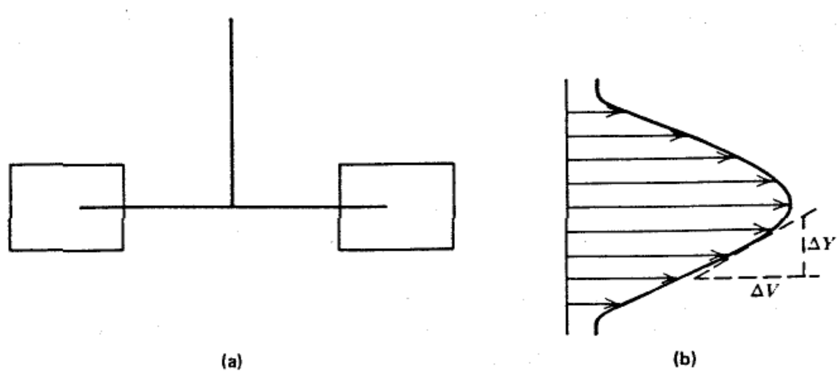

Figure 2. Velocity discharge from (a) a radial-flow impeller and (b) the definition of fluid shear rate. Shear rate = $\Delta V/\Delta Y$.

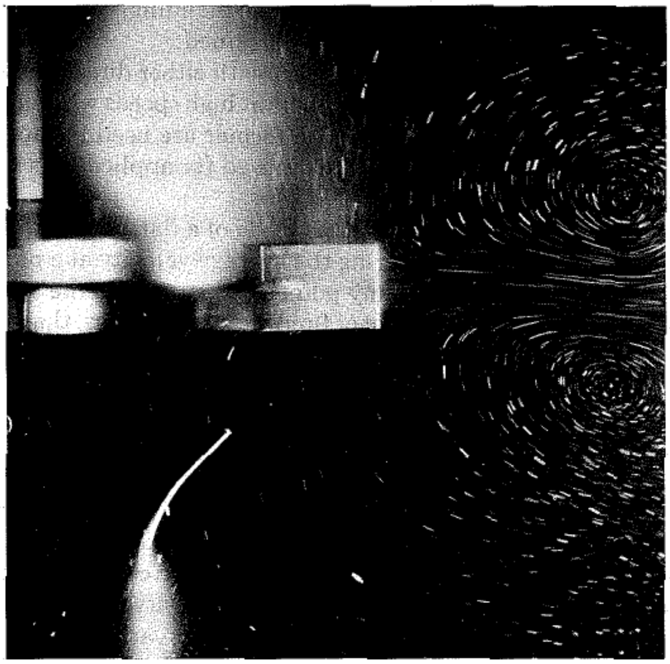

Figure 3. Radial-flow impeller in a baffled tank in the laminar region. A thin plane of light is passed through the center of the tank.

To evaluate the effect of macroscale shear rates, the average point velocity is used. Four values give a good approximation to describe the shear rate profile, maximum and average impeller-zone shear rates, and average and minimum tank-zone shear rates.

Microscale shear rates are used as velocity fluctuations at a point and can be expressed as $\sqrt{(\mu')^2}$.

The power put into the mixer has to be dissipated as heat through the mechanism of viscous shear—regardless of the viscosity—in order to obtain shear rates small *enough to dissipate their velocity energy in terms of the viscous shear rates. It has been* estimated that the transition size is ca 500 μm between particles subject primarily to macroscale shear rates and microscale shear rates (11–12).

Impellers

Impellers are either radial flow or axial flow (13). Figure 6 illustrates a radial-flow disk turbine. Flow patterns in baffled and unbaffled tanks are shown in Figures 7 and 8.

The flat-blade turbine is normally placed 0.5–1 impeller diameter off bottom and has a coverage of 1–2 impeller diameters. The spacing between multiple impellers is somewhere between 1.5 and 3 impeller diameters.

Axial-flow impellers include the square-pitch marine-type propeller shown in Figure 9 and the axial-flow turbine shown in Figure 10. The former has a variable angle

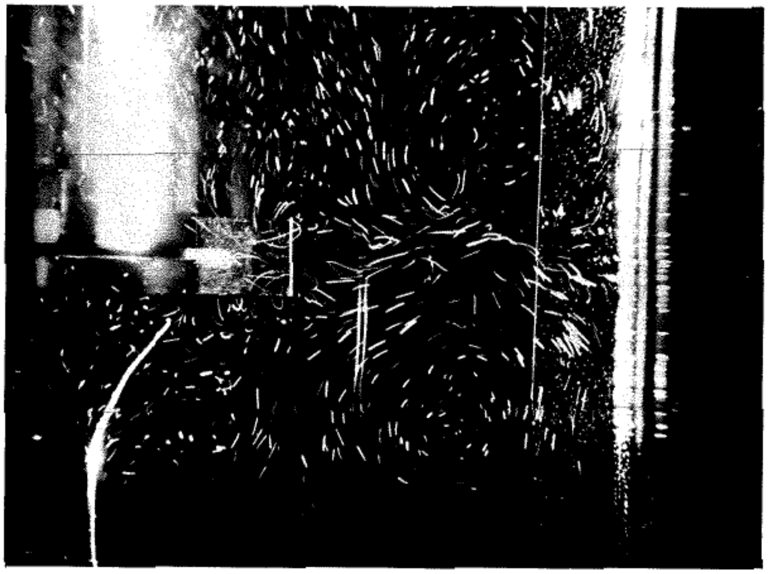

Figure 4. Flow patterns in a mixing tank in the turbulent region. A thin plane of light is passed through the center of the tank.

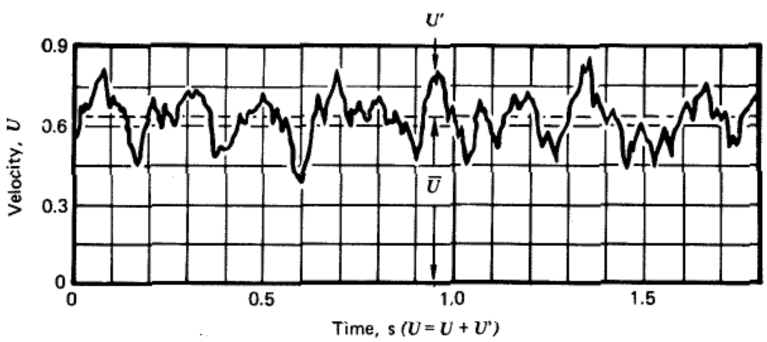

Figure 5. Typical velocity fluctuation pattern obtained from high frequency velocity probe placed at a point in the mixing vessel.

and, therefore, an approximately constant pitch across the impeller face. This gives the most uniform flow pattern across the impeller periphery, and the propeller tends to have the highest pumping capacity per unit power.

The axial-flow turbine has a constant blade angle and, therefore, a variable pitch across the surface. It has an effective flow pattern but is not quite as efficient as the propeller. Axial-flow turbines are used in large-size equipment primarily because of their low cost. The performance of axial-flow impellers in baffled tanks is illustrated in Figure 11. In general, for applications requiring primarily pumping capacity, such as blending and solid suspension, axial-flow turbines are the choice. For applications requiring gas–liquid or liquid–liquid mass transfer, or in multistage columns, radial-flow turbines are preferred.

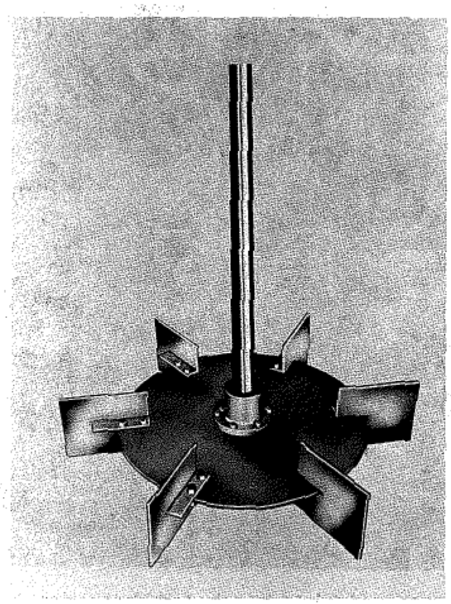

Figure 6. Radial-flow, flat-blade, disk turbine.

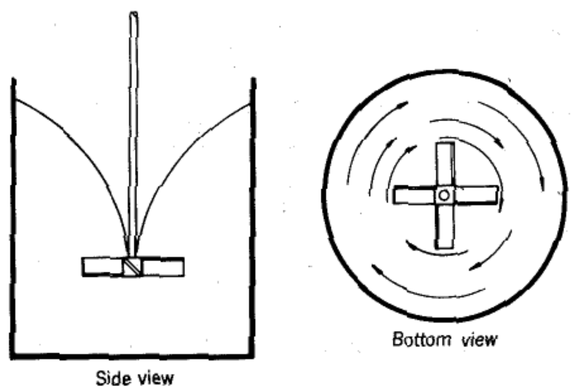

Figure 7. Vortexing flow pattern obtained with any type of unbaffled impeller.

The fluid force acting in a radial direction on the impeller tends to deflect the shaft owing to the many variables of impeller design, tank geometry, and flow patterns. The manufacturer may supply relationship estimates.

Various types of mixer drives are illustrated in Figures 12–15. Portable mixers may have direct or gear drive, usually operated with propellers (Figs. 12–13). They are commonly used in angular, off-center position to achieve a good top-to-bottom flow pattern without the use of tank baffles. The fixed-mounted portable mixer has a mounting base and may either be used in open tanks or in a stuffing box for closed tanks. Since the mounting base is rigid, the shaft's critical speed and length must be designed with care.

For large equipment, heavy-duty fixed-mounted mixer drives are used (top-

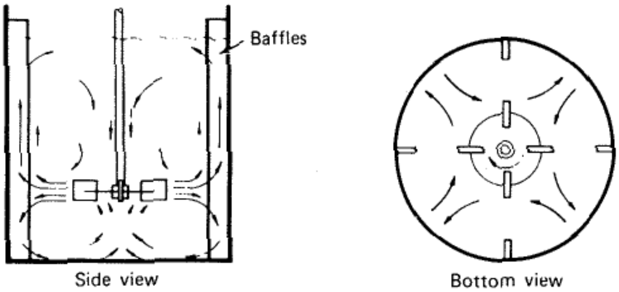

Figure 8. Typical baffled flow pattern of radial-flow impeller.

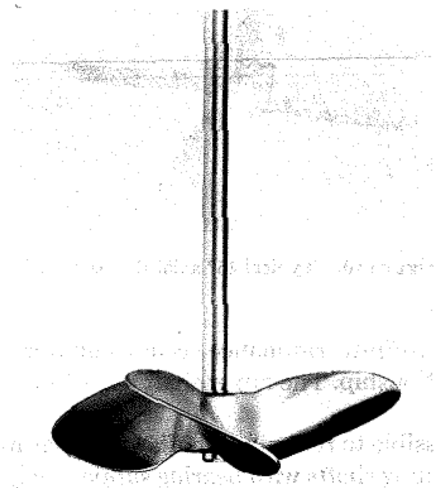

Figure 9. Square-pitch marine-type impeller.

entering, Fig. 14). They must operate below the first critical speed of the mixer shape and most commonly use axial-flow or radial-flow turbines (Figs. 10 and 6, respectively).

The side-entering mixer (Fig. 15) normally uses propellers, has relatively short shafts, and is an extremely cost-efficient device for blending homogeneous fluids. The shaft enters the tank via the mechanical seal or stuffing box. Severe abrasive or corrosive conditions impede operation and cause leakage.

Bottom-entering drives shown in Figure 16 also may be used. The mixer is easily accessible for maintenance but leakage at the seal or stuffing box may have serious consequences. These drives are more suitable for single impellers near the bottom of the tank, since long shaft extensions into the upper part of the tank cause design problems.

In general, large impellers running at slower speeds need less power. However, depending upon the exponential relations of power and D/T, the torque required for the mixer drive typically increases. Torque is the ratio of power to impeller speed. It is the main criterion of cost of the mixer drive assembly. Mixer drives are normally rated on their torque capacity, and output energy is a function of output speed.

A mixer shaft has a natural vibration frequency. When it is rotated at that speed,

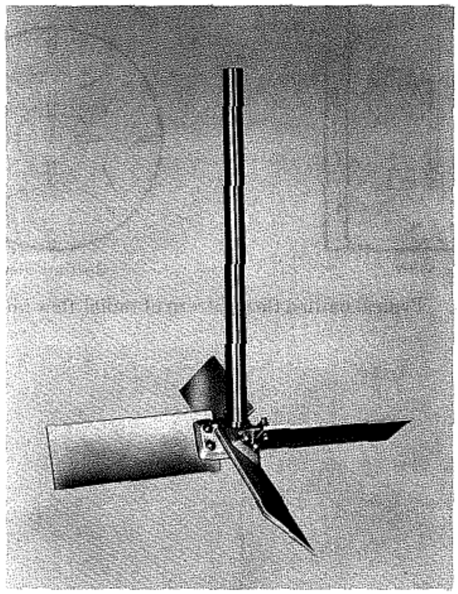

Figure 10. Typical 45° axial-flow turbine.

shaft deflection becomes infinite, resonance occurs, and it may be destroyed. Figure 17 shows the general relationship. The mixer shaft must be operated at a speed below its critical speed.

It is theoretically possible to run mixer shafts above natural frequency, but when combined with long overhung shafts with bearing support only at the upper end, they cannot be used on large equipment.

Portable mixers, such as shown in Figure 12, run above and below the critical speed because of the instability of many portable installations.

Mixing

The mixing requirement may be expressed as pumping or circulating capacity. The impeller flow is defined through the impeller peripheral-discharge area, or total flow in the tank (14). In draft-tube circulators, shown in Figure 18, the impeller is enclosed in a draft tube. Head, flow, and pumping efficiency are measured as in a pump.

In an unbaffled tank (see Fig. 7), a swirl and a vortex are obtained which can be useful for certain kinds of mixing processes. However, the swirl has a tendency to suck gases into the vortex and is troublesome for many mixing applications. Baffles (see Figs. 10–11) give the top-to-bottom flow pattern which is often desired but require more power than the unbaffled tank. If the swirling flow pattern, with or without a detrimental accompanying vortex, is satisfactory for the process, it gives motion in the mixing tank at the lowest power requirement. Baffles give a better overall result, but lead to higher fluid shear rates in the tank. The improvement in process result has to justify the additional energy required.

In square and rectangular tanks, the effect of the corners and the shape provides

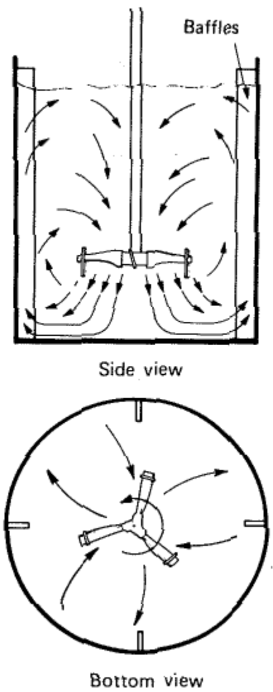

Figure 11. Typical baffled-flow pattern of axial-flow impeller.

a certain amount of baffle action (see Fig. 19). Therefore, at low power levels, eg, in storage, paint-blending, or milk tanks, a satisfactory flow pattern may be obtained without the use of baffles. The shape of the bottom is not of great importance. There are minor differences within the general classification of flat bottom, ASME dish bottom, or shallow cone but they can be handled on an individual mixing-specification basis.

When concerned with elliptical heads, spherical heads, or deep cones, it is necessary to consider the insertion of some kind of baffle in the tank bottom to prevent a residual swirling action.

Power Consumption. The power drawn by an impeller at a given speed and diameter is independent of process performance. The power number N_P is a function of the Reynolds number $ND^2\rho/\mu$:

$$N_P = P/(\rho N^3 D^5)$$

This results in the series of curves shown in Figure 1 which are for standard baffles defined as four baffles each one twelfth of the tank diameter in width.

In turbulent or viscous areas special equations are employed. In the turbulent region, power varies with $\rho N^3 D^5$, whereas in the viscous area, it varies with $\mu N^2 D^3$. Figure 1 gives curves for a variety of different impellers, and interpolation to other impeller shapes can be made readily.

Solid–Liquid Contacting. The settling velocity of solid particles is a critical factor in the process design of many mixing applications. Figure 20 gives the approximate settling velocity for spheres in water. It is, however, desirable to obtain the settling

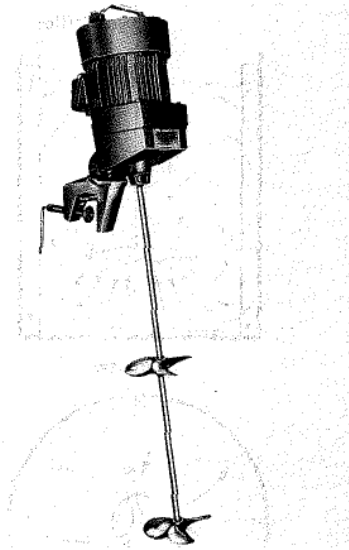

Figure 12. Portable propeller mixer.

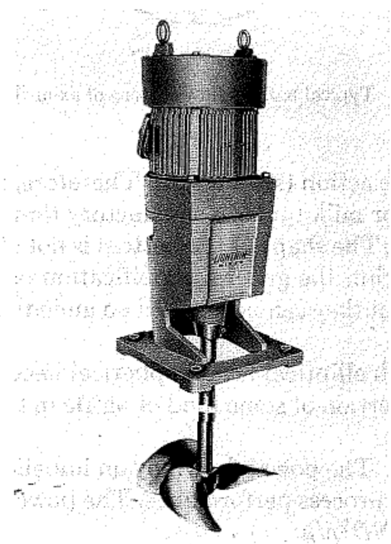

Figure 13. Fixed-mounted propeller mixer.

velocity experimentally by dropping individual particles of various sizes into a graduated cylinder and timing the settling velocities.

In a free-settling system, the settling velocity is above 300 mm/min and is a main factor in the design of the equipment. If the settling velocity is less than 300 mm/min, there is a fair degree of uniformity in the tank, once there is motion.

There are three types of solid suspension (15–21): complete motion on the tank bottom; all the particles have an upward motion (off-bottom suspension); or complete

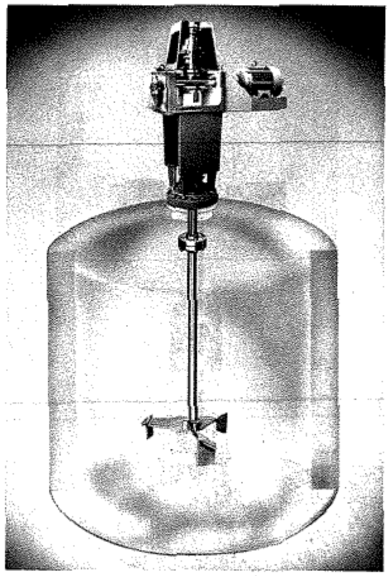

Figure 14. Top-entering drive with axial-flow turbine.

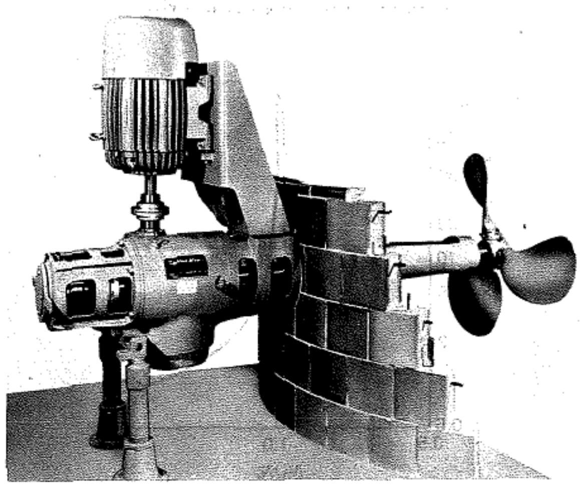

Figure 15. Side-entering propeller mixer.

uniformity. The last is a relative term since the particles have a vertical settling velocity and the top of the tank has a horizontal velocity across the surface. It is, therefore, not possible to obtain complete uniformity in the upper layers of the tank. The relationship between these three types depends upon the settling velocity of the system (see Table 1). High concentrations of solids require increased power. Since the settling velocity of the particle is lower in the hindered settling range, an inverse function is needed to correlate the percentage of solids and the effect of settling velocity on power.

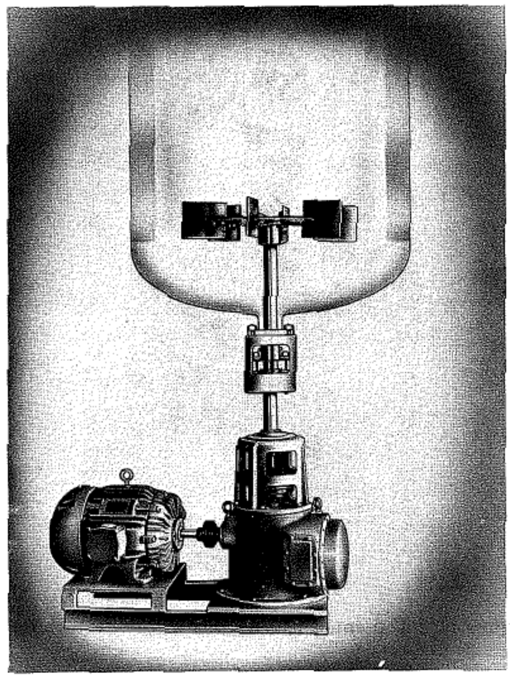

Figure 16. Bottom-entering drive.

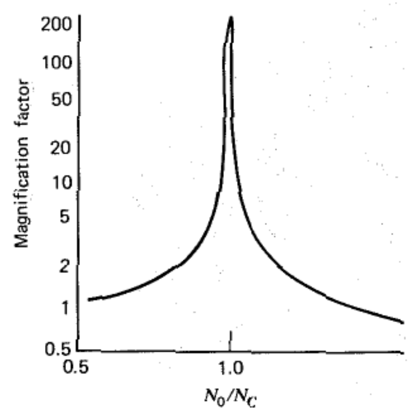

Figure 17. Shaft deflection has a ratio to total indicated run-out, TIR, as a function of the operating speed to critical speed ratio.

Figure 21 gives an approximate selection of the power required for a solid suspension as a function of percent solids and settling velocity of the heaviest particle in the system. This relationship is not suitable when precise design is required for a given application.

The effect of power on solid-liquid mass transfer also can be measured by considering various suspension definitions (22). In the range up to complete off-bottom

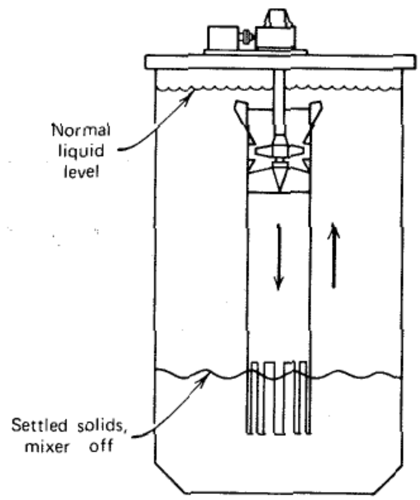

Figure 18. Draft-tube circulator.

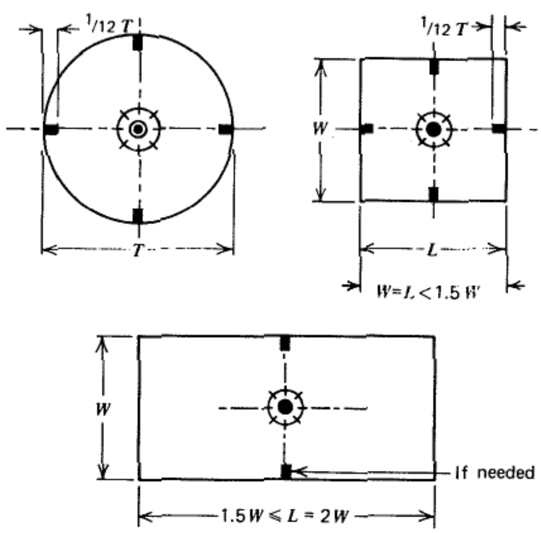

Figure 19. Suggested baffles for square and rectangular tanks.

suspension, the effect of power is typically given by a slope of 0.2 or 0.3 on a logarithmic plot. As shown on Figure 22, above that point, the slope changes to about 0.1. In this range, only the slip velocity of the particles affects the film coefficient. Much greater gains for a given power level are expended up to that point than up to complete uniformity.

In general, axial-flow impellers are superior to radial-flow impellers for solid suspension. However, for gas–liquid–solid systems, the effect of the upward gas velocity on the downward pumping capacity of the axial-flow turbine almost completely neutralizes the impeller's flow pattern, and causes the power consumption for a given

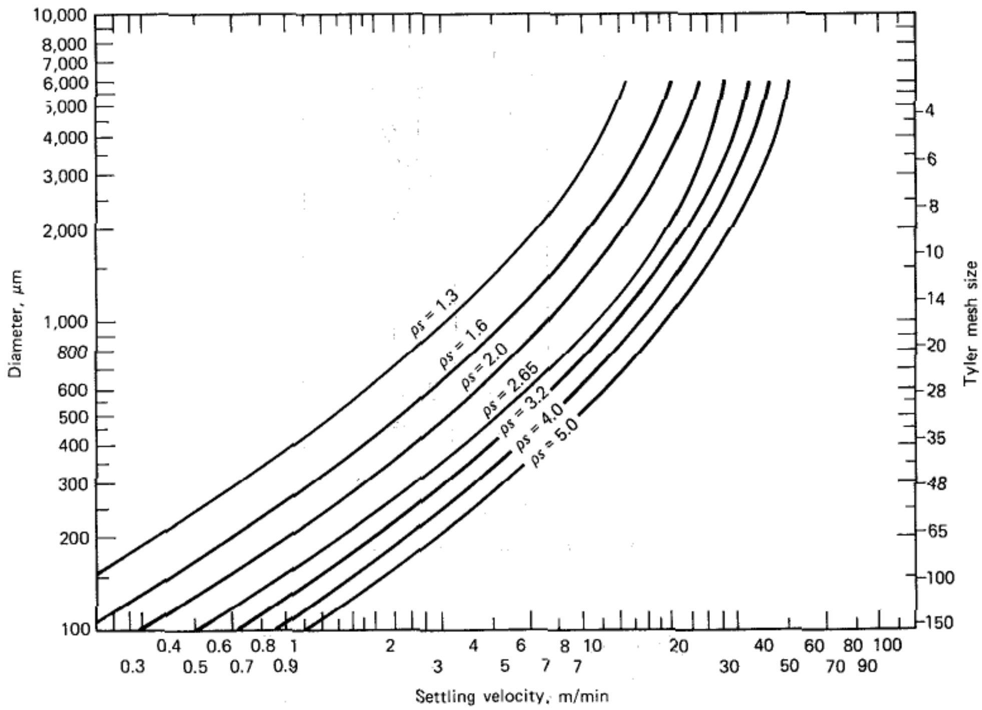

Figure 20. Settling velocity of spheres of various sizes in water as a function of particle size (ps) and specific gravity of the solids.

Table 1. Process Power Ratios

Process type	N^a	Power ratio, Settling velocity: 4.9–18.3[b]	1.2–2.4[b]	0.03–0.18[b]
complete uniformity	2.9	25	9	2
off-bottom suspension	1.7	5	3	2
on-bottom motion	1.0	1	1	1

[a] Impeller rotational speed, rps.
[b] *Settling velocity is in m/min.*

degree of suspension to be much higher with axial flow than with the radial flow (see also Extraction, liquid–solid).

Gas–Liquid Operations. Gas–liquid processes are affected by changes in power, speed, pumping capacity, and shear rate (23). Specification should give the process conditions required, such as mass-transfer and reaction rates. The dispersion specification alone, eg, 1000 m^3/min, is not sufficient.

For any gas–liquid process selection, the superficial gas velocity is needed. This is defined as the average volumetric gas-flow rate in and out of the vessel divided by its cross-sectional area at the temperature and pressure at the midpoint of the tank expressed in meters per second. Normally a value of about 0.1 m/s is the boundary between normal mixer applications and those which must be carefully designed for

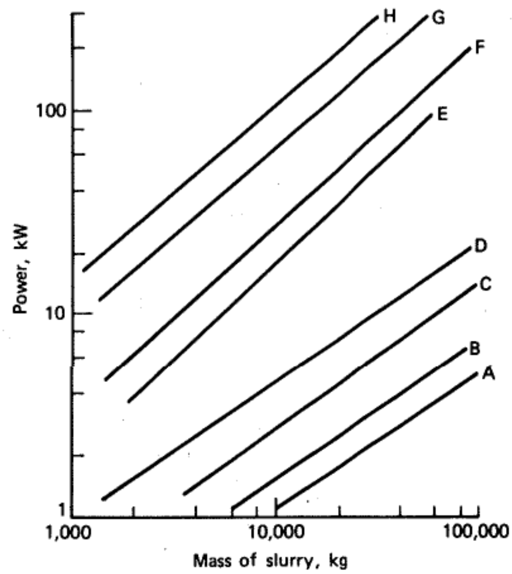

Figure 21. Approximate selection of mixer power, D/T ratio of 0.4, for two different settling velocities, two different percent solids concentration, as a function of mass of mother liquor in the tank. A, 2% off bottom, 5 m/min; B, 30% off bottom, 5 m/min; C, 2% uniform, 5 m/min; D, 30% uniform, 5 m/min; E, 2% off bottom, 15 m/min; F, 30% off bottom, 15 m/min; G, 2% uniform, 15 m/min; H, 30% uniform, 15 m/min.

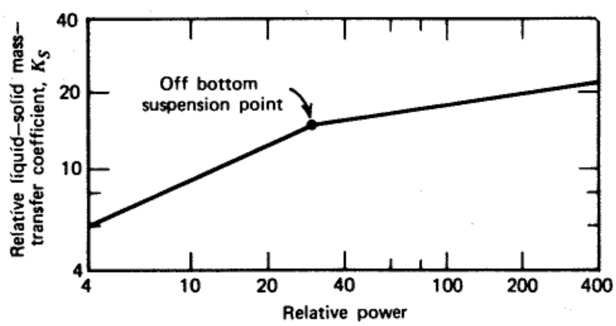

Figure 22. Effect of liquid–solid mass transfer.

process considerations such as liquid interface foaming and splashing, and fluid forces on mechanical equipment.

Intimate dispersions are controlled by the mixer flow patterns accompanied by power levels on the order of three times more energy in the mixer than in the expanding gas stream.

Minimum dispersion occurs at a point where the energy content of the mixer and gas stream are nearly equal. This usually results in a gas-controlled flow pattern in which uniform bubble dispersion leaves the area of the impeller. However, the overall flow pattern is governed by the rising expanding gas stream. Below that point are the areas where gas tends to form geysers and large splashing brings bursts of liquid to the surface.

As a general rule, the flow pattern from axial-flow impellers is governed by the upward velocity of the gas stream. Thus, axial-flow impellers do not operate satisfactorily unless their energy input is five to ten times higher than that of the gas stream.

Radial-flow impellers are most commonly used and also disk-type impellers. The latter prevent the gas from rising through the low shear zone around the hub.

Sparge rings or open inlets permit the admission of the gas at a somewhat smaller diameter than the impeller periphery below. Thus, the gas can rise into the maximum impeller-zone shear rate.

Liquid–Liquid Dispersion. Stable emulsions include household products, cosmetics, pharmaceuticals, and a wide variety of other combinations (see Emulsions). A minimum of fluid shear rate is usually required for the production of a uniform stable emulsion. When this fluid shear rate is produced by the mixer, the composition can either maintain its dispersion or the particles coalesce. If, on the other hand, the mixer is not capable of producing the shear rate required, the product will not be satisfactory. However, when the mixer produces the emulsion at high impeller shear rate, eventually all particles achieve the mininum particle size necessary for a satisfactory product.

The mixing time depends on the pumping capacity of the unit and on the number of times the particles pass through the high shear zone of the impeller and the length of time each particle spends in that shear zone.

Thus a variable known as the shear work, which is the product of shear rates and the time of contact, gives a measure of the total amount of energy expended into the dispersion.

Figure 23 shows a multistage mixer column in which the dispersed phase travels countercurrent to a continuous phase, either heavy or light (24). Mixer columns are similar in performance to packed, spray, or plate columns but can handle solids in either one or both phases. Since the dispersion produced is a function of the impeller relationships, scale-up is extremely reliable (see also Extraction, liquid–liquid).

Blending

Miscible Liquids. When blending miscible liquids, two distinct mechanisms are involved. In general, one material is run into the vessel with the mixer, whereas the second is injected into the tank. The uniformity of some particle or a physical or chemical property is then measured. A sampling point has to be chosen and the uniformity required for blending has to be defined.

Larger impellers at slow speeds reduce blend times. For a propeller, the blending time is proportional to D^{-2}, whereas for a turbine the blending time is proportional to D^{-1}. To obtain the same circulating time on scale-up, the ratio P/V increases with the square of the tank diameter. However, this relationship cannot be used for design. Therefore, an increase in blending time is incorporated into the overall plant design (25).

In another blending method, the tank is initially stratified. After mixing is started, the time required to eliminate the stratified condition is measured. This is typical of large petroleum storage tanks (26–28). Using hot and cold water and the associated temperature differences, the blending equation is

$$\theta \propto P^{-1}(D/T)^{-2.3}\left(\frac{\Delta\rho}{\rho}\right)^{0.9}$$

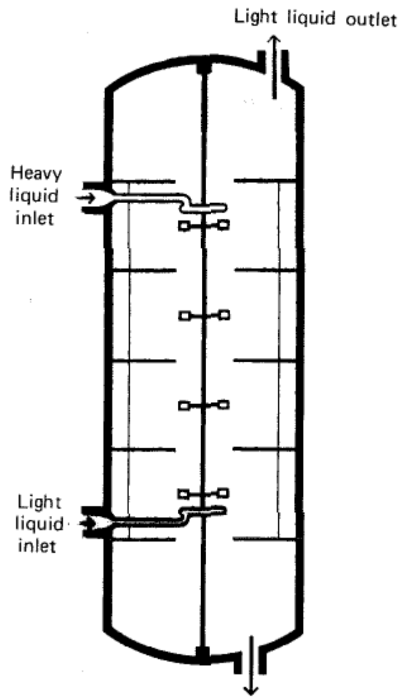

Figure 23. Oldshue-Rushton extraction column, using radial-flow turbines in each compartment.

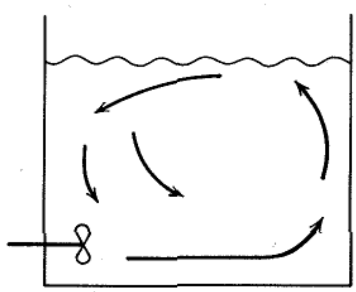

Figure 24. Typical flow pattern of properly positioned side-entering drive.

Other injection-technique studies give different exponents. In fact, blend time is proportional to power and not speed. It takes several times more power to achieve blending in a stratified tank than in an injection tank, but this difference is reduced on scale-up since P/V is more typically constant in the case of stratified blending.

For homogeneous fluids, side-entering mixers offer the most economic combination of capital and operating costs. Their effectiveness is based on the fact that the

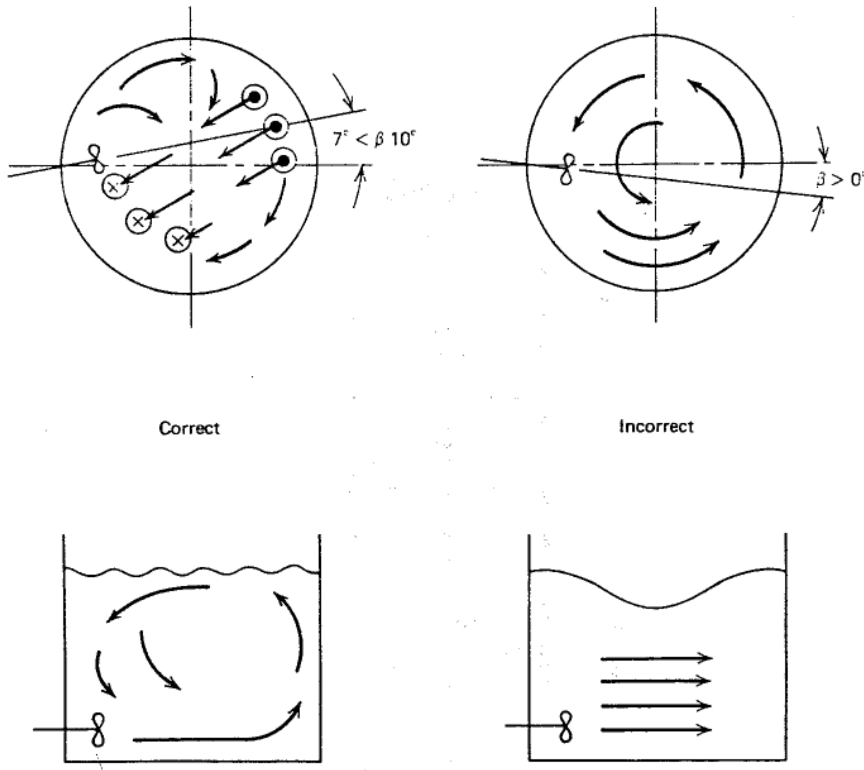

Figure 25. Flow pattern when proper angle of entry of the impeller shaft is or is not maintained.

fluid in the tank, as shown in Figure 24, makes a gentle series of 90° turns, and not the sharp 180° turns occurring in top-entering systems. However, there are quiet zones at the sides of these tanks, 90° away from the impeller position. In large gasoline and petroleum storage tanks this can cause the deposition of so-called bottom sediment and water. To alleviate this condition the angle of the mixer or the rotation is changed to keep the solids deposition moving around the tank at an acceptably controllable level. Figure 25 illustrates the flow pattern with and without proper angle of the impeller shaft.

Viscous Fluids. There are several areas of viscous blending. In large industrial tanks, low viscosity is defined as 5 Pa·s (50 P). From 5 to 50 Pa·s (50–500 P), which is the medium-viscosity region, either open-axial or radial-flow turbines may be considered or the close-clearance anchor or helical impellers.

The area above 50 Pa·s is defined as high viscosity mixing, in which typically an anchor or helical impeller should be used (see Figs. 26–27). Figure 28 gives typical data obtained in measuring the circulation time of helical-flow mixers in both Newtonian and non-Newtonian fluids (3,29).

Anchor impellers do not have any tendency for top-to-bottom flow and, therefore, do not provide effective blend time or achieve effective temperature uniformity in heat transfer applications. However, for many processes anchor impellers are used.

The helical impeller has a strong axial-flow component and can give effective blending and circulation times in both Newtonian and non-Newtonian fluids. Typical

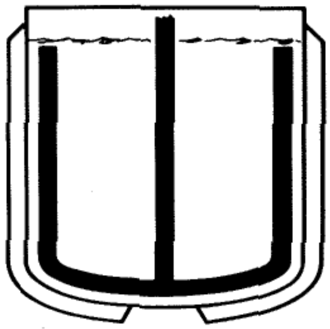

Figure 26. Anchor impeller.

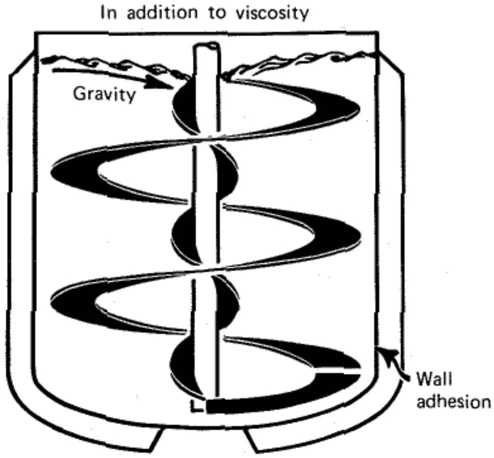

Figure 27. Helical impeller.

are a pitch of 0.5 and a blade width of $1/12$ to $1/6$ T. The inner flight is only effective on non-Newtonian fluids but gives superior performance over noninner flight.

A single outer flight gives the lowest torque and is completely satisfactory for large equipment. However, if a double outer flight is desired, it can be easily produced but results in a lower operating speed and a larger mixer drive for the same process requirement.

In the plant, the nominal clearance of 25–100 mm between the blade and the tank wall may be 25–100 mm, which requires no machining of either component. On the other hand, a clearance of 5–25 mm requires manufacture of the tank first. A template is prepared and the impeller machined to suit. Scraper blades also can be used which may be product loaded or spring loaded.

Scraper blades normally increase both power consumption and the heat-transfer coefficient by a factor of two.

Heat input from the mixer may be an appreciable part of the total heat load in the process.

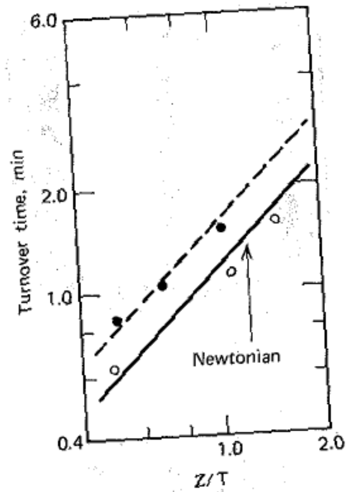

Figure 28. Turn-over time vs helical impeller speed Z for a variety of viscosities. $N = 20$ rps; pitch ratio = 0.5; 46 cm tank; 43 cm helix. Non-Newtonian fluid: o, single outer, single inner; ●, single outer.

In-Line Mixers. Figures 29–31 illustrate several of the in-line mixing elements used in pipelines (see Pipelines). For viscous fluids, elements are preferred that can twist and cut the streams or actually force materials through channels or tubes. Mixer performance is expressed as the width of disbursement, ie, the number of cuts per element which can be typically either 2, 3, or 4, raised to the nth power where n is the number of elements. A prediction can be made of the size of the final dispersion produced in the unit. Pressure-drop data generally are available.

In low viscosity operations, fluid turbulence and shear stresses are introduced by the elements as the flow passes through the unit. For a certain pressure drop, the same degree of shear rate and mixing effect are obtained. Thus, different types of elements require different lengths, depending on their specific pressure drop relationship.

In-line mixer elements provide primarily transverse uniformity but not time-interval uniformity (see Fig. 32). The latter requires a type of mixing volume that is accomplished either by a vessel in the pipeline, or a suitable mixing tank of the required size. In general, to effect time-interval uniformity, the residence time has to be about the same order of magnitude as the length of time it takes to blend the elements.

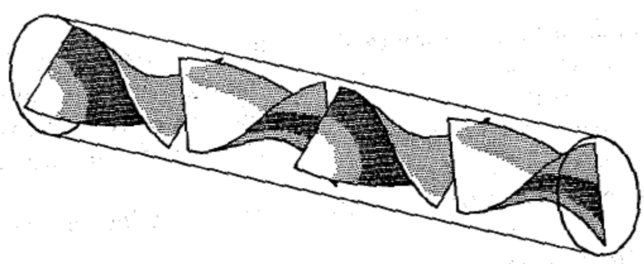

Figure 29. Kenics in-line mixer.

Figure 30. Koch in-line mixer.

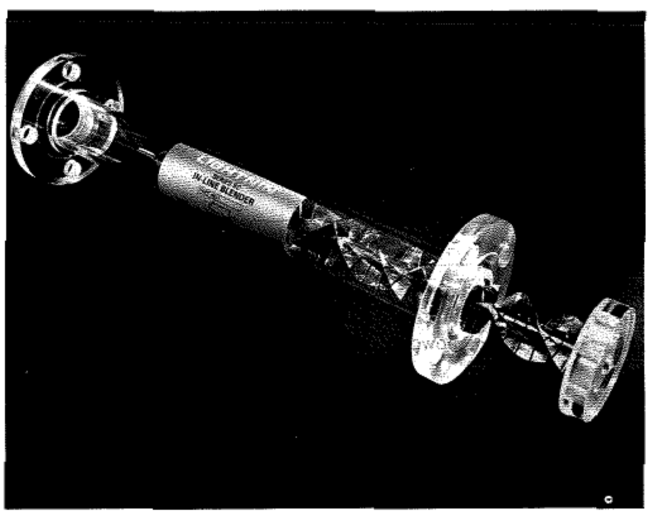

Figure 31. Lightnin in-line mixer.

Scale-Up

Table 2 shows the change of typical mixing parameters on a scale of 100–12,500 L capacity. The linear-scale ratio is 5:1. In calculating scale-up, these parameters are considered in correlation. Another consideration is geometric similarity, which has been used to derive Table 2. Figure 33 shows the maximum impeller-zone shear rate tends to increase, whereas the average impeller shear rate tends to decrease with increasing scale. Thus, the distribution of shear rates in a large mixing tank differs from

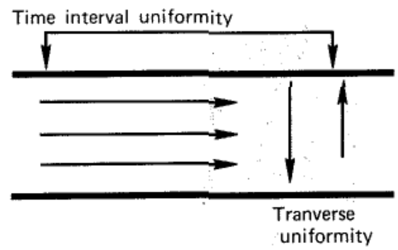

Figure 32. In-line mixers can provide transverse uniformity but not time-interval uniformity.

Table 2. Properties of a Fluid Mixer on Scale-Up

Property	Pilot scale, 0.1 m^3	Plant scale, 12.5 m^3			
P	1.0	125	3125	25	0.2
P/V	1.0	1.0	25	0.2	0.0016
N	1.0	0.34	1.0	0.2	0.04
D	1.0	5.0	5.0	5.0	5.0
Q[a]	1.0	42.5	125	25	5.0
Q/V	1.0	0.34	1.0	0.2	0.04
ND[b]	1.0	1.7	5.0	1.0	0.2
$\dfrac{ND^2\rho}{\mu}$	1.0	8.5	25.0	5.0	1.0

[a] Impeller pumping capacity.
[b] Tip speed.

Table 3. Dry Mixers

Type of mixing	Mixers used[a]
bakery mixes, foodstuffs	A, B, E
dry chemicals, crystalline, moderately fine to coarse	A, B, C, D, F, G, H
dry chemicals, crystalline, fine powders	A, B, C, D, F, G, H
pharmaceuticals	A, D
fertilizers	A, D, G
glue chips or powder	A, B, C, D, F, G, H
resin chips or powder	A, B, C, D, F, G, H
powdered metals	A, D, G
tobacco	A, D, G
feed and grain	A, B, C, G
foundry sand mixes	A, F

[a] A, tumble type, drum, container, V, cone; B, ribbon; C, paddle; D, centrifugal; E, planetary; F, pan, wheel, plow; G, spiral elevator; H, fluid bed.

those in a small tank. For a dynamic, heterogeneous system, in which bubbles and drops are dispersed and coalesced, the particle size distribution in a large tank differs from that in a small tank. This might or might not affect the overall process objective.

It is difficult to combine the two shear rates in big tanks as in small tanks. How-

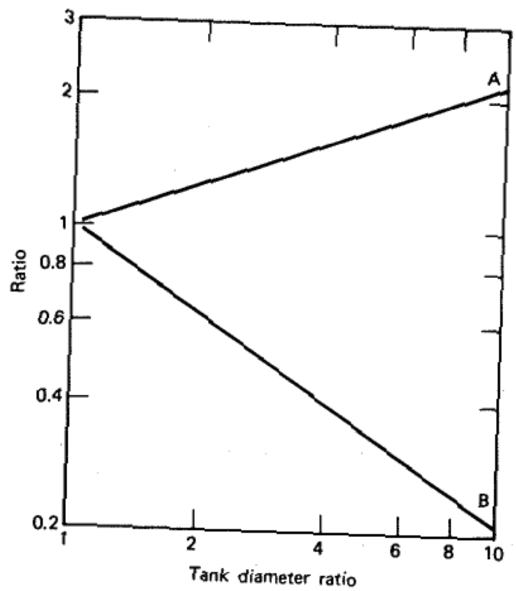

Figure 33. Maximum impeller-zone shear rates, A, increase while average impeller-zone shear rates, B, decrease when using geometric similarity and constant power per unit volume on scale-up.

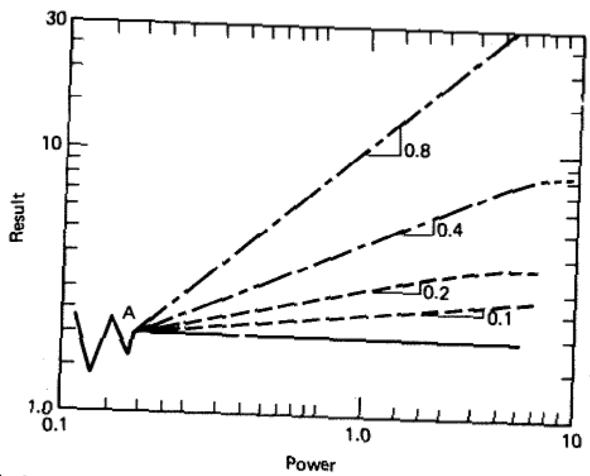

Figure 34. Relationship of process result and impeller power consumption, obtained by varying impeller speed. The slope can be used as an indicator of controlling factors.

ever, it is possible to control the increase or decrease of either of these shear rates by nongeometric scale-up or scale-down techniques.

In general, a small pilot-scale mixing tank designed to perform a particular plant-scale process is not a good model for a full-scale process because the pumping capacity and the maximum impeller-zone shear rate are too high, and the blend time is too short.

Geometric, kinematic, and dynamic similarity govern many scale-up consider-

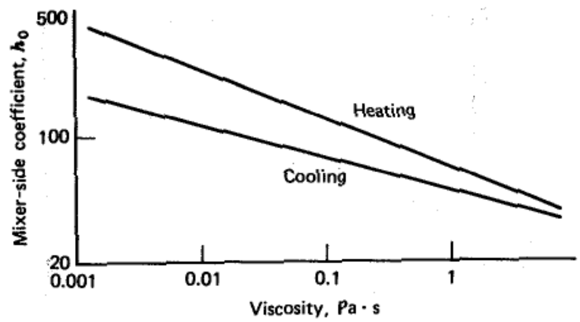

Figure 35. Design curves for the mixer-side coefficient as a function of viscosity for heating and cooling. To convert Pa·s to P, multiply by 10.

ations (30–31); eg, dynamic similarity requires that the ratio of the four main fluid forces in the tank, ie, the inertia force put in by the mixer F_i; the opposing force of viscosity F_v; the surface tension F_σ; and the gravity F_g in both the model $M$and the prototype P be equal to a common constant ratio R shown below. μ = viscosity, g = force of gravity, σ = surface or interfacial tension.

Geometric:

$$\frac{X_M}{X_P} = X_R$$

Dynamic:

$$\frac{(F_i)_M}{(F_i)_P} = \frac{(F_v)_M}{(F_v)_P} = \frac{(F_g)_M}{(F_g)_P} = \frac{(F_\sigma)_M}{(F_\sigma)_P} = F_r$$

Force ratios:

$$\frac{F_i}{F_v} = N_{Re} = \frac{ND^2\rho}{\mu}$$

$$\frac{F_i}{F_g} = N_{Fr} = \frac{N^2D}{g}$$

$$\frac{F_i}{F_\sigma} = N_{We} = \frac{N^2D^3\rho}{\sigma}$$

Unless the fluid in the model and the prototype is changed, dynamic similarity of all four fluid forces cannot be obtained. The viscosity is considered the main force but the Reynolds number and the ratio of inertia force to viscous force are often used in mixing correlations.

It is difficult to write an appropriate process group in dimensional terms which corresponds in principle to the dynamic similarity ratio of the Reynolds number. The usual requirements of pumping capacity, shear rates, diffusion rates, and many other quantitative and qualitative factors preclude the use of a single dimensionless group as a suitable indicator of process result. Exceptions are the Nusselt number for heat transfer and the blend number for well-defined blending criteria.

For scale-up, the following method is used: Some experimental data can give an approximate estimation of the principal controlling process steps; and the pertinent mixing parameters relating to these controlling steps have to be identified.

In general, the power in the pilot-plant experiment should be varied in order to change flow and fluid shear rates markedly. This is normally accomplished by changing

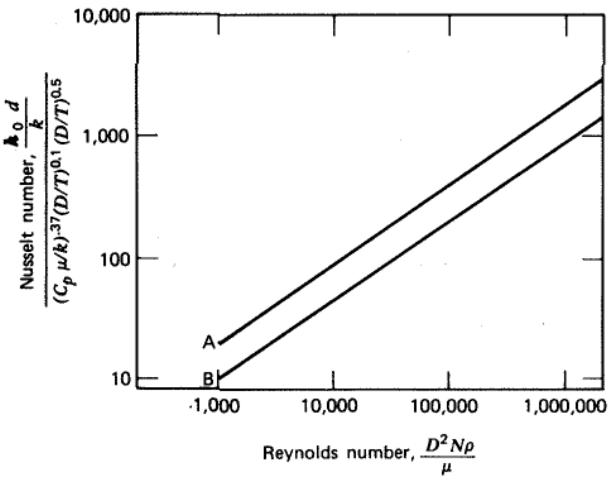

Figure 36. Correlation of Nusselt no. vs Reynolds no. for A, radial-flow flat-blade turbines, 6 blades; and B, propellers, 3 blades, pitch ratio 1.0. Tank, 1.2 m; liquid, 1.2 m level; baffles, 10 × 10 cm wide at wall.

the speed. Figure 34 is a quantitative plot of the process result, although it can be a qualitative estimation. A zero slope indicates a homogeneous chemical reaction mechanism, and mixing does not have a decisive effect on the process. On the other hand, a slope, on the order of 0.5 to 0.8, typically indicates gas–liquid or liquid–liquid type of mass-transfer-controlled processes (see Mass transfer).

A pilot process can determine the controlling factor, and thus indicate the best scale-up technique. However, most mixing-process designs do not require a pilot-plant model. Mostly, sufficient data are available to permit designers and vendors to select the equipment required. If not, the proper scale-up technique can be determined on the basis of some operating data (see Pilot plants and microplants).

In order to maintain equal blend time on scale-up, the ratio P/V must be increased with the square of the tank size. This usually is not feasible and, therefore, blending times are allowed to increase in larger tanks.

Constant P/V is usually too conservative for many processes. It is, however, a good criterion for homogeneous chemical reactions, and for scale-up over volume ranges of 5:1 or less. Parameters of big tanks are different in many respects and must be evaluated as to their ultimate effect.

Constant tip speed equals constant torque per unit volume, and is normally not predictable on scale-up. The value P/V drops in direct proportion to scale ratio, and this creates problems with macro- and micro-scale shearing effects as well as large increases in blend and circulation time. It does, however, work well for draft-tube circulators.

Heat-transfer correlations with Nusselt and Reynolds numbers allow accurate prediction of performance on different scales.

Heat Transfer. The source of turbulence affecting the heat-transfer coefficient in a mixing vessel is the fluid flow around and across the heat-transfer surface. The impeller affects only the flow that actually reaches the heat-transfer surface.

For design purposes, usually only a gross overall approximate coefficient is needed

(32). Figure 35 shows typical heat-transfer coefficients as a function of viscosity for organic fluids with various viscosities. The material with 1 mPa·s (= cP) viscosity is an aqueous solution.

In general, propellers and turbines give similar heat-transfer performance. For example, Figure 36 shows a correlation of Nusselt vs Reynolds numbers with other dimensionless groups (33). However, the propeller draws only 3% of the power of the flat-blade turbine at the same Reynolds number. Therefore, when it is accelerated to draw the same power, it has about the same heat-transfer coefficient.

The slope of heat-transfer coefficient vs mixer power in the forced convection region is very low, $h \propto P^{0.22}$. Typically, once forced convection is achieved, the heat-transfer coefficient is not raised by increasing power. Usually it is calculated, and the overall heat-transfer performance is modified by design changes (see Heat-transfer technology).

Mixing of Dry Solids

Table 3 gives an overview of dry-mixing applications. Bulk blenders are used for the complete gamut of dry-solids blending, from coating of solids to paste mixing.

Tumbling. Gentle mixing by a tumbling action causes materials to cascade from the top of the rotating vessel. Common types offer various vessel configurations including drum, container, V, and cone. Intake and discharge of materials take place through an opening in the vessel end. Dry and partly dry powders, granules, and crystalline substances are readily mixed in such equipment. Liquid feed and intensive blending are possible with the addition of a liquid-feed or high speed mix bar.

Ribbon Type. Spiral or other blade styles transfer materials from one end to the other or from both ends to the center for discharge. This mixer can be used for dry materials or pastes of heavy consistency. It can be jacketed for heating or cooling.

Paddle Type. This type is similar to the ribbon type except interrupted flight blades or paddles transfer materials from one end to the other, or from both ends to the center for discharge. The paddle-type mixer can be used for dry materials or pastes of heavy consistency. It can be jacketed for heating or cooling.

Centrifugation. Materials are passed through a rotor consisting of disks spaced ca 2.5 cm apart and held together by rod-type supports that act as impacters. Centrifugal forces throw the material against the rods and inside walls or the housing, from which it drops by gravity to the discharge (see Centrifugal separation).

Planetary Type. Paddles or whips of various configurations are mounted in an off-center head that moves around the central axis of a bowl or vessel. Material is mixed locally and moved inward from the bowl side, causing intermixing. This mixer handles dry materials or pastes.

Pan Type. Mulling action of the wheel type is similar to the action of a mortar and pestle. Scrapers move the materials from the center and sides of a pan into the path of rotating wheels where mixing takes place. The pan may be of the fixed or rotating type. Discharge is through an opening in the pan. The flow type uses rotating plows in a rotating pan to locally mix and intermix by the rotation of plows and pan, respectively (Fig. 37).

Spiral Elevator. Materials are moved upward by the centrally located spiral-type conveyor in a cylindrical or cone-shaped vessel. Blending occurs by the downward movement at the outer walls of the vessel. The vessel serves the dual purposes of blending and storage.

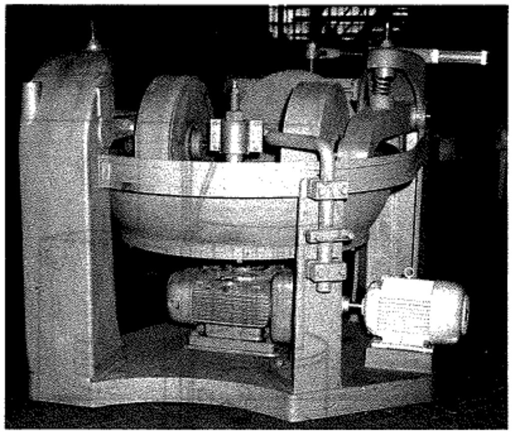

Figure 37. Model 507 Clearfield mixer.

Fluidized-Bed Type. Particles suspended in a gas stream behave like a liquid. They are mixed by turbulent motion and intimate contact of components (see Fluidization). This mixer is used for mixing and drying, or mixing and reaction.

Equipment for Pastes and Viscous Materials

Blending of viscous materials and pastes can be achieved either by the age-old art of dividing and recombining, or by layering the various components to some predefined striation thickness until adequate uniformity is attained. Moving agitators may have to come close to walls or stationary baffles in order to provide the high shear required to separate agglomerates and to reduce the size of regions occupied by one component. The energy requirements may be very high because of the work involved in dividing and shearing the material. Allowable heat rise frequently imposes a limit at which power can be applied.

Mixing machinery is selected according to its capacity to shear material at low speed and to wipe, smear, fold, stretch, or knead the mass to be handled. Mixers with intermeshing blades are sometimes required to keep the material from clinging unmixed to the lee side of the blade. Wiping of heat-transfer surfaces promotes addition or removal of heat.

Batch Mixers. Batch rather than continuous mixing is still preferred when batch identity must be maintained, eg, in pharmaceutical preparations; frequent product changes would require too much off-specification production between products, eg, in dye and pigment manufacture; a multitude of ingredients is required, each of which can be accurately weighed and charged without relying on absolute constancy of flowmeters, eg, in some adhesives and caulking compounds; various changes of state may be involved, eg, the pulling of a vacuum on the end product to drive off volatiles after reaction of a granular material which would not permit a fluid seal in a continuous mixer; and very long mixing and reacting times are required.

Change-Can Mixers. In change-can mixers one or more blades cover all regions of the can either by a planetary motion of the blades or a rotation of the can (see Fig.

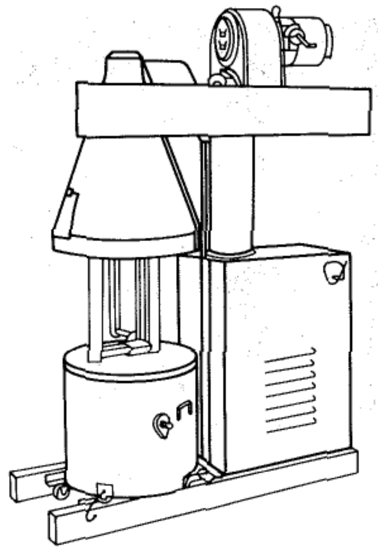

Figure 38. Change-can mixer. Courtesy of Chas. Ross & Son.

38). The blades may be lowered into the can or the can may be raised to the mixing head. Separate cans allow the ingredients to be measured carefully before the mixing operation begins, and can be used to transport the finished batch to the next operation while the next batch is being mixed.

Stationary-Tank Mixers. Stationary-tank mixers are recommended when the particular advantages of the change-can mixer are not required. The agitator may be particular to a specific industry, like the soap crutcher, or for general use as with anchor mixers (Fig. 26) or gate mixers. In the latter, some part of the agitator moves in close proximity to the vessel walls or stationary bar baffles.

The impeller may also consist of a single- or double-helical blade (Fig. 27) to promote top-to-bottom turnover while minimizing the amount of hardware that must be moved through the viscous mass.

Double-Arm Kneading Mixers. These mixers have been used for a long time (see Fig. 39). The material is carried by two counterrotating blades over the saddle section of a W-shaped trough. Randomness is introduced by the difference in blade speed and end-to-end mixing by differences in the length of the arms on the Σ-shaped blades. Other blade shapes are used for specific end purposes, such as smearing or cutting edges on the blade faces. Discharge is usually by tilting the trough or by a door in the bottom of the trough. Double-arm kneading mixers also are available with a screw centrally located to discharge the contents (Fig. 40).

Intensive Mixers. Intensive mixers, such as the Banbury (Fig. 41), are similar in principle to the double-arm kneading mixers, but are capable of much higher torques. Used extensively in the rubber and plastics industries, the Banbury mixer is operated with a ram cover so that the charge can be forced into the relatively small volume mixing zone. The largest of these mixers holds only 500 kg but is equipped with a 2000-kW motor.

Roll Mills. When dispersion is required in exceedingly viscous materials, the large surface area and small mixing volume of roll mills allows maximum shear to be

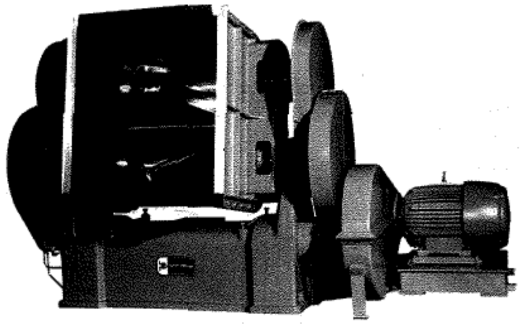

Figure 39. Sigma-blade mixer. Courtesy of Baker Perkins Inc.

maintained as the thin layer of material passing through the nip is continuously cooled. The rolls rotate at different speeds and temperatures to generate the shear force with preferential adhesion to the warmer roll.

Ribbon Blenders. Ribbon blenders (Fig. 42) provide end-to-end mixing and overall lifting of the blender contents by the pitch of the helical blades. Variations include closed vessels with a plow-shaped head on a horizontal rotor operating at relatively high speed to scrape the trough wall as well as to hurl the contents throughout the free space.

Similar applications also are handled in vertically mounted mixers equipped with a high speed blade. This blade scoops material from the bottom of the vessel, hurling it upward along the vessel walls, where baffles may be located to provide variation in turnover. Such mixers may be used for dispersion in thin slurries, or in wetting or coating of granular materials.

Cone-and-Screw Mixer. A more gentle blender is the Nauta-type cone-and-screw mixer (Fig. 43). This mixer uses an orbiting screw to provide bottom-to-top circulation. Reversing the screw aids discharge. Other variants are available that provide either an epicyclic orbit or two screws to provide greater volumetric coverage. The mixing action of cone-and-screw mixers is independent of the degree of fill.

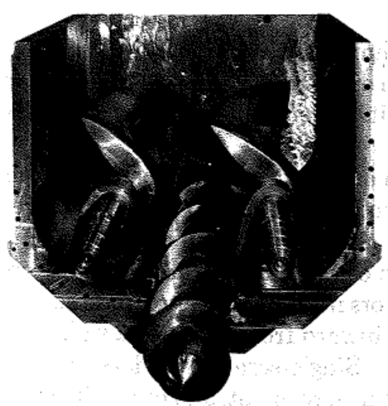

Figure 40. Sigma-blade mixer with screw discharged. Courtesy of B. P. Guittard.

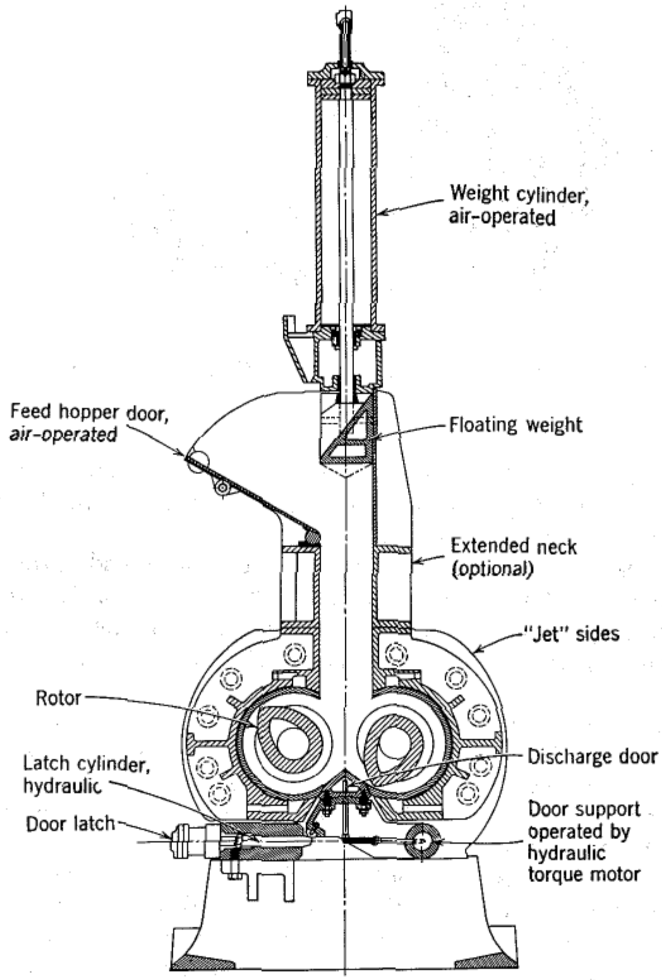

Figure 41. Banbury mixer. Courtesy of Farrel Corp.

Pan Muller. A mixer similar to mortar and pestle is the pan muller, in which plows bring material into the path of the rolling mullers. Such mixers are used for applications where the final blend is neither very fluid nor very sticky, such as foundry sand, clay, or chocolate.

Continuous Mixers. In most continuous mixers one or more screw or paddle rotors operate in an open or closed trough. Discharge may be restricted at the end of the trough to control holdup and degree of mixing. Some ingredients may be added stagewise along the trough or barrel. The rotors may be cored to provide additional heat-transfer area. The rotors may have interrupted flights to permit interaction with pins or baffles protruding inward from the barrel wall.

Single-Screw Extruders. Single-screw extruders (Fig. 44) incorporate ingredients, such as antioxidants, stabilizers, pigment, and other fillers into plastics and elastomers. In order to provide a uniform distribution of these additives, the polymer is brought to a fusion state primarily by the work energy imparted in the extruder, rather than

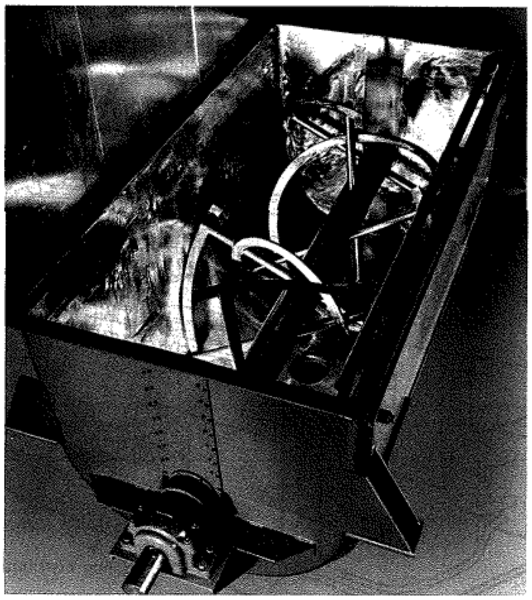

Figure 42. Spiral-ribbon mixer. Courtesy of Teledyne Read Co.

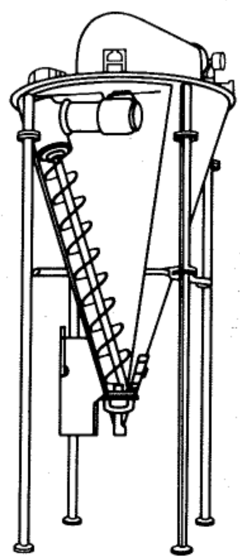

Figure 43. Nauta mixer. Courtesy of J. H. Day Co.

by heat transfer through the barrel wall. In addition to melting the polymer, the extruder is used as a melt pump to generate pressure for extrusion through a die, shaping the molten product in a specific profile, strands, or pellets. The extruder screw drags the polymer through the barrel, generating shear between screw and barrel. In addition, some axial mixing occurs in response to a back flow along the screw channel caused by the pressure required to get through the die.

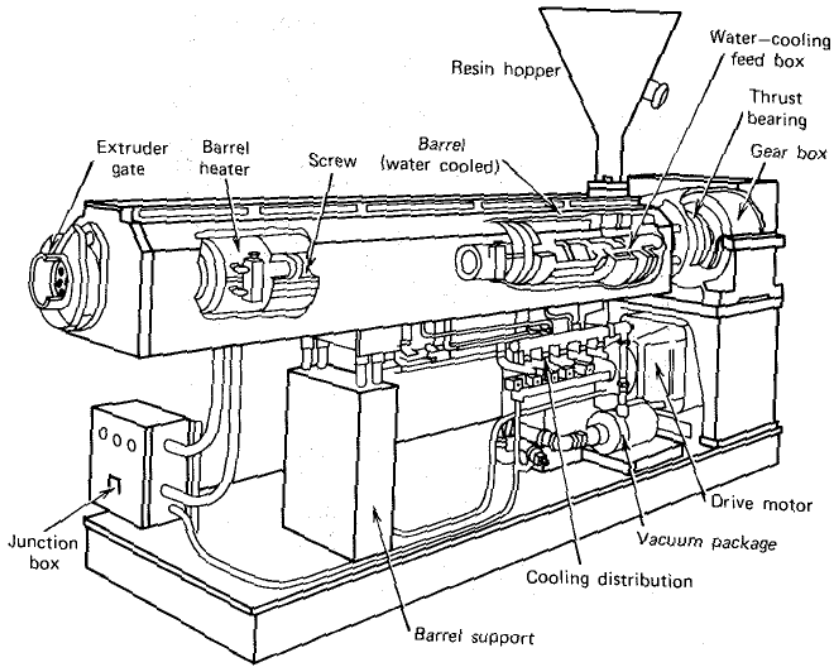

Figure 44. Sectional view of a typical extruder.

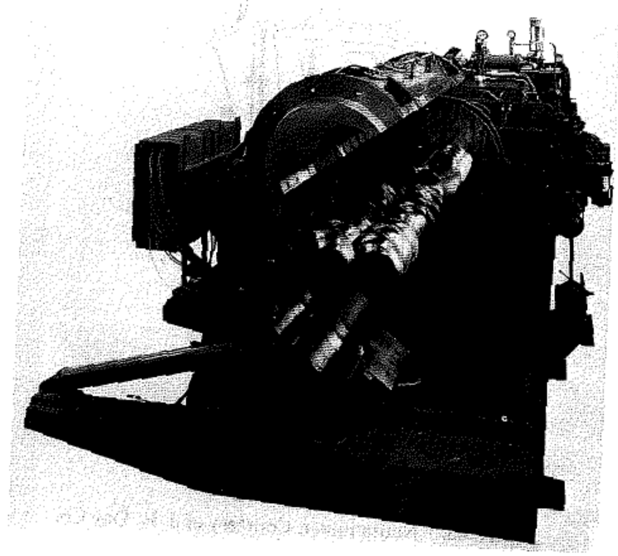

Figure 45. Twin-screw mixer. Courtesy of Baker Perkins Inc.

Mixing in a single-screw extruder can be enhanced by interrupting the flow pattern within the screw flight channel. This can be done by variations in the screw flight-channel width or depth, or by causing an interchange with spiral grooves in the barrel, or by interengaging teeth.

Twin-Screw Mixers. Twin-screw continuous mixers provide more radial mixing by interchange of material between the screws, rather than just acting as screw conveyors.

Intermeshing co-rotating twin-screw mixers (Fig. 45) have the additional advantage that the two rotors wipe each other as well as the barrel wall. This action eliminates any possiblity of dead zones or unmixed regions. In addition to variations in screw helix angles, these mixers can be fitted with kneading paddles that interactively cause a series of compressions and expansions to increase the intensity of mixing. Such mixers are used for a variety of pastes and doughs, as well as in plastics compounding.

Nomenclature

a = constant
b = constant
C = impeller off bottom distance
D = impeller diameter, m
d = tube diameter, m
F = superficial gas velocity, m/s
H = velocity head, J/kg
h_0 = mixer side coefficient
K_S = mass-transfer coefficient
N = impeller rotational speed, rps
N_{Fr} = Froude number
N_P = power number
N_{Re} = Reynolds number
N_{We} = Weber number
P = power, W
Q = impeller flow; pumping capacity, kg/s
T = tank diameter, m
TIR = total indicated run-out
U = liquid velocity, m/s
V = volume
Y = distance, m
Z = helical impeller speed
θ = blend time
μ = viscosity
Φ = power number
ρ = fluid density
σ = surface tension

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