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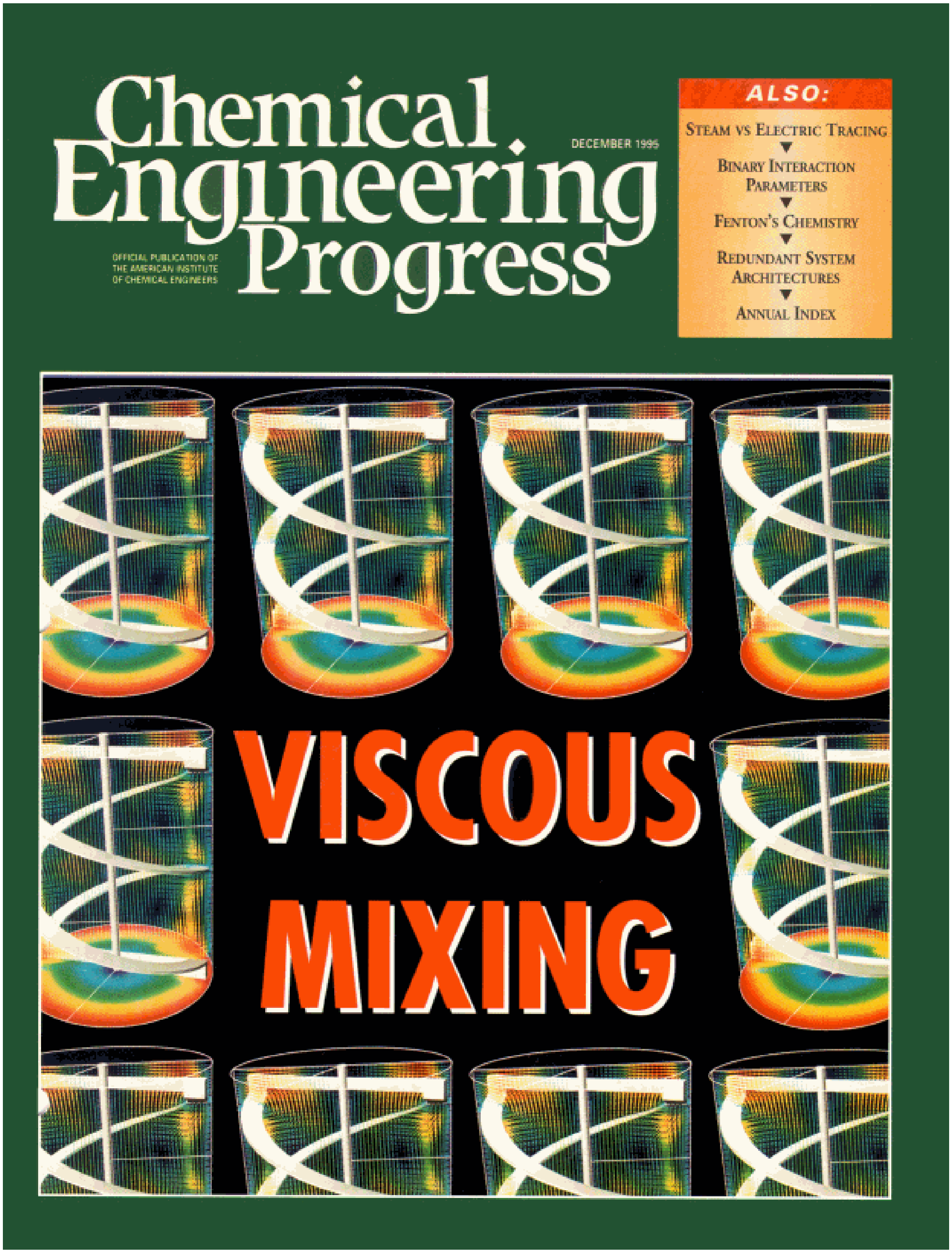

Properly Choose Mechanical Agitators for Viscous Liquids

Turbine and helical-ribbon mixers are the most commonly used mixers. Calculations, modeling, and economics are all factors in selection.

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High-viscosity mixing applications occur in most chemical process industries (CPI) plants. For instance, the polymer production industry must blend high-viscosity reaction masses to thermal and chemical uniformity. This industry must also blend small amounts of low-viscosity antioxidants and colorants into polymer streams. The personal-care products industries encounter many high-viscosity mixing applications in the preparation of creams, lotions, pastes, and drugs. Other high-viscosity applications occur in the production of food, paint, drilling mud, and greases, to name a few. Viscosities can be in the range from about 1 Pa-s all the way up to 25,000 Pa-s in some extreme cases. The materials to be mixed in these applications can be very important economically.

Mixing can occur in pipeline systems with motionless mixers, or in vessels using mechanical agitators, depending on the application and the process requirements. A wide variety of both motionless mixers and mechanical agitators is available to handle specific mixing problems and fluid types. One or more agitators can be used to blend a tank. Each agitator can have one or more shafts, with one or more impellers that do not necessarily have to be of the same type. Impellers can be top-entering, side-entering, or bottom-entering.

This article gives an overview of designing the most commonly used agitator for blending applications: a top-entering agitator with a single shaft. The agitator can be equipped with multiple turbine-style impellers of different design, or with helical-ribbon or anchor-style impellers to optimize the agitator for the specific application and blending problem on hand.

Industrial-scale helical-ribbon mixer with supporting cross-bars.

Various overviews of liquid blending can be found in the literature, see for example, Fasano and Penney *(1)*, Bouwmans *(2)*, and Fasano *et al. (3)*. These articles focus mainly on blending of liquids in the turbulent regime: impeller Reynolds numbers of around 10,000 and higher.

Although turbulent blending will be briefly discussed here also, this article will focus on blending in the laminar and transitional regimes. Also, we will discuss the special requirements for blending non-Newtonian fluids, with and without yield stress. We will first discuss the flow patterns and applicability of different impeller types and will then present some design guidelines.

Turbine agitators

A wide variety of turbine-style impellers is available. Figure 1 lists the most-common ones: the pitched-blade turbine, high-efficiency impeller, disc turbine, and simple straight-blade turbine. Well-designed turbine impeller systems can be used up to viscosities of about 50 Pa-s, depending on the scale, application, and process requirements.

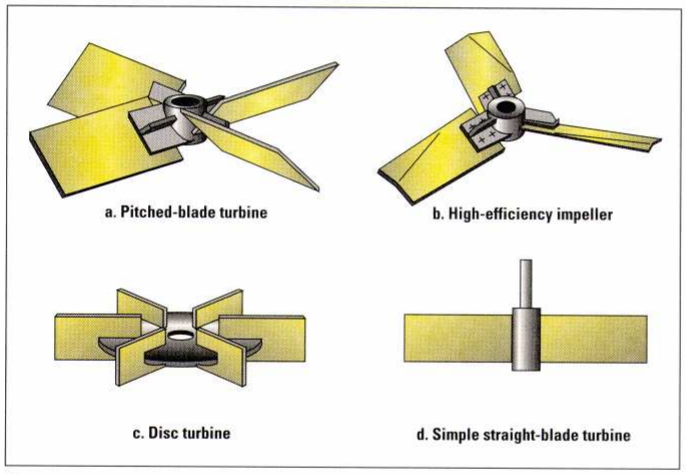
Each impeller has its own characteristics and its own area of application. Straight-blade impellers and disc turbines tend to be used to create zones with high shear rates, such as in dispersion. High-efficiency and pitched-blade impellers, and pitched-blade turbines tend to be used at Reynolds numbers larger than about 100, where a good overall circulation flow is important. The impeller Reynolds number is defined as:

$$N_{Re} = \frac{\rho N D^2}{\mu_{eff}} \qquad (1)$$

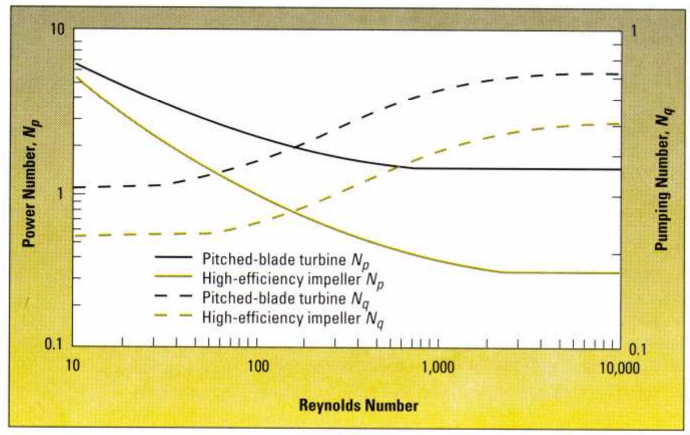
For Newtonian fluids, the effective viscosity μ_{eff} is simply the dynamic viscosity. For non-Newtonian fluids, the viscosity will vary with shear rate throughout the volume of the tank. To calculate the average viscosity experienced by the impeller, an effective shear rate S_{eff} is calculated by Metzner and Otto *(4)*:

$$S_{eff} = KN \qquad (2)$$

Here, N is the impeller rotational speed in s^{-1} and K is an impeller constant. Table 1 lists K-values for several different impeller types. Substituting the effective shear rate in the viscosity equation for the particular fluid then gives the effective viscosity, which can be used to calculate the impeller Reynolds number. Several other methods of calculating the effective shear rate have been reported in the literature. Most of these are more complicated than the method used here, while not always more accurate.



■ ***Figure 1. A wide variety of turbine agitators is available to handle viscosities to about 50 Pa-s.***



■ ***Figure 2. Power number and pumping number as functions of Reynolds number for a pitched-blade turbine and high-efficiency impeller.***

Two other dimensionless numbers that are often used to characterize an impeller are the power number and the pumping number. The power number N_p is implicitly defined by the following equation:

$$P = N_p \rho N^3 D^5 \qquad (3)$$

When the power number of an impeller is known, the power draw of the impeller can be calculated using this equation, given liquid density, impeller speed, and diameter.

Table 1. *K*-values for effective shear-rate model (Eq. 2).

Impeller Type	*K*
High efficiency	10
Pitched blade	11
Straight blade	11
Disc turbine	11.5
Anchor (D/T = 0.98)	24.7
Helical ribbon (D/T = 0.96, P/D = 1)	29.4

The values for the anchor and helical ribbon impellers are for the specified standard geometries only. For other geometries, Eqs. 12 and 17 should be used.

The pumping number N_q is implicitly defined by:

$$Q_l = N_q N D^3 \tag{4}$$

When the pumping number of an impeller is known, the pumping rate of the impeller Q_l can be calculated for a given diameter and speed.

Both the impeller power number and pumping number depend on a variety of factors, such as the ratio of impeller to tank diameter D/T, the impeller off-bottom clearance ratio C/T, and the impeller Reynolds number. Figure 2 shows both power and pumping number correlations for both the high-efficiency impeller and pitched-blade turbine, using a standard geometry: $D/T = C/T = 1/3$.

For Reynolds numbers above 10,000, both N_p and N_q show little variation with Reynolds number. In this regime, the flow is considered to be fully turbulent. When the Reynolds number decreases, the pumping number decreases, while the power number goes up. This regime is usually called the transitional regime.

At very low Reynolds numbers ($N_{Re} < 10$), the power number is inversely proportional to the Reynolds number. In this Reynolds number regime, the flow is considered laminar. The exact Reynolds numbers at which the transitions occur are a function of impeller type, size, and number. Grenville *et al.* *(5)* correlated the Reynolds number at the boundary between the turbulent and transitional regime as follows

$$N_{Re,\,t} = \frac{6{,}370}{N_p^{1/3}} \tag{5}$$

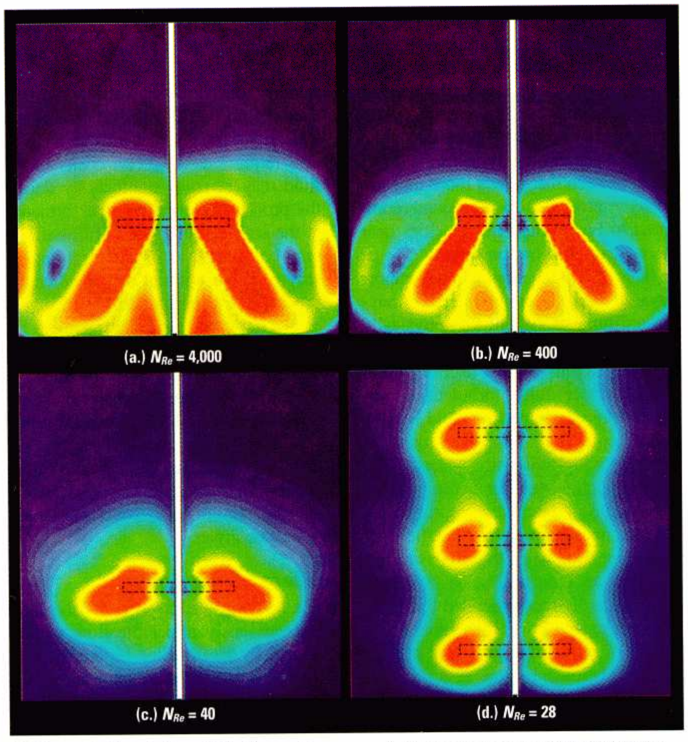
Flow patterns

The rate of mixing in a tank is related to the overall flow pattern. Flow patterns can be effectively studied using modern techniques, such as computational fluid mixing (CFM), digital particle image velocimetry (DPIV), and laser Doppler velocimetry (LDV).

As an example, we will discuss the flow patterns created by the pitched-blade turbine. These patterns will be illustrated with results of CFM simulations. In these simulations, the impellers were modeled using LDV data obtained by Wang *et al.* *(6)*. The results of the simulations are shown in Figure 3. Color denotes the local velocity magnitude. The fluid in the red regions moves fast, while that in the blue regions moves slow.

When the flow is turbulent, the pitched-blade turbine creates an axial jet (Figure 3a). Although most axial-flow impellers can be used pumping upwards also, they are mainly used pumping downwards as shown in the figure. The impeller creates a typical axial flow pattern, with one main circulation loop. A smaller circulation loop with weak flow forms near the shaft, under the impeller. The highest velocities are found in the outflow of the impeller, near the blade tip. Lower velocities are found far away from the impeller, for example, near the liquid's surface.

When the Reynolds number is decreased, for example, by lowering the



■ *Figure 3. Simulation of the flow patterns created by a single pitched-blade turbine in (a.), (b.), (c.), and by a three-impeller system in (d.).*

speed or by increasing the liquid viscosity, the outflow from the impeller changes. The flow pattern gradually becomes more radial and, as a result of the decreasing pumping capacity, the average velocity in the tank decreases also. Figures 3b and 3c show the flow pattern at Reynolds numbers of 400 and 40.

At the lowest Reynolds number shown here, the impeller pumps mainly in the radial direction. The liquid velocities are very weak farther away from the impeller. Obviously, such a flow pattern is not desirable to achieve good overall mixing.

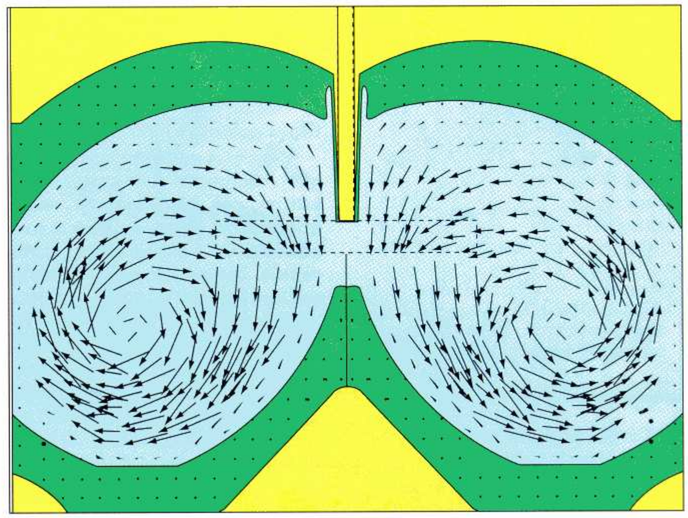
To increase the rate of mixing in the tank under these circumstances, a different impeller system must be used. The overall mixing in the tank can be improved by using more impellers. Also, the diameter of the impeller should be large enough to create significant liquid movement at the wall.

Figure 3d shows a redesigned impeller system for this tank. The number of impellers has been increased to three. The impeller speed has been decreased such that power draw remains the same. It can be seen that now a more-uniform velocity profile is being found compared to Figure 3c. The velocities near the wall are still low, however.

This example stresses the importance in high-viscosity systems of having an impeller system that physically agitates the whole volume of the tank. In a tank with a low-viscosity fluid where, under fully turbulent conditions, only one impeller would be sufficient, multiple impellers may be needed for high-viscosity systems.

When agitating viscous fluids in the transitional or laminar regime, it may not be necessary to use baffles, as should be done when the flow is turbulent. On an industrial scale, full baffling is usually recommended for fluids with a viscosity lower than about 5 Pa-s, to convert swirling into axial and radial circulation.

Full baffling means the use of four baffles with a width of $T/12$ and offset from the tank wall a distance of $T/72$. For more-viscous fluids, a smaller number of narrower baffles may be used, or no baffles may be used at all. This, however, also depends on the process requirements. The viscosity of the fluid may increase significantly during the process. For optimum performance at the start of the process, full baffling may be needed, while the performance may be better without baffles when the process nears completion.



■ ***Figure 4. Simulation of the flow pattern in a tank filled with paper pulp, a yield-stress fluid.***

An optimum baffle arrangement for the process as a whole thus needs to be determined. An alternative might be to mount the impeller off-center or to use a rectangular tank to create an effect similar to baffling, without creating the slow-moving zones in the wake of the baffles. In such cases, special testing may be required and exact guidelines cannot be given.

Effect of rheology

Changing the rheology of the fluid will obviously affect the flow pattern created by the impeller. Figure 4 shows a similar tank and agitator as in the previous figures, but now filled with paper pulp, which has a yield stress (see Bakker and Fasano *(7)*). In the blue region around the impeller, the fluid is turbulent. The flow is laminar in the blue-green region. In the areas that are colored yellow, the shear stresses are below the yield stress and the fluid is stagnant.

The impeller creates a cavern in which the fluid is moving relatively fast and where the flow is turbulent. However, in the bulk of the fluid, where the shear stresses are below the yield stress, the fluid is not moving at all. Increasing the diameter of the impeller or the impeller speed will increase the size of the cavern.

With yield-stress fluids, the engineer must make sure that the shear stresses in the whole volume of the fluid are larger than the yield stress. A calculation procedure for sizing impellers for such applications is beyond the scope of this article. However, there is sufficient information available in the literature to calculate cavern sizes for various impeller systems (Solomon *et al. (8)*, Elson *et al. (9)*, and Etchells *et al. (10)*).

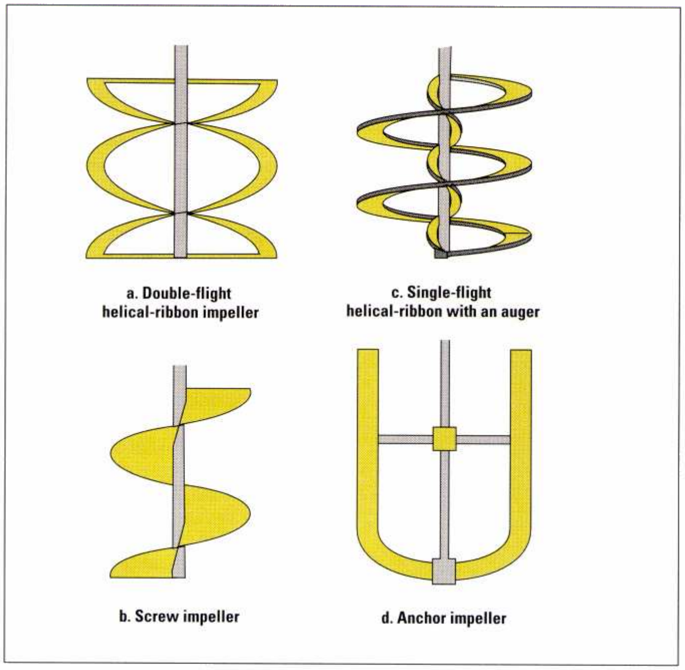
In low-Reynolds number ($N_{Re} \leq 400$) laminar flow, it is not always sufficient to create movement throughout the whole volume. It is important to avoid the creation of self-circulating or segregated zones, in which the fluid only slowly mixes with the rest of the tank. The existence of such segregated regions in viscous systems has been found for various impeller systems, as mentioned by Nagata *(11)*.

If segregated zones form in an industrial reactor, the product quality and consistency may deteriorate. Muzzio and Lamberto *(12)* suggest varying the impeller speed to create time-dependent flow patterns. This prevents the formation of steady self-circulating flow loops.

In general, at low Reynolds numbers, when the use of helical impellers is not yet necessary or not possible, it is best to use multiple, large-diameter axial flow impellers. Multiple feed-pipes adding directly into the impellers are preferred. This gives the best guarantee that the fluid is evenly mixed throughout the whole liquid volume. Both laboratory tests and CFM modeling can be used to study the flow patterns in detail.

The above description of mixing mainly applies to steady-state flows, or when the overall fluid flow is a lot slower relative to the time-scale at which mixing occurs. Although this description is fairly accurate under laminar flow conditions, the situation is more complicated when the flow is turbulent.

In turbulent flow, eddies are present with a wide variety of time-scales that are shorter than the blade-passage frequency. Also, large-scale chaotic variations in the mean flow can occur with a period that is much longer than the impeller rotational frequency, as shown by Myers *et al. (13)*. The high-frequency turbulent eddies provide mixing action termed turbulent diffusion. In contrast, long time-scale fluctuations in the overall mean-flow field provide additional, quasi-random stretching and folding action in the fluid that can be very effective in decreasing large-scale concentration gradients.



■ ***Figure 5. Helical-ribbon and anchor impellers provide an alternative to turbine impellers.***

Helical-ribbon and anchor impellers

An alternative to selecting a turbine impeller system is the use of a helical-ribbon impeller or an anchor impeller (Figures 5a and 5d). These impellers sweep the whole wall surface of the tank and physically agitate most of the fluid batch. As a result, they can be used at much lower Reynolds numbers ($N_{Re} \leq 400$) than the open-style turbines.

Helical ribbons are primarily used when very viscous materials are to be mixed, starting at viscosities of about 20 Pa-s and higher, depending on the scale. Helical ribbons have been successfully used for viscosities up to 25,000 Pa-s. These impellers often have a diameter that is very close to the inside diameter of the tank, and, therefore, are sometimes called close-clearance agitators. They guarantee liquid motion all the way to the wall of the tank, even for these very viscous materials.

Helical impellers can also be manufactured in a different configuration, with the helical impeller blade wrapped closely around the shaft (Figure 5b). Such impellers are also called helical-screw impellers, and are sometimes used in combination with a draft tube.

When these screw impellers are combined with a close-clearance helical ribbon as shown in Figure 5c, the screw is sometimes called an auger. The power draw from an impeller is proportional to the diameter to the 5th power, as given in Eq. 3. Since the diameter of the auger is much smaller than the diameter of the outside helical ribbon, the power draw of an auger is often less than 1% of that of a helical ribbon. When correctly designed, an auger can be used to improve the mixing near the shaft, without significantly increasing the power draw and torque requirement for the agitator. If the auger is incorrectly designed, and the pumping capacity of the auger does not match the flow created by the helical ribbon impeller, the auger may actually block the liquid flow near the shaft and hamper the mixing process.

The cover photo shows a flow pattern created by a helical-ribbon impeller at N_{Re} = 10, as calculated using CFM. Again, the velocity vectors point in the direction of the local liquid velocity. Both the length of the vectors and the color show the local velocity magnitude. Red denotes a large velocity and blue, a small one. Only the axial and radial velocity components are shown.

The helical ribbon is rotating clockwise (seen from the top of the tank) and is pumping downward at the wall, although these devices can be built to pump upward at the wall also. Circulation loops form around the helical blade. Part of the liquid that is pumped downwards by the top helix reaches the bottom helix. The rest of the fluid is circulated back through the center of the tank.

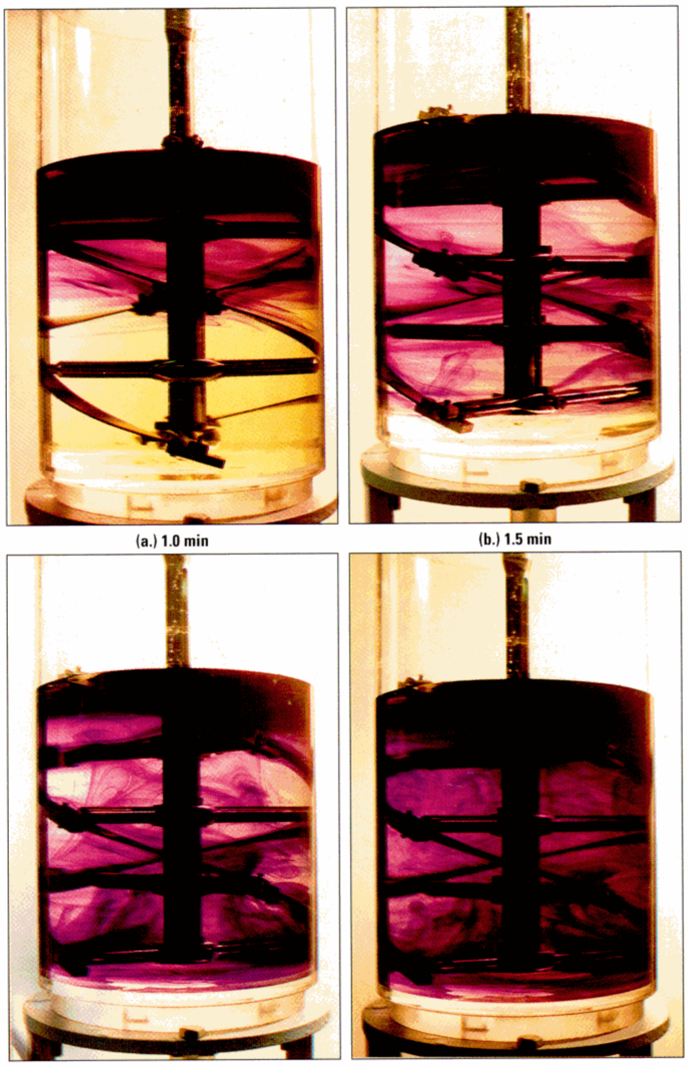
Even at this low Reynolds number, the fluid in the tank is moving at sufficiently high velocities to guarantee good liquid movement and top-to-bottom mixing. The flow pattern of the helical ribbon depends on many geometrical parameters, such as pitch, blade width, and helix diameter. For example, if the gap between the helix and the tank wall is too large, liquid may circulate back upwards at the tank wall, thus decreasing the velocities in the center of the tank. CFM models can be used to accurately predict such effects and to optimize the mixer design for the application on-hand.

Figure 6 shows the result of a laboratory-scale blending test, using a double-flight helical ribbon. A tracer fluid is added at the liquid surface and slowly mixed in. The photographs were taken after 1.0, 1.5, 3.5, and 5.0 min. The helical ribbon is rotating at 6 rpm.

Fluid is pulled down at the wall and circulated back up through the center. The complicated flow pattern generated by the helix is clearly visible from the intricate shape of the striations and surfaces seen in the figure.

The helix is usually mounted with several supporting cross-bars (see photo on p. 25). These cross-bars also physically sweep the liquid, adding to the mixing action. It is recommended to extend the helix all the way to the liquid surface and to make the clearance between the bottom cross-bar and the tank bottom as small as possible.

Another option is to add anchors at the top and bottom. This prevents the formation of slow-moving or stagnant



(a.) 1.0 min (b.) 1.5 min (c.) 3.5 min (d.) 5.0 min

■ ***Figure 6. Results of a blend time experiment with a laboratory-scale double-flight helix.***

Design example: Helical-ribbon or turbine?

A chemical reaction vessel requires an agitator that will blend the process liquids to 95% uniformity in less than 5 min. Both a helical-ribbon impeller and a turbine impeller system will be designed to meet this criterion, and will be compared. The reactor information is as follows:

Tank dia.	$T = 2.5$ m
Liquid level	$Z = 2.5$ m
Liquid viscosity	$\mu = 25$ Pa-s
Liquid density	$\rho = 1{,}200$ kg/m^3

A standard helical ribbon impeller with $D/T = 0.96$ and $W/D = 0.1$ will be designed. The required impeller rotational speed can be calculated from the required blend time and Eq. 13:

$$N = \frac{75}{t_{b,\,95\%}} = \frac{75}{300} = 0.25\ \mathrm{s}^{-1}$$

The closest impeller speed that is commercially available is $N = 0.275$ s^{-1} = 16.5 rpm. This is the speed that will have to be used in the reactor, and the blend time is then:

$$t_{b,\,95\%} = \frac{75}{0.275} = 4\ \mathrm{min},\ 33\ \mathrm{s}$$

To verify that the Reynolds number is in the range of validity of the blend-time correlation, the Reynolds number is calculated as:

$$N_{Re} = \frac{1{,}200 \times 0.275 \times 2.4^2}{25} = 76 < 100$$

The power number is then calculated from Eq. 9 to be:

$$N_p = \frac{350}{76} = 4.61$$

And the power draw is then:

$$P = 4.61 \times 1{,}200 \times 0.275^3 \times 2.4^5 = 9{,}160\ \mathrm{W}$$

The required torque is calculated as:

$$M = \frac{P}{2\pi N} = \frac{9{,}160}{2\pi\left(0.275\right)} = 5{,}301\ \mathrm{N\text{-}m}$$

Designing a turbine impeller system that meets the same blend-time criterion at this low Reynolds number requires making several iterations through the equations presented in this article, and using the blend-time correlations presented by Fasano and Penney *(1)*, and Fasano *et al.* *(3)*. A suitable system was found to consist of three pitched-blade turbines with a dia. of 1.45 m, operating at 0.933 s^{-1} = 56 rpm.

This system has a calculated blend time of 4 min, 44 s. The Reynolds number is 94, which is pushing the lower limit at which these impellers can still be used. The power draw of this impeller system is $P = 41{,}230$ W, and the torque requirement is $M = 7{,}033$ N-m.

Comparing the two systems reveals that both have approximately the same blend time. Because of the larger power draw and torque requirement of the turbine agitator it will require a more expensive motor and gear box. A helical ribbon impeller itself will be more expensive than three pitched-blade turbines. The total capital cost of the whole agitator with the helical ribbon (including shaft, motor, and gear box) is about 10% higher than when pitched-blade impellers are used. However, the power draw is less than a fourth of the pitched-blade impeller system, resulting in a significant reduction in electricity and operating expenses.

The final decision on which impeller system is to be used may also depend on future plans. If it is anticipated that, at some point, liquids with a viscosity higher than the design viscosity of 25 Pa-s will be used, then the helical-ribbon impeller will be the best choice. Conversely, if it is anticipated that future processes may use lower-viscosity materials, the pitched-blade impeller system may be used.

zones near the liquid surface and the vessel bottom.

When it is important to have high velocities at the tank wall, for example, in heat-transfer applications, it is recommended to make the clearance between the helix and the tank wall as small as possible. Further, the helix can be equipped with scrapers that physically remove the fluid from the tank wall.

Agitator design

The previous sections gave qualitative discussions about the flow patterns generated by the different impeller styles and the effects on mixing in the tank. This section gives more quantitative design guidelines, both for designing turbine agitators and close-clearance agitators such as helical ribbons and anchor impellers. The above sidebar presents a worked-out example in sizing both types of mixers.

The required agitator size is often determined, based on previous process experience, by means of laboratory tests, or by means of computer simulations. An often-used design aid is the scale of agitation S_A defined by Hicks *et al.* *(14)*:

$$S_A = 32.8\,\frac{N_q N D^3}{\frac{\pi}{4} T_{eff}^2} \tag{6}$$

where:

$$T_{eff} = \left(\frac{4}{\pi} V_l\right)^{1/3} \tag{7}$$

Here, T_{eff} is the effective tank diameter and V_l is the total volume of liquid. S_A provides a 1–10 range of agitation intensity with 1 being mildly mixed, and 10 being very violently mixed. Agitation levels of 1 and 2 are characteristic of applications that only require low fluid-velocities to achieve the process result. Levels of 3–6 are most common in the CPI. Values of 7–10 are characteristic of applications requiring high fluid velocities, such as in critical reactors. Gates *et al.* *(15)* present guidelines on how to relate S_A to specific process applications.

The scale of agitation is not suited for use with fluids with a large yield-stress or which are viscoelastic. Viscoelastic fluids tend to develop flow normal to the direction of shear. This can lead to "counterintuitive" effects in the flow field, such as fluids climbing up a rotating shaft, and fields in which the direction is reversed when compared to Newtonian fluids. Such fluids require special testing and design considerations when being agitated, and are not discussed in detail in this article.

When laboratory tests are performed, the scaled-down system is usually geometrically similar to the large-scale industrial vessel. To achieve the same mixing intensity in both the laboratory and industrial vessel, S_A is usually kept constant. To achieve the same qualitative flow pattern, the Reynolds number, as defined in Eq. 1, should be kept the same on both scales. If possible, both S_A and the Reynolds number should be kept constant by using a model fluid with a lower viscosity on the laboratory scale.

When studying vortex motion at the liquid surface, it is necessary to keep both the Reynolds number and the Froude number constant at both scales. The Froude number is defined as:

$$N_{Fr} = \frac{N^2 D}{g} \tag{8}$$

The impeller speed at the laboratory scale can be calculated from the impeller speed at the large scale and the Froude number. A laboratory model-fluid has to be chosen such that the Reynolds number is the same on both scales. Other commonly used scaling criteria are constant power-per-unit-volume, which is often used in mass-transfer and dispersion applications, and constant tip-speed, which is similar to a constant S_A, provided that the Reynolds number is kept constant also.

The power draw of the pitched-blade turbine and high-efficiency impeller is easily calculated using Figure 2. In multiple-impeller systems, the power draw of each impeller is usually calculated as if the impeller were the only one in the tank. Then the individual power draws are added to obtain the overall draw. In situations where the impellers are mounted closely together and pump in the same direction, the overall power draw may actually be less, but adding the power of each impeller gives a usable, somewhat conservative estimate of the power-draw and torque requirements.

In power-draw calculations for non-Newtonian fluids, the effective viscosity μ_{eff} is substituted for the dynamic viscosity. The effective viscosity is calculated from the particular viscosity vs. shear-rate curve and the effective shear rate S_{eff} from Eq. 2.

When testing non-Newtonian fluids, care should be taken that the viscosity measurements are done in the appropriate shear-rate range. For example, the popular power-law model is often only valid over a range of about two decades of shear rate. This should be taken into account when doing lab testing at a small scale and a relatively large shear rate, and building the final plant on a large scale, at much lower shear rate. To avoid costly errors, viscosity curves should always be measured in the shear-rate range and temperature range in which they will be used in the plant.

Nomenclature

C	= impeller to bottom clearance, m
D	= impeller dia., m
f_{geo}	= geometry correction factor in power draw equations, dimensionless
g	= gravitational acceleration, m/s²
H	= impeller or helix height, m
K	= impeller constant, dimensionless
M	= torque, N-m
n_f	= number of flights for a helix
N	= impeller rotational speed, 1/s
N_{Fr}	= Froude number, dimensionless
N_p	= impeller power number, dimensionless
N_q	= impeller pumping number, dimensionless
N_{Re}	= Reynolds number, dimensionless
$N_{Re,t}$	= Reynolds number at boundary between transitional and turbulent flow, dimensionless
P	= impeller power draw, W
P_i	= pitch of a helical ribbon impeller, m
Q_l	= impeller pumping rate, m³/s
S_{eff}	= effective shear rate experienced by the impeller, 1/s
S_A	= scale of agitation, dimensionless
W	= blade width, m
$t_{b,95\%}$	= blend time to 95% uniformity, s
T	= tank dia., m
T_{eff}	= effective tank dia., m
V_l	= liquid volume, m³
Z	= liquid height in tank, m
Greek letters	
ρ	= density, kg/m³
μ	= dynamic viscosity, Pa-s
μ_{eff}	= effective dynamic viscosity in power-draw calculations, Pa-s

Blend time

Blend time is another important quantity, often used in the design of chemical reactors and mixing tanks. An extensive discussion of blend time calculations for axial flow impellers such as the high-efficiency impeller and the pitched-blade turbine is presented by Fasano *et al.* *(3)*. The algorithms developed are limited to impeller Reynolds numbers larger than 100. For lower Reynolds numbers, the blend time increases so fast with increasing viscosity that accurate correlations are not available. This is also the regime where the use of helical-ribbon impellers should be strongly considered. If it is nonetheless necessary to estimate the blend time for such a highly viscous system equipped with turbine agitators, it is recommended to perform a laboratory test. The blend time at the large scale can then be calculated, assuming that the product of impeller speed and blend time is the same at both scales.

The power draw of a *helical-ribbon impeller* is similarly calculated from the power number, as follows.

For $N_{Re} \leq 100$:

$$N_p = \frac{350}{N_{Re}} f_{geo} \tag{9}$$

For $N_{Re} > 100$:

$$N_p = 1.12 \exp[6.15 - 1.37 \ln N_{Re} + 0.0613(\ln N_{Re})^2] f_{geo} \tag{10}$$

where:

$$f_{geo} = \left(\frac{D}{P_i}\right)^{1/2} \left(\frac{H}{D}\right) \left(\frac{W/D}{0.1}\right)^{0.16} \left(\frac{D/24}{T-D}\right)^{1/2} \left(\frac{n_f}{2}\right)^{1/2} \tag{11}$$

In Eq. 11, pitch P_i is the vertical climb of a single flight of the helix in 360 deg. rotation. The N_p curve is shown in Figure 7 for a standard design with $f_{geo} = 1$.

The Reynolds number is calculated using the viscosity at the effective shear rate. For helical ribbon impellers the effective shear rate is given by:

$$S_{eff} = 25\left(\frac{D}{T}\right)^{1/2}\left[\frac{P_i}{\left(\pi^2 D^2 + P_i^2\right)^{1/2}}\right]^{-0.152} N \tag{12}$$

Accurate blend-time correlations are scarce, and there is a significant amount of scatter in the data presented in the literature, especially when it comes to the effect of changes in the geometry. For a standard double-flight helical-ribbon impeller ($P/D = 1$, $W/D = 0.1$, and $D/T = 0.96$) the blend time for 95% uniformity can be estimated from the equation below.

For $N_{Re} \leq 100$:

$$t_{b,\,95\%} = \frac{75}{N} \tag{13}$$

The power number for *anchor impellers* is given by the equation below.

For $N_{Re} \leq 10$:

$$N_p = \frac{400}{N_{Re}} f_{geo} \tag{14}$$

For $10 < N_{Re} < 10{,}000$:

$$N_p = 1.05\exp[5.64 - 0.783\ln N_{Re} - 0.0523(\ln N_{Re})^2 + 0.00674(\ln N_{Re})^3] f_{geo} \tag{15}$$

where:

$$f_{geo} = \left(\frac{H}{D}\right)\left(\frac{W/D}{0.1}\right)^{0.16}\left(\frac{D/49}{T-D}\right)^{1/2} \tag{16}$$

The N_p curve is shown in Figure 7, for a standard design with $f_{geo} = 1$. For the helical ribbon, $D/T = 0.96$ and $P_i/D = 1$. For the anchor impeller, $D/T = 0.98$. The effective shear rate for anchor impellers is given by:

$$S_{eff} = 25\left(\frac{D}{T}\right)^{1/2} N \tag{17}$$

The blend time for 95% uniformity for anchor impellers of standard geometry ($W/D = 0.1$, $D/T = 0.98$, and $Z/T = 1$) can be estimated by Eq. 18.

For $100 < N_{Re} < 10{,}000$:

$$t_{b,\,95\%} = \exp\left(12.9\, N_{Re}^{-0.135}\right) \tag{18}$$

Comparing Eq. 13 with Eq. 18 shows that at a Reynolds number of 100, it will take an anchor impeller more than 13 times as long to achieve 95% uniformity as a helical-ribbon impeller operating at the same speed.

Both with helical-ribbon impellers and with anchor impellers, the power input can be so high that cooling is required to remove the excess heat. Heat-transfer calculations are beyond the scope of this article. A good source of heat-transfer correlations is Nagata *(11)*.

Conclusions

This article gives an overview of the major considerations when mixing viscous fluids in industrial situations. Much information is available in the lit-

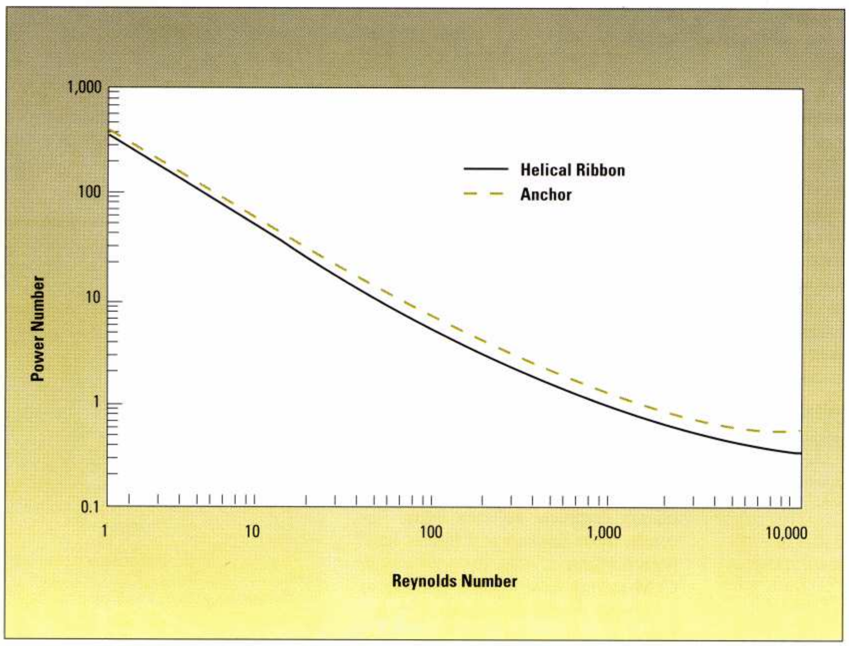


Figure 7. Power number as a function of Reynolds number for a helical-ribbon impeller and an anchor impeller, both of standard geometry.

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erature on mixing of low-viscosity fluids, but less is offered on mixing of higher-viscosity fluids. This is due to the smaller number of applications, the complexity of studying high-viscosity mixing, and the often-unusual rheological characteristics of viscous process fluids. While the number of applications is smaller, the dollar value of equipment for high-viscosity applications is higher because mixing equipment costs rise quickly with increasing viscosity.

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The technology base for design of high-viscosity mixers is much more limited than for design of low-viscosity mixers. While this article has summarized important design information for high-viscosity mixing, the authors strongly encourage the use of computer simulations, and laboratory and pilot-plant studies to supplement and confirm the recommended approaches given here. Careful analysis of all relevant mechanical and process design parameters, and capital and maintenance cost is required in all cases where viscous materials are to be mixed on an industrial scale.

Historically, mixing used to be a mainly empirical field of study. Recently, more fundamental studies have been performed. Also, the advances in CFM mixing have made it possible to accurately predict flow fields, generating information that can be used in evaluating the mixing performance of an agitator to a level of detail that was unheard of a decade ago. CEP

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