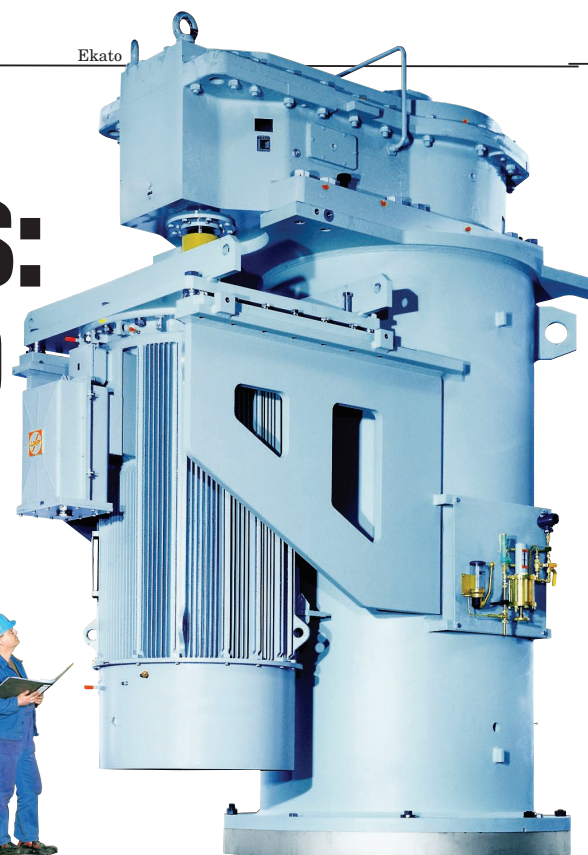## Cover Story

Mixing Systems:
Design and Scale Up

**Many options are available to meet
the mixing challenges confronted by the CPI.
Both theoretical and empirical methods can help
the engineer to fit the right system to the task**

**Werner Himmelsbach, David Houlton,
Wolfgang Keller, Mark Lovallo**
EKATO Mixing Technology

Fundamental to the chemical process industries (CPI) — whether specialty or bulk chemicals, pharmaceuticals, food products, minerals processing, environmental protection or other products or activities — is the need for mixing. The wide variety and complexity of mixing tasks encountered in industrial applications require careful design and scale up to ensure that effective mixing is achieved efficiently. Designs based on a small range of traditional agitators are no longer economically acceptable. Modern impellers and the use of physical or computer modeling can greatly enhance performance and reduce costs.

Mixing tasks fall into six main categories: 1. blending of miscible liquids; 2. blending of mixtures with "difficult" rheologies (such as non-Newtonian properties); 3. suspension of solids; 4. liquid-liquid dispersions; 5. heat transfer; and 6. gas-liquid dispersion. Different mixing behaviors and rules govern each basic mixing task. To optimize a design, or to scale-up reliably, these behaviors and rules need to be understood and defined. Complex tasks that involve two or more of the above categories require special attention. The controlling task must be identified to determine the design and scale-up rules to be applied.

This article addresses the first five mixing tasks that are listed above. Gas-liquid dispersion is a complex subject in itself and has had many significant advances in recent years. Rather than give a very abbreviated summary here, we refer the reader to Reference [2] for this topic.

In addition to agitator design and power requirements, which are fundamental to mixing systems, many other considerations also play a part in maximizing performance. These considerations include mechanical aspects, seal selection, materials of construction and surface finishes to prevent fouling or aid cleaning. Figure 2 outlines the process and factors involved in designing a suitable system.

BLENDING OF MISCIBLE LIQUIDS

During processing, inhomogeneities of concentration or temperature often arise. This typically happens during process steps such as addition of chemicals, mass transfer, heat transfer and chemical reaction. Inhomogeneities lead to non-uniform processing and can negatively impact product quality. The objective of blending is to maintain the required degree of homogeneity.

Degree of homogeneity

The homogeneity at a given time, $M(t)$, is given by the change in concentration of a component from c_0 to $c(t)$ with time:

$$M(t) = \frac{c_0 - c(t)}{c_0 - c_\infty} \qquad (1)$$

FIGURE 1. This 1900 kW agitator-drive unit is for a continuously operated bulk-chemicals reactor
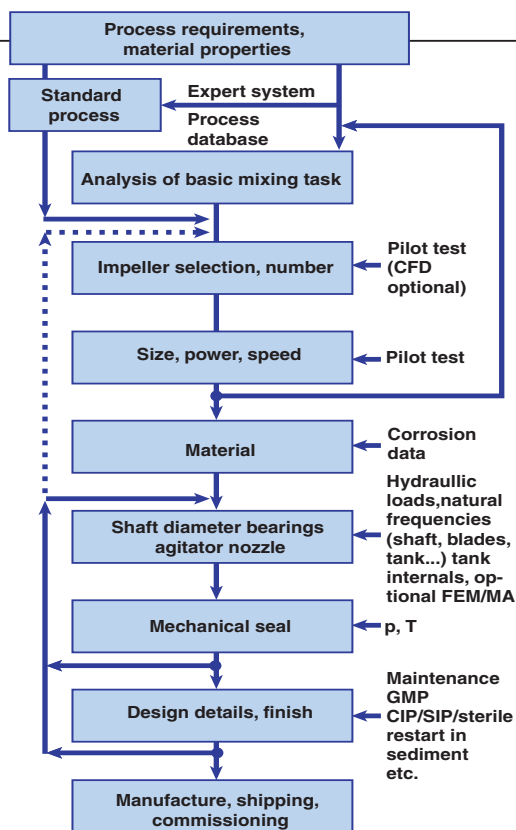
where c_∞ is the concentration after an infinite period of time. The same relation also holds for temperature homogeneity. A standard target homogeneity is 95%. The blend time needed to achieve a higher homogeneity can be calculated from the equation for standard blending time since inhomogeneities generally decline exponentially:

$$t_{m,x} = t_{m,95} \cdot \ln(1-x) / \ln(0.05) \qquad (2)$$

where x is the desired degree of homogeneity. When comparing the performance of blending equipment, it is particularly important to compare both the blending time and the degree of homogeneity achieved.

Turbulent blending

The dimensionless blend time characteristic (Nt_m) is constant for geometrically similar, agitated, baffled vessels that operate in the turbulent-flow regime. Appropriate baffling is always required in the turbulent regime, not only to achieve efficient blending, but for all mixing tasks. Without proper baffling, fluids will tend to rotate in the vessel and blend times will increase. The value of Nt_m depends mainly on the type of impeller

NOMENCLATURE

A	m^2	heat transfer area
c_w	kg/kg	concentration (mass fraction)
c_v	m^3/m^3	concentration (volume fraction)
C	-	impeller specific factor used in Equ. (12) and (17)
D	m	impeller diameter
d_p	m	particle size
f_c	-	correction factor for hindered-settling velocity
g	m/s^2	acceleration due to gravity
H	m	vessel filling height
h_P	W/m^2K	heat-transfer coefficient, process side
k_{MO}	-	Metzner-Otto constant
k, k_p	W/mK	thermal conductivity, process side
K	$Pa \cdot s^m$	viscosity "consistency factor"
m	-	viscosity "flow index"
M_t	$N \cdot m$	shaft torque
$M(t)$	-	homogeneity (time)
n_{js}	s^{-1}	minimum shaft speed for suspension
N	s^{-1}	agitator speed
Nt_m	-	dimensionless mixing time

Nu	-	Nusselt number $Nu = hPT/kP$
P	W	impeller power (transmitted to fluid)
Po	-	impeller power number, $Po = P / (\rho N^3 D^5)$
Pr	-	Prandtl number $Pr = \mu c/k$
P_{settle}	W	settling power
q	m^3/s	impeller pumping rate
Q	W	rate of heat
Re	-	impeller Reynolds number, $Re = N D^2 \rho / \mu$
s	-	Zwietering constant, impeller specific
T	m	vessel internal diameter
t_m	s	mixing time
V	m3	volume of fluid
$v_{s,sh}$	m/s	settling / hindered settling velocity
$v(r)$	m/s	local velocity at radius r
x	-	degree of homogeneity
z	-	number of recirculations to achieve homogeneity
μ, μ_W	$Pa \cdot s$	mixture viscosity, at wall
ψ	m^2/s	kinematic viscosity
$\rho_{l,s}$	kg/m^3	density liquid, solid
γ	s^{-1}	local shear rate
$\tau_{0, w}$	Pa	yield stress, stress at wall

FIGURE 2. Many factors are involved with designing an agitator system for a given application

and the diameter ratio. Mersmann [1] evaluated the measurement results of various authors and determined a simple correlation that characterized the performance of many mixers, with $H/T = 1$ and a single-stage impeller, in terms of the impeller power number and ratio of impeller to vessel diameter. In dimensionless form his equation becomes:

$$N t_m = 6.7 \cdot \left[\frac{D}{T} \right]^{-5/3} \cdot Po^{-1/3} \qquad (3)$$

Our own measurements show that the above equation forms a good basis for design calculations for impellers with large diameter ratios ($D/T > 0.5$), whereas the following equation is more accurate for predicting blend times for axial-pumping impellers with diameter ratios in the range of 0.1 to 0.5. Accuracy is about ±10% [2]:

$$N t_m = 5.5 \cdot \left[\frac{D}{T} \right]^{-5/3} \cdot Po^{-1/3} \qquad (4)$$

Circulation rate

When comparing the mixing performance of different agitators, many users and suppliers discuss the concept of "circulation rate," which expresses the number of times that it is necessary to circulate the vessel volume to achieve a given homogeneity. The blend time is calculated from the assumption that vessel contents will be homogeneous after they have been recirculated z times by the mixer, where z is usually taken to be 4 or 5:

$$t_m = z \cdot \frac{V}{q} \qquad (5)$$

where V is the total liquid volume in the vessel, and q is the impeller discharge or pumping rate, which can be measured or estimated by calculation. This theoretical method is not as reliable as the dimensionless mixing time method, embodied by Equations (3) and (4), which is based on direct measurements and scale-up rules. The discrepancy arises because mixing is not just a function of the main impeller discharge flow, but also of the flow patterns generated.

Mixing for chemical reactions

For a chemical reaction to proceed, it is necessary to intimately mix the reactants down to the molecular scale. According to turbulence theory [3], however, there is a theoretical minimum size of vortex, called a micro-vortex, that is generated in a turbulent liquid. The size of the micro-vortices is a function of the viscosity of the liquid and the average specific power input. Turbulence cannot help to mix the reactants on a scale smaller than the micro-vortex. Mixing down to the molecular scale therefore relies on molecular diffusion of the chemicals within the micro-vortices.

The time for micro-mixing to occur is, in most cases, not significant compared to the macro-blend time. For example, an aqueous solution might typically have micro-vortices of 36 μm and a diffusion rate of approximately 2×10^{-9} m^2/s, so the micro-mixing time would be about 0.08 s. Engineers should, however, always consider whether micro-mixing might be an issue. In some cases it can be decisive, such as when competing consecutive reactions occur extremely quickly.

VISCOUS AND NON-NEWTONIAN FLUIDS

For high-viscosity liquids, the mixing regime changes from one in which turbulence dominates to one in which viscous drag forces dominate, and agitation throughout the bulk is by no means uniform. This non-uniformity can be made significantly worse if the

high-viscosity mixture also exhibits the anomalous flow properties of non-Newtonian rheology.

The most frequently encountered anomaly is shear thinning, where viscosity decreases with increasing shear rate according to:

Local viscosity, $\mu = K \cdot \gamma^{m-1}$ or **(6)**

Local shear stress $\tau = K \cdot \gamma^m$ **(7)**

where γ is the local shear rate, K is the "consistency factor" and m is the "flow index" which describes how strongly the apparent viscosity changes with shear rate.

Another frequently encountered anomaly is the so-called Herschel-Bulkley anomaly, in which case the mixture does not flow at all until a certain yield stress is exceeded. The rheological behavior of such complex mixtures can usually be described by the following correlation:

Local viscosity, $\mu = \dfrac{\tau_0 + K \cdot \gamma^m}{\gamma}$ **(8)**

where τ_0 is the yield stress at which flow begins.

For Newtonian fluids, τ_0 is zero, m is 1, and Equations (6) and (8) reduce to: viscosity, $\mu = K$, which is constant throughout the vessel. For non-Newtonian liquids, the effective viscosity of the mixture varies throughout the vessel. Shear rate will be highest near the impeller, and lowest near the vessel walls and the liquid surface. Therefore, for shear-thinning fluids, the apparent viscosity will be lowest at the impeller and highest at the wall (Figure 3). If an agitator is not correctly designed for these fluids, the mixture in regions near the walls can be completely stagnant. This is called the "cavern effect".

Designing for anomalies

The first step in assessing a design is to calculate the power and torque absorbed by the proposed agitator. To do this, it is necessary to know the local viscosity, which may vary with shear rate. The approach of Metzner and Otto [4] can often be used to predict local shear rate, γ, at the impeller region. They discovered that γ in the direct vicinity of an impeller is propor-

tional to the shaft speed.

$\gamma = k_{MO} \cdot N$ **(9)**

where k_{MO}, the Metzner-Otto constant, depends on the impeller design. Using the impeller shear rate, the local viscosity can be calculated from Equation (6) or (8). The local viscosity can then be used to determine the Reynolds number, Re, from which the impeller's power number, P_o can be derived. P_o is a function of Re. The absorbed power can then be calculated using:

$P = P_o \rho N^3 D^5$ **(10)**

Finally, the impeller torque, M_t, can be calculated from its speed and the absorbed power:

$M_t = \dfrac{P}{2 \pi N}$ **(11)**

The above describes mixing near the impeller. To be thoroughly blended, every part of the mixture must be in motion. The regions that move least, or not at all, are those most remote from the impeller. Fluid flow at the walls can be assessed using the "torque balance" assumption, which is that the torque transmitted from the impeller to the mixture must be balanced by the shear stress at the vessel wall, τ_w. This assumes that there are no other internals to complicate the model. Using this assumption, τ_w can be calculated for a cylindrical tank by:

$\tau_w = \dfrac{M_t}{C \cdot V}$ **(12)**

where V is the volume of the mixture and C is a factor that is impeller-specific and determined experimentally. For shear-thinning fluids without a yield stress, Equations (12), (6) and (7) can be used to calculate a local viscosity close to the tank wall or liquid surface. If the mixture has a yield stress, τ_0, it is also necessary to check whether the predicted shear stress at the wall is large enough to create motion ($\tau_w > \tau_0$). If it is not, there will be stagnant areas in the tank.

The above approach can also be applied to empirical scale up from pilot-scale trials to plant design. Trials in a vessel with a transparent wall
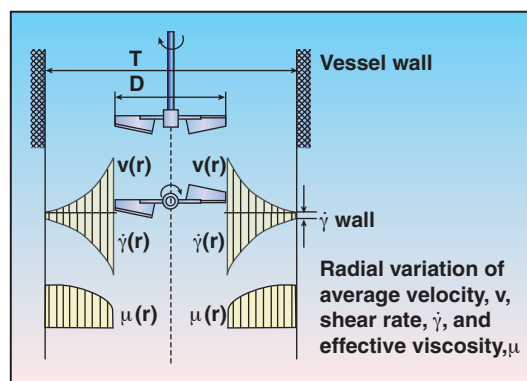
FIGURE 3. For shear-thinning fluids, the apparent viscosity will be lowest at the impeller and highest at the wall

and base can be used to observe the agitator speed at which good, overall movement is achieved. The required, minimum shear stress at the walls can then be calculated. This value can be used to scale up to a geometrically similar plant-agitator design that provides a similar flow pattern to that on the pilot scale.

Near-wall impellers

For Reynolds numbers as low as 50, it is possible to ensure good mixing using correctly designed, large-diameter impellers with a steep pitch. At lower Reynolds numbers, however, it is generally necessary to resort to impellers that run near to the vessel wall. Three common types are illustrated in Figure 4.

The anchor agitator is the simplest form of near-wall agitator and is commonly in use for high-viscosity applications. The vertical arms run close to the walls and shear the mixture as they pass through it. Satisfactory homogeneity is not, however, efficiently achieved since anchors create mostly tangential displacement of the fluid. Liquid in the center of the vessel experiences little movement and the zone around the agitator shaft is poorly mixed.

Helical ribbon and optimized-ribbon impellers achieve much improved mixing by displacing the liquid up (or down) the wall of the vessel. The liquid flows to the center of the vessel and is then drawn in the opposite direction along the axis to replace liquid that is displaced by the agitator. Even at extremely high viscosities of up to 1,000 Pa·s, the fluid can be circulated throughout the entire vessel. The mix-
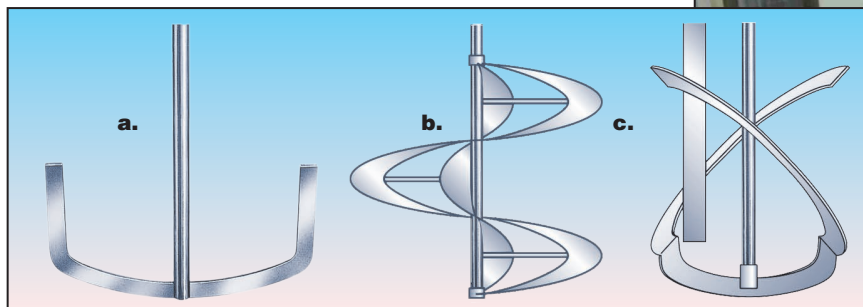
FIGURE 4. Agitators suited for low Reynolds numbers include the a) Anchor, b) Helical ribbon and c) Optimized ribbon
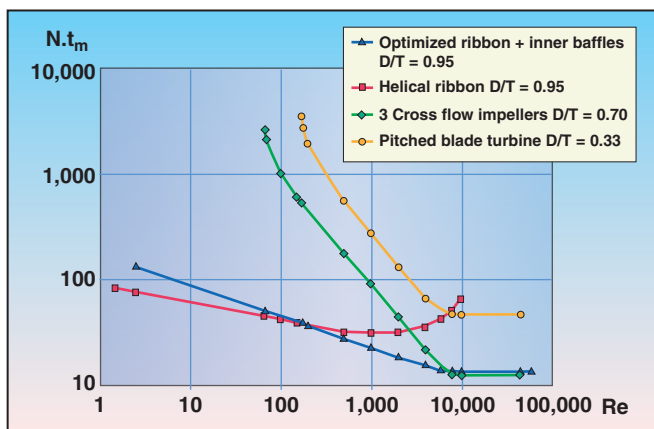
FIGURE 5. The blending time is dependent on agitator type

FIGURE 6. Pilot plant testing of high-viscosity slurries, such as this mineral-ore slurry with a yield stress, is highly recommended for good mixer-design performance

ture is stretched and folded, and is also sheared local to the impeller blades. Once sufficient stretching and folding has taken place, local molecular diffusion completes the micromixing.

The optimized ribbon has a simpler construction than the helical ribbon and is suitable for a wider range of viscosities due to its flexibility. This optimized impeller can be fitted with inner baffles that allow it to maintain good mixing performance even at high Reynolds numbers (low viscosities). Helical ribbons are poor mixers in non-laminar-flow conditions because the viscous drag of the mixture on the walls becomes insufficient to prevent bulk rotation. Figure 5 compares the dimensionless-blending-time, Nt_m, performance of several open and near-wall impellers over a wide range of Reynolds numbers.

The rule that the dimensionless mixing time is constant in geometrically similar systems applies not only for turbulent but also for laminar systems, when a positive-displacement impeller is used. To explain this phenomenon, one can consider that in both turbulent- and laminar-flow regimes, the amount of liquid displaced by the agitator is proportional to the number of times it turns and the circulation rate of the impeller. Hence, the blending time for a large-scale plant is calculated from the number of impeller revolutions found to be necessary in the pilot scale. In fact, larger vessels will generally have higher Reynolds numbers than smaller vessels containing the same mixture. The rule is, therefore, somewhat conservative.

High-viscosity slurry processing
Slurries that contain a high-solids concentration of small particles can exhibit non-Newtonian flow behavior. Such slurries are commonly encountered in the mineral-processing industry, where typical solids volume concentrations are 35 to 50%. In these applications, the settling velocities of the suspended solids are very low and the key mixing task becomes the blending of the highly-viscous, non-Newtonian mixture, rather than solids suspension. Tanks for such ore slurries can be up to 10,000 m³, and agitator powers can be correspondingly very large. It is not possible to accurately predict flow behavior in such a system by using theoretical correlations based on solids concentration and particle-size data. For high-viscosity slurries, tests with original product are indispensable to ensure good performance without overdesign (see Figure 6).

SUSPENSION OF SOLIDS
There are various objectives when using an agitator for suspension tasks. In washing processes or for solids dissolution, off-bottom suspension can be sufficient. In a process with continuous operation or where several vessels are arranged in series, more-stringent specifications must be made for the homogeneity of the suspension.

Figure 7 illustrates three degrees of suspension. To progress from one degree to the next requires an increase in power input via higher agitator speeds. In the first case, "on-bottom motion," there may be temporary local deposition of solids on the base, with occasional movement of the solids — large clear zones can occur in the

upper part of the vessel. In the second case of "off-bottom suspension," no particle comes to rest for longer than one second on the bottom of the vessel. This criterion, often used for large-scale, mineral-processing applications, defines the lowest mixing power at which the entire surface area of the particles is exposed to the liquid phase for chemical reaction or mass transfer. Zwietering [5] produced the following well-known correlation for minimum shaft speeds, n_{js}, at which "off-bottom suspension" occurs:

$$n_{js} = s \cdot v^{0.1} \cdot \left(\frac{g \cdot (\rho_s - \rho_l)}{\rho_l} \right)^{0.45} \cdot d_p^{\,0.2} \cdot c_w^{\,0.13} \cdot D^{-0.85}$$

$$(13)$$

The third case is "visually uniform suspension," which is defined as having no large clear zones. Settling of coarse particles may still occur for solids with a wide range of particle sizes.

Occasionally, for very demanding duties such as continuous-overflow operation, or manufacture of dispersions as end products, "uniform suspension" is required. In this case, particle concentration and size distribution are uniform throughout the vessel. This is very difficult to achieve unless the particle settling velocity is very low.

Design fundamentals

In addition to agitator parameters and the vessel geometry, the properties of both the liquid and the solid particles influence the fluid-particle hydrodynamics and, thus, the suspension. The important physical properties for agitator design are: the liquid density, the density difference between solids and liquid, the liquid viscosity, the average particle size and the volumetric concentration of the solids.

A single particle's free-settling velocity, v_s, is calculated by methods given in the relevant literature. The hindering effect on the settling process due to the presence of several particles is quantified by the following relation, where the exponent m is a function of the particle Reynolds number, and varies between 2.33 and 4.65:

$$v_{sh} = v_s \,(1 - c_v)^m \qquad (14)$$

where v_{sh} is the hindered settling velocity, v_s is the free-settling velocity and c_v is the volume fraction of solids.
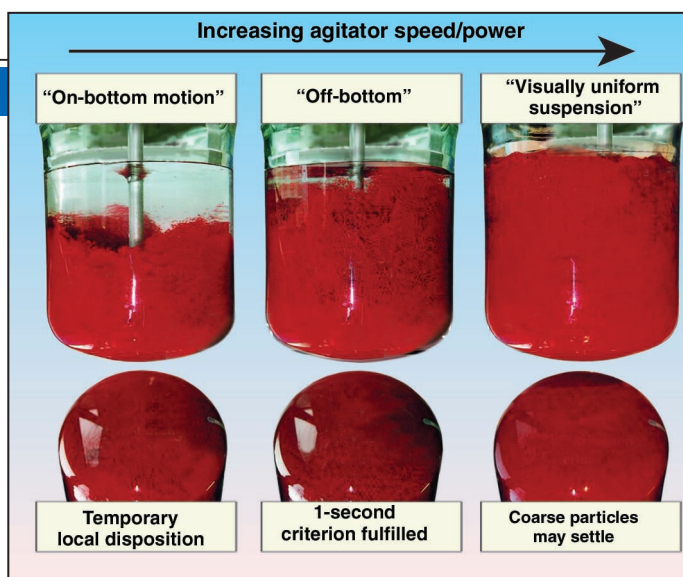
FIGURE 7. The degree of suspension desired for a given application depends on the objective. Here, three levels of suspension are shown, each requiring more power input than the one before
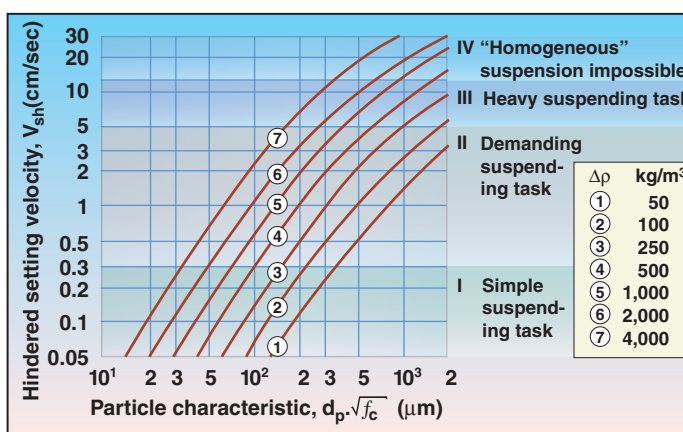
FIGURE 8. To describe the complex nature of suspension tasks, suspension duties are classified into four broad groups

If it is assumed that all solid particles in the liquid are distributed uniformly, and all simultaneously begin to settle under the effect of gravity, they release a "settling power", which can be quantified by the relation:

$$P_{settle} = v_{sh} \cdot c_v \cdot \Delta\rho \cdot g \cdot V \qquad (15)$$

where $\Delta\rho$ is the difference in density between solid and liquid. In order to maintain a defined degree of uniformity in the suspension, the agitator must provide a power input to the liquid that counteracts this settling power. The agitator power always amounts to a multiple of the settling power.

When one is using the above Equations (14) and (15), the choice of particle size that is used to calculate the free-settling velocity, v_s, is very important. In powders or slurries, the individual particles vary in size and shape. Choosing the largest particle size can result in a much higher agitator power than is required. From experience, reliable results are obtained with a design particle size that corresponds to a

value where between 80 and 90% pass through the mesh size.

Scale up

Scale up of suspension duties can be very complex. Various scale up criteria have been proposed based on the type of suspension needed, as discussed earlier. Specific process or product requirements can impose additional criteria for consideration. Some common complications include these:

• Solids with extremely wide particle size distributions — the fine particles affect the suspension of the large particles
• Very high-solids concentrations — particle interactions affect the apparent rheology
• Presence of small amounts of extremely large particles — impossible to suspend but must be moved around on the base of the vessel
• The presence of significant quantities of extremely small particles — these essentially behave as part of the fluid

To accommodate these considerations, solid-suspension duties are generally
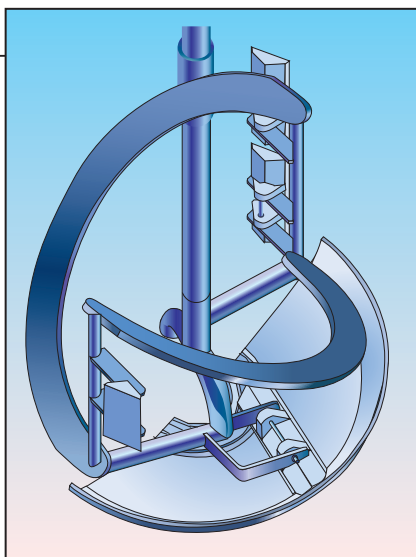
FIGURE 9. High viscosities can hinder heat transfer. Wall scrapers added to an agitator, such as this optimized ribbon, can help increase heat transfer substantially

classified into four broad categories on the basis of hindered-settling velocity (see Figure 8).

Type I tasks are simple suspending duties that are readily predicted because the liquid flows around the particles in simple, laminar flow. Type II are demanding suspension tasks where the fluid flow is more complex but predictable from empirical correlations — this category covers the majority of industrial applications. Type III are difficult or "heavy" suspension tasks, which probably involving large or heavy particles. In this category, scale-up is usually based on pilot-scale tests. For type IV tasks, uniform suspension is no longer attainable, as they require very high liquid velocities that cannot be achieved economically.

In general, for solids suspension, the agitator power requirement is scaled up as a function of tank diameter according to the following equation:

$$\frac{P}{V} \propto D^X \qquad (16)$$

where X can vary from 0 to -1 depending on the type of suspension duty. For example, if particles with a high hindered-settling velocity must be kept in visually uniform suspension (Type III in Fig. 8), a criterion close to constant specific power input P/V must be used ($X = 0$). For solids with low hindered-settling velocities, a criterion closer to constant tip speed ($X = -1$) can be used (Type I in Fig. 8). These examples illustrate how the specific power inputs required to achieve "suspension" can vary hugely on an industrial scale.

To achieve homogeneous suspension of true suspensions, high-efficiency, axial-pumping impellers are generally used because of the lower power inputs required.

Special considerations
Some considerations for specific suspension applications are as follows:
- In many solids-suspension tasks, especially in the minerals-processing industries, abrasion can be a significant issue. In this case, lower velocities may be required to limit abrasion. An increased impeller size can compensate for the lower velocities.
- During power failures, sediments can quickly build up in suspension tanks. Impellers are often designed to withstand attempted restarts, while submerged in a densely settled slurry. In some instances, "restart in slurry" becomes the key design criterion.
- Solids suspension and gas dispersion commonly occur simultaneously in the chemical- and minerals-processing industries. The presence of gas affects the performance of the impeller and the ability of the fluid to suspend solids. Likewise, the presence of solids affects gas dispersion. Both must be taken under consideration.

Additional factors that can have a severe impact on the agitator design include: the shape of the base of the vessel (dished, conical or flat); the location of the draw-off point; and the aspect ratio (H/T) of the tank. Continuous processes are most sensitive to such factors.

LIQUID-LIQUID DISPERSIONS
Many industrial processes require the dispersion of one liquid into another, immiscible liquid. Unstable dispersions are created in processes such as solvent extraction (to provide a large surface area for mass transfer), and in dispersion polymerizations (to create the required size distribution of polymer particles or provide a heat sink).

In the case of solvent extraction, the dispersion must be fine enough to allow rapid mass transfer and keep the size of the extractor down, while at the same time, the dispersion should be coarse enough to enable rapid

separation of the phases afterwards. Mixer-settler systems can generally use finer dispersions than solvent-extraction columns. The former systems are generally better suited to slower extractions, and the latter to fast extractions. Typical industrial applications of solvent extraction are refining of metals from acid leach liquor and purification of pharmaceutical products.

Surfactants are often used to aid formation of the correct dispersion size and to hinder recoalescence, in both temporary and stable emulsions. The interaction of the various factors affecting dispersion processes is very complex and nearly always necessitates mixing tests. The main factors affecting dispersion include these:
1. Physical properties of the liquids (especially interfacial tension)
2. Effects due to mutual solubility or reaction
3. The presence of surfactants, and
4. The turbulence and shear created by the agitator.

Most liquid-liquid dispersions are formed in the turbulent regime and droplet sizes of between 5 and 500 μm can be achieved. Typically, for a specific power input of 1 W/kg, droplet sizes between 100 and 150 μm can be expected.

Droplets break when shear stresses induced in the droplets by turbulence cause sufficient deformation to overcome the stabilizing effect of surface tension. Minimum droplet size is governed by the size of the micro-vortices, and is generally between 3 and 5 times the size of the micro-vortices. When finer dispersions are required, rotor-stator devices that produce a zone of very high energy dissipation can be used. Stabilizers can be added to ensure that fine droplets produced in this high-energy zone do not recoalesce in the bulk. For even finer emulsions (from 5 μm to 50 nanometers), high-pressure homogenizers with local energy dissipations of over 500 W/kg are used.

Scale up
Due to the complexity of the interacting factors that affect droplet formation, actual performance is usually measured in pilot trials. Because of
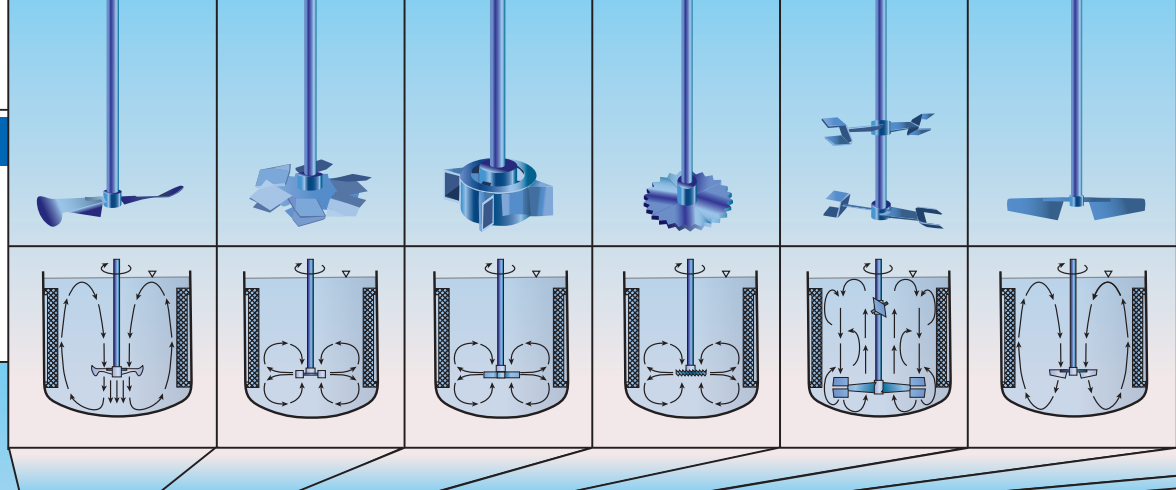
Preferred arrangement and primary flows							
Agitator duty:							
Blending	T				Tr	NN Tr T	L NN Tr T	L NN Tr T
Suspension	T				Tr	T		NN Tr T
Dispersion gas/liquid		T	T					
Dispersion liquid/liquid				Tr T		Tr		L NN Tr T
Dispersion solids/liquid				Tr T		Tr		L NN Tr T
Heat transfer	T	T	T		Tr	NN Tr T	Tr L NN	L NN Tr T
Flow regime/ viscosity	T = Turbulent, low viscosity L = Laminar, high vicosity Tr = Transition region, medium viscosity NN = Non-Newtonian flow properties							

FIGURE 10. Modern-day impeller systems offer a great range of choice to help meet the mixing challenges facing the CPI

the sensitivity of surface properties to minor levels of contaminants, pilot trials should use actual chemicals from the process plant rather than laboratory-quality reagents. Scale up to production should follow rules that are based on the mechanism (bulk or local turbulence) which was used to achieve the required performance.

HEAT TRANSFER
Stirred vessels are rarely used purely for heat transfer because equipment with much more-efficient heat exchange is available. Heat transfer is, however, a critical unit operation that stirred vessels must be capable of performing. A typical batch process, for example, could comprise the heating of the bulk to reaction temperature, blending and cooling during the reaction to remove the reaction heat, further heating to evaporate a solvent, and finally cooling down to near ambient temperature before the product is discharged.

Heat transfer performance is governed by:
- The flowrate and temperature of the utilities, and heating/cooling medium
- The heat-transfer coefficients on the product side and the utility side
- The type and contact area of heat exchange surfaces

Some of these factors may need to be modified in order to achieve required heat fluxes. In many reactors, the vessel, itself, provides insufficient surface area and it is necessary to install additional heat-transfer surfaces.

Design
Heat-transfer coefficients in jackets, coils and plate heat exchangers can be predicted from well established correlations [2]. Since the range of utility fluids encountered is quite small and their physical and thermodynamic properties are generally well documented, the accuracy of predicted film coefficients is often very good.

On the process side, however, prediction of the heat transfer coefficient is based on a general equation of the form:

$$Nu = \frac{h_P \cdot T}{k_P} = C \cdot \text{Re}^{2/3} \cdot \text{Pr}^{1/3} \cdot \left(\frac{\mu}{\mu_W}\right)^{0.14} \quad (17)$$

where the constant C, which depends on the impeller type and size, can be found in the literature or derived by measurements. The viscosity

References
1. Mersmann, K., *Chemie Ingenieur Technik*, Vol. 23, pp. 953-956, 1975.
2. EKATO (Ed) "Handbook of Mixing Technology", EKATO Rühr- und Mischtechnik, Schopfheim, Germany, 2000.
3. Kolmogorov, A. N., Die lokale Struktur der Turbulenz in einer inkompressiblen zähen Flüssigkeit bei sehr großen Reynoldsschen Zahlen. Goering, H. (Ed) "Sammelband zur statistischen Theorie der Turbulenz", Akademie-Verlag, Berlin, Germany, 1958.
4. Metzner A.B. and Otto R.E., Agitation of non-Newtonian fluids, *AIChE J*, Vol. 3, No. 1, pp. 3-10, 1957.
5. Zwietering T.N., *Chemical Engineering Science*, Vol. 8, pp. 244-253, 1958.
6. DeLaplace, G., others, Numerical simulation of flow of Newtonian fluids in an agitated vessel with a non standard helical ribbon impeller, "Proceedings 10th European Mixing Conference", Elsevier 2000.
7. Zlokarnik, M., Rührtechnik: "Theorie und Praxis", Springer-Verlag, Berlin, Germany, 1999.
8. "Autorenkollektiv, Mischen und Rühren, Grundlagen und moderne Verfahren für die Praxis," VDI-GVC, 1998
9. Perry, R.H. and Green, D.W., "Perry's Chemical Engineers' Handbook," 7th ed., McGraw-Hill, 1997.
10. Paul E.L. and others (Eds.). "Handbook of Industrial Mixing: Science and Practice," John Wiley Inc, New Jersey, 2004.
11. Kraume, M. and Zehner, P, Experience with experimental standards for measurement of various parameters in stirred tanks, *TransIChemE*, Vol. 79, 2001.
12. Mezaki, R., others, "Engineering Data on Mixing", Elsevier 2000.
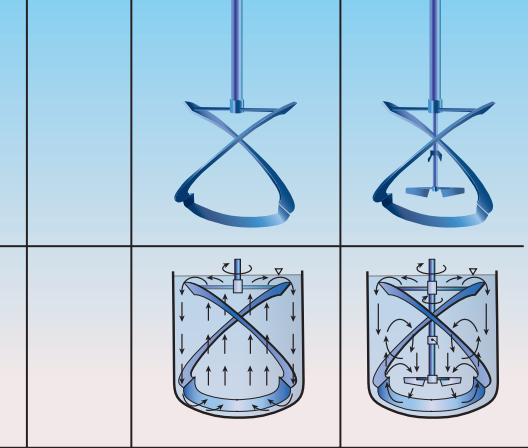
term represents the effect of viscosity changes in the boundary layer at the heat transfer surface. For internal components, such as coils, the value of C differs from that for the vessel wall. A further complication is that physical properties of the vessel contents are often changing during processing — due not only to changing operating conditions, but also to physical or chemical processes that are occurring.

The design engineer is often faced with the problem that the required heat flux cannot be achieved with existing conditions such as heat-exchange area or utility temperatures. Here are some common reasons and suggested remedies:

Effect of scale: Heat generated by a reaction increases proportionally with the volume of the mixture ($Q_R \propto V$). For geometrically similar equipment, however, the area for heat exchange increases proportionally to the volume raised to the power 2/3 ($A \propto V^{2/3}$). A reaction that was simple to control through wall cooling at the pilot scale may therefore require additional heating/cooling surfaces in order to increase the surface area lost on scale-up. Tube bundles or coils mounted in the reactor are often used. In cases where very high surface areas are required, vertical plate heat exchangers mounted approximately radially in the reactor can provide as much as 25 m²/m³.

Viscosity: The material property that commonly governs heat exchange is viscosity, μ. In industrial applications, the viscosity of mixtures can range widely, such as from 0.1 to 10^6 mPa·s. Because the heat transfer coefficient is proportional to $\mu^{-1/3}$, the coefficients for high-viscosity fluids are much lower than those for low-viscosity applications. An increase in viscosity can also mean that a liquid is no longer in the turbulent regime, so temperature differences within the mixture

would increase. These issues are best addressed by careful selection of the agitator to ensure good homogeneity across the range of operating conditions. Consideration should be given to close-clearance impellers, such as the helical ribbon and the optimized ribbon, which increase shear and therefore heat transfer at the wall. In difficult cases, the use of wall scrapers (Figure 9) can further increase heat transfer by a factor of up to 10.

Wall fouling: Fouling is a risk in many processes. In cooling crystallization, for example, the liquid can become supersaturated in the wall-boundary layer. Scrapers may help, but they are subject to wear in the solid layer. In this case, it is better to change the process to use cooling by evaporation. This process may require operation under vacuum, which brings another consideration — flow velocities at the liquid surface must be fast enough to avoid increased local over-concentrations that can lead to crystal nucleation. Impellers with good axial-pumping efficiency are used to maintain concentration and temperature homogeneity and ensure uniform crystal growth.

Pressure vessels: Wall thicknesses can become the limiting factor for heat transfer, especially with stainless steel vessels that contain a low-viscosity fluid. It is not possible, for example, to achieve an overall heat transfer coefficient above 300 W/m²K if the wall thickness is 50 mm. This overall heat-transfer coefficient cannot be improved by more intense agitation. Similarly, deposits on the utility side of the jacket or heat-transfer surfaces can build up and cause poor thermal conductivity. Only a few millimeters of deposit can have a detrimental effect on the heat transfer. Regular cleaning procedures may be required in these cases.

Agitator-power input: The power input, *P*, has a relatively small influence on heat transfer. The process-side coefficient is proportional to $P^{0.22}$, so doubling power will increases the film coefficient by only 16%. When cooling viscous fluids, increasing agitator power can actually have a negative effect on cooling rate because the agitator-power input to the fluid, which is converted to heat, can be quite significant.

SUMMARY

Careful consideration of operating conditions and fluid characteristics is needed to effectively design and scale up mixing systems. A broad body of information in this field is available for guidance, and a wide range of modern impellers (Figure 10) is available to meet mixing challenges. ∎

Edited by Dorothy Lozowski

Author

Werner Himmelsbach is manager of EKATO RMT's R&D Department (Käppelemattweg 2, 79650 Schopfheim, Germany; Phone: +49 7622 29227; Fax: +49 7622 29454; Email: him@ekato.com). He has 25 years experience in process design and development, plant design and maintenance having previously worked for major international manufacturers of speciality chemicals and pharmaceuticals. Himmelsbach holds a masters degree in chemical engineering from the University of Karlsruhe (Germany).

Wolfgang Keller is senior process engineer in EKATO RMT's R&D Department (Käppelemattweg 2, 79650 Schopfheim, Germany; Phone: +49 7622 29468; Fax: +49 7622 29454; Email: kew@ekato.com). He has over 10 years experience as plant engineer and in process development, having previously worked for an international manufacturer of speciality polymer films. Keller holds a masters degree in process engineering from the University of Karlsruhe (Germany).

David A. Houlton is senior process engineer responsible for reaction consultancy and design in EKATO RMT's R&D Department (Käppelemattweg 2, 79650 Schopfheim, Germany; Phone: +49 7622 29512; Fax: +49 7622 29454; Email: hda@ekato.com]. He has over 25 years experience in process plant design and research, having previously worked for major international manufacturers of speciality and bulk chemicals. Houlton holds B.Eng. and M.Phil. degrees in chemical engineering from the University of Bradford (U.K.). He is a fellow of the U.K. IChemE and a member of the German VDI.

Mark Lovallo is the Technology Manager for EKATO Corporation (Ramsey, NJ 07446; Phone 201-825-4684). He heads the North American Technology center for the EKATO Group also in NJ. Mark has previously worked for Union Carbide as a research engineer in the polyolefins catalyst division. He holds a Ph.D. in chemical engineering from the University of Massachusetts Amherst.