Sizing Impellers for Agitated Aerobic Fermenters

GREGORY T. BENZ, P.E.
BENZ TECHNOLOGY INTERNATIONAL, INC.

Optimize fermenter performance and cost by sizing the proper impeller system for your application.

PROPER DESIGN OF IMPELLERS FOR INDUCING gas dispersion in fermenters is a widely published topic that has typically emphasized such issues as agitator power, liquid holdup and flooding *(1)*. Recent articles have focused on the advantages that various impeller configurations may have in fermenters *(2, 3)*. This article describes the merits of different relative impeller sizes and explains how design parameters, such as the ratio of impeller diameter (D) to tank diameter (T) or D/T, affect a fermenter's performance and cost. Illustrative examples are used to quantify these differences.

Mass-transfer principles

In an aerated tank, the purpose of the impeller (or agitator) is to disperse gas into the liquid while providing enough mixing to prevent high gradients of oxygen and localized concentration of nutrients. Normally, the resistance to mass transfer occurs primarily at the liquid/gas interface, although, there are also resistances within the gas and within the bulk liquid. A simple relationship can be written that relates the gas mass-transfer rate to the interfacial area, liquid film resistance and driving force:

$$M = k_l a\,(C_{sat} - C_l) \tag{1}$$

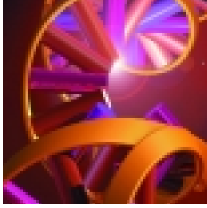
where: M is the mass transfer rate in mg/L-s; $k_l a$ is the overall mass-transfer coefficient in s^{-1}; C_{sat} is the concentration of dissolved oxygen (*DO*) in the liquid at saturation in mg/L ; and C_l is the concentration of *DO* in the liquid in mg/L. For a tall vessel, it is more realistic to use the log-mean driving force of oxygen concentrations, as both C_{sat} and C_l may vary substantially from the top to the bottom of the fermenter. However, the expansion of Eq. 1 to the log-mean form is beyond the scope of this article.

The overall mass-transfer coefficient, $k_l a$, may be correlated to agitator power and gas flowrate as follows:

$$k_l a = A(P_g/V)^B(U_s)^C \tag{2}$$

where: A, B and C are dimensionless empirical constants; P_g is the power required from the agitator to achieve a desired M in an aerated fermenter (or impeller power draw in a "gassed" system) in kW; V is the liquid volume in m^3; and U_s is the superficial gas velocity in m/s (U_s may also be defined as the actual gas volumetric flowrate (Q_g in m^3/s) at the midpoint of the liquid level, divided by the tank cross-sectional area).

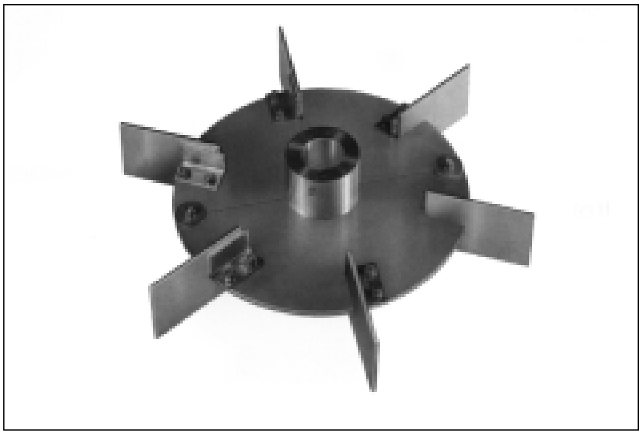
If sufficient data are available, proper agitator design would consist of defining the required M, performing a mass balance to determine the available concentration driving force for mass transfer based on an assumed Q_g , calculating the required $k_l a$, and solving



for required P_g. This could be done for multiple gas flowrates to optimize P_g *(4)*. It is worth noting that Q_g and M may vary during fermentation. Thus, P_g is normally determined by the maximum required M.

Impeller systems — practice and principles

Aerobic fermentation was commercialized in the 1940s, with its first product being penicillin. At that time, radial impellers, such as the Rushton impeller (Figure 1), were used and remained popular well into the 1970s. These impellers generally had six flat blades affixed to a disc at a 90-deg angle. As fermenters became larger in size, the staged flow pattern created by the use of multiple radial turbines resulted in excessive *DO* and nutrient concentration gradients. In one case where a 150-m^3 fermenter was equipped with four radial turbines, the resulting *DO* was about two orders of magnitude higher at the bottom of the fermenter than at the top.



■ Figure 1. The Rushton turbine — a first-generation radial impeller — is used for gas dispersion. It has lower gas-handling and gas-loading characteristics than succeeding impellers in this family.
Photocredit: Chemineer, Inc.

■ Figure 2. The pitched-blade turbine — a first-generation pitched-blade impeller — is used primarily for pumping. It has problems in gas-dispersion services because it allows gas bubbles to coalesce quickly. More recently, it has been replaced by hydrofoil axial-flow impellers.
Photocredit: Chemineer, Inc.

Nomenclature

A = coefficient of Eq. 2, dimensionless
a = interfacial (gas–liquid) area per unit volume, m^{-1}
B = exponent for P/V in Eq. 2, dimensionless
C = exponent for U_s in Eq. 2, dimensionless
C_l = concentration of dissolved oxygen in liquid, mg/L
C_{sat} = concentration of dissolved oxygen in liquid at saturation, mg/L
D = impeller diameter, m
DO = dissolved oxygen concentration, mg/L
g = gravitational acceleration; 9.8 m/s^2 at sea level
k_la = overall mass transfer coefficient, s^{-1}
M = mass transfer rate, mg/L-s
N = agitator-shaft rotational speed, s^{-1}
N_a = aeration number, Q_g/ND^3, dimensionless
N_{Fr} = Froude number, N^2D/g, dimensionless
N_{Re} = impeller Reynolds number, $D^2N\rho/\mu$, dimensionless
N_P = impeller power number, $P_g/\rho N^3D^5$, dimensionless; N_P is also $P_u/\rho N^3D^5$ for ungassed conditions as in Eq. 4
N_q = impeller pumping number, Q/ND^3, dimensionless
P_g = agitator power required to achieve a desired mass-transfer rate in an aerated vessel, kW (*i.e.*, "gassed" agitator power)
P_u = agitator power required to achieve a desired mass-transfer rate in an unaerated vessel, kW (*i.e.*, "ungassed" agitator power)
Q = impeller liquid pumping rate, m^3/s
Q_g = gas volumetric flowrate, m^3/s
S = impeller tip speed, πND, m/s
T = tank inside diameter, m
U_s = superficial gas velocity, m/s
V = ungassed liquid volume, m^3
VVM = volume of gas per unit volume of liquid per min at 1 atm and 0°C, min^{-1}

Greek letters
ρ = liquid density, kg/m^3
μ = liquid viscosity, Poise (kg/m-s)
θ = blend time, s or min

Because of such gradients, mixer manufacturers and end users began to experiment with impeller systems that combined axial and radial impellers. At first, the pitched-blade turbine (Figure 2) was used as an upper impeller. But, this design was not successful due to a tendency of the gas to coalesce at the top of the vessel and because there was a general lack of knowledge among the designers about proper power distribution. For example, the gas would coalesce because the pitched blades, while properly positioned above the hydrodynamic stall angle (*i.e.*, the angle above which turbulent boundary layers separate from the laminar sublayer), were not at a high enough angle to impart sufficient power to adequately redisperse the gas.

Later, various hydrofoil impellers were tried. Hydrofoils are axial flow impellers that operate below the stall angle of the blade. Some installations with axial impellers were very successful; others were not. Power distribution was found to be a critical parameter in the impeller system's mass-transfer performance. Too little power in the axial impellers caused gas to coalesce near the top of the fermenter. Too much power in the top impellers resulted in larger gas bubbles at the bottom and, consequently, a failure to take full advantage of the potentially higher driving force.

Later, high-solidity ("solidity" refers to the degree of occlusion of an impeller's swept circle by its blades) axial flow (HSAF) impellers (Figure 3), which typically occlude 70% or more of their swept circle, were added to the arsenal. These impellers were more forgiving of power split (proportion of total power allocated to each impeller) and high Q_g.values *(3).* In a down-flow arrangement (*i.e.*, when liquids are pumped downwards along the shaft axis and then upwards along the walls of the vessel), HSAF impellers are typically used in conjunction with a radial turbine. In an up-pumping arrangement (*i.e.*, when liquids are pumped upwards along the shaft axis and then downwards along the walls of the vessel), they can be used with or without a radial-flow turbine, depending on Q_g.

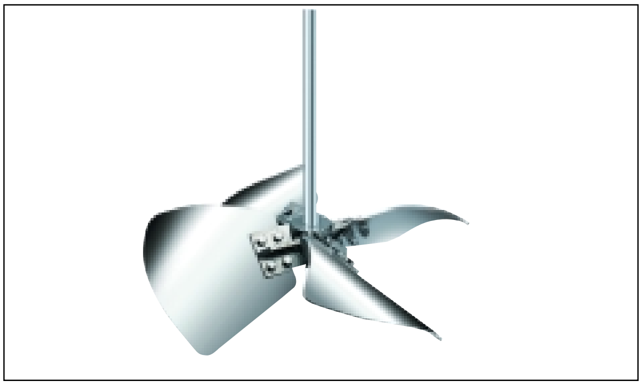
Meanwhile, the design of radial impellers was improved, starting with the invention of concave radial turbines (Figure 4). Concave radial turbines demonstrate an overall higher gas-handling capacity, less sensitivity of P_g to Q_g, and an increase in the fermenters' mechanical stability *(2, 5, 6),* compared with non-concave impellers.

All of the impellers mentioned can be fabricated for small and large fermenters. The use of axial and radial impellers on the same shaft has been successfully tried in fermenters as large as 450 m^3 *(6).* Up-pumping axial impellers have also been advocated in large vessels, due to their higher gas-handling capacity and mechanical stability *(7).* In addition, these impellers can be fabricated of any machinable material, such as the popular Type 316 stainless steel. More corrosion-resistant alloys are used when necessary.

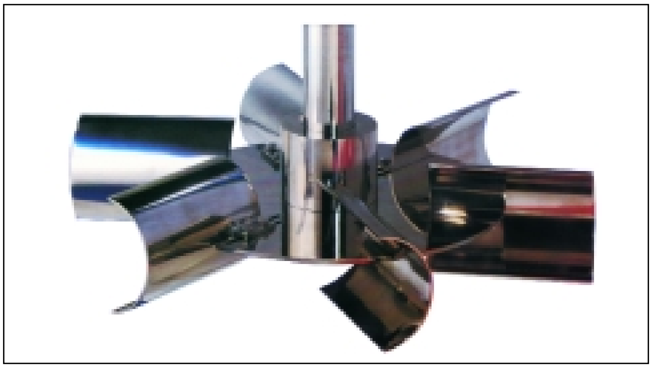
Current fermenter design practices include the use of lower-radial and upper-axial impellers *(3).* The lower-radial impeller has a concave blade design, either of proprietary shape or a shape similar to that of the impellers tested by John Smith *(8).* Upper narrow-blade hydrofoil impellers create more pumping motion and will not cause as much of a drop in P_g at low gas velocities (U_s < 0.03–0.04 m/s). An advantage of the HSAF impellers is that they will not suffer a significant drop in P_g at higher Q_g values. The measurable drop in P_g upon introduction of gas into a rotating impeller system is commonly referred to as the gassing factor, P_g/P_u, defined as the ratio of the agitator power in an aerated vessel (P_g or "gassed" agitator power) to the agitator power in an unaerated vessel (P_u or "ungassed" agitator power).

Most of the time, the upper impellers are up-pumping. This arrangement is more mechanically stable than down-pumping impellers. In more tangible terms, the vibration velocities created by up-pumping impellers are 50–60% of the vibration velocities created by down-pumping impellers. Also, at low-to-medium values of Q_g, up-pumping upper impellers have a lower drop in P_g due to the presence of gas. Up-pumping impellers also have a shorter gassed blend time (θ) — or time required to reach a certain degree of solute concentration attenuation (typically defined as ±1%) after the solute is dosed.

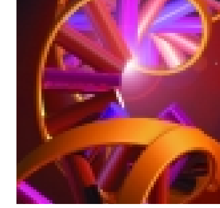
As mentioned earlier, the introduction of gas into a rotating impeller system generally causes the power required to achieve a desired M to drop. Thus, the gassing factor must be taken into account during fermenter design in order to: (1) size the impellers so that they draw the power required (*i.e.*, meet P_g specification) to achieve optimal mass transfer in the gassed condition; (2) size the gear drive and motor (*i.e.*, specify P_{motor}) for whatever mini-



■ Figure 3. The high-solidity axial-flow impeller — a third-generation axial impeller — enhances axial mixing in a fermenter in the presence of very high gas flowrates.
Photocredit: Chemineer, Inc.



■ Figure 4. The concave-blade radial impeller — a second-generation radial impeller — is ideal for gas dispersion. With higher gas-handling capabilities and gas-loading characteristics than its predecessors, this impeller has been the subject of numerous design configurations that have produced even more efficient gas-dispersion impellers.
Photocredit: Chemineer, Inc.



mum Q_g is required. Different impeller styles and sizes have different degrees of power drop (different losses in P_g) as a function of Q_g.

Power split (the amount of power invested in upper vs. lower impellers) is another crucial factor in fermenter design. If too little power is invested in either the upper axial or lower-radial turbines, gas will coalesce in the region of low power. Large gas bubbles reduce the $k_l a$ and thus, impede mass transfer. For competitive reasons, the exact power split is normally not disclosed by agitator manufacturers. For very viscous fermentation solutions, uniform blending of the liquid may be a paramount concern. For such cases, a multiplicity of up-pumping axial impellers with large *D/T* values may be required.

Impact of *D/T* at constant Q_g

The effects of *D/T* on fermenter performance are often calculated for comparative purposes under the conditions of constant Q_g vs. variable Q_g. The outcomes can differ significantly.

For the case of constant Q_g, several observations may be made regarding impeller systems: If impellers of different sizes (*i.e.*, different *D/T* ratios) but equal style and P_g are used, *N* would be lower for the larger *D/T* systems. As *D/T* increases at constant P_g, the impeller tip speed (*S* in m/s) decreases, resulting in a lower maximum shear rate imposed by the agitator.

More specifically, the author has found that:

1. If one ignores changes in the power number ($N_P = P_g/\rho N^3 D^5$) and P_g/P_u, *S* is proportional to *D/T* raised to the power of –2/3 ($S \propto D/T^{-2/3}$). If the organism is shear-sensitive, the effect of lower shear due to lower *S* may lead to a lower incidence of cell death. If the organism tends to form clumpy colonies, the added shear associated with a higher *S* or a smaller *D/T* may actually help mass transfer by breaking the colonies into smaller units.

2. Higher *D/T* ratios result in lower *N* values. Ignoring changes in N_P and the gassing factor, *N* is proportional to *D/T* raised to the –0.6 power ($N \propto D/T^{-0.6}$). This lower *N* results in lower average shear imparted to the fluid.

More importantly, a lower *N* at constant P_g means higher torque, which translates into a more-expensive gear drive, larger shaft and impellers, and more mechanical support for the agitator system. The author has found that agitator cost is roughly proportional to torque raised to the 0.8 power in large fermenters. "Large" as used here means a fermenter working volume above approximately 200 m³ or an impeller motor larger than 200 kW.

3. Larger impellers pump more liquid. Higher turnover may aid in the discharge of CO_2 at the surface of the liquid. The impeller pumping number, N_q, decreases with increasing *D/T*. The net effect is roughly that the liquid flowrate created by the impeller(s), *Q* in m³/s, is proportional to *D/T* raised to the 1.9 power ($Q \propto D/T^{1.9}$).

4. θ will be slightly shorter at larger *D/T* ratios. This is a generic observation. Correlations involving θ are, in fact, specific to impeller type; so are the effects of *D/T* on impeller performance. However, over the range of *D/T* ratios usually found in fermenters (*D/T* = 0.2– 0.6), θ at constant P_g will be up to10% faster at large *D/T* ratios than at small *D/T* ratios.

5. The heat-transfer coefficient (*htc*) also increases with *D/T*. Just how much it increases depends on the design of the heat-transfer surface (*e.g.*, helical coils, jacket, vertical tubes or panel coils). Generally, the *htc* is proportional to *D/T* raised to a power ranging from 0.95–1.05. The reader is cautioned against designing the agitator specifically for heat-transfer requirements. It is better to increase the heat-transfer surface area or the temperature differential if heat transfer must be improved. Furthermore, large *D/T* ratios may prevent use of extensive coils or vertical tube bundles in the vessel, due to mechanical interference.

6. Gas-handling capacity, quantified in terms of Q_g and defined as the highest gas flowrate tolerated by the system prior to flooding, increases with *D/T*. Sensel's work *(5)* suggests that at constant P_g, the gas-handling capacity before flooding is proportional to *D/T* raised to the –0.4 power ($Q_g \propto D/T^{-0.4}$). If large gas flowrates are anticipated, Q_g may dictate the required *D/T*.

7. Mechanical parts, such as gears, roller bearings, steady bearings and shaft seals, will last longer at lower *N* values associated with large *D/T* ratios, assuming the same service factors and stresses are used.

From the above, one may deduce that large *D/T* ratios are favored at constant gas flowrates. The higher cost of such systems is the main drawback.

Impact of *D/T* at variable Q_g

In real fermenters, variable Q_g values are used to avoid wasting compressor power during the early stages of fermentation. Q_g typically begins at a low value because the cell population is low and not much mass transfer is needed. As the cell population increases, Q_g is ramped up. This is followed by a period of constant Q_g. Late in the batch, Q_g may be reduced for various reasons, including the need to control product distribution. A change in Q_g directly affects P_g and the impeller's mass-transfer performance.

P_g/P_u is a function of the impeller Reynolds number ($N_{Re} = D^2 N \rho \mu$), the aeration number ($N_a = Q_g/ND^3$) and the Froude number ($N_{Fr} = N^2 D/g$) *(5)*. As *D/T* increases at constant P_g, N_a and N_{Fr} decrease. P_g/P_u changes more rapidly as a function of Q_g at low values of N_a and N_{Fr} than it does at high N_a and N_{Fr} conditions. This means that P_g is more constant at low *D/T* ratios, which can affect the impeller's mass-transfer performance and the required motor power, P_{motor}. This concept is best illustrated by example *(9)*.

Putting principles into practice

Normal practice when specifying fermenter agitators is to define motor size, usually by specifying P_{motor} and the maximum and minimum values of Q_g. In the fermentation industry, Q_g is frequently expressed in terms of the volume of gas per volume of liquid per minute (*VVM*, min^{-1}) at 0°C and 1 atm. This is equivalent to specifying the molar air flowrate per unit volume of liquid. The stoichiometric *VVM* requirement depends on the metabolic requirements of the microorganisms and their population density in the broth. The actual *VVM* requirement must be higher than the stoichiometric flowrate, since it is not possible to transfer 100% of the available oxygen. Common fermentations require a peak *VVM* of approximately 0.5–1.0 min^{-1}, with allowance for about 50% oxygen transfer efficiency. However, some fermentations require a peak *VVM* that is smaller than the approximated range, while others require much higher values.

N and *D*/*T* are normally selected by the agitator manufacturer. To avoid overloading the motor, impellers are designed to consume a maximum of 90% of P_{motor} at the minimum Q_g. In other words, P_g/P_{motor} should be no more than 0.9. At the maximum Q_g, the power draw of the impeller when gas is flowing through the vessel,, P_g in kW, is less than it is at the minimum Q_g. In addition, when comparing two different *D*/*T* ratios at the maximum Q_g, the larger *D*/*T* will have a lower P_g Hence, a smaller $k_l a$ value will exist at the maximum Q_g, even though P_g will be the same at the minimum Q_g.

An alternate practice is to specify the required P_g at the maximum Q_g, and let the vendor specify the *D*/*T* and P_{motor}. In this case, the impeller's mass-transfer performance ($k_l a$) and P_g at the maximum Q_g will be the same, but a large *D*/*T* will require a higher P_g at the minimum Q_g, necessitating a larger P_{motor}, or perhaps specifying a minimum Q_g that is higher than that originally specified. Either option potentially uses more energy.

The following examples quantitatively describe a typical fermenter and compare calculated values of P_g and *D*/*T*. The calculations were performed using proprietary software provided by Chemineer, Inc. The sidebar illustrates how this procedure may be done by hand.

For the fermenter, $T = 5$ m and $V = 200$ m^3. The process temperature = 40°C. The fluid has sufficient ionic strength to be essentially non-coalescing for mass-transfer calculation purposes. Its specific gravity is 1.02 and its viscosity, μ, is 5 mPa-s.

The impeller system consists of a lower concave-bladed disc turbine (180-deg concavity) and two up-pumping, high-solidity hydrofoil impellers. For simplicity, the impellers' diameters are assumed to be equal. But this is not always the case for commercial systems. Moreover, while commercial *D*/*T* values typically range from 0.2 to 0.6, the values selected for calculations in this article are (the more common) $D/T = 0.3$ and 0.5, respectively referred to as "smaller" and "larger" impeller systems.

Calculation of impeller power under gassed conditions.

Most agitator manufacturers use proprietary impellers for which specific power-draw characteristics have not been published. However, the same general calculation methods are used for all impellers, with specific correlations dependent on specific impeller designs. In general, the procedure consists of three steps:

1. Calculate the ungassed power, P_u, from the impeller power number relationship:

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

Eq. 3 may be rearranged to solve for the ungassed power:

$$P_u = N_P \rho N^3 D^5 \qquad (4)$$

2. Calculate the gassing factor using Eq. 5 This ratio is a function of $N_a = Q_g/ND^3$, the *D*/*T* ratio, N_{Re} and N_{Fr} and varies with each impeller type.

$$P_g/P_u = f(N_a, D/T, N_{Re}, N_{Fr}) \qquad (5)$$

3. Calculate the actual invested power required by the impeller in a gassed vessel using Eq. 6. The term in parenthesis is treated as a single variable and is calculated using Eq. 5.

$$P_g = P_u(P_g/P_u) \qquad (6)$$

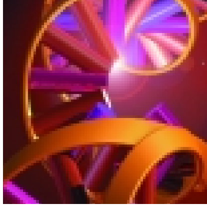
The methodology described above will be used to determine the power draw, P_g, for a Rushton turbine of standard design. Assume $D = 1.0$ m, $N = 100$ rev/min (1.67 rev/s), $T = 3.0$ m, $\rho = 1{,}000$ kg/m^3 and $\mu = 1$ mPa-s (1 cP). The "gas" is air and $Q_g = 2.6$ m^3/min (actual) or 0.0433 m^3/s.

1. N_P is not defined as a particular value; so, it must be experimentally determined for the particular impeller type. According to Sensel *(5)*, $N_P = 5.5$, for a standard Rushton impeller under the turbulent conditions in this example, and is not markedly dependent on *D*/*T*.

2. Using Eq. 4, $P_u = 5.5 \times 1{,}000$ kg/m$^3 \times (1.67/s)^3 \times (1.0\ m)^5$ = 25,616 kg-m^2/s^3 or 25.6 kW. According to Sensel *(5)*, P_g/P_u for the above impeller in a 1.0-cP liquid may be correlated as per Eq. 5: $P_g/P_u = 1 - 0.742\tanh(20.301 \times N_a) \times (N_{Fr})^{0.272}$. For this case, $N_a = Q_g/ND^3 = (0.433\ m^3/s)/((1.67/s)(1.0\ m)^3) = 0.259$; $N_{Fr} = N^2D/g = (1.67/s)^2(1.0\ m)/(9.8\ m/s^2) = 0.285$. Per Eq. 5, $P_g/P_u = 1.0 - 0.0742\tanh(20.301 \times 0.259)(0.285)^{0.272} = 0.473$.

3. Using Eq. 6, $P_g = P_u\,(P_g/P_u) = 25.6\ kW \times 0.473 = 12.1$ kW.

Thus, the calculation of P_g is straightforward when *N* and certain process conditions are known. But, achieving a target P_g by choosing *D* or *N* is an iterative process, because P_g/P_u changes with *D* and *N*.



Example 1. In the first case, P_{motor} is specified as 420 kW. Q_g is specified over a range of VVM = 0.1–0.7 min^{-1}. At standard conditions, this translates to Q_g = 20 m^3/min when VVM = 0.1 min^{-1}, and Q_g = 140 m^3/min when VVM = 0.7 min^{-1}. The appropriate value of N is selected such that P_g/P_{motor} is no greater than 0.90 at VVM = 0.1 min^{-1} for D/T ratios of 0.3 and 0.5. At this N, calculations are made for P_g and k_la at VVM = 0.1, 0.3, 0.5 and 0.7 min^{-1}. The calculated values of P_g and k_la are displayed in Table 1 and are respectively plotted in Figures 5 and 6 for different values of VVM.

The data show that at the minimum Q_g, P_g is the same for impeller systems with D/T ratios of 0.3 and 0.5. However, at VVM = 0.7 min^{-1}, P_g drops more for the larger impeller. More explicitly, P_g of the larger impeller is 82% that of the smaller impeller; and k_la is about 91% that of the smaller impeller. Consequently, if the overall process is oxygen-limited, less product will be made in the larger system. If the fermentation is not mass-transfer limited, then specifying a larger system would mean that the agitator is bigger than it needs to be. This case is clearly one where a smaller D/T may be better, at least in terms of mass-transfer performance. The larger D/T agitator also costs about 2–2.5 times as much as a smaller design.

Example 2. In the second case, P_g is specified at the maximum Q_g (corresponding to VVM = 0.7 min^{-1}) and P_{motor} must be chosen based on the minimum Q_g (corresponding to VVM = 0.1 min^{-1}). Thus, the shaft power is specified as P_g = 350 kW at VVM = 0.7 min^{-1} and P_{motor} will be calculated at VVM = 0.1, 0.3, 0.5 at 0.7 min^{-1} and at D/T ratios of 0.3 and 0.5. The calculated values of P_g and k_la are displayed in Table 2 and are respectively plotted in Figures 5 and 6 for different values of VVM.

The data indicate that at VVM = 0.7 min^{-1}, for both D/T = 0.3 and D/T = 0.5, P_g and k_la are the same as they would be at VVM = 0.7 min^{-1}. However, at VVM = 0.1 min^{-1} for D/T = 0.3, the smallest standard motor size required to ensure P_g/P_{motor} does not exceed 0.9 is 450 kW. At VVM = 0.1 min^{-1} for D/T = 0.5, the smallest standard motor size required to ensure P_g/P_{motor} does not exceed 0.9 is 550 kW.

Alternatively, the larger system (D/T = 0.5) could use a 450-kW motor if the minimum Q_g were equivalent to VVM = 0.3 min^{-1} because at this VVM, P_g/P_{motor} would not exceed 0.9. However, more compressor power would be used by the larger design during the early growth stage of the batch, compared with the smaller D/T configuration at the same P_{motor}. Also, the larger D/T design would cost 2–3 times as much as the smaller one if it uses a 550-kW motor and 1.5–2.5 times as much if it uses a 450-kW motor.

Conclusions

Agitator power may be invested using high N values and low D/T ratios or low N values and high D/T ratios. Although the high D/T case has some advantages at a fixed Q_g, in a real fermenter, Q_g varies, and a more constant P_g, which is characteristic of a smaller impeller system, may

Table 1. Effects of *D/T* on fermenter k_la for Example 1.

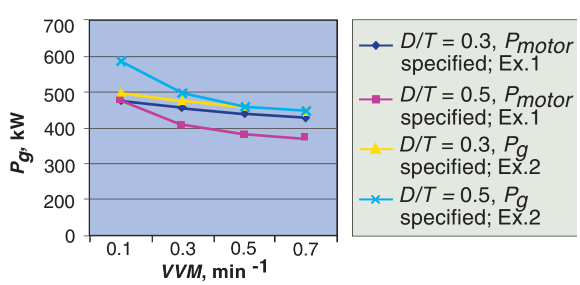
P_{motor} = 420 kW, 90% loading at VVM = 0.7 min^{-1}.

	***D/T* = 0.3**		***D/T* = 0.5**	
***VVM*, min^{-1}**	**P_g, kW**	**k_la, s^{-1}**	**P_g, kW**	**k_la, s^{-1}**
0.1	379	0.172	378	0.173
0.3	355	0.29	308	0.271
0.5	339	0.365	278	0.33
0.7	328	0.424	270	0.385

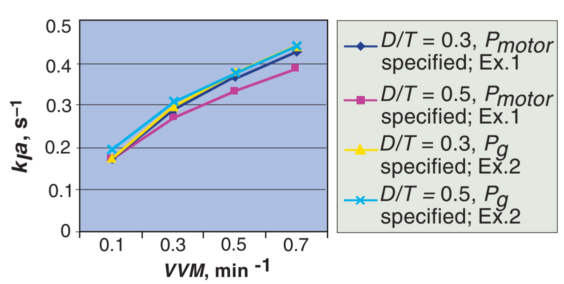
Table 2. Effects of *D/T* on fermenter k_la for Example 2.

P_g = 350 kW at VVM = 0.7 min^{-1}.

	***D/T* = 0.3**		***D/T* = 0.5**	
***VVM*, min^{-1}**	**P_g, kW**	**k_la, s^{-1}**	**P_g, kW**	**k_la, s^{-1}**
0.1	403	0.177	487	0.197
0.3	378	0.298	398	0.307
0.5	362	0.377	360	0.375
0.7	350	0.438	350	0.438



■ Figure 5. This graph shows the effect of *D/T* on P_g for examples 1 and 2 using data in Tables 1 and 2.
Photocredit: Benz Technology International, Inc.



■ Figure 6. This graph shows the effect of *D/T* on k_la for examples 1 and 2 using data in Tables 1 and 2.
Photocredit: Benz Technology International, Inc.

lead to higher mass-transfer performance and energy savings. Therefore, one should seriously consider a small D/T design if the following criteria are met:

- the maximum S is within acceptable limits for the organism
- Q_g is sufficient
- heat transfer is sufficient (hint: add more area if it is not, as opposed to changing the agitator)
- the shaft design is feasible
- Q_g has an adequate safety margin (20% or more) before flooding can occur.

If the above criteria are met, higher performance can be realized at a lower equipment cost. CEP

Acknowledgements

The data used to prepare Tables 1 and 2 and Figures 1 and 2 were calculated with the help of Chemineer, Inc.'s (Dayton, OH; www.chemineer.com) proprietary software. Though the specific results are based on Chemineer impeller styles, the conclusions regarding the effects of D/T on P are generic. The author wishes to thank Eric Janz, the custodian of Chemineer's software, for procuring illustrative materials.

GREGORY T. BENZ is president of Benz Technology International, Inc. (2305 Clarksville Road, Clarksville, OH 45113; Phone: (937) 289-4504; Fax: (937) 289-3914; E-mail: benztech@mindspring.com), specializing in general engineering, mixing and bioreactor design consultation, including equipment specification and bid evaluation. Benz has over 27 years experience in the design of agitation systems and has taught numerous courses on their design for liquid motion, solids suspension and gas dispersion; as well as fermenter design; non-intrusive mixing design and sanitary design concepts for pharmaceutical applications. Benz received his BSChE from the Univ. of Cincinnati and has taken a course on fermentation biotechnology from the Center for Professional Advancement. He is registered professional engineer in Ohio and a member of ISPE and AIChE.

Literature Cited

1. **Rautzen, R., *et al.*,** "Design and Scale-up of Agitators for Gas Dispersion in Fermenters," Presented at the Engineering Foundation Conference on Mass Transfer and Scale-up of Fermentors, Henniker, NH (July 1977).
2. **Myers, K. *et al.*,** "A Comparative Study of Alternative Gas Dispersion Impellers," Presented at Gas-Liquid Systems session, Mixing XVI, Williamsburg, VA (June 1997).
3. **Benz, G.,** "Enhancement of Fermentation with Effective Impeller Systems," Proceedings of New Horizons in Microbiology and Biotechnology 2001 International Symposium, Seoul Korea, pp. 153–154 (June 2001).
4. **Benz, G.,** "Optimize Power Consumption in Aerobic Fermenters," *Chem. Eng. Prog.*, pp. 100–103 (May 2003).
5. **Sensel, M.,** "Gas Dispersion at High Aeration Rates in Low to Moderately Viscous Newtonian Liquids," Master's Thesis, Univ. of Dayton (Apr. 1992).
6. **Bakker, A., and G. Benz,** "Mixing in Industrial Fermenters," Presented at Biotechnology Conference, Sydney, Australia (1996).
7. **Nienow, A.,** "Gas Dispersion Performance in Fermenter Operation," *Chem. Eng. Progress*, pp. 66–71 (Feb. 1990).
8. **Ulbrecht, J., and G.K. Patterson, Eds.,** "Dispersion of Gases in Liquids," in Mixing of Liquids by Mechanical Agitation, Gordon and Breach Science Publishers, pp. 139–202 (1985).
9. **Bakker, A., *et al.*,** "How to Disperse Gases in Liquids," *Chem. Eng.*, 12 (**1**), pp. 98–104 (1994).