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82. SOME SCALE-UP CONSIDERATIONS FOR FERMENTATION MIXERS

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ABSTRACT

Mixing is an integral part of fermentation process. This paper will describe some of the current techniques in use in analyzing the quantitative and qualitative performance of mixers in this operation.

KEYWORDS

Mixing; scale-up; mechanically-agitated fermentors; mass transfer; agitation power

INTRODUCTION

Two common agitation problems encountered in the fermentation industries are the selection of a suitable mixer for a larger-sized tank (scale-up) or the improvement of productivity of an existing tank by using a different combination of mixer horsepower and gas rate. Mixers are often specified on the basis of one of the following criteria:

1. Change in productivity requirements based on production data with a particular size fermenter in the plant.
2. New production capacity based on pilot plant studies.
3. Specification of agitator based on the sulfite absorption rate in aqueous sodium sulfite solution.
4. Specification of the oxygen uptake rate (OUR) in the actual broth for the new system.

SOME GENERAL RELATIONSHIPS IN LARGE SCALE MIXERS COMPARED TO SMALL SCALE

In general, a large scale mixing tank will have a lower pumping capacity per unit volume than will a small tank. This means that its blend time and circulation time will be much larger than in a pilot tank.

There is also a tendency for the maximum impeller shear rate to go up while the average impeller zone shear rate will go down on scale-up. In addition, the average tank zone shear rate will go down as well as the minimum tank zone shear rate. Therefore, there is a much greater variety of shear rates in the larger tank, and in dealing with pseudoplastic slurry it will have a quite different viscosity relationship around the tank in the big system compared to the smaller system.

On the question of micro-scale shear rates, these operate in the range of 300 microns or less, and are governed largely by the power input.

The power input from the gas per unit volume will increase on scale-up. This is because there is a greater head pressure on the system and also increased gas velocity. The mixer power may be reduced since the energy from the gas going through the tank is higher in order to maintain a particular mass transfer coefficient, K_Ga. The capacity for a blending type flow pattern is not affected in the same way with changes in the mixer power level as is the gas-liquid mass transfer coefficient.

SCALE-UP BASED ON DATA FROM EXISTING PRODUCTION PLANT

If data are available on a fermentation in a production-size tank, scale-up may be made by increasing, in a relative proportion, the various mass transfer, blending and shear rate requirements for the full-scale system. For example, it may be determined that the new production system is to have a new mass transfer rate x% of the existing mass transfer rate. There may be specifications put on maximum or average shear rates and changes in blend and circulation times may be considered. In addition, there may be a desire to look at the relative change in the CO_2 stripping efficiency in the revised system.

At this point, there is no reason not to consider any size or shape of tank. Past tradition for tall, thin tanks, or short, squat tanks, or elongated horizontal, cylindrical tanks does not mean that those traditions must be followed in the future. To illustrate the principle involved, consider gas-liquid mass transfer.

Gas-Liquid Mass Transfer

The mass transfer coefficient normally is related to the superficial gas velocity power per unit volume and to geometric variables such as the D/T ratio of the impeller.

A thorough analysis takes a look at every proposed tank shape,looks at the gas rate range required, calculates the gas phase mass transfer driving force, and then calculates the required K_Ga. Reference is made to data on the mixer under the condition specified to obtain the right mixer horsepower level for each gas rate in combination with various D/T ratios.

Viscosity and Shear Stress

At this point, the role of viscosity must be considered. Figure 1 and Table 1 give typical data on the effect of viscosity on mass transfer.

TABLE 1

TYPICAL EFFECT OF VISCOSITY ON MASS TRANSFER

Transfer rate @ 10 g/L	
10 mol O_2/MJ - 0.15¢/mol O_2	
Transfer rate @ 20 g/L	
6.4 mol O_2/MJ - 0.23¢/mol O_2	

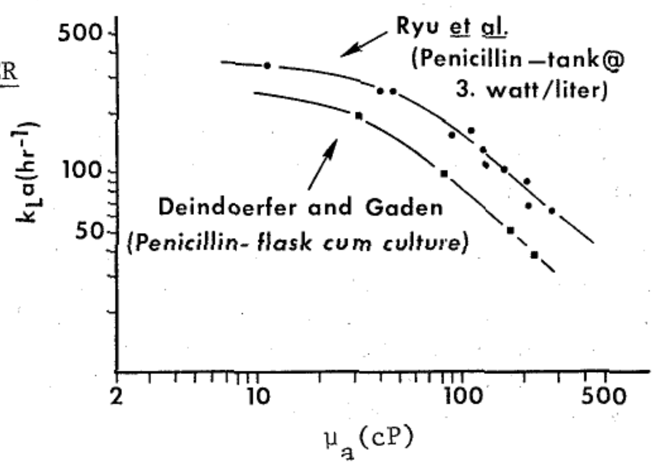

Fig. 1. Illustration of the decrease in K_La with increase in apparent viscosity.

It is necessary to measure viscosity with a viscosimeter which mixes while it measures viscosity. The Stormer viscosimeter is one device that can be used to measure viscosity under mixing conditions with known shear rates.

Estimates should be made of the shear rate profile around the system, and then of the shear stress [Shear Stress = μ(Shear Rate)] that actually carries out the process. Estimates can then be made of how different the proposed new tank may be compared to the known performance of the existing production tank.

SCALE-UP BASED ON PILOT PLANT DATA

To keep the ratios of impeller, gas bubble and solid clump sizes in the fermentation related to full scale, the impeller size and blade width in the small scale must always be two or three times bigger than the biomass particle size.

It is possible to model the fermentation biological process from a fluid mechanics standpoint, even though the impeller is not properly related geometrically to the gas-liquid mass transfer step. One pilot plant scale might be suitable to model the liquid-solid mass transfer step or a liquid-phase chemical reaction step (e.g., sulfite oxidation) but might not be suitable for analysis of the gas-liquid mass transfer step. The decision then is based on how suitable existing data are for any steps which are not modeled properly in the pilot plant.

Ideally, data should be taken during the course of the fermentation of gas-flow rate, gas absorption rate, dissolved oxygen level, dissolved carbon dioxide level, yield of desired product and other parameters which might influence the decision on the overall process.

If the pilot plant is to duplicate certain properties of fluid mixing, then it may be necessary to use non-geometric impellers and tank geometries to duplicate mixing performance rather than applying similarity. As a general rule, geometric similarity does not control any mixing scale-up property whatsoever.

It may also not be possible to duplciate all of the desired variables in each run, so a series of runs may be required changing various relationships systematically and then a synthesis made of the overall results. One variable in particular is important: the linear superficial gas velocity should be run in a few cases at the levels expected in the full-scale plant. This means that foaming conditions are more typical of what is going to happen in the plant and the fermenter should always be provided with enough head space to make sure the foam levels can be adequately controlled in the pilot plant. As a general rule, foam level is related to the square root of the tank diameter on scale-up or scale-down.

If it is desired to duplicate maximum impeller zone shear rates on a small scale, there may be a very severe design problem in the mechanics of the mixer, or the shaft speed, mechanical seals and other things. This means that careful consideration must be given to the type of runs to be made and whether the design of the pilot plant or the semi work-scale equipment must be capable at all times of duplicating the maximum impeller zone shear rates in the plant, and whether this sort of data will be obtained on a different type of unit dedicated to that particular variable.

Figure 2 shows what often happens in the pilot plant in terms of correlating the mass transfer coefficient, K_Ga, with power and gas rate in the pilot plant. This curve is then translated to a suitable relationship for full scale. It is possible to consider that with the higher superficial gas velocity, the power level may be reduced in the full scale to keep the same mass transfer coefficient. The box on the right in Figure 2 shifts to the box on the left. This should be considered, but it should be borne in mind that this changes the ratio of the mixer power to gas power level in the system, which in turn changes the blend time, the flow pattern and the foaming characteristics. It can also markedly affect the liquid-solid mass transfer rate, if that is important in the process.

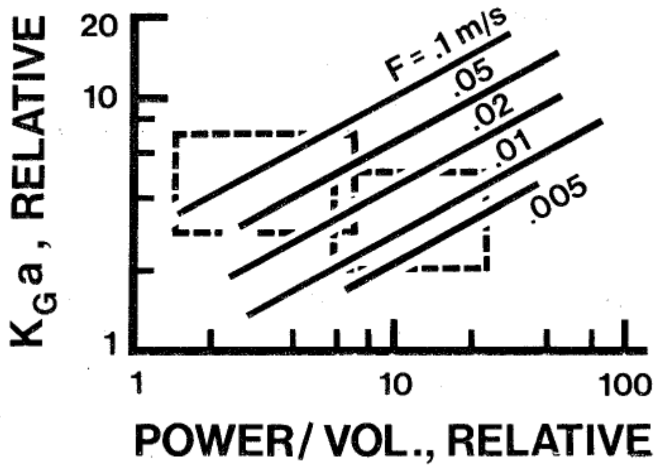

Fig. 2. Typical K_Ga dependence on horsepower and gas velocity. Box on the right indicates typical pilot plant experiments; box on the left indicates typical full-scale range.

In all cases, a suitable mass transfer driving force must be used. There is a marked difference between the average driving force, the log-mean driving force, and the exit gas driving force. For a larger fermenter, experience indicates that gas concentrations are essentially step-wise stage functions, and a log-mean average driving force is the most representative. For small laboratory fermenters, depending on the power level, an estimate must be made of the gas mixing characteristics and an evaluation made of the suitability of using the exit gas concentration for the driving force compared to the log-mean driving force. This is one area which needs to be explored in the pilot program and the calculation procedures.

In the waste-treating operations, it is quite common to run an unsteady state reaeration test. Initially, the tank is stripped of oxygen; air is started with the mixer started and the dissolved oxygen level increase is monitored up to the saturation level. At that point, the DO level is usually between the saturation value at the top and the saturation value at the bottom of the tank. This means that there must be absorption at the bottom but stripping at the top; a very peculiar mass transfer situation results compared to what normally occurs during waste treatment or fermentation operations. Running experimental tests and basing a lot of calculations on that peculiar driving force gives a marked difference in the way the mixer will operate under normal conditions.

SULFITE OXIDATION DATA

Data may be obtained on small and large size fermentation tanks using excess sodium sulfite with suitable catalysts which keep the dissolved oxygen level at zero. One caution is that this system should be used only when the tank is completely clean of anti-foams which may be residual from the fermentation process. The presence of trace amounts of anti-foam can cause marked differences in the mass transfer coefficient.

If a relationship between the sulfite oxidation number and the performance required in the fermenter has been established, this is a perfectly valid way to specify equipment and tests can be run to give an indication of the overall mass transfer rate ensuing.

OXYGEN UPTAKE RATE IN THE BROTH

If it is desired to relate fermenter performance to oxygen uptake rate in the broth, this number can be specified along with suitable desired gas rates and the mixer estimated based on this performance. Again, the link between this particular mass transfer specification and the actual performance of the fermentation must be known.

If the specified oxygen uptake rate is based on pilot plant data, then the effect of the different shear rates and different blend times on scale-up on the mass transfer relationship, viscosity and the resulting fermentation must be considered.

IMPELLER SELECTION

Multiple Impellers

The lower impeller does the major work for dispersing gas in the system and it is typical practice today to put a high proportion of the power into the lower impeller, somewhat similar to what is shown in Figure 3. Multiple impellers do have zoning action in terms of blending; this is not a great factor in a fermentation which takes several days, but there are instantaneous differences in the mass transfer, blending and concentration profiles in a tank with multiple impellers.

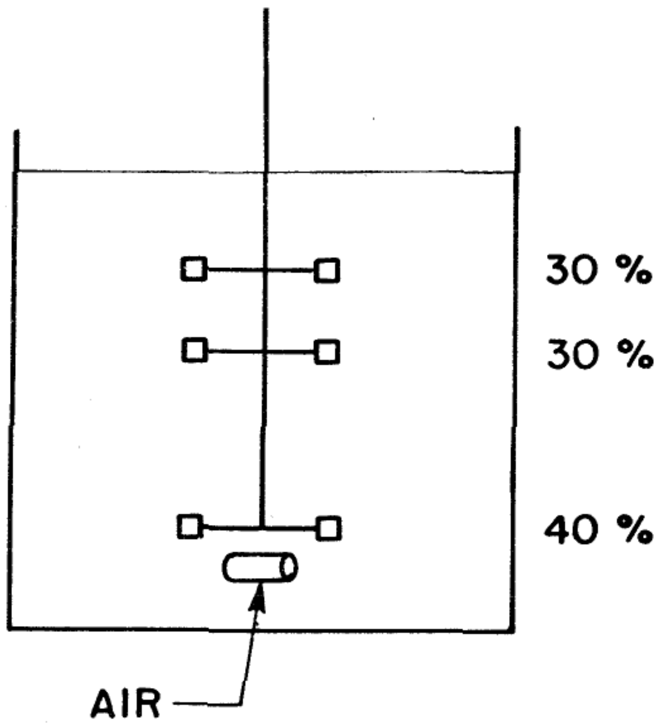

Fig. 3. Typical power consumption relations for triple impeller installation, giving higher horsepower in proportion to the lower impeller.

Axial Flow Impellers

If it is desired to consider axial flow impellers in a gas-liquid system for any reason, it should be remembered that the upward flow of gas tends to negate the downward action of the pumping capacity of the axial flow turbine. A radial flow turbine must have three times more power than the power in the gas stream for the mixer power level to be fully effective. On the other hand, the axial flow impeller must have eight to 10 times more power than in the gas stream for it to establish the axial flow pattern.

Reverse Rotation Dual Power Impellers

In gas-liquid systems, one of the reasons that the power of the impeller is lower with the gas on than with the gas off, is that the gas bubbles collect behind the impeller blade. This streamlines the blade, reducing the power. Looking at Figure 4, we can speculate what would happen if we were to fill up and streamline the back of the impeller with solid material.

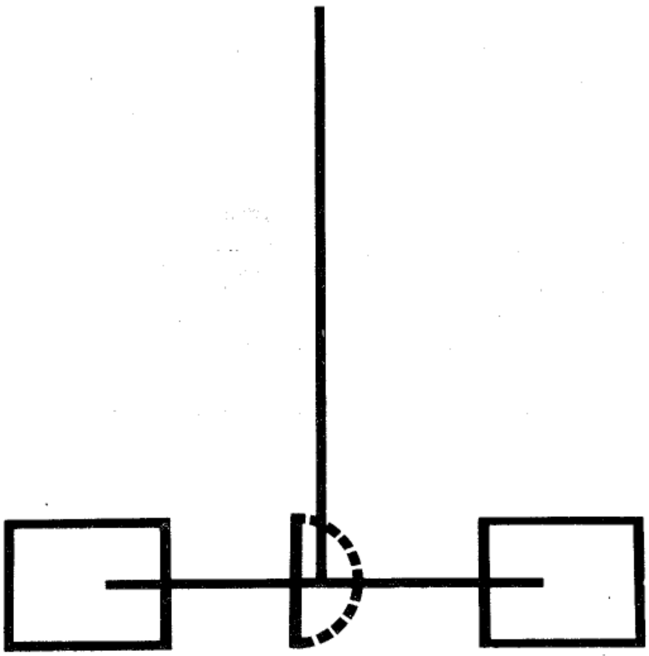

Fig.4. Typical dual power number impeller with streamlined back of blade.

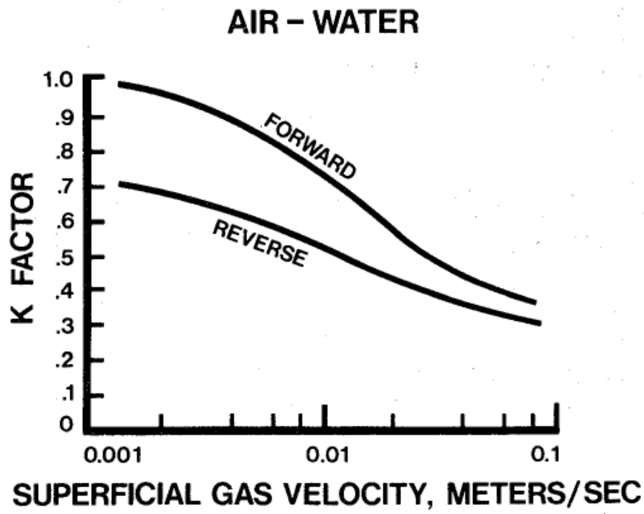

Fig. 5. Power consumption of dual power number impellers.

With no gas sparging, the impeller draws less horsepower with the back of the blade streamlined than with the back of the blade flat. With gas sparging, the flat impeller blade has a K factor which changes markedly with gas rate, typical of impellers of that type, while the impeller with the streamline back of the blade has a much less change in power with gas rate (Figure 5). A wide variety of power consumption availabilities without and with gas sparging can be obtained by having the mixer and motor capable of being reversed electrically; this opens up a wide variety of process options.

SUMMARY

A mixing tank exhibits an extremely complicated combination of shear rates, blend times, flow patterns and mass transfer coefficients. Fermentations cover a wide spectrum of conditions, ranging from low viscosity material types to the higher viscosity mycelial type on up to the Xanthan gum fermentations which become very viscous. Thus, the particular mixing characteristics required in a given fermenter must be considered and there is no one "standard" design of fermentation mixers for all kinds of processes. If the aeration of the biological waste-treating system is included, the range of variables and requirements becomes even more extreme.

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