SECOND WORLD CONGRESS OF CHEMICAL ENGINEERING

MONTREAL, CANADA, OCTOBER 4-9, 1981

SCALE-UP OF FLUID MIXERS

James Y. Oldshue*, Mixing Equipment Company, Rochester, N.Y., USA

Mixing processes can be described in relation to the pumping capacity of the impeller and various shear rates existing in the tank. These shear rates have both a macro-scale and a micro-scale characteristic. Using geometric similarity causes changes to occur in relationship between many of the pertinent mixing variables. The use of non-geometric relationships allows control of some of the key components involved. Some guidelines for looking and the qualitative and quantitative effect of scale-up on process results will be presented. Comments and relationships on pilot planting and scale-up in the mixing area, covered by preceding speakers, will give a summary of where the scale-up techniques will be in the future. In general, there is considerable latitude in the choice of mixing scale-up parameters as long as they are used for correlation and not forced to be constant without experimental of theoretical adjustment

SCALE-UP OF FLUID MIXERS

James Y. Oldshue*, Vice President, Mixing Technology
Mixing Equipment Co., Rochester, N.Y., U.S.A.

The design of a fluid mixer almost always involves scale-up from a different configuration on which data are available. Each mixing operation usually has its own individual process performance characteristics, so scale-up relationships for several different kinds of mixing criteria are often involved in final selection.

The characteristic of modern-day scale-up techniques is to evaluate the change on scale-up of different kinds of process requirements and to use the variables of impeller geometry and tank geometry to give the type of process characteristics desired.

As a starting point, let us define some of the characteristics of mixer scale-up. It is well to refresh our memory that there are two basically different types of process criteria which differ in measuring techniques and requirements for fluid mixer relationships. One criterion involves uniformity of one or more components in the mixing system. Examples are the description of solid suspension in a tank, blend time requirements, or physical characteristics of a liquid-liquid emulsion.

The other process characteristic involves mass transfer, heat transfer and chemical reaction. In these cases we are dealing with quantitative observations of changes in composition of one or more phases in the tank and can often express the results in quantitative mass transfer relationships. The relationship between fluid mixer variables is quite different in each of these two ways of evaluating process result.

Table I lists five basic pairs of materials, then points out that each of them has these two different process objectives making a general category of 10 separate mixing process operations.

There are many different parameters which can be examined in looking at mixing scale-up. Table II lists a number of these and shows the comparison in going from a small scale tank to a large tank seven times the linear dimension. It is also based on geometric similarity.

Parameters listed in Table I are the power consumption of the impeller, P, impeller diameter, D, impeller speed, N, power per unit volume, P/V, pumping capacity of the impeller, Q, pumping capacity of the impeller per unit volume of the tank, the fluid upward velocity, the flow to the impeller divided by the area of the tank, peripheral speed of the impeller, ND, Reynolds number, N_{Re}, Froud number, N_{Fr}, and Weber number, $\frac{N^2D^3\rho}{\sigma}$.

The properties of the fluid remain constant on scale-up. For comparison purposes, some of these parameters are held constant successively and reference is made to the other parameters to see how they change on scale-up. All these parameters are given an initial value of 1 to examine the effect.

The first column for the scale-up mixer uses constant P/Vol. Some of the key characteristics when this is used are that the operating speed drops; the flow per unit volume decreases; the peripheral speed increases and the N_{Re} increases.

The pumping capacity per unit volume decrease indicates that blend time and circulation time in the large tank will be longer. The operating speed decrease says that the maximum impeller zone macro-scale shear rate will decrease. The peripheral speed increase says that the maximum impeller zone macro-scale shear rate will increase. The N_{Re} increase says that we will move out on the N_{Re}-Power curve toward the turbulent region. These characteristics will normally change the way the full-scale mixer performs in the process.

It often happens that the mixer process responds to a trade-off of these many kinds of variables and the process result in the large tank will be the same as observed in the pilot tank.

On the other hand, there are a large number of processes where these changes will markedly affect the process result.

Usually there will not be sufficient data on every possible change in the mixing regime, so experience and approximation must be used to arrive at a satisfactory overall design and installation.

It may be that only a general qualitative evaluation of these effects is possible and the consequences of that on the accuracy of the process result to be achieved must be assessed.

In the second column we have maintained equal pumping capacity per unit volume. This normally requires the power per unit volume to go up with the square of the tank diameter. This does achieve similar blend time and circulation time on the larger system, but power consumption is much higher than would be practical for most installations, so this particular technique is seldom used.

The third column shows constant impeller peripheral speed. This has the characteristic of decreasing the power per unit volume in inverse proportion to the tank diameter. This means that circulation time will increase in direct proportion to the tank diameter. This particular criterion, in my experience, is a very risky procedure and can lead to serious underdesign of mixing equipment on full scale. This particular criterion is also the same as equal linear velocity of the fluid. Two points can be made from the above:

1. Keeping one particular parameter constant normally changes many others.
2. Is there a constant scale-up parameter for every one of the several thousand mixing processes handled every year?

The answer to question 2 is that it is much more rational and reliable to think of these parameters as correlating parameters and let them change on scale-up if need be to accomplish certain types of process results and to let them be a part of process correlations without the restriction that they must be constant. If it turns out that some of them are constants in particular applications, then it makes the correlation much simpler.

A correlary phenomenon in Table II is shown in Fig. 1 in which the various macro-scale shear rates in the tank are shown as a function of equal power per unit volume and geometric similarity. The macro-scale shear rates, which operate on particles of 500 microns and larger, tend to have a greater variety in the larger tank, since the maximum values go up and the average values go down.

The micro-scale shear rates in the tank, operating on particles from 200 microns and less, do not change very much at this particular power level. The question of scale-up then involves having sufficient data, correlations and experience to have the right relationship of the parameters for all of the many types of mixing processes to be considered.

USE OF GEOMETRIC AND DYNAMIC SIMILARITY

The principle of dynamic similarity says that the ratio of each of the four individual fluid forces, inertia, gravity, viscosity and surface tension, in the model and the prototype must be equal to a common constant ratio, Table III. This cannot be accomplished in a mixing vessel without changing the fluid properties.

There are four dimensionless ratios which are suggested for these four forces, each one of them involving the input inertia force of the mixer divided by, in turn, the opposing forces of viscosity, gravity and surface tension resulting in the Reynolds number, Froude number and Weber number shown on Table IV. It also can be seen mathematically that only one of these groups can be maintained constant if fluid properties are not changed. There have been many attempts made to use these kinds of correlations in process design. One of the oldest, well-known and extremely satisfactory is the Reynolds number/Power number curve. This works so well because these forces are the ones which determine the power consumption of the impeller.

This often holds true in other engineering fields in which we are often looking for the forces on boats, structures, airplane wings, etc., and fluid force ratios are the ratios which give us the information we want.

In a mixing vessel, however, the fluid force ratios may describe the fluid mechanics of the vessel, but the particular requirements of the mixing process, of which there are many thousand different variations and requirements, do not lend themselves to dimensionless groups.

There are two process correlations, however, which have worked out very well. One is heat transfer[1]. We say to ourselves that the heat transfer coefficient divided by the thermal conductivity of the fluid with another length term to make it dimensionless, would be a logical relationship to describe the process performance. Figure 2 shows this as an excellent correlation over different tank sizes and fluid properties.

Another correlation is blend time[2], in which blend time multiplied by the operating speed of the mixers turns out to be a good correlating group as shown in Figure 3.

However, there is no way today to write process groups around the many numbers of mixing applications, such as polymerization, crystallization, nitration, aeration, or many other types of suspensions and reactions, so we must look for other techniques.

The major goal of a pilot plant or plant study is to find out the way the process responds to a change in mixing variables so we can evaluate the important parameters on the process. This means two things: 1) that there will be some data taken; and 2) that they will be sufficient to describe important parameters on the mixing process result. Figure 4 shows the effect of a change in power on process result showing several different possibilities. Power is most conveniently changed by changing the speed of the impeller at constant diameter. When we do this, we change the flow rate and the shear rate and if increases in both of these do not cause a change in process result there is little likelihood that other variables will. It is always possible, however, that an increase in flow rate might help and an increase in shear rate might hurt the process,so they neutralize each other and there is little effect of power on the process. To be absolutely sure a study needs to be made with a different impeller diameter to see whether it was just a peculiarity of the particular combination used in the first experiment.

Looking at Figure 5, if the exponent on the slope is high,(A), that normally means a mass transfer process is involved, and quite typically gas-liquid mass transfer is the most sensitive to mixer power. Liquid-liquid mass transfer can also be involved, but liquid-solid mass transfer usually has a much lower slope, more like that shown on Curve D of that figure. If the slope is zero,(E), it often is caused by a chemical reaction that is controlling in which mixer variables are important. The jagged line shown on the left side of the curve indicates that the power levels below those required to provide a satisfactory blend time in the tank and process results may be quite erratic.

If the slopes are somewhere in the middle range, Curves B and C, there is less of a clear-cut definition as to what the controlling step may be, so either further data must be available or obtained, or other things done in experimentation.

SIZE OF PILOT SCALE TEST

In general, the blade width and the overall flow pattern in the plant size unit will be much larger in size than any of the particles, bubbles or fluid clumps entering into the reaction.

If we are to get an exponential type relationship for a chosen scale-up parameter it is usually necessary for the impeller in the pilot plant scale to also be larger in physical dimension than these fluid elements. The important dimension, usually, is the blade height which should be two or three times larger than the biggest particles and clumps in the process.

You can certainly run experiments where this ratio is not maintained, but that means that the interpretation of the effect of scale-up on the parameter chosen is likely to be different in the small scale tanks compared what would be obtained in a pilot scale tank where the ratios are maintained.

EXAMPLE 1. GAS-LIQUID MASS TRANSFER

It is quite typical in the pilot plant to vary speed over two or three levels and then vary gas rate over two or three levels measuring mass transfer rate and suitable concentrations in the gas and liquid phases. It is possible,then, to calculate the K_La from each run. A plot similar that shown in Figure 5 is obtained[3]. In general the same volume of gas per volume of liquid per minute is suggested in both scales, and it turns out that this specification means the linear superficial gas velocity goes up, which is the flow of gas divided by the cross-section of the tank, increases on scale-up.

Since this linear gas velocity has a marked effect on the mass transfer coefficient, it is possible to show that K_Ga required for the plant may be achieved at these higher superficial velocities at a lower power level for the mixer. This is further enhanced by the fact that the greater liquid depth in the full-scale tank often results in a higher gas-liquid mass transfer driving force, which also tends to require lower power levels at a given mass transfer rate.

This means the pilot plant is often run in the area shown on the right of Figure 5, while the pilot plant unit tends to operate in the area to the left. The pilot plant should be run at a superficial velocity equal to the general area of the superficial expected in the plant, even though this is not required for a satisfactory pilot plant experiment. Also, we need to look at blend time, possible liquid-solid mass transfer rates and other things to see what effect they will have on the process. In addition this higher gas liquid velocity will change the percentage of gas held up in the liquid medium and will also affect the foaming and geysering characteristics of the installation. The pilot plant study requires some evaluation both quantitative and qualitative, and predictions made on what that will do in the full-scale unit. Since blend time will increase anyhow on scale-up, the reduction of power suggested by the mass transfer coefficient will further accentuate that situation. This must be very carefully evaluated as to the effect that will have on concentration gradients, dissolved gas gradients in the liquid and other characteristics important to the process. This will also affect liquid-solid transfer coefficients and other parameters.

· ZINC PURIFICATION PROCESS

In a process study made in conjunction with American Zinc[4], an illustration of interaction of manv variables in the process is shown. In this particular process, it is desired to retain the cadmium at a high level in the purification process and to remove cobalt and arsenic. The process involves adding zinc dust to a solution from a leach system. There is usually a sudden drop in the cadmium as well as the cobalt and arsenic, and then if sufficient time is possible, the cadmium seems to be regenerated in the process and comes back to acceptable recovery level.

It was desired to see whether mechanical mixing can affect the time of this reaction and also affect the recovery in the cadmium and the rejection of the arsenic and cobalt.

Figure 6 shows the results in a 1016 mm diameter pilot plant study. In this study for any given run there were typical curves shown in Figure 6, and at a particular optimum horsepower level there was a good recovery of the desired cobalt in the process. On that scale, Figure 7 shows that the 355 mm diameter impeller was much more effective than the 230 mm impeller indicating that flow-to-shear ratios are important.

Data were also obtained on a 560 mm diameter scale and while similar process profiles were obtained with power and time, it turned out that the two impeller diameters tested, 122 mm and 203 mm, gave the same result (Figure 7).

The interpretation here is that on the small scale, 560 mm, there is so much extra pumping capacity and short blend time compared to 1016 mm diameter tank, and of the full-scale tank in the plant, that any reduction in the pumping capacity due to the 122 mm diameter impeller is not sufficient to affect the process result.

On the other hand, on the 1016 mm scale, any reduction in the pumping capacity and increase in shear can affect the process and it would be expected that D/T would be an important ratio in going up to a full-size system.

If it were desired to obtain this kind of information on only the 560 mm diameter scale, it would have been necessary to drastically change the impeller blade width ratio to markedly cut the pumping capacity down into the range where it will probably be in the full-scale plant to show us whether this pumping capacity in full scale will be a serious detriment to the process performance.

Plant scale data reported indicate that by using suitable D/T ratios in the plant, plant scale performance was predicted by looking at the overall scale-up effect between the 1016 mm and 560 mm diameter tanks, and taking into account the optimum D/T ratio needed to carry out this process.

Flocculation is a process which is extremely sensitive to mixing variables. In a series of publications, Oldshue and Mady[5] showed that for two different impellers the G factor for the minimum turbidity, which is a measure of optimum flocculation performance decreased with tank diameter as shown in Figure 8. The G Factor is the square root of the P/Vol. divided by viscosity. The study found that this minimum value of turbidity, which corresponds to optimum process performance, occurred at a particular impeller speed; speeds above and below this gave poor performance. Flocculation particles are increased in tank size by fluid shear rate which gets particles from various parts of the tank together, but the particle size is reduced by the shear stress which results so there is an optimum speed for every particular geometry.

Another observation about Figure 8 shows there is a decreasing power per unit volume required as tank size increases to obtain this optimum flocculation performance.

NOMENCLATURE

D	=	impeller diameter
D/T	=	impeller diameter-to-tank diameter ratio
G Factor	=	$\sqrt{(HP/Vol)/\mu}$
hd/k	=	blend time number
KGa	=	gas-liquid mass transfer coefficient
N_{Re}	=	Reynolds number, $\frac{ND^2\rho}{\mu}$
N_{Fr}	=	Froude number, $\frac{N^2D}{g}$
N_{We}	=	Weber number, $\frac{N^2D^3\rho}{\sigma}$
P	=	power
P/V	=	power per unit volume
Q	=	pumping capacity of impeller per unit volume of the tank
T	=	tank diameter
Z	=	liquid level
ρ	=	density
σ	=	surface tension
μ	=	viscosity

BIBLIOGRAPHY

1. Oldshue, J.Y. and Gretton, A. T., Chem.Eng.Prog. 50, 615 (1954).
2. Levenspiel, Octave, and Soon J. Kang, Chem. Eng. 83, 141 (1976).
3. Oldshue, J. Y., (Presented Internat'l Fermentation Symp., Waterloo, Can. July 1980
4. Carpenter, R.K., and Painter, L.A. (Presented 1955 Annual Mtg AIMME) Feb. 1955
5. Oldshue, J.Y. and O.B. Mady, Chem.Eng.Prog. 74,103 (Aug.1978)
6. Oldshue, J.Y. and O.B. Mady, Chem.Eng.Prog. 75, 72 (May 1979)

TABLE I

MIXING PROCESSES

PHYSICAL PROCESSING	APPLICATION CLASSES	CHEMICAL PROCESSING
SUSPENSION	LIQUID-SOLID	DISSOLVING
DISPERSIONS	LIQUID-GAS	ABSORPTION
EMULSIONS	IMMISCIBLE LIQUIDS	EXTRACTION
BLENDING	MISCIBLE LIQUIDS	REACTIONS
PUMPING	FLUID MOTION	HEAT TRANSFER

TABLE II

PROPERTIES OF A FLUID MIXER ON SCALE-UP

PROPERTY	PILOT SCALE 80 LITERS	PLANT SCALE - 27,400 LITERS			
P	1.0	343	16,800	49	0.14
P/VOL	1.0	$\boxed{1.0}$	49	.14	0.0004
N	1.0	0.27	1.0	.14	0.02
D	1.0	7.0	7.0	7.0	7.0
Q	1.0	93	343	48	7.0
Q/AREA	1.0	1.9	7.0	1.0	0.14
Q/VOL	1.0	0.27	$\boxed{1.0}$	.14	.02
ND	1.0	1.9	7.0	$\boxed{1.0}$	0.14
$\frac{ND^2\rho}{\mu}$	1.0	13.2	49	7.0	$\boxed{1.0}$
$\frac{N^2D}{g}$	1.0	.4	7.0	1.0	.003
$\frac{N^2D^3\rho}{\sigma}$	1.0	8580	117,650	2300	47

TABLE III

HYDRAULIC SIMILITUDE

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

DYNAMIC $\frac{(F_I)_M}{(F_I)_P} = \frac{(F_\mu)_M}{(F_\mu)_P} = \frac{(F_g)_M}{(F_g)_P} = \frac{(F_\sigma)_M}{(F_\sigma)_P} = F_R$

TABLE IV

FORCE RATIOS

$$\frac{F_I}{F_V} = N_{R_e} = \frac{ND^2\rho}{\mu}$$

$$\frac{F_I}{F_g} = N_{F_r} = \frac{N^2D}{g}$$

$$\frac{F_I}{F_\sigma} = N_{W_e} = \frac{N^2D^3\rho}{\sigma}$$

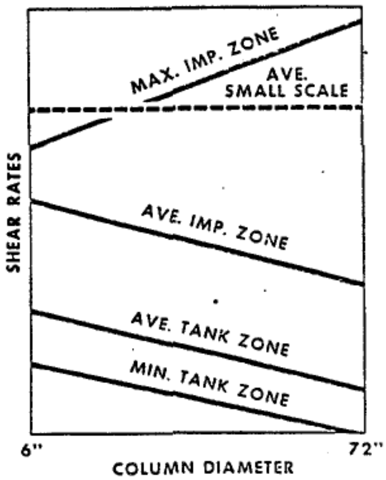

FIG. 1: CHANGE IN SEVERAL TYPES OF SHEAR RATES ON SCALE-UP.

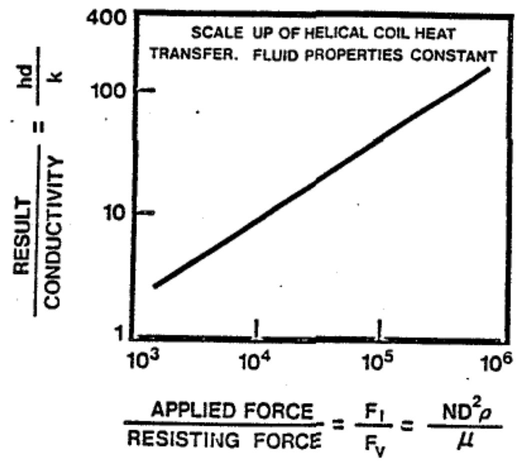

FIG. 2: DIMENSIONLESS GROUP CORRELATION FOR HEAT TRANSFER.

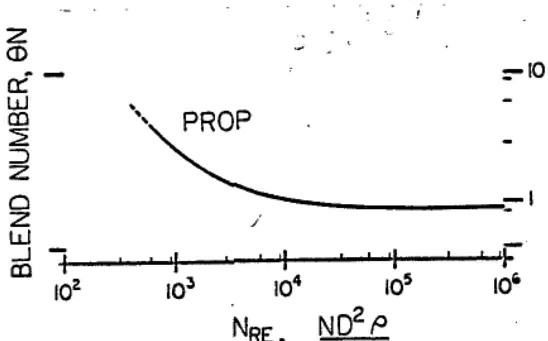

FIG. 3: DIMENSIONLESS BLEND NUMBER CORRELATION FOR BLENDING WITH AN AXIAL FLOW TURBINE.

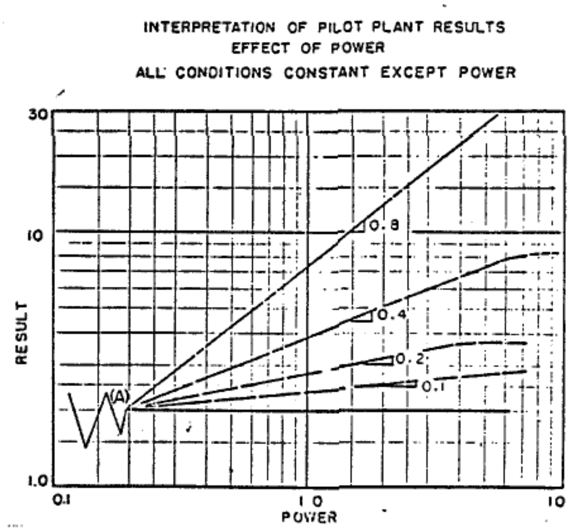

FIG. 4: CHANGE IN PROCESS RESULT WITH MIXER POWER OBTAINED BY VARYING SPEED.

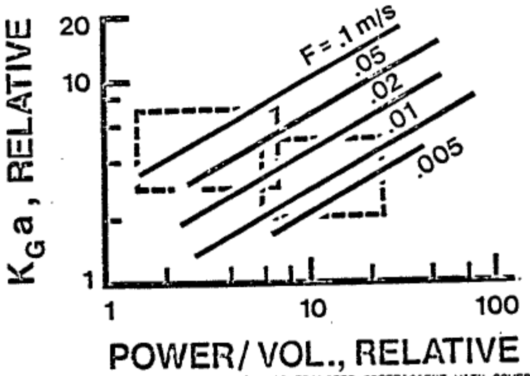

FIG. 5: TYPICAL CORRELATION OF MASS TRANSFER COEFFICIENT WITH POWER AND GAS VELOCITY. BOX ON RIGHT REPRESENTS TYPICAL PILOT PLANT CONDITIONS. BOX ON RIGHT REPRESENTS TYPICAL PLANT-SCALE CONDITIONS.

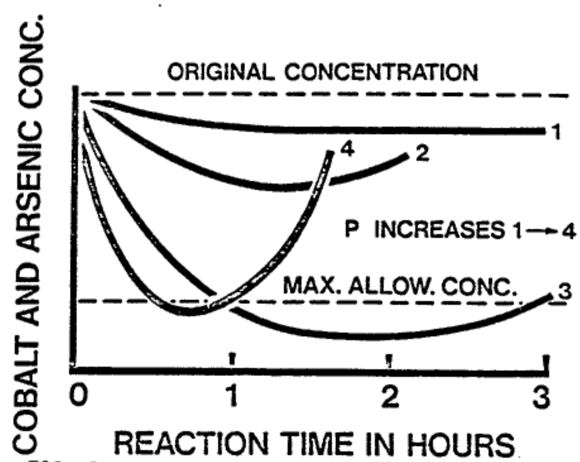

FIG. 6: A SERIES OF CURVES FOR A ZINC PURIFICATION PROCESS, SHOWING POWER LEVEL #3 GAVE AN OPTIMUM PERFORMANCE IN TERMS OF RETAINING COBALT AND ARSENIC AT A MINIMUM CONCENTRATION VALUE.

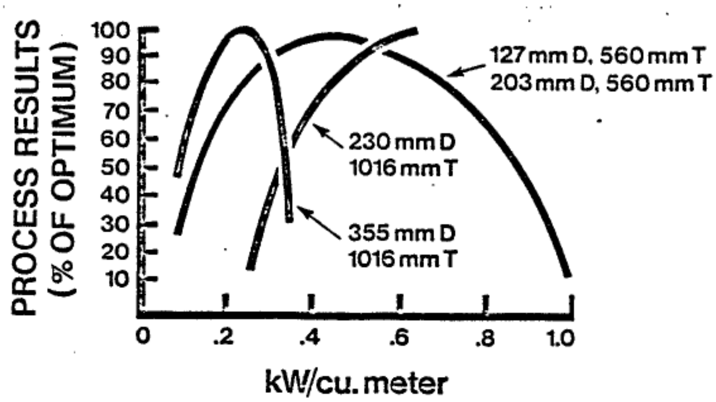

FIG. 7: EXTENSION OF THE CONCEPT IN FIG. 6, SHOWING THE PROCESS RESULT AT THE OPTIMUM CONCENTRATION OF COBALT AND ARSENIC AT TWO DIFFERENT IMPELLER DIAMETERS IN EACH OF TWO DIFFERENT TANK SIZES.

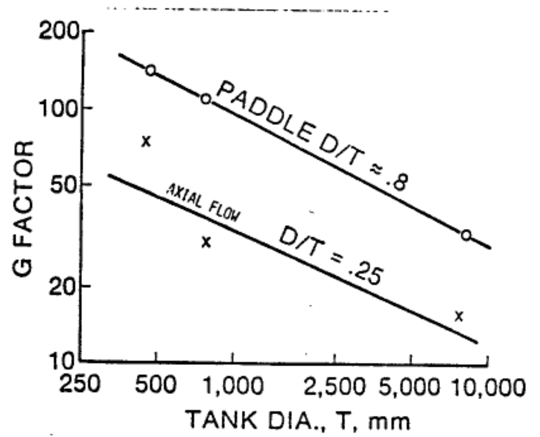

FIG. 8: SCALE-UP CORRELATION SHOWING G FACTOR REQUIRED IS LOWER AS TANK SIZE INCREASES.

JYO:jka
4/13/81