Fluid Mixing

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When considering fluid mixing, separating process requirements into pumping capacity or fluid shearing effects allows a preliminary analysis of possible mixing effects. If fluid shear is thought to be important, then considering the six types of shear rates present in the tank allows proper scale-up relationships to be considered. The author describes how pilot-plant experiments can be designed to vary power, pumping capacity and the two major shear rates independently to allow interprepretation of mixing requirements.

ONE of the most powerful concepts in analysing the effect of mixing on the many different mixing applications that exist is the relationship that the power supplied by the mixing impeller produces a pumping effect and a velocity head. This can be expressed as:

$$P \propto Q\,H \qquad \ldots\ldots\ldots\ldots (1)$$

A large, slow-moving impeller produces a large pumping capacity and a low value of impeller head, while a small impeller at a high speed produces a lower volume of fluid pumping but a higher velocity head.

$$(Q/H)_P \propto (D/T)^{2.66} \qquad \ldots\ldots\ldots\ldots (2)$$

This can be used for analysing those mixing processes which are largely dependent upon the pumping capacity of the impeller. The majority of industrial mixing applications are more sensitive to pumping capacity than impeller head, thus many principles of fluid mechanics can be employed in designing mixers for those applications.

Circulating capacity

The impeller head is related to the velocity head leaving the blades of the impeller, $V^2/2\,g$. The fluid shear rate is the velocity gradient. Referring to Fig. 1, which is the velocity profile in a given plane in the discharge from the impeller, the fluid shear rate may be obtained at any point by taking the slope of that velocity profile. It is important to realise that fluid shear stress is the major quantity determining the degree of dispersion or the particle size in the fluid. It equals the viscosity of the fluid times the shear rate, as given below:

$$(\text{fluid shear stress}) = \mu\ (\text{fluid shear rate}) \qquad \ldots\ldots (3)$$

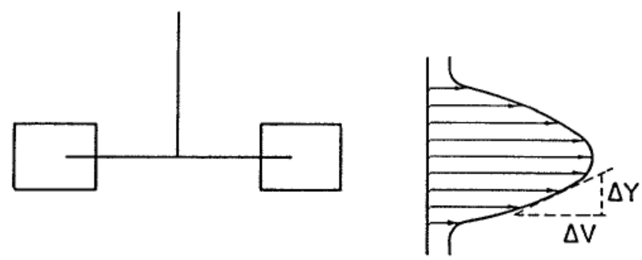
Six types of shear rates may be defined in a mixing system:

(*i*) maximum impeller shear rate (based on a mean point velocity);
(*ii*) average impeller shear rate (based on a mean point velocity);
(*iii*) average shear rate (based on a mean point velocity);
(*iv*) maximum impeller turbulence intensity and associated shear rates;
(*v*) average impeller turbulence intensity and associated shear rates; and
(*vi*) average tank turbulence intensity and associated shear rates.

There are many other shear rates in the tank, but this discussion is restricted to these six.

No matter whether the flow from the impeller is laminar or turbulent, the first three shear rates can be defined by using the mean point velocity to plot the velocity profile and to calculate the velocity gradient.

The results of studies on this concept,[1] have yielded the following principles: the maximum impeller shear rate is related to tip speed for a geometric series of one impeller type; the average impeller shear rate is a function only of the impeller speed for a geometric series of a given impeller type; the average tank shear rate is an order of magnitude below the average impeller shear rate; and the turbulent shear stresses occur at a very high



$$\text{SHEAR RATE} = \frac{\Delta V}{\Delta Y}$$

Fig. 1. Schematic drawing of velocity profile from turbine impeller

frequency, in the order of 1–100 c/s, and have a relatively small scale compared to particles that are in the visible range from 10–300 mesh.

At high frequencies, in the order of 5–100 c/s, turbulent velocity fluctuations can be measured by high-frequency velocity probes. These frequencies are apparently too high or the scale of the shear rates too small to markedly affect particles in the above visible range. They are, however, very important for micro-scale chemical reactions and other similar phenomena. Turbulence measurements are normally made in terms of the root mean square of the velocity fluctuations, and are related more to the energy dispersion at that point than to the actual shear rates that exist. There are shear rates associated with these velocity fluctuations but these have not been extensively calculated. Fig. 2 shows what happens to the maximum impeller sear rate and the average impeller shear rate when scaling-up with geometric similarity. As a general statement, maximum shear rates tend to increase while average shear rates tend to decrease on scale-up. Therefore it is difficult to maintain the same magnitude of fluid shear effects on scale-up.

The maximum impeller shear rate can be considered to determine what the ultimate effect of shear rate on the process can be if the particles get into that shear zone for a sufficient length of time. The pumping capacity and the volume of this shear zone are important, since they determine how long the shear stress can operate on the fluid materials. If the particles are affected irreversibly by fluid shear, then with sufficient time all the particles in the tank could achieve a condition governed by this maximum shear rate. If, on the other hand, there is tendency to coalescence in parts of the tank away from the maximum shear rate zone, then the overall condition of the particles in the tank is a function of the average shear stresses that the fluid or particle 'sees' during its sojourn in the vessel.

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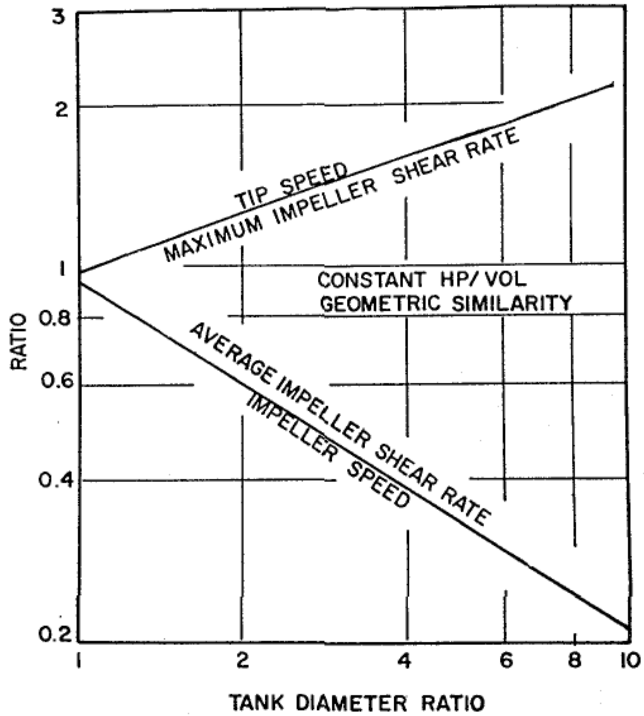

Fig. 2. Typical changes on scale-up. Power per unit volume is held constant and geometric similarity maintained

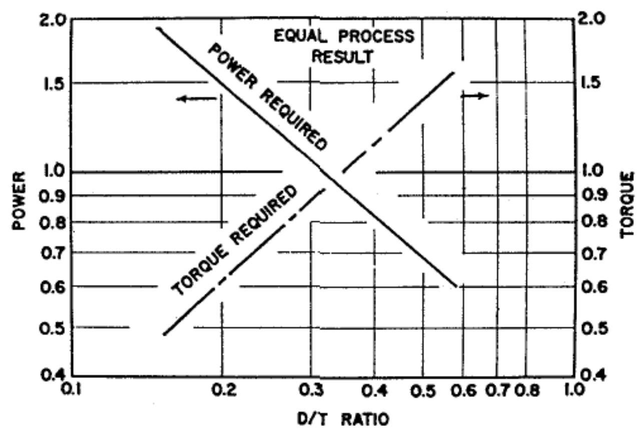

Fig. 3. Effect of D/T *ratio on typical blending process*

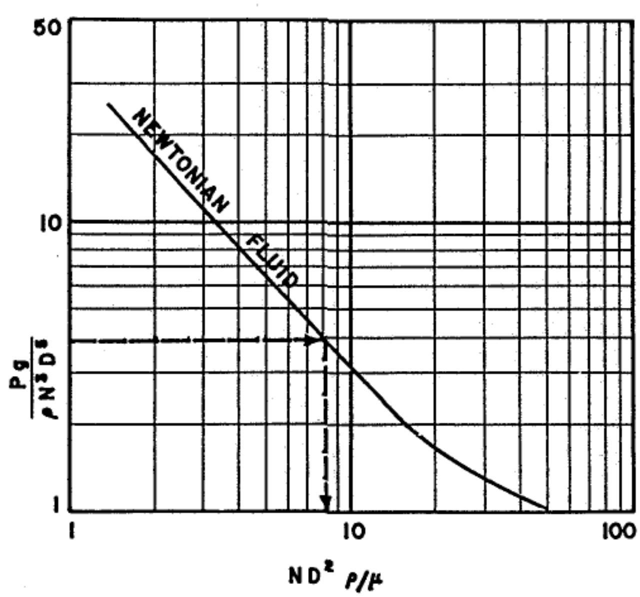

Fig. 4. Curve of power consumption in Newtonian fluids

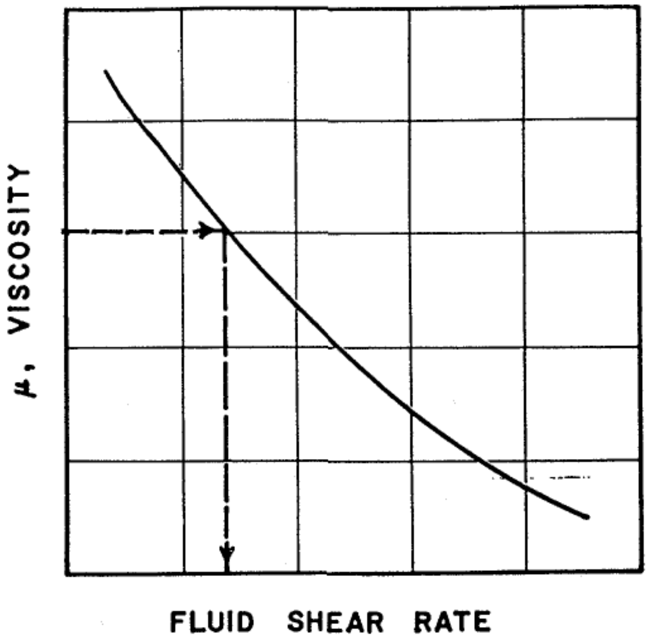

Fig. 5. Plot of viscosity versus fluid shear rate

NOMENCLATURE

- D = impeller diameter
- g = gravitational acceleration
- H = velocity head
- P = total power
- Q = flow from the impeller/unit time
- T = tank diameter
- V = volume
- μ = viscosity
- $\Delta v/\Delta y$ = velocity gradient
- v = velocity

Three examples of important processes that depend on fluid shear rates are: *(i)* in non-Newtonian fluids, the average shear rate as represented by the impeller speed is sufficient to estimate the viscosity that the impeller 'sees'; *(ii)* in emulsion polymerisation, bubble degradation is controlled by maximum shear rates, while particle size distribution and the average particle size can be adjusted by other mixing parameters; and *(iii)* fermentation involves control of the maximum fluid shear rate to give optimum liquid/gas dispersion, yet prevent biological solids damage, while other mixing parameters are adjusted to get the proper overall mass transfer, blending and suspension.

Scale-up

In scaling up a mixing system, the list of possible parameters is almost endless. Table 1 indicates some of these parameters and how they change when holding any one of them constant. Generally, it is not possible to keep more than one variable constant on scale-up when using geometric similarity. In fact, it is not normally desirable to believe that there is a parameter that should be kept constant on scale-up. It is better to think of a variable as a correlating parameter, and find out how to change it on scale-up to get the desired process result. There is one exception to this general rule. That is, the maximum impeller shear rate which does determine in a major way the ultimate shear rate to which a particle can be exposed. There is, of course, some effect of tank volume and length of time the particle is exposed to the shear rate, that would determine the actual condition of the materials. If it is desired to control the maximum shear stresses in a system so that the shear rate can be kept either below a critical value to prevent damage, or at a certain

Table 1. Properties of a fluid mixer on scale-up

Property	*Pilot scale (20 gal)*	*Plant scale (2,500 gal)*			
P	1·0	125·0	3,125·0	25·0	0·2
P/vol.	1·0	**1·0**	25·0	0·2	0·0016
N	1·0	0·34	1·0	0·2	0·04
D	1·0	5·0	5·0	5·0	5·0
Q	1·0	42·5	125·0	25·0	5·0
Q/vol.	1·0	0·34	**1·0**	0·2	0·04
ND	1·0	1·7	5·0	**1·0**	0·2
$ND^2\rho/\mu$	1·0	8·5	25·0	5·0	**1·0**

Table 2. Properties of a fluid mixer on scale-down

Property	*Plant scale (2,500 gal)*	*Pilot scale (3·4 gal)*			
P	1·0	0·00137	0·0022	0·0022	0·0022
P/vol.	1·0	1·0	1·6	1·6	1·6
N	1·0	4·3	5·1	6·4	10·1
D	1·0	0·11	0·11	0·097	0·097
Q	1·0	0·006	0·007	0·006	0·004
Q/vol.	1·0	4·3	5·1	4·3	1·7
ND	1·0	0·48	0·56	0·62	1·0
$ND^2\rho/\mu$	1·0	0·07	0·08	0·06	0·09
D/T	0·35	0·35	0·35	0·30	0·30
D_w/D	1·0	1·0	1·0	1·0	0·25

Table 3. Cost for 75 hp compared to 60 hp mixers (75-hp mixer is £600 less than 60-hp mixer)

	1 Year 8 hr/day	*5 Years 24 hr/day*	*1 Year 24 hr/day*	*10 Years 24 hr/day*
Initial saving	+ £600	+ £600	+ £600	+ £600
Power cost	− £150	− £450	− £2,000	− £3,000
	+ £450	+ £150	− £1,400	− £2,400
	Saving for 75 hp	Saving	Loss	Loss for 75 hp

level to ensure that the particles would achieve an ultimate particle size, this parameter must be carefully considered. Since, on scale-up, maximum shear rates tend to increase, it is necessary to go to non-geometric scale-up when it is required for any reason to keep shear rates at the same value they had in the pilot-scale equipment.

Scale-down

As a general rule, maximum impeller shear rates in the pilot plant will be one-half to one-third the values they are in the homologous plant scale equipment. For many processes this is of no concern. However, if a process is known to be governed by fluid shear effects, then the pilot unit must be non-homologous geometrically to the full-scale unit in order to achieve similarity to the fluid shear stresses. Dimensionless similarity is similarity of dimensions and not similarity of pumping capacity and fluid shear effects. Table 2 gives one example of how a pilot unit can be constructed to give comparable shear rates to those which would exist in the plant.

Pumping capacity

Solid suspension and blending are two examples of processes that depend largely upon the pumping capacity of the impeller. This means that velocity head, velocities and velocity gradients are normally high enough that total pumping capacity of the impeller is a predominant criterion. In these processes, larger diameters give more flow per horsepower than smaller-diameter impellers, and have a constant process result; a curve similar to Fig. 3 is often obtained.

At first glance, this suggests the use of larger impellers to cut horsepower to the lowest point. Whether this is economical or not depends on the total cost of the mixer. It turns out that large, slow-speed impellers require more torque even at their low horsepower than do the higher-speed, higher-horsepower units. Table 3 indicates a comparison between two different mixers capable of doing identical process jobs. When compared for various operating times, depending on the time picked for evaluation and the actual power cost involved, one or the other would be the most logical choice for the installation.

Pseudoplastic fluids

Pseudoplastic fluids have lower viscosities as shear rate increases. The average impeller shear rate has also been determined by measuring the power consumption of impellers in non-Newtonian fluids. This has been done as follows:

(*i*) a curve of power consumption in Newtonian fluids in the viscous region is experimentally determined in the laboratory and correlated as the traditional power number/Reynolds number plot (Fig. 4);

(*ii*) a pseudoplastic fluid whose viscosity versus shear rate relationship has been determined in a capillary viscosimeter, is placed in the mixing tank and its power consumption measured at various speeds (Fig. 5); and

(*iii*) from the power number measured experimentally, the corresponding Reynolds number from Fig. 4 is obtained and therefore the viscosity that existed at the average *impeller shear rate. Referring to Fig. 5, the average shear* rate can be estimated from this viscosity value.

The average impeller shear rates calculated in this manner agree quite well with the average impeller shear rates calculated from velocity probe profiles in the turbulent fluid region. Therefore, the concept of the average impeller shear rate being related only to the impeller speed is a valid observation.

Since the average shear rate in many commercial viscosimeters is much lower than the average shear rate in a mixing tank; it is desirable to obtain viscosities for design purposes in a calibrated mixing impeller dynamometer viscosimeter so that viscosity data at the correct shear rates will be available for design of industrial equipment.

REFERENCES

1. Oldshue, J. Y., *Biotech. and Bioeng.*, 1966, 8 (1), 3.