# Mixing Processes In the Protective Coating Industry

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Mixing is an integral part of the manufacture of protective coatings. General mixing principles and a classification of the various types of mixing criteria and various ranges of viscosities are important considerations.

Theoretical considerations are applied to practical mixing technology. The role of fluid shear rate and fluid shear stress in mixing operations is covered, as well as the use of data from commercial viscosimeters. In a mixing vessel there are a variety of shear rates throughout the vessel and it is important to distinguish which viscosities are to be used under various design conditions.

## INTRODUCTION

Mixing encompasses a tremendous variety of operations. For the most meaningful discussion, we should go one step further and ask, "What type of mixing result are we trying to achieve?"

To talk in the same terms, *Table* 1 gives a definition of the kinds of mixing jobs that can be achieved. In the center of the table is the breakdown into the types of fluids being handled. Within each type of fluid system there are two very different mixing objectives.

(1) *Physical Uniformity,* which includes such things as blending and solid-suspension, in which the criterion of success or failure is the uniformity achieved.

(2) *Mass Transfer* and *Chemical Reaction* processes which involve the transfer of material from one phase to another, and include such things as gas absorption and liquid-liquid extraction.

In the protective coatings industry, we are most often involved with the first case above, producing a certain uniformity of liquids or solids which can be measured by an evaluation of their physical properties.

Three types of operations encountered frequently are:

(1) Dispersion of pigment in a vehicle. This may, or may not, involve reduction of agglomerate sizes and/or reduction of actual particle size.

(2) Paint letdown, in which a small amount of high viscosity material is blended into a large volume of low viscosity material.

(3) Blending of materials that have relatively similar viscosities.

Industrial protective coatings often involve developing full color in an oil vehicle which requires dispersing pigments and/or dispersing high viscosity materials into lower viscosity fluids.

In water base household protective coatings, the various stocks used are often highly pseudoplastic which involves special mixing considerations.

## VARIABLE BATCH SIZES

Before proceeding to a discussion of the basic principles and practice of fluid mixing, it would be well to treat one of the most important aspects of obtaining satisfactory fluid mixing in an actual installation. If there are a variety of batch sizes and/or various batch viscosities to be handled in a single mixing vessel, the requirements must be carefully specified. Normally there is only one mixer drive and mixer shaft. There may be one, two or more, impellers, and there may, or may not, be variable speed.

At a constant mixer speed, the power drawn by a single impeller is approximately constant once it is adequately covered. If there are two or three distinct

Table 1—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 |

Presented at the 1967 Ray Vonesh Memorial Lecture Series sponsored by the Chicagoland Paint Industry.
* Box 1370, Rochester, N. Y. 14603.

A Journal of Paint Technology Reprint

batch levels to be used, then it is often possible to use two or more impellers so that each impeller is properly covered for a particular batch depth. By varying the diameter, blade dimensions, or impeller type of each impeller, it is possible to get widely different degrees of mixing at constant speed at two or more distinct batch levels. It is very difficult to eliminate splashing and spraying when the final or transient liquid level is right at the impeller level.

If the requirements of the different batch levels are markedly different, it may be desirable to have two-speed drives, four-speed, or variable speed. It is also possible to install different numbers and kinds of baffles at various batch levels to take care of particular requirements.

In essence, there is hardly any combination of batch levels, viscosities and mixing requirements that cannot be handled if they are completely specified when initially designing the mixer. All the principles discussed in this paper apply to any particular batch level and the impellers that are fully operable at that condition.

## MIXING PROCESS DESIGN

From a practical standpoint, there has to be an accurate description of what mixing is required before a mixer selection can be made. Quite often this description is based on presently operating equipment, either a presently satisfactory operation, or the desire to make a change or improvement. It would not be practical to try and list absolute conditions for the tremendous variety of mixing requirements encountered. The approach taken in this paper is predicated on the assumption that there is a knowledge about a particular operation being carried out, or that there is a similar operation available, and the question is, "What mixer is required for a different batch size, incorporating any desired changes or improvements?"

## TYPES OF MIXERS

The first classification that can be made is between top-entering and side-entering. Side-entering mixers use propellers and operate at such typical speeds as 380, 420 and/or 1150 RPM. Side-entering mixers are particularly effective for blending processes in which there are no appreciable amounts of free-settling solids present. The flow pattern is shown in *Figure* 1. They are usually the most economical choice for relatively large volume blending. Their main operating drawback is in the submerged seal or stuffing box where the shaft enters the tank. Extremely low liquid level operation is difficult.

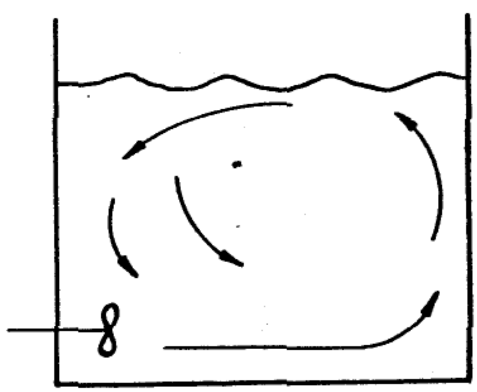


**Figure 1—Typical flow pattern of side-entering propeller type mixer**

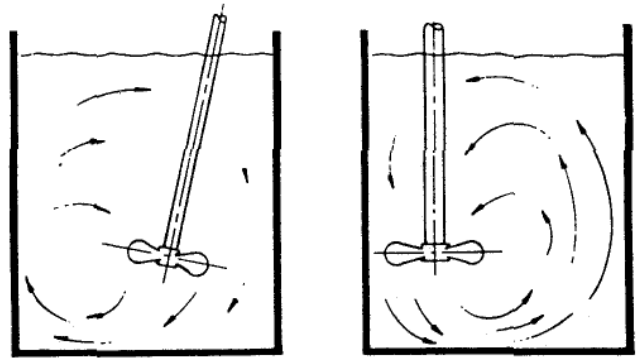


**Figure 2—Typical flow pattern from top-entering propeller type mixer in off-center position**

Top-entering equipment is usually propeller type in the smaller sizes, up to three horsepower, and turbine type above three horsepower. Top-entering pro-

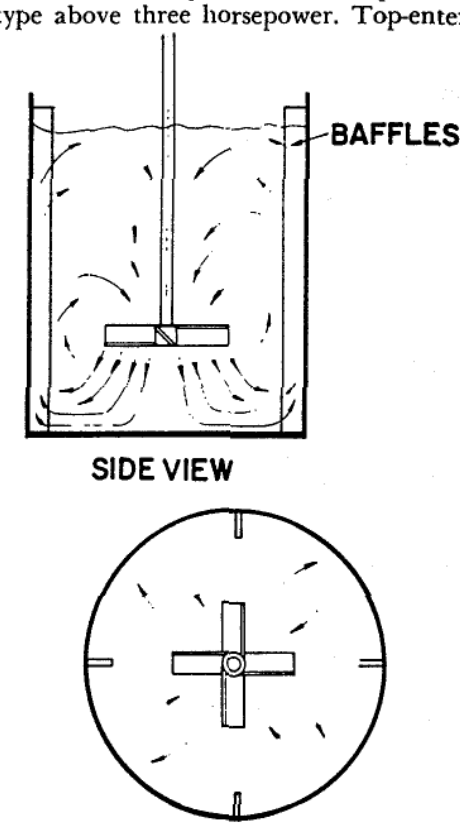



**Figure 3—Typical flow pattern from axial flow turbine in baffled tank**

**Table 2—Elements of Mixer Design**

| | | |
|---|---|---|
| (1) | Process design | Fluid mechanics of impellers<br>Fluid regime required by process<br>Scale up: Hydraulic similarity |
| (2) | Impeller power characteristics | Relate impeller HP; speed & diameter |
| (3) | Mechanical design | Impellers<br>Shafts<br>Drive assembly |

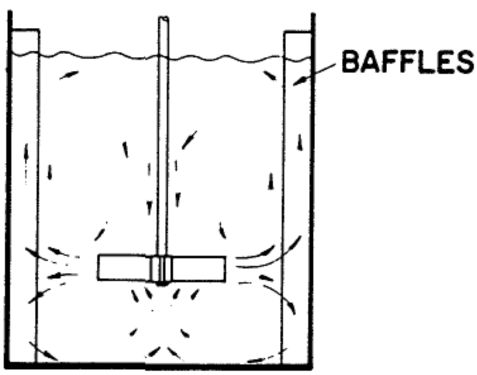
peller units are often mounted angularly off-center, *Figure* 2, to give effective top-to-bottom turnover.

For general blending and suspension operations, the axial flow turbine, *Figure* 3, is normally used.

## FLOW PATTERNS

Referring to *Table* 2, a distinction is made between fluid mechanics of flow from mixing impellers and the requirements of a given process for a particular flow pattern. There is no way to judge whether a given flow pattern is good or bad in the abstract. There must be a specific process requirement in order to evaluate the suitability of a particular flow pattern.

Measuring or predicting the flow patterns from various kinds of impellers is quite important in understanding mixing results. For example, the flat blade turbine can be made with or without a disc. It is



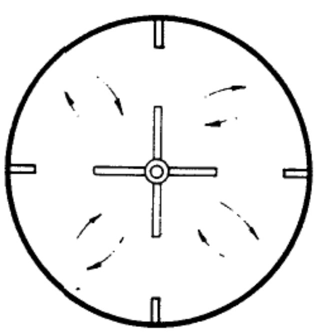




Figure 4—Usual flow pattern assumed for radial flow turbines

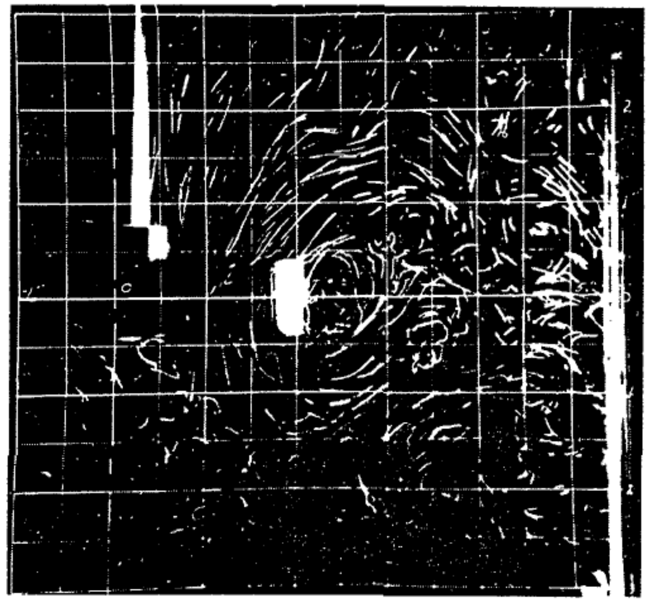


Figure 5—Streak photograph showing axial pumping characteristics for open flat blade turbine

sometimes assumed that the flow pattern for a turbine without a disc is radial out to the tank wall, as shown in *Figure* 4. In waterlike materials it seldom does act this way. Placed close to the bottom of the vessel, the impeller pumps downward like the axial flow turbine in *Figure* 5. Placed farther off the tank bottom, it pumps upward like an upward-thrusting axial flow turbine. This can make the prediction of process results quite difficult if a radial flow pattern has been assumed.

## BAFFLES

Tanks may be classified as baffled or unbaffled in terms of their general over-all fluid flow characteristics. A baffled tank has the following characteristics:

(1) Top-to-bottom turnover with complete intermixing of all fluid streams throughout the entire vessel.

(2) The absence of a vortex through which air is drawn down into the fluid.

(3) The absence of a swirling flow pattern as contrasted to top-to-bottom turnover.

In most mixing cases, a baffled flow pattern is desirable. There are times, however (when it is desirable to draw down powders from the surface or introduce gas from the surface) that an unbaffled flow pattern, *Figure* 6, is useful.

Unbaffled flow patterns at high power levels can often be unstable and cause severe fluid forces which act on the impeller and the shaft and are erratic in scale-up performance.

A baffled flow pattern may be produced in several different ways. In a vertical, cylindrical tank the most common way is the use of four baffles, each one-twelfth the tank diameter in width, placed equally around the

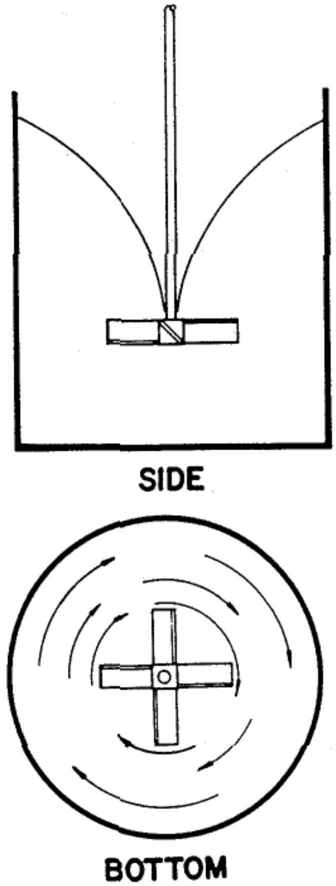



Figure 6—Typical unbaffled flow pattern

tank, *Figure* 7. This baffle width may often be modified when dealing with non-Newtonian fluids. With top-entering propeller type mixers up to about three horsepower capacity, the mixer can be placed in an angular off-center position and achieve the characteristics of a baffled flow pattern.

In a rectangular tank, the shape can supply a baffling action at low power levels. A baffled flow pattern in a square tank is produced with two baffles placed opposite the impeller on opposite sides, each baffle being one-twelfth the tank width. This system is also used in tanks in which the length is less than one-half times the width. The corners act as partial baffles, and in some cases, baffles can be eliminated and still retain the top-to-bottom flow pattern. In rectangular tanks, in which the length is one-and-one-half times the width, it is possible to eliminate the use of baffles. At high horsepower levels baffles must be provided even in rectangular tanks. These are illustrated in *Figure* 7.

## IMPELLERS

Impellers may be divided into two general classifications, axial and radial flow. Typical flow patterns are shown in *Figure* 3.

Many modifications can be made to the blade shapes of paddles and turbines, but their overall flow pattern in a baffled tank is basically very similar.

Considering the impellers, mention should be made that it is essential that an impeller produce the desired process result under conditions which allow sound mechanical design. Many of the proportions and designs that are used commercially were selected to *achieve the desired process result under conditions of* sound mechanical operation.

## FLOW AND SHEAR RATE

The pumping capacity, Q, of propeller and turbine impellers for any given geometric series of impellers is given by the proportionality

$$Q \propto N D^3$$

where N is impeller speed, RPM
D is impeller diameter
Q is volumetric fluid displacement of impeller.

The power drawn by the impeller produces this circulating capacity against the impeller head across

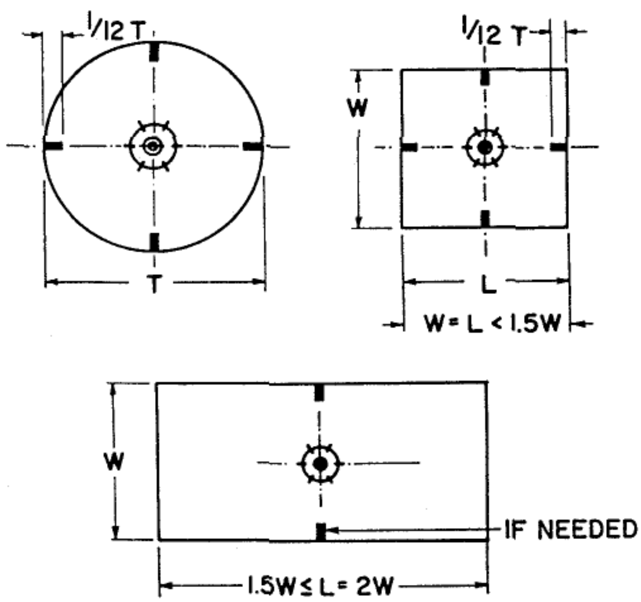



*Figure 7—Schematic illustration of tank baffles*

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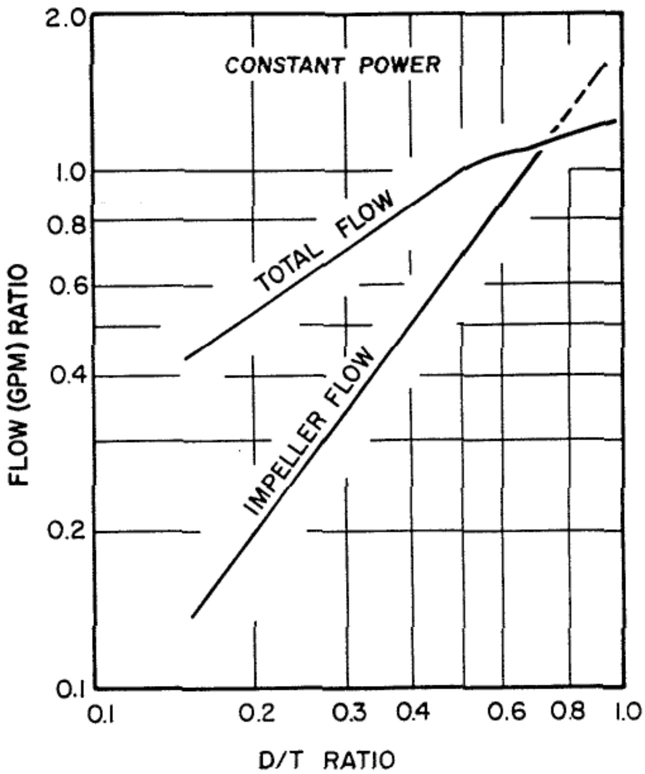



**Figure 8—Impeller flow and total flow as a function of D/T ratio at constant power**

the system. This impeller head is related conceptually to the fluid shear rate existing in the tank.

At a constant power level, the pumping capacity of various diameter impellers of the same geometric proportions in a given tank is related by the proportionality

$$Q \propto D^{4/3} \text{ (constant power)}$$

However, the flow from the impeller entrains fluid in the tank, so that the total flow circulating through the system can be several times greater. As the impeller gets larger and larger, it is pumping more flow at less head. It also has less entrainment distance due to its proximity to the tank wall. The total flow in the system does not increase as fast as the above equation, and at about 0.6 D/T ratio there is no further advantage gained in increasing the impeller diameter, *Figure* 8.

For high viscosity applications, it is often necessary to go to close-clearance impellers which have a D/T ratio of .95 or greater, in order to positively pump fluid through the vessel and off of the tank walls.

The discussion so far holds most exactly for low viscosity fluids. As viscosity increases, and/or if there are appreciable pseudoplastic effects present, there are certain minimum D/T ratios that must be used to get motion throughout the tank. Within the range of D/T's that can be used, most blending and suspension processes require less power with large diameter impellers. *Figure* 9 illustrates typical performance. Even though power decreases with large impellers, the low speed of these impellers actually requires an increase in torque for the mixer drive. A balance between first cost and operating cost of the equipment is required.

Turning now to fluid shear rates, a mixing vessel has a complete spectrum of shear rates. Large size eddies and swirls transfer their energy to smaller eddies and down to still smaller size eddies, and eventually through viscous shear into heat. The essential requirement when examining the fluid shear required for a particular mixing process is to identify the size and frequency of the shear rates that affect that particular process result. We can illustrate the point by mentioning that we do not use an ultrasonic drill to drive a rivet, nor do we use a rivet hammer to drill a tooth.

The velocity profile from a radial flow turbine impeller is illustrated in *Figure* 10. The slope of a line tangent to the velocity profile gives the shear rate

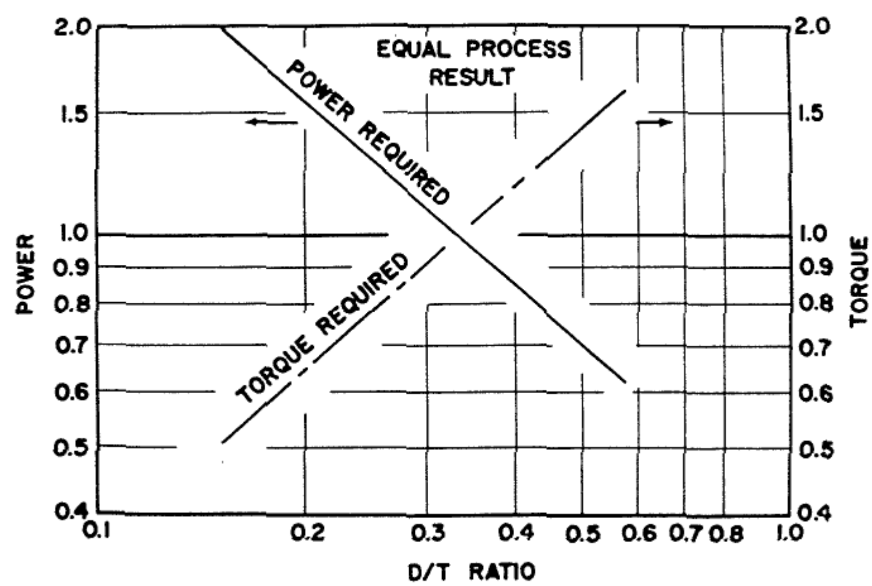



**Figure 9—Effect of D/T ratio on typical blending process**

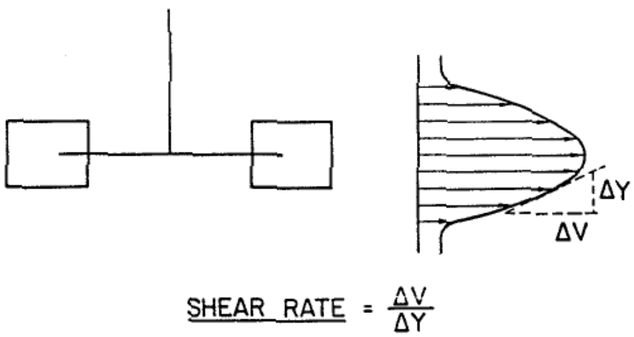



**Figure 10—Schematic drawing of velocity profile from turbine impeller**

at that point. Multiplying the shear rate by the viscosity of fluid at that point gives the shear stress. This shear stress breaks up particles and makes dispersions.

In a later section some examples of some of the recent research in this area are given.

If a particular fluid has a constant proportionality between the fluid shear stress in the fluid and the shear rate generated in the fluid, it is a Newtonian fluid and consideration of its viscosity is quite straightforward.

If, however, the viscosity of the fluid varies with shear rate, we then must know what viscosity does the impeller "see" at shear rates existing around the impeller, and what viscosity does the process "see" which best describes the viscosity at the shear rates in the remainder of the tank which affect the overall process result.

Pseudoplastic fluids can be relatively Newtonian at very high shear rates and at very low shear rates, while being very non-Newtonian at shear rates existing in a mixing tank or in a commercial viscosimeter. There is no limit to the complexity or the number of arbitrary constants in a mathematical equation used to describe this condition. However, for shear rates between one and 30 seconds$^{-1}$, which are the order of magnitude of shear rates in a mixing tank, the viscosity usually plots as a straight line on a log-log plot with shear rate, and that the "Power Law" relationship with one arbitrary constant, n, is adequate.

$$\text{Shear Stress} = K\,(\text{Shear Rate})^{n}$$

at a given shear rate,

$$\mu = \frac{\text{Stress}}{\text{Rate}} = K'\,(\text{Shear Rate})^{n-1}$$

## CENTIPOISE AND KREBS

The basic definition of viscosity requires that it have the units of shear stress over shear rate. The most common unit used is centipoise.

Several publications,[1-6] as well as data at the end of this paper, show that the average shear rate in the vicinity of an impeller is related only to the impeller speed and not to the impeller diameter for a geometric series of impellers.

Table 3—Rheology Definitions

| Definition | Effect on Viscosity, $\mu$, of Time Increase | Effect on Viscosity, $\mu$, of Shear Rate Increase |
|---|---|---|
| Newtonian | No effect | No effect |
| Non-Newtonian | | |
| Pseudo plastic | No effect | Decrease |
| Dilatant | No effect | Increase |
| Thixotropic | Decrease | No effect |
| Rheopectic | Increase | No effect |

A Krebs impeller must rotate at approximately three times the speed of the axial flow turbine, shown in *Figure 3* to give the same shear rate that this particular impeller "sees." Other impeller types and other viscosimeters have different ratios, and it is essential in translating viscosity data that these ratios be available for both the viscosimeter and the industrial mixing impeller to be used.

It should also be emphasized that the average shear rate in the mixing tank is considerably less than the shear rate at the impeller and the proper viscosity to use for process discussions and correlations would have a higher value than the viscosity at the impeller.

Mention should be made of the various viscosity definitions shown in *Table* 3. Since there is no uniformity among various technical groups about nomenclature, the term "pseudoplastic" used in this paper refers only to shear stress effects, and the term "thixotropic" refers only to time effects.

At the shear rates existing in a mixing vessel, time effects are normally not significant, so that pseudoplastic effects are the main consideration. If a fluid had both shear stress and time effects, then in this paper, it would have to be called a pseudoplastic, thixotropic fluid.

It might be well to mention viscoelastic effects since they are more than interesting curiosities. Some fluids exhibit normal shearing stresses which cause them to rise up a mixer shaft during agitation. On small scale pilot plant experiments, these effects can be so pronounced that they get a reverse flow pattern, in which the fluid flows up the side of the tank and down the outside into the impeller, practically the opposite of the flow pattern shown in *Figure* 4. This effect largely disappears on commercial size tanks.

Krebs is a unit of viscosity that is widely used in the paint industry. By measuring the seconds per 100 revolutions of the Krebs paddle for various weights on the pully, a chart, *Table* 4, gives the Krebs value. The Reynolds number of the Krebs paddle is such that it operates essentially in the viscous region for any fluid over 500 cps. The grams of weight to drive the paddle are proportional to viscosity and speed, as shown below:

$$\text{Grams} \propto \mu N$$

Table 4—Krebs' Stormer Chart

| Seconds For 100 Revolutions | 75 | 100 | 150 | 200 | 250 | 300 | 350 | 400 | 500 | 600 | 700 | 800 | 900 | 1000 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| 24 | 42 | 52 | 65 | 75 | 83 | 90 | 95 | 99 | 108 | 115 | 122 | 128 | 132 | 136 |
| 26 | 47 | 56 | 68 | 78 | 85 | 91 | 96 | 101 | 110 | 117 | 123 | 130 | 134 | 138 |
| 28 | 51 | 59 | 70 | 80 | 87 | 93 | 98 | 102 | 112 | 118 | 124 | 130 | 134 | 139 |
| 30 | 54 | 61 | 72 | 82 | 89 | 95 | 100 | 104 | 112 | 120 | 125 | 131 | 136 | 140 |
| 32 | 56 | 63 | 74 | 83 | 90 | 96 | 101 | 105 | 113 | 120 | 126 | 132 | 136 | 140 |
| 34 | 58 | 64 | 75 | 84 | 91 | 97 | 102 | 106 | 114 | 122 | 127 | 132 | 137 | 141 |
| 36 | 60 | 66 | 76 | 85 | 92 | 98 | 103 | 107 | 115 | 122 | 128 | 133 | 137 | 142 |
| 38 | 62 | 68 | 78 | 87 | 93 | 99 | 104 | 108 | 116 | 123 | 129 | 134 | 138 | 142 |
| 40 | 63 | 69 | 79 | 88 | 94 | 100 | 104 | 108 | 116 | 124 | 130 | 134 | 138 | 143 |

For a non-Newtonian fluid the following relationship exists:

$$\text{Grams} \propto (\text{Seconds}/100\ \text{Rev.})^{-n}$$

In setting up the Krebs table, typical paints were used, largely oil base. Krebs units were selected to reflect both practical viscosity and pseudoplastic characteristics.

Referring to *Figure* 11, at Krebs 65, pseudoplastic properties are not usually evident in oil paints and any Newtonian fluid that has $n = 1.0$ will have a constant Krebs over the allowed speed range.

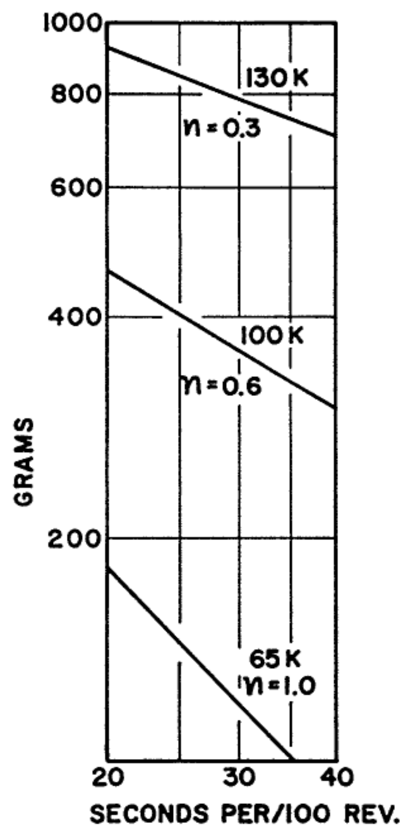



**Figure 11—Grams versus seconds per 100 revolutions at constant Krebs, from Table 4**

At 100 Krebs, pseudoplastic properties are more common, and to have a constant Krebs value at all allowable speeds, n would have to equal 0.6. For a Krebs value of 130, n must equal 0.3 to have a constant Krebs.

Most water base paints will have lower n values, and will not give constant Krebs.

In typical tests, an alkyd flat enamel showed a value of $n = 0.5$ at Krebs $= 90$. A rubber base interior paint showed an $n = 0.4$ at Krebs $= 70$. An interior water base paint showed $n = 0.2$ at Krebs 100.

A Newtonian fluid will not have a constant Krebs so it is not possible to have an absolute relationship between centipoises and Krebs. *Figure* 12 shows an approximate conversion between Krebs and centipoise.

To obtain exponent n to express the degree of pseudoplasticity, it is usually more convenient to use a viscosimeter that is calibrated in centipoises and plot

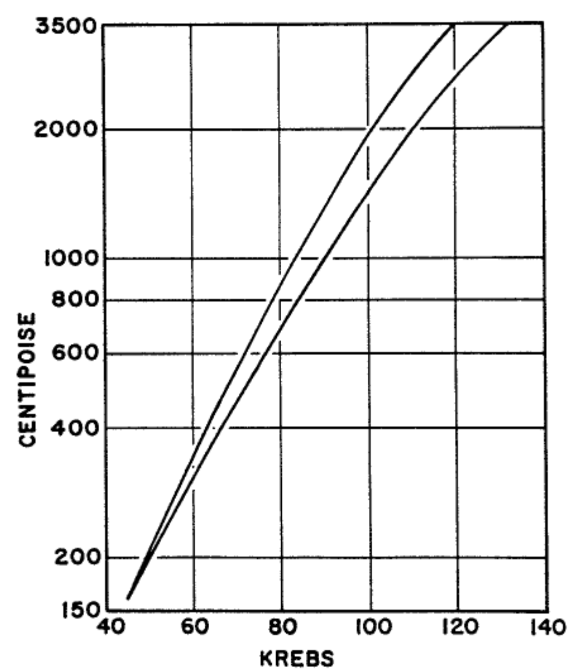



**Figure 12—Approximate relationship between Krebs and centipoise**

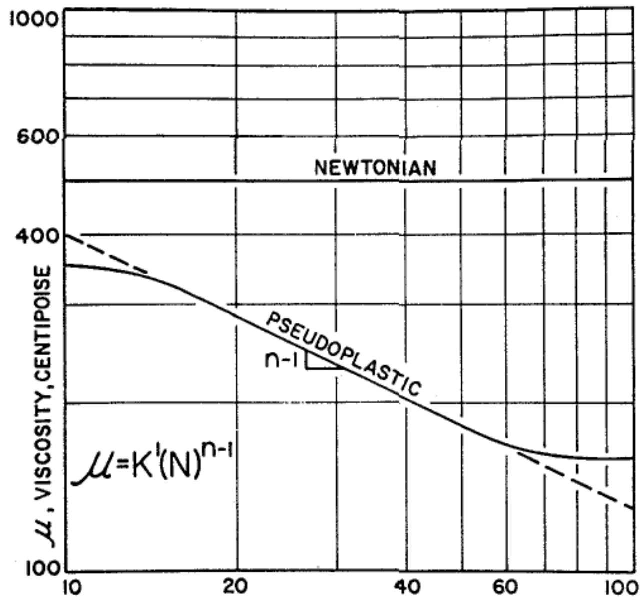



**Figure 13—Typical curve of viscosity versus impeller speed for pseudoplastic paint**

three or more viscosity and speed points as shown in *Figure* 13.

## BLENDING

Blending of similar low viscosity materials can be a relatively easy operation. However, if we are interested in analyzing in detail what blending is, or are interested in the degree of uniformity in blending high viscosity materials, the "scale of scrutiny" of the uniformity of blending is critical.

*Figure* 14 shows the progress of blending on a batch basis from a tank was initially stratified with hot and cold water. The impeller was in the lower cold phase as shown by Legend A. Legend B was about at the initial interface and C, D and E were in the hot layer. Since the ability to detect blending shows uniformity at 105 seconds, this could be the blend time. If, however, we had a more sensitive measuring device or were interested down in the molecular scale, the blend time is considerably longer.

No matter what scale is used, there is always an absolute blending that is somewhat less than complete. Of course the ultimate definition of the blend or the molecular level depends upon whether we use a random or ordered mathematical model.

Most blending operations follow the principles shown previously in *Figure* 9, in which less horsepower is required with larger diameter impellers. Thus there is always the relationship between operating cost and first cost to be considered.

For non-Newtonian fluids we, first of all, have to be sure that the entire batch is in motion. If this condition is satisfied, then blending progresses in the same fashion that it does in Newtonian fluid and it has been found that blend time correlates well with the viscosity at the average shear rate of the tank with blend times obtained in Newtonian fluids.

Assuming that the blend time for a particular product is known, mixer power and blend time will be approximately inversely proportional.[7] If the difference in specific gravities is changed, blend time goes up in proportion. Blend time is approximately directly proportional to viscosity changes in the continuous phase and to the ratio of viscosities between the phases.

If viscosity varies during a run, the mixer must actually be sized for the highest viscosity expected. At lower viscosities the mixer power level may be exces-

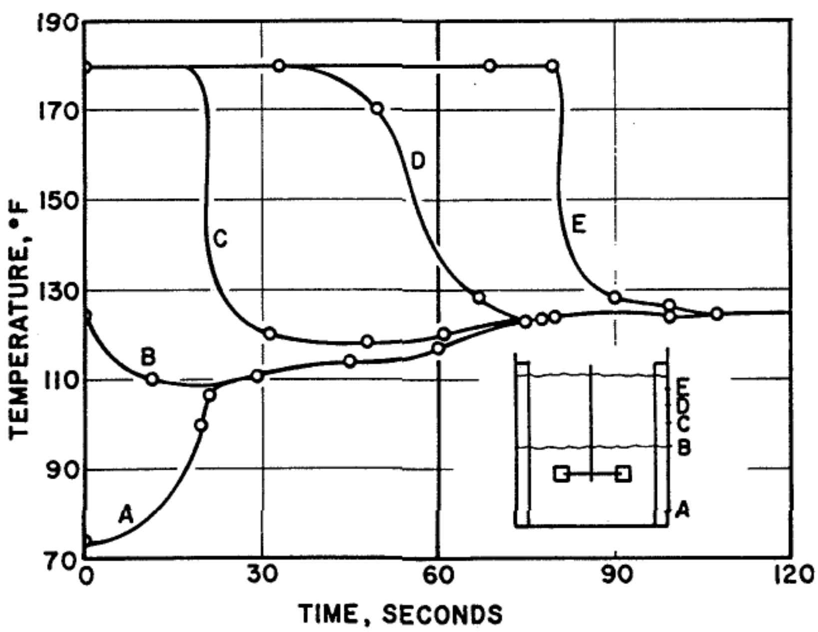



**Figure 14—Progress of blending in an initially stratified tank**

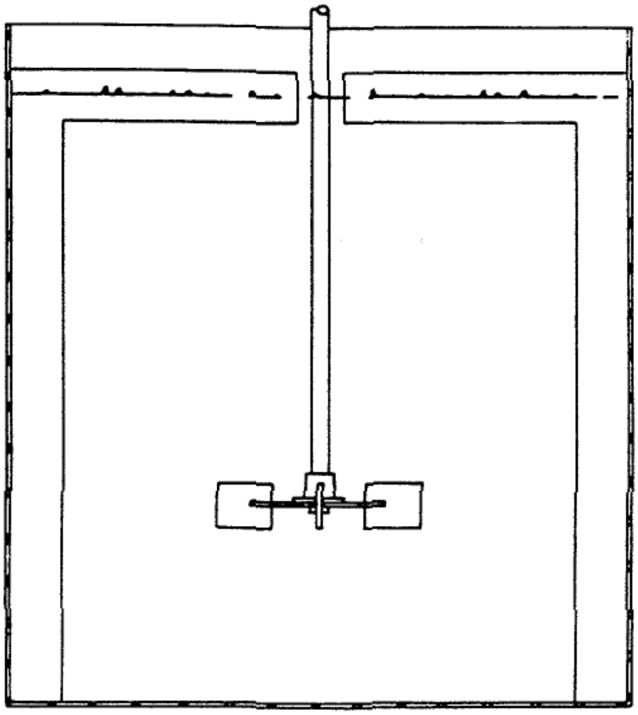


**Figure 15—Horizontal surface baffles to eliminate air entrainment**

sive and splashing and excessive turbulence may be a problem. In cases where entrainment is a problem, a slower speed may be required at the low viscosity stage.

In case it is desired to keep air entrainment completely out of a tank in which mixing is required, placing baffles across the surface of the tank, such that no splashing can occur over the top of the baffle, will eliminate this, *Figure* 15. These baffles should be high enough so that liquid will not wash over them and extend low enough so that the wave troughs do not extend below the baffle.

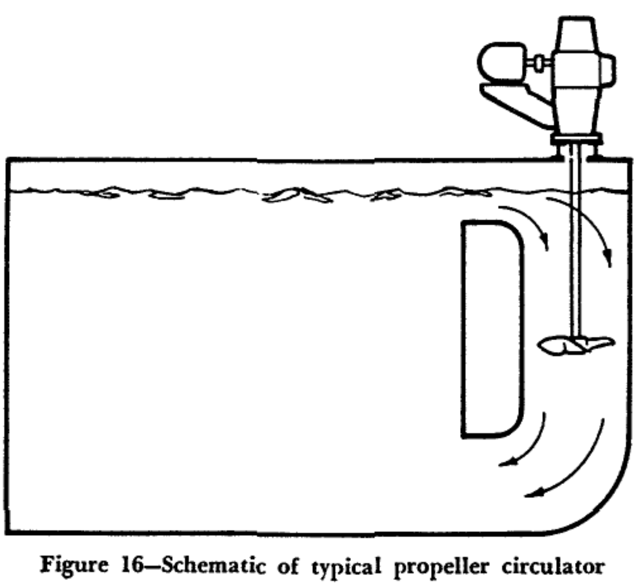


**Figure 16—Schematic of typical propeller circulator**

## CIRCULATORS

In some applications it is desirable to get very high and direct pumping capacity throughout the vessel. In these cases, it is possible to use draft tubes which can be used to circulate the fluid at a controlled velocity on the discharge from a draft tube. By placing these units on the sides of a tank and putting in curved sections, a reliably controlled velocity pattern can be achieved, *Figure* 16.

## POWER RELATIONSHIPS

At the viscosity and power levels in most mixing systems, power will vary with the cube of the impeller speed and fifth power of the impeller diameter for a geometric proportion in a baffled tank. In unbaffled tanks, or in higher viscosities, power will vary with speed to some exponent between the square and the cube and some exponent on the diameter between a cube and a fifth power.

## SCALE-UP AND SCALE-DOWN

As an example of scale-up, let's take a process that is operating satisfactorily in a 160-gallon batch. It is desired to do the same operation in a 2500-gallon batch. *Table* 5 shows what will happen if we keep some of the common parameters constant. At this point it should be mentioned that there is no theoretical or practical reason for keeping any parameter constant on scale-up unless it has been proven by experience or experiment that this is the proper correlation. It is necessary to know how to change a particular chosen scale-up parameter to yield equal process result. On the table it is shown, for example, that keeping power per unit volume constant varies the ratio of everything else if geometric similarity is maintained However, geometric similarity is not required for scale-up. By changing the impeller size to tank size ratio in the last column of *Table* 5, it is possible to keep power per unit volume, flow per unit volume and tip speed constant.

In analyzing any particular process, the important thing is to try to relate one or more of these parameters

**Table 5—Properties of a Fluid Mixer on Scale Up**

| Property | Pilot Scale 160 Gallons | Plant Scale 2500 Gallons | | | |
|---|---|---|---|---|---|
| P | 1.0 | 15.6 | 98 | 6.2 | 15.6 |
| P/Volume | 1.0 | 1.0 | 6.2 | 0.4 | 1.0 |
| N | 1.0 | 0.54 | 1.0 | 0.4 | 0.26 |
| D | 1.0 | 2.5 | 2.5 | 2.5 | 3.9 |
| Q | 1.0 | 8.5 | 15.6 | 6.2 | 15.6 |
| Q/Volume | 1.0 | 0.54 | 1.0 | 0.4 | 1.0 |
| ND | 1.0 | 1.35 | 2.5 | 1.0 | 1.0 |
| $\frac{ND^2\rho}{\mu}$ | 1.0 | 3.4 | 6.2 | 2.5 | 1.57 |
| D/T | 0.33 | 0.33 | 0.33 | 0.33 | 0.52 |

**Table 6—Properties of a Fluid Mixer on Scale Down**

| Property | Plant Scale 2500 Gallons | Pilot Scale 3.4 Gallons | | | |
|---|---|---|---|---|---|
| P | 1.0 | .00137 | .0022 | .0022 | .0022 |
| P/Volume | 1.0 | 1.0 | 1.6 | 1.6 | 1.6 |
| N | 1.0 | 4.3 | 5.1 | 6.4 | 10.1 |
| D | 1.0 | .11 | .11 | .097 | .097 |
| Q | 1.0 | .006 | .007 | .006 | .004 |
| Q/Volume | 1.0 | 4.3 | 5.1 | 4.3 | 1.7 |
| ND | 1.0 | .48 | .56 | .62 | 1.0 |
| $\frac{ND^2\rho}{\mu}$ | 1.0 | .07 | .08 | .06 | .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 |

to the process objective and determine how it should be varied to obtain the desired results.

Scale-down presents a unique problem as illustrated by *Table* 6. A process is operating satisfactorily in a 2500-gallon tank. However, some new materials are being evaluated and it is desired to run a pilot experiment to see what effect they will have on the process. A scale-down from the 2500-gallon tank to a 3.4-gallon tank, which is approximately 10 inches in diameter, by maintaining geometric similarity and equal power per unit volume gives a much higher circulation rate and a much lower level of maximum fluid shear rates.

The second column shows the changes required to increase the power per unit volume which is typically required on scale-down. Then, two changes are made to increase the maximum fluid shear and reduce *the pumping capacity to* levels more similar to those used in the plant. By decreasing the impeller size to tank size ratio, and then decreasing the blade width, the last column, *Table* 6, shows that the impeller tip speed can be maintained at the same level as was used in the plant, while maintaining many of the other ratios in reasonable proportion.

*Notice, however, Table 6, that the* Reynolds number of the small unit is only one-tenth as large as it was in the plant unit and therefore the pilot unit will be much more sensitive to viscosity effects than will the plant size unit.

Pilot planting is a valuable tool but should not be entered into lightly.

## MIXERS AND DISPERSERS

For a given circulation rate, the higher the fluid shearing rate in the vessel, the more power it takes. This higher power introduces heat into the batch at the rate of 2545 BTU per hour per horsepower. Thus a 30 horsepower unit in a 500-gallon tank will add 38,000 BTU in a period of 30 minutes.

It is important to evaluate the proper requirement of a process for flow and fluid shear. For typical "grinding type" operations, the conventional high speed dispersion units are required. However, many of the new pigments and pastes used in letdown operations for industrial coatings can be dispersed adequately at much lower shear rates. These can be obtained by increasing the speed and horsepower levels of conventional turbine type units illustrated in *Figure* 3.

It is also possible to modify the blade proportions on turbine type units so that they have shear rates intermediate between those shown in *Figure* 3 and high speed dispersers. Power savings of one-half to two-thirds can often be realized by properly designed turbine type units over standard high speed dispersers.

Much work is being devoted at the present time to evaluate whether a given process requires high shear dispersion equipment or whether by increasing the power levels and shear rates of turbine mixers, satis-

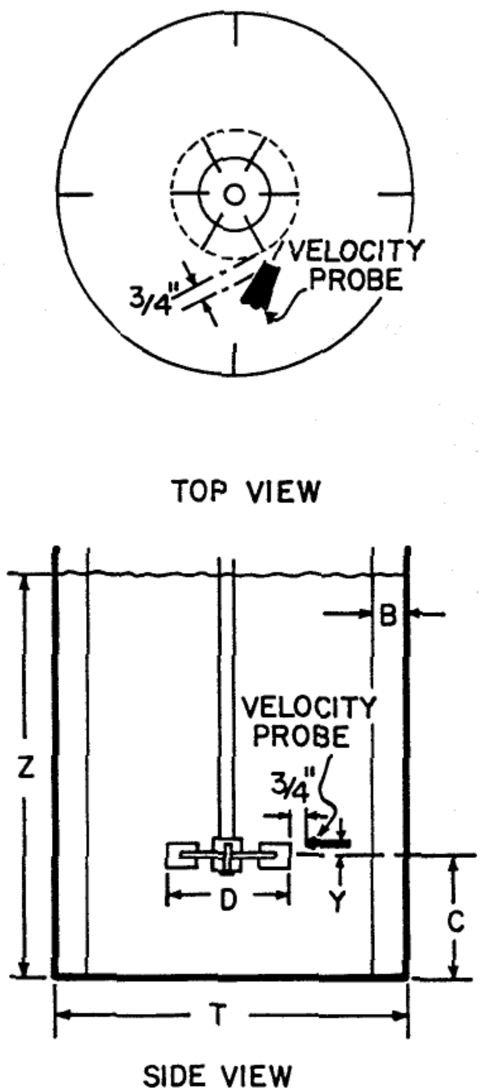
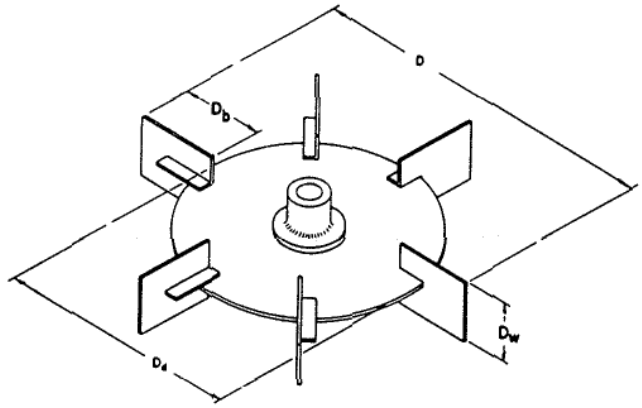



**Figure 17—Typical turbine installation in baffled tank, showing location of hot-wire velocity probe**




$D : D_b : D_w : D_d = 1 : \frac{1}{4} : \frac{1}{5} : \frac{2}{3}$

**Figure 18—Typical dimensions of flat blade turbine impeller**

factory letdown and blending can be achieved. Carefully conducted pilot plant or plant scale tests can reveal the fluid shear requirements of any particular process and allow the prediction of full scale performance.

## SOME EXPERIMENTAL DATA ON FLUID SHEAR RATES IN MIXING TANKS

If the objective were to find only the mean velocity at a given point, then some modification of a pitot tube would be the most practical method. However, it was desired to determine the rapid fluctuations in a stream, and so the hot-wire velocity meter was used as shown in *Figure* 17. The primary purpose here is to give the magnitude and ratios of some of the effects, and to avoid becoming involved in complicated mathematical discussions of flow and turbulence. For simplicity, only the measurements made when the probe was positioned as shown in *Figure* 17 will be considered here.

**Figure 19—Flow pattern, side view, 4-inch diameter flat blade turbine, 12-inch diameter tank, water. Streak photography showing particle velocities in a ¼-inch wide vertical plane**

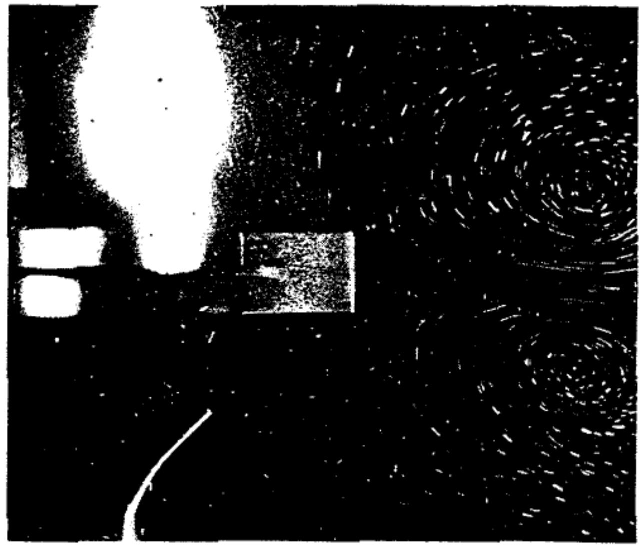


**Figure 20—Typical flow pattern from flat-blade turbine in baffled tank. Side view, 4-inch diameter turbine, 12-inch diameter tank, water. ¼-inch wide horizontal plane in impeller centerline**

The pumping capacity of flat-blade turbines has been measured by various investigators. The flat-blade disc turbine referred to here is of the type shown in *Figure* 18. It has six blades, and the dimension ratios given on the figure.

One of the most complete works on pumping capacity is that of Sachs,[8] who used the photographic technique of analysis for velocity patterns. To illustrate the type of flow patern, *Figures* 19 and 20 are included from his work. They show the overall type of flow developed by a flat-blade turbine.

To obtain more information on the nature of flow from a turbine impeller, a hot-wire velocity meter was developed for use in mixing systems. The measuring element consists of a small wire suspended between two electrodes. The wire was 0.100 in. long and 0.0007 in. in diameter. The wire was heated electrically to about 20 F (7 C) above the tank temperature. As the water flows by the probe, it tends to cool it, and thus the amount of current required to maintain constant temperature is related to the velocity of the fluid.

The electronic arrangement of the equipment is as described by Hubbard.[4] A hot-wire probe gives the maximum flow rates when the wire is at right angles to the flow stream. In this study, the wire was always positioned in a horizontal plane and was rotated at various angles to a radius from the centerline of the tank until maximum velocity readings were obtained. From this study, it appeared that the angle for maximum velocity corresponded to a tangential velocity component at the impeller periphery.

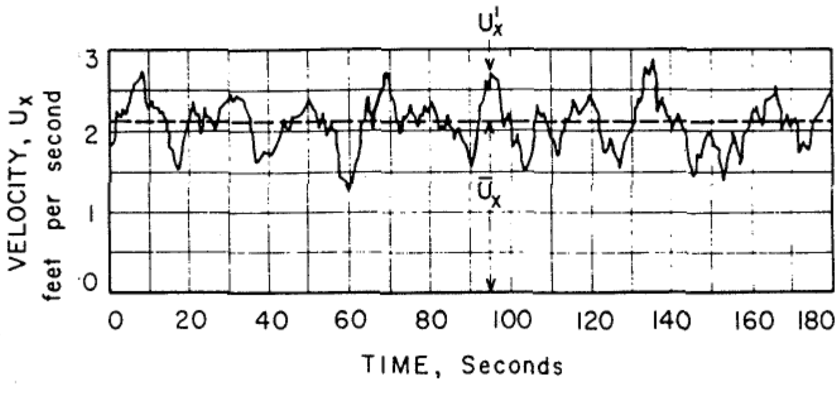



Figure 21—Typical velocity from hot-wire velocity probe. This is not a chart from this experimental program

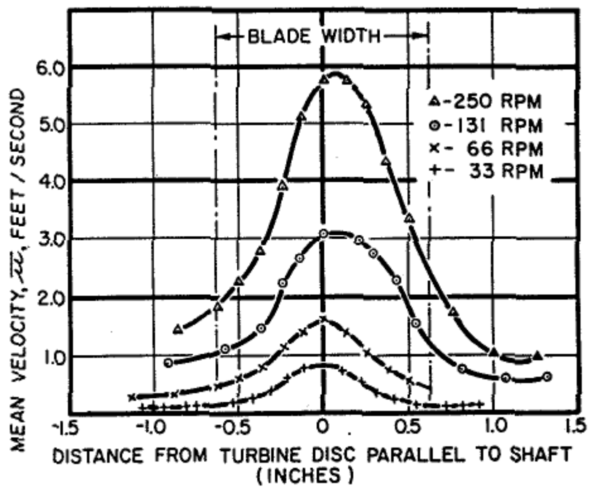



Figure 22—Mean velocity, ū at various positions above and below a 6-inch diameter turbine, 6 flat blades

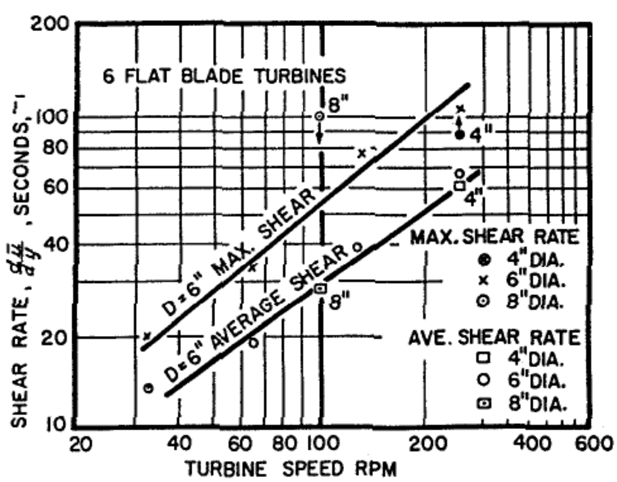



Figure 23—Maximum and average shear rates obtained from Figure 6

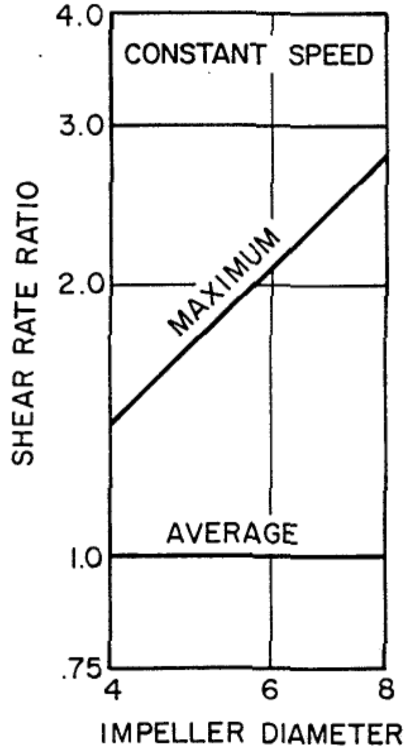



Figure 24—Ratio of maximum and average shear rates for 4-inch, 6-inch, and 8-inch flat blade turbine impellers, compared to average shear rate for 6-inch diameter impeller

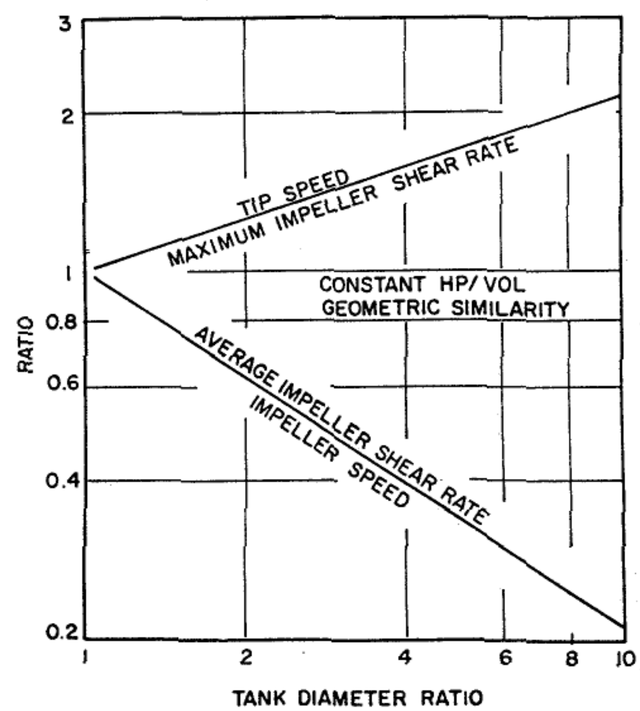



Figure 25—Typical changes on scale-up. The power per unit volume is held constant and geometric similarity is maintained

In the data reported here, the probe was positioned ¾ of an inch away from the turbine periphery, directed along a tangent to the impeller periphery, and was positioned at various distances above and below the turbine centerline.

The tank diameter, T, used in this work was 18 inches and the liquid level, Z, was 20 inches. Impeller diameters, D, used were 4, 6 and 8 inches. Four baffles were used in the tank, each 1.5 inch wide.

*The hot-wire velocity meter has a very fast response*, so that the velocity, u, at a point at any time *can be expressed as a mean velocity over a time interval*, $\bar{u}$, plus a fluctuating velocity component, u′, so that

$$u = \bar{u} + u'$$

The probe was calibrated by placing it into the flow fom an orifice in a tank with a constant head for the period of the calibration. The orifice was constructed very carefully in accordance with fluid mechanics standards, and *the flow through the orifice* calculated from a knowledge of the static head and the orifice coefficient.

A typical trace from a hot-wire probe is shown in *Figure* 21. This is *not* an actual recording from this experiment.

*Figure* 22 shows the results at three different speeds as the probe was moved up and down from the horizontal plane through the impeller center.

The average value of the slope of the mean velocity, $\bar{u}$, versus distance line is plotted in *Figure* 23. The maximum slope measured is also plotted in *Figure* 23. With the 6-inch flat-blade turbine, the maximum shear rate was approximately twice the average

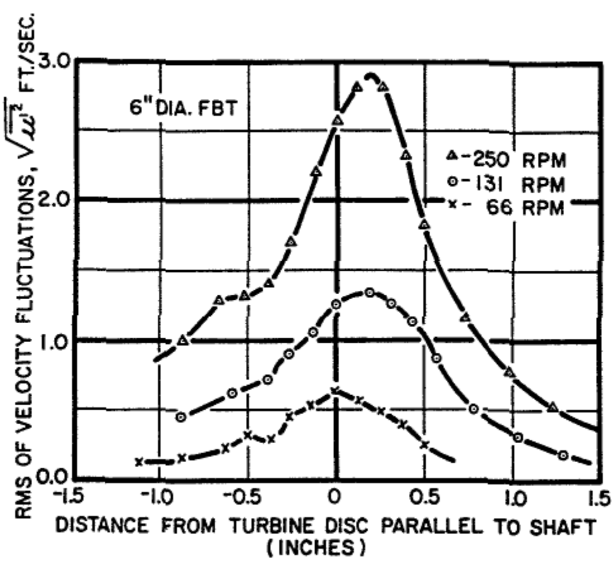



**Figure 26—RMS of velocity fluctuations for 6-inch diameter flat blade turbine as shown in Figure 22**

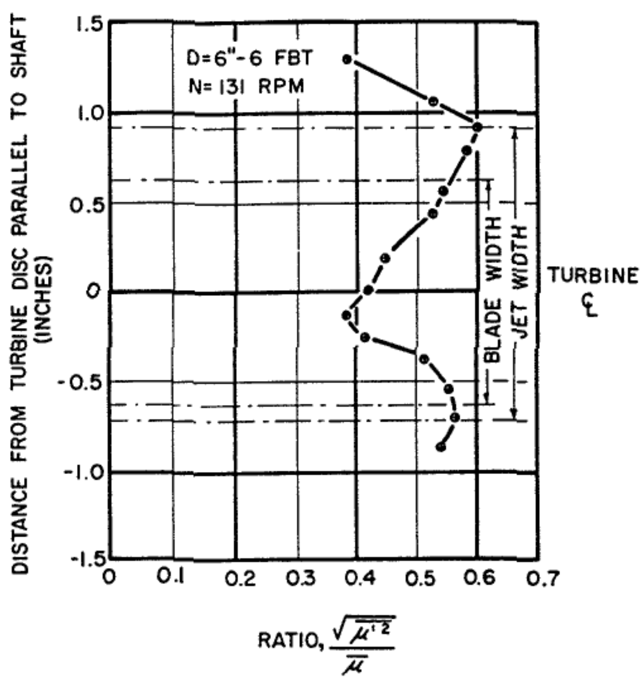



**Figure 27—Ratio of RMS velocity fluctuations to mean velocity at corresponding points**

shear rate. However, as the impeller size was decreased, or increased, the average shear rate remained about the same at the same impeller speed, *Figure* 24. The maximum shear rate increased with impeller diameter at the same impeller speed. This indicates that, as scale-up progresses, lower average shear rates will be encountered in the stream from the impeller due to lower impeller speeds, but higher maximum shear rates, due to the higher peripheral velocity of the impeller.

The next question, then, is: What kinds of velocity fluctuations are found in this stream?

By taking the root mean square of the velocity fluctuations, one obtains a measure of the intensity of the turbulent fluctuations. This particular probe (and electronic equipment) considers all fluctuations having a frequency of 100 cps or higher.

To obtain a complete picture of the turbulent fluctuations and energy decay and dissipation, it would be necessary to actually analyze the complete spectra of frequency and size of the turbulent currents.

All the horsepower applied to the tank appears as heat, and since this is generated by viscous shear, when the eddy becomes small enough, its energy is dissipated by that mechanism. Therefore, processes that occur at the molecular level may well be a function only of energy input per unit volume.

As can be seen from *Figure* 26, the intensity of turbulence follows a pattern similar to the mean velocity. The ratio of the root mean square of the fluctuations to the mean velocity is between 0.4 and 0.6 *within the jet from the impeller, Figure* 27. ♦

## Nomenclature

B ........ Baffle width
D ........ Impeller diameter
$D_d$ ........ Disc diameter
$D_w$ ........ Blade width
$D_w/D$ ........ Blade width to impeller diameter ratio
D/T ........ Impeller diameter to tank diameter ratio
HP ........ Horsepower
N ........ Impeller rotational speed, RPM
$N_{Re}$ ........ Reynolds number, ratio of inertia force to viscosity force ND /
P ........ Power
P/V ........ Power per unit volume of liquid
Q ........ Flow from the impeller per unit time
Q/V ........ Flow per unit time per unit volume of tank
RMS ........ Root mean square
RPM ........ Revolutions per minute
T ........ Tank diameter
u ........ Velocity
$\bar{u}$ ........ Mean velocity over a time interval
$u'$ ........ Fluctuating velocity component
V ........ Volume
y ........ Height above impeller centerline horizontal plane
Z ........ Liquid depth
Z/T ........ Liquid depth to tank diameter ratio
$\mu$ ........ Viscosity

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