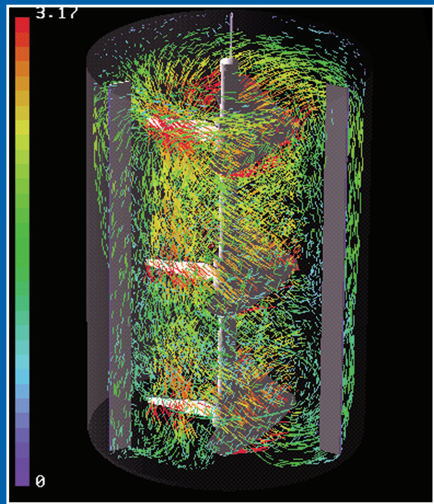


A GUIDE TO FLUID MIXING was written for those who desire a quick, concise approach to the principles of fluid mixing. Written primarily as an introductory text or primer for chemical engineering students, it is also useful to those in the Chemical Process Industries (CPI) who have found the need to acquaint themselves with this topic. Unfortunately, due to heavy academic loads and time constraints imposed by the typical undergraduate curriculum, many chemical engineers have not had the opportunity to explore the many facets of this exciting field. Fluid mixing is indeed a topic which deserves more attention than it has received if, for no other reason, one considers its tremendous impact on almost every production process and every industrial and consumer product.

In addition, this book also contains a laboratory manual which covers a broad selection of fluid mixing experiments.

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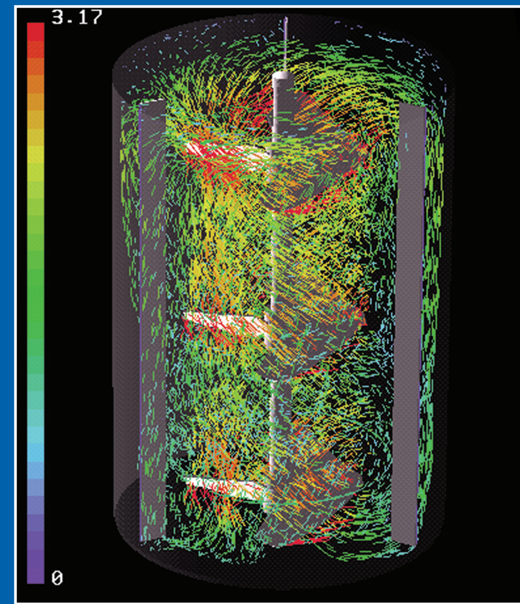
A Guide To Fluid Mixing by J.Y. Oldshue, N.R. Herbst and T.A. Post




# A Guide To Fluid Mixing

by J.Y. Oldshue, N.R. Herbst and T.A. Post

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# A GUIDE TO FLUID MIXING

The cover photograph shows up-pumping mixing technology utilized in multi-phase processes. Using impellers that pump upwardly, rather than radically or downwardly, can provide a superior solution.

# A GUIDE TO FLUID MIXING

J.Y. Oldshue, N.R. Herbst, T.A. Post

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# PREFACE

A GUIDE TO FLUID MIXING was written for those who desire a quick, concise approach to the principles of fluid mixing. Written primarily as an introductory text or primer for chemical engineering students, it is also useful to those in the Chemical Process Industries (CPI) who have found the need to acquaint themselves with this topic. Unfortunately, due to heavy academic loads and time constraints imposed by the typical undergraduate curriculum, many chemical engineers have not had the opportunity to explore the many facets of this exciting field. Fluid mixing is indeed a topic which deserves more attention than it has received if, for no other reason, one considers its tremendous impact on almost every production process and every industrial and consumer product. This book is devoted to presenting the basics and whetting the appetite for future study. Compiled from decades of research and development and industrial experiences, it is documented in six easy-to-assimilate chapters.

Chapter 1 is a brief introduction to THE MIXING PROCESS, THE MIXER, ELEMENTS OF MIXER DESIGN and FLUID MIXING VARIABLES. Chapter 2 discusses important BASIC CONCEPTS including DIMENSIONLESS GROUPS, FLUID REGIMES, POWER RESPONSE, IMPELLER DESIGN and the IMPELLER SPECTRUM. Chapter 3 examines FLOW-CONTROLLED OPERATIONS which include FLUID BLENDING, SOLIDS SUSPENSION and HEAT TRANSFER. Chapter 4 explains the types of GAS-LIQUID OPERATIONS. Chapter 5 introduces PILOT PLANT OPERATIONS and SCALE-UP and Chapter 6 explores basic MECHANICAL DESIGNS.

In addition, a laboratory manual (Chapter 7) is included. It contains experiments covering POWER CONSUMPTION OF MIXING IMPELLERS, POWER NUMBER/REYNOLDS NUMBER DETERMINATION, BAFFLE DESIGN, SOLIDS SUSPENSION, LOW VISCOSITY BLENDING and GAS-LIQUID MASS TRANSFER.

This book is presented as the first step in examining the truly challenging field of fluid mixing. It is hoped that the reader will respond to this text with a desire to further enhance his or her knowledge of mixing. To that end we recommend the book, FLUID MIXING TECHNOLOGY by J.Y. Oldshue (Chemical Engineering, McGraw-Hill Publishing Company, 1983).

## THE AUTHORS

The late Dr. J.Y. Oldshue was the author of FLUID MIXING TECHNOLOGY (Chemical Engineering, McGraw-Hill Publishing Company, 1983) and numerous papers about the mixing process. He was President of Oldshue Technologies International, having retired from LIGHTNIN, a unit of General Signal, after four decades of service. While at LIGHTNIN he held several key technical positions, serving as Vice-President of Mixing Technology prior to his retirement.

N.R. Herbst was Senior Technical Marketing Engineer upon retiring from LIGHTNIN where he held the positions of Test Engineer, Marketing Services Engineer and Technical Marketing Engineer over the past several years. In addition, he has had broad experience in Research and Development, and Application Engineering.

Dr. T.A. Post was Vice-President of Mixing Technology at LIGHTNIN and responsible for many aspects of new technology development.

# one

## INTRODUCTION

### THE MIXING PROCESS

Fluid mixing plays a major role in processing a multitude of products. Most consumer and industrial products require some kind of fluid mixing in their preparation and application. Fluid mixing is no doubt the most common operation encountered in the Chemical Process Industries (CPI) and allied industries as well. Products obtained from food, petroleum, mining, pharmaceutical, pulp and paper and chemical industries would be unavailable were it not for fluid mixing equipment and fluid mixing technology. Municipal and industrial waste treatment and water treatment would suffer as well without fluid mixers, aerators and flocculators. Furthermore, fluid mixers are essential in helping to protect the environment. For example in Flue Gas Desulfurization (FGD) - a process which helps control air pollution (Acid Rain) - fluid mixers play an important role in sulfur dioxide absorption.

Mixing process requirements vary and include (but are not limited to) blending low viscosity fluids with high viscosity fluids, suspending solids in liquids, dispersing gases or solids in liquids, mass transfer and heat transfer. The specific type of mixing for a given process also varies from relatively low energy-intensive, simple fluid motion to high energy-intensive mixing which requires high levels of fluid shear and turbulence.

A general classification of mixing applications as they apply to physical and chemical processes is given in Table 1-1. Here we see that a specific application class is associated with both physical processing and chemical processing. There are also many processes which involve more than one application class - for example, liquid-solid-gas and liquid-liquid-solid-gas.

TABLE 1-1 MIXING PROCESSES

| Physical Processing | Application Class | Chemical Processing |
|---|---|---|
| Suspension | Liquid-Solid | Dissolving |
| Dispersion | Liquid-Gas | Absorption |
| Emulsification | Immiscible Liquids | Extraction |
| Blending | Miscible Liquids | Reactions |
| Pumping | Fluid Motion | Heat Transfer |

## THE MIXER

A fluid mixer is a device usually consisting of a drive mechanism, a shaft and one or more impellers mounted on the shaft. A mixing impeller is any device used on a rotating shaft to circulate materials in a mixing vessel. Rotating mixing impellers promote an interchange of materials within the system in order to satisfy the process requirement. There are also "static-type" mixers which have stationary mixing elements and no moving parts. They are normally used in process pipelines where power and flow are derived from the process stream.

There are various kinds of mixing impellers including propellers, shrouded turbines, disk turbines, rakes, gates, paddles, pitched-blade turbines, etc. In general, mixing impellers can be assigned to two broad categories - axial flow impellers and radial flow impellers.

Axial flow impellers have a principal direction of discharge which coincides with the axis of impeller rotation. Fig. 1-1 shows examples of axial flow impellers and Fig. 1-2 illustrates the typical axial impeller flow pattern.

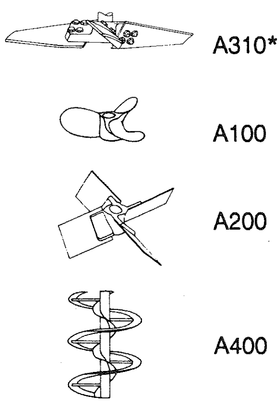



Fig. 1-1
Axial Flow Impellers

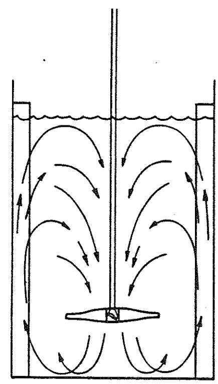


Fig. 1-2
Typical Axial Flow Pattern

Radial flow impellers have a principal direction of discharge normal to the axis of rotation. Figs. 1-3 and 1-4 illustrate examples of radial flow impellers and their typical flow patterns.

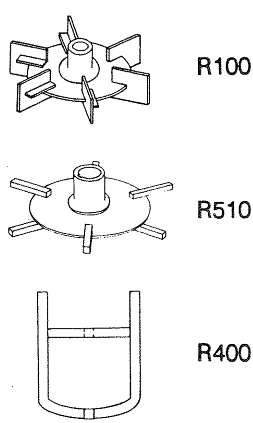



Fig. 1-3
Radial Flow Impellers

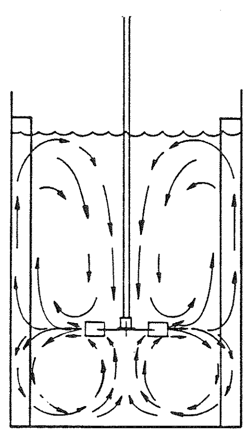


Fig. 1-4
Typical Radial Flow Pattern

* See note on impeller designations, page 32.

Industrial fluid mixers are manufactured in a wide range of sizes and types to accommodate mixing requirements from bench scale to large production scale:

1. Laboratory Mixers - usually less than 1/8 hp (0.1 kw), variable speed devices mounted on a ring stand with impeller sizes suitable for laboratory beakers and small vessels. (Fig. 1-5).

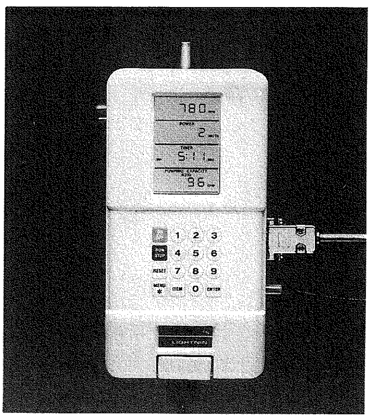


Fig. 1-5 Variable speed Laboratory Mixer

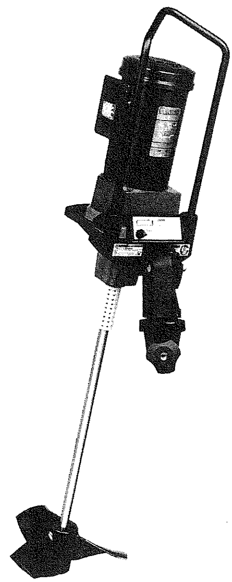


Fig. 1-6 Gear Drive Portable Mixer

2. Portable Mixers - typically 1/4 to 3 hp (0.2 to 2.2 kw) mixers equipped with a clamping device suitable for mounting on the rim of a vessel. They are called portable because they are not permanently supported and are often not permanently wired to a starter. They can simply be unplugged, unclamped and moved quickly from one vessel to another (Fig. 1-6).

3. Top-Entering Mixers - normally 5 to 250 hp (3.75 to 187 kw) or greater, mounted on permanent supports above or on top of the mixing tank (Fig. 1-7). These mixers can be used on open or closed tanks. The shaft is vertical, with one or more impellers. A shaft sealing device is sometimes required depending on the application.

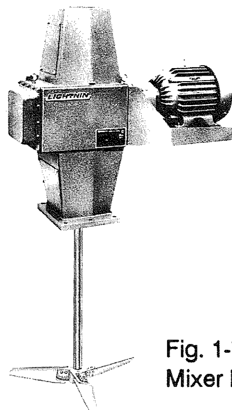



Fig. 1-7 Top-Entering Mixer Installation

4. Side-Entering Mixers - typically from 1 to 75 hp (0.75 to 56 kw), mounted on a flange on the side of a tank normally near the bottom. The shaft is horizontal, extending into the tank and equipped with a single axial flow impeller (Fig. 1-8). A shaft sealing device is always required with side-entering mixers.

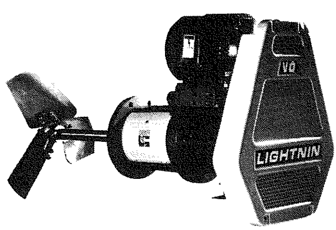



Fig. 1-8 Side-Entering Mixer

5. Bottom-Entering Mixers - mounted permanently under the bottom of the tank. A shaft sealing device is always required with this kind of mixer (Fig. 1-9).

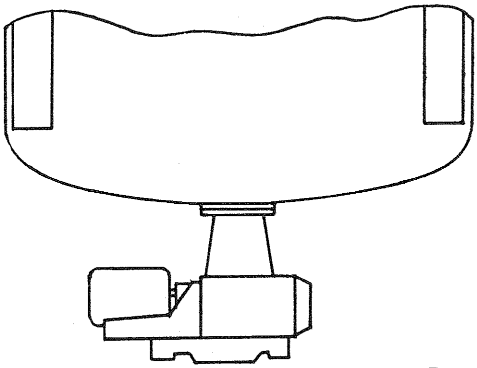


Fig. 1-9
Bottom-Entering Mixer

6. Static or Motionless Mixers - these are constructed with stationary, consecutive, alternately pitched mixing elements encased in a conduit. Power is supplied from a flowing stream derived from a pump, gravity or pressure source. Mixing is achieved by stream splitting, reversing stream rotation and re-combining the process stream components (Fig. 1-10).

Fig. 1-10 "Motionless" Mixer

## ELEMENTS OF MIXER DESIGN

There are three major elements of mixer design - process design, impeller power characteristics and mechanical design criteria. (Table 1-2).

**Table 1-2 MAJOR MIXER DESIGN ELEMENTS**

| | |
|---|---|
| 1. Process Design | - Fluid Regime Required by the Process |
| | - Fluid Mechanics of Impellers |
| | - Scale Up |
| 2. Impeller Power Characteristics | - Impeller Speed, Power & Diameter Relations |
| 3. Mechanical Design | - Impellers |
| | - Shafts |
| | - Drive Assembly |

Process design is concerned with the fluid regime required (e.g., turbulent, transitional or laminar), impeller fluid mechanics (flow, head and shear) and the translation to various vessel sizes (scale-up).

Impeller power characteristics relate the combined effects of fluid properties, impeller geometry, impeller diameter and speed to power consumption.

Mechanical design elements include the drive (speed reducer or gearbox), output shaft, impeller, shaft sealing device (when required) and the prime mover (the initial agent that puts a machine in motion).

## FLUID MIXING VARIABLES

There are a number of fluid mixing variables and design parameters which may affect the process result. (The term process result means what we hope to achieve in the mixing vessel - e.g., a 3 minute blend time, an 80% yield, a uniform suspension, rapid mass transfer, etc.). However, those variables and parameters that are significant for one process may not be significant for another. Furthermore, it is important to note that *power characteristics of mixing impellers are independent of the process result.* Thus an understanding of basic fluid mixing technology and the criteria by which processes are scaled up or down is essential.

**Variables Associated With Fluid Mixing / Reference Conditions**

The following notations are widely used to define fluid mixing variables with key variables in boldface type:

**1. NOTATION**

A = transfer area
B = baffle width
C = impeller distance from the tank bottom to the horizontal centerline of the impeller blades
Cp = specific heat of the liquid
**D = impeller diameter**
d = diameter of heat transfer tube
g = gravitational acceleration
gc = gravitational acceleration constant
H = impeller velocity head
h = heat transfer coefficient
k = thermal conductivity of liquid
**N = impeller speed**
**P = power**
P/V = power per unit volume
**Q = impeller pumping capacity**
q = average heat transfer rate
S = spacing between multiple impellers
**T = tank diameter**
u = heat transfer coefficient
V = volume
**Z = fluid depth in vessel**
**$\rho$ = density of fluid or solid**
**$\mu$ = fluid viscosity**
$\mu_s$ = viscosity of fluid at fluid film temperature at heat transfer surface
$\sigma$ = surface tension

**SUBSCRIPTS**

l = liquid
s = solid
u = ungassed
v = per unit volume

## 2. Reference Conditions and Reference Geometry

In order to compare the performance of different impeller designs, the following reference and geometric conditions apply (Fig. 1-11).

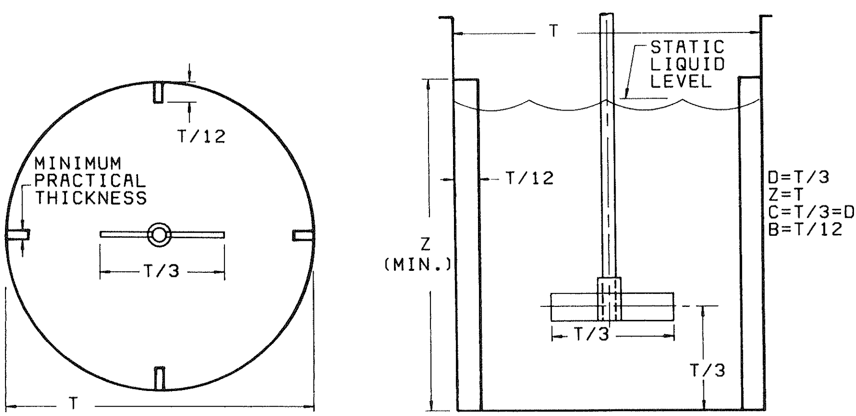



Fig. 1-11 Reference geometry where vessel is vertical, cylindrical, flat bottom and reference conditions where fluid specific gravity = 1.0 and viscosity = 1.0 cP (0.001 Pa · s). Four vertical, flat wall baffles required as shown.

Although reference conditions serve to quantify the performance of one impeller design with respect to another, actual *process* conditions may vary widely from reference conditions and reference geometry. Therefore, reference conditions and reference geometry represent an arbitrary standard set of conditions used as a basis for comparison.

# two

## BASIC CONCEPTS

### DIMENSIONLESS GROUPS

In fluid mixing technology, force ratios presented as dimensionless groups are used in correlating scale-up parameters. Relating inertia force (the input force from the mixer) to other forces yields the following ratios:

1. The ratio of inertia force, $F_I$ , to viscous force, $F_V$ , is the Reynolds number, $N_{Re}$ .

$$\frac{F_I}{F_V} = N_{Re} = \frac{ND^2 \rho}{\mu} \qquad (2\text{-}1)$$

2. The ratio of inertia force, $F_I$ , to gravitational force, $F_g$ , is the Froude number, $N_{Fr}$

$$\frac{F_I}{F_g} = N_{Fr} = \frac{N^2 D}{g} \qquad (2\text{-}2)$$

3. The ratio of inertia force, $F_I$ , to surface tension force, $F_\sigma$, is the Weber number, $N_{We}$

$$\frac{F_I}{F_r} = N_{We} = \frac{N^2 D^3 \rho}{\sigma} \qquad (2\text{-}3)$$

Other dimensionless groups define Power number, $N_P$, and Flow number, $N_Q$ :

$$N_P = \frac{Pg}{N^3 D^5 \rho} \qquad (2\text{-}4)$$

(ratio of applied force to mass times acceleration)

$$N_Q = \frac{Q}{ND^3} \qquad (2\text{-}5)$$

(where Q = primary impeller pumping capacity)

## Power Number vs. Reynolds Number Relations

If we were to run experiments in the laboratory to determine the relationships of impeller power consumption, impeller speed and diameter, fluid density and viscosity for a specific impeller design, here is what we would find:

1. Impeller power consumption in a fluid of constant viscosity and density is proportional to the impeller speed cubed (Fig. 2-1).

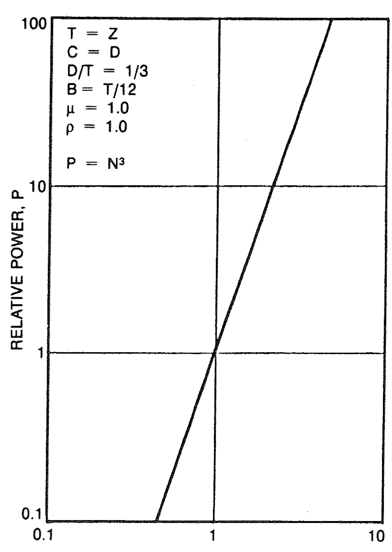



Fig. 2-1
Impeller power consumption as a function of speed

2. If we repeated the experiment for several different impeller diameters, maintaining reference conditions and the same impeller geometry, we would find that power consumption is proportional to the impeller diameter raised to the 5th power (Fig. 2-2)

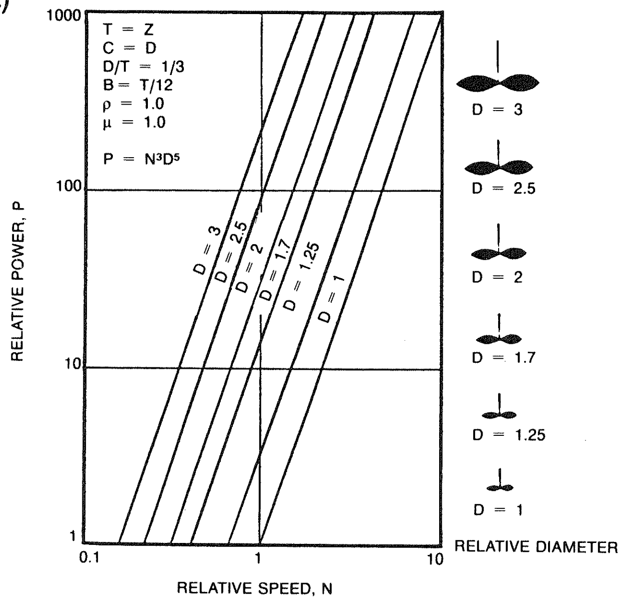



Fig. 2-2 Impeller power consumption as a function of speed and diameter

3. Furthermore, if we varied the fluid density we would find that power consumption is directly proportional to density,

$$P \propto \rho \qquad (2\text{-}6)$$

and for the total effect including density, speed & diameter

$$P \propto N^3 D^5 \rho \qquad (2\text{-}7)$$

The ratio of P to $N^3D^5\rho$ is proportional to the power number (dimensionless).

$$N_P \propto \frac{P}{N^3 D^5 \rho} \qquad (2\text{-}8)$$

## FLUID REGIMES OF MIXING IMPELLERS

Mixing impellers operate in various fluid regimes depending on the mixing Reynolds number. For each impeller design, $N_P$ vs $N_{Re}$, data can be generated to produce a curve characteristic of the specific impeller design. When this is done, typical curves are obtained as shown in Fig. 2-3.

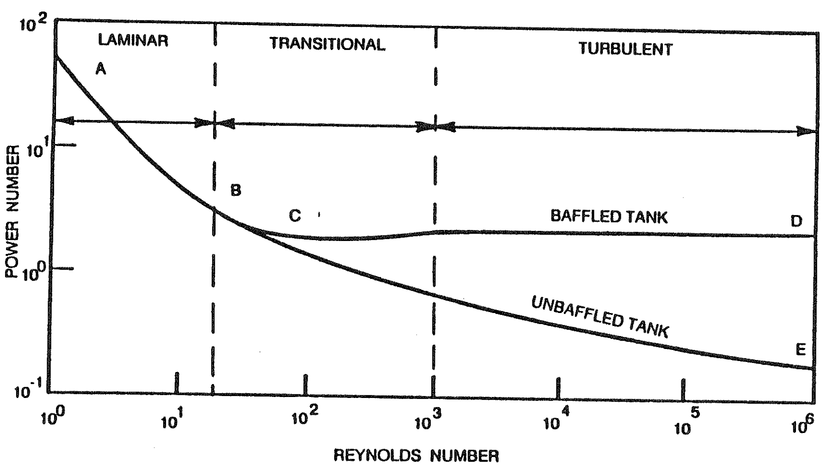



Fig. 2-3 Typical Power No. vs Reynolds No. Curves at Reference Geometry

Note that the curves are divided into three distinct regimes or regions - laminar, transitional and turbulent flow. This is similar to the friction factor curves for flow in pipes. In the LAMINAR region, power number varies inversely as Reynolds number. In the TRANSITION region, slope varies continuously and in the TURBULENT region (in a baffled tank) slope is constant (zero) therefore power number is also constant in the turbulent region.

If we had three completely different impeller designs, three separate curves could be generated as symbolized in Fig. 2-4.

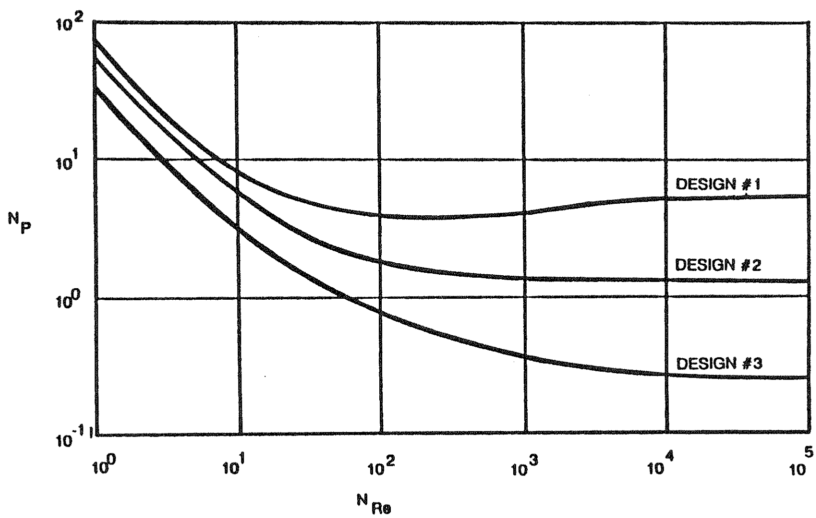



Fig. 2-4 $N_P$ vs $N_{Re}$ for Three Different Impeller Types

When comparing power numbers of different impeller designs, the Reynolds number must be held constant. But in the turbulent regime $N_P$ is constant for a given impeller design through a wide range of Reynolds numbers. This is often the basis for comparison. For example, if it is said that the power number for impeller R is 5.0 and for impeller A is 1.0 the turbulent regime is always implied unless otherwise specified. Because the turbulent regime begins at different Reynolds numbers for different impellers, we define the turbulent regime for *all* impellers at $N_{Re} > 10^5$. At this value and beyond, all mixing impellers operate in turbulence.

## APPLIED POWER

The power applied by any mixing impeller produces a pumping effect and a velocity head which is expressed as:

$$P \propto QH \qquad (2\text{-}9)$$

where,

$P$ = power input
$Q$ = flow or pumping capacity
$H$ = velocity head

The agitation intensity in the immediate discharge zone of the impeller is defined by the fluid head which is proportional to shear rate or turbulence. In fluid mixing technology the terms "head" and "shear" are used interchangeably. Either the flow component, Q, or the head component, H, can be emphasized such that a large flow/small head or a large head/small flow can be produced for the same power input, P. For a given power input, an operation requiring a high flow will need a large impeller running at a low speed, while an operation requiring a high head (or shear) will use a small impeller at a high speed. These two operations are called flow-controlled and shear-controlled respectively. The pumping capacity of the mixing impeller controls the circulation rate in the mixing vessel.

The power input of a rotating mixing impeller depends on its geometry, diameter, D, speed, N, its location in the mixing vessel, and the physical properties of the fluid. The major fluid properties upon which impeller power depends are the density, $\rho$ , and the viscosity, $\mu$ . The typical relationship between the dimensionless power number, $N_P$, and mixing Reynolds number, $N_{Re}$, has been shown in Fig. 2-3 and Fig. 2-4.

## BAFFLES AND IMPELLER LOCATION

Baffles are flat metal plates, usually four in number, equally spaced and attached to the vessel wall. Without baffles, swirling and vortexing results and very little real mixing takes place. An unbaffled tank has virtually no vertical currents and no velocity gradients, both of which are necessary for mixing. Furthermore, the vortex formed in the absence of baffles may incorporate air bubbles producing an undesirable chemical effect (e.g., oxidation) or an undesirable physical effect (e.g., frothing, foaming). The power consumption of a mixing impeller is also reflected by the degree of baffling. Referring once again to Fig. 2-3, the slope of the curve (AB) in the laminar or viscous range is approximately minus one,

$$N_P \propto (N_{Re})^{-1} \qquad (2\text{-}10)$$

where power consumption is much more sensitive to small changes in Reynolds number and the need for baffles gradually diminishes to none at all.

The curve labeled CD of Fig. 2-3 shows power response in a baffled tank, while curve BE is for an unbaffled tank. The turbulent range describes the mixer input to the system in low viscosity (high Reynolds number) fluids. Baffles are almost always required for low viscosity operations with top entering mixers.

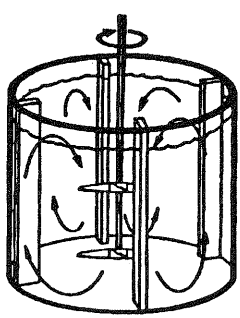


Fig. 2-5 Typical Mixing Tank with Baffles

The proper design of tank baffles is important in obtaining satisfactory mixing results. Adequate baffling assures good mixing by providing a flow pattern that carries throughout the entire batch. Excessive baffling (also called over-baffling) reduces the mass-flow and localizes the mixing, and may result in poor process performance. An appropriate mixing environment includes proper tank and baffle designs, and efficient use of applied power.

## Top-Entering Mixers Mounted on the Tank Centerline

Baffles must be provided for low viscosity mixing when top entering mixers are located on the tank centerline. This also includes fixed-mounted (portable-size) mixers mounted vertically on the tank centerline. Baffles assure that the entire batch will pass through the zone of the impeller where there is maximum agitation intensity. Other effects of using baffles are that they:

....promote the flow pattern required for the process,

....direct flow from the impeller, producing the required vertical currents and velocity gradients,

....change the flow from a rotary (essentially static) pattern to a mixing pattern,

....prevent excessive swirling, vortexing and air induction,

....greatly improve mixer loading accuracy,

....assure a stable, consistent power draw,

...produce more uniform, radial shaft loads (less buffeting) - a mechanical design consideration, which allows longer shaft lengths.

An unbaffled system would produce wide power variations, making efficient loading difficult and motor underloading probable. Without baffles, only the simplest agitation could be accomplished - requiring an extremely low power input.

## Power Response in Low Viscosity Fluids

For low viscosity fluids (fluid viscosity <50,000 cP {50 Pa · s}) vertical-cylindrical tanks are normally equipped with four baffles, 1/12 the tank diameter in width, extending vertically along the straight side of the tank and located 90° apart. Wider baffles produce slightly stronger vertical mixing currents but they may also act as flow dampeners by reducing mass flow and prohibitively reducing rotary motion (some rotary motion is always required). While using *four* baffles has been the generally accepted standard for many years, the latest high efficiency axial flow impeller designs require only three baffles.

Fewer baffles or narrower baffles allow more rotary motion, or tangential mass flow, but also reduce impeller power draw. Reducing the power draw limits the energy that can be applied to the batch. At higher power levels swirling is developed when baffles are inadequate or when there are no baffles.

The relative power response as a function of the number and width of baffles is shown in Figure 2-6. A value of 100% power response has been assigned to standard baffles. (Standard baffles are 1/12 the tank diameter in width and four in number, equally spaced.)

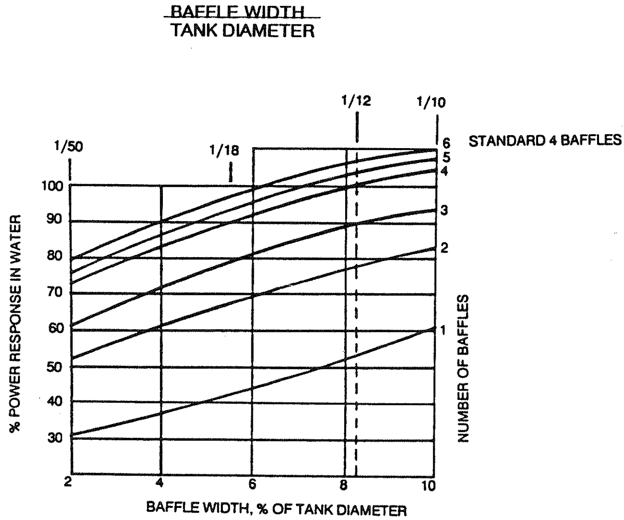



Fig. 2-6 RELATIVE POWER RESPONSE AS A FUNCTION OF BAFFLE SIZE AND NUMBER

Note that a reduction in number and/or width of baffles (from the standard) produces a rapid loss in power drawn by the impeller, whereas an increase in the number and/or width (beyond the standard) has little effect on increasing power. Baffles, as with any other tank internals, should be only as large as necessary. Figure 2-6 is valid for viscosities between 1 and 100 centipoise. It is presented to show the effect of underbaffling on power consumption and to show how mixers may be underloaded.

## Viscosity Effect - Newtonian Fluids

As batch viscosity increases, so does the viscous drag on the tank wall or other internals, thereby reducing the need for tank baffles for proper loading. For normal or Newtonian fluids (where viscosity is independent of fluid shear rate), tank baffles for both radial and axial flow impellers must be reduced as viscosity increases (Fig. 2-7).

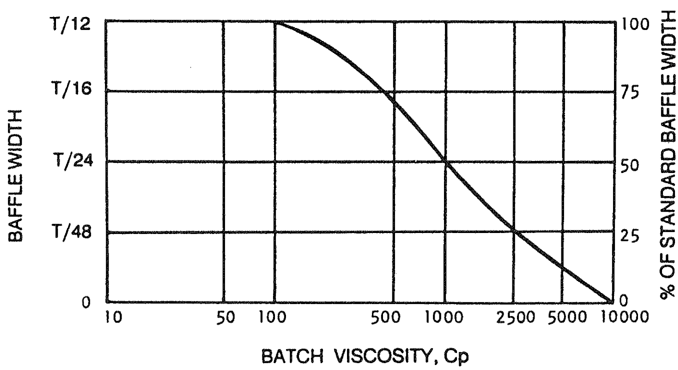



Fig. 2.7 BAFFLE WIDTH AS A FUNCTION OF VISCOSITY

## Viscosity Effect - Non-Newtonian Fluids

Non-Newtonian fluids encountered in mixing are usually pseudoplastic (the apparent viscosity decreases with increased shear rate). Since highest shear rates are encountered in the region of the impeller, it is common for these non-Newtonian materials to be fluidized around the impeller, while remaining stagnant in remote parts of the tank (on the walls and at the surface of the batch). Tank baffles increase the tendency for stagnation and must be reduced in size or in some cases, eliminated when mixing non-Newtonian fluids.

In viscous media the power characteristics of mixing impellers are much different from the power characteristics in lower viscosity fluids. This is evidenced by the slope in the laminar regime of the power number vs Reynolds number curve (see Fig. 2-3 - as a rule of thumb, viscosities of 50,000 cP (50 Pa·s) or greater are usually considered to be in the laminar or viscous region).

Baffles are not normally required in very viscous systems, since the desired vertical flow currents and overall blending flow pattern can be obtained without them. The viscous drag of the fluid acts as a dynamic baffle. In highly viscous systems, the presence of baffles, internal tank fittings (e.g., heating or cooling coils) and impeller stabilizing fins may actually impede the mixing process. Conversely, high intensity mixing in low viscosity fluids requires the presence of baffles to develop the localized turbulence that produces high-shear stresses (see Chapter 3, Flow-Controlled Operations, for a discussion of shear stress and shear rate.)

Rectangular tanks don't normally require baffles. The corners of the rectangular tank act to disrupt swirling, producing the desirable vertical blending currents. Only for high-shear mixing would additional baffles be needed in a rectangular tank. *Viscous* mixing operations should *not* be carried out in rectangular tanks because of the natural baffling effect of the vessel.

## IMPELLER SPACING AND LOCATION

When using multiple impellers it is necessary to place them in the proper position in order to achieve an efficient flow pattern. Specific guidelines apply to radial and axial flow impellers.

### Radial Flow Impellers

With multiple radial flow impellers each impeller must develop its own flow pattern. When it does, we have impeller spacing within the desired range. If spacing is too close, then the impeller flow patterns merge and we tend to get a single flow pattern. This is particularly true with disc impellers. If impeller spacing is too great, there may be quiescent or stagnant zones between the impellers.

## AXIAL FLOW IMPELLERS

Axial flow impellers do not normally interfere with each other, and if they are placed close together the flow from one impeller merely reinforces the flow from the other impeller with less tendency toward developing separate impeller agitation zones.

It is normally best to place the lower axial flow impeller between 2/3 and 1 impeller diameter off the tank bottom for efficient mixing. However, for drawing down (draining the tank while mixing), it may be necessary to locate the impeller much closer to the tank bottom.

## GEOMETRIC SERIES OF IMPELLERS & THE IMPELLER SPECTRUM

When a series of geometrically similar impellers is operated in the turbulent regime, certain basic and predictable characteristics are expressed by the following:

$$P \propto QH \quad (2\text{-}9)$$

$$P \propto N^3D^5 \quad (2\text{-}11)$$

$$Q \propto ND^3 \quad (2\text{-}12)$$

and, at constant power

$$Q/H \propto D^{8/3} \quad (2\text{-}13)$$

(In fluid mixing technology fluid shear rate and velocity head, H, can be used interchangeably). It is important to note that *power characteristics of mixing impellers are independent of the process result*, i.e., a rotating impeller will consume a fixed amount of power in a process fluid *whether or not it is satisfying the process requirement.* Impeller power is split between fluid flow and fluid shear rate. Once the specific process requirements are known, the correct impeller type can be selected. Fig. 2-8 shows the Impeller Spectrum which indicates the relative power devoted to Q and H for various impeller types.

## IMPELLER SPECTRUM

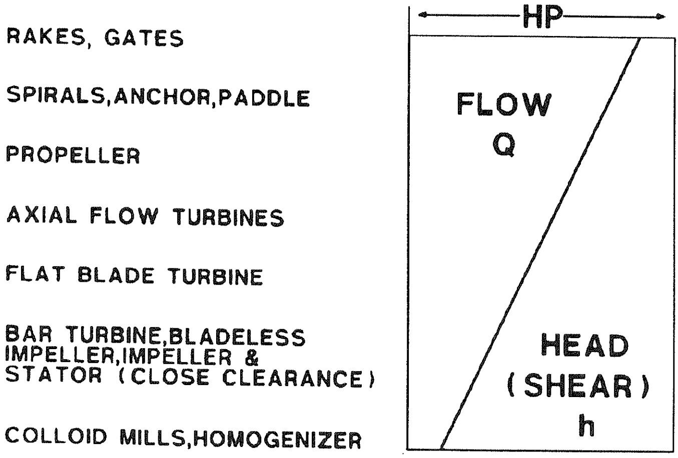



Fig. 2-8 IMPELLER SPECTRUM

**IMPELLER DESIGN CRITERIA**

The function of any mixing impeller is to dissipate energy in the form of flow and head. Impeller characteristics must be consistent with the demands of a given process or operation. The majority of industrial mixer applications are more sensitive to flow than to impeller head. Rotating mixing impellers transmit mixing energy in both forms (flow and head). The portion devoted to either of these two major components of mixing is largely controlled by impeller design.

The most important factors in the design of a mixing impeller are:

1) Size (diameter) relative to vessel size
2) Speed, RPM
3) Geometry, (number, width and angle of blades, shape, pitch, etc.)

All of these factors are required in defining the mixer input to the system. In very viscous fluids, which offer a high resistance to flow, very large, low-speed impellers (high D/T ratios) are generally required to ensure that the entire tank content is blended. The other end of the spectrum uses thin-bladed, small diameter impellers at very high speeds (tip speeds of 5 or 6 thousand feet per minute (1500 to 1800 m/min), for example, for paint pigment dispersion and overall circulation).

Comparison of impellers for a particular operation is made on the basis of power input requirements. Blending very viscous materials is normally accomplished with a large, low-speed impeller but could also be done with a small high-speed impeller at a *considerable* additional expenditure of power. The small, high-speed impeller produces flow with high turbulence which is not required for viscous blending, thus energy would be wasted. The reverse is also true. A large impeller primarily designed for flow production would waste an enormous amount of power in trying to meet the shear rate level required for pigment dispersion.

Both shear rate and flow are required in all mixing operations. A proper balance of the two is necessary for an optimum impeller design. Therefore, a large impeller operating at a low speed must produce the mass flow required for circulation while maintaining a low level of turbulence which is required for blending. Conversely a dispersing impeller must provide the high intensity input required for the dispersion while also providing enough flow for recirculation through the high intensity zone of the impeller.

Another way to achieve the proper balance of flow and shear rate is to combine different impeller types on a single mixer shaft; one impeller is used for flow production, the other to provide the required fluid shear rate.

## FLOW PATTERNS

The first demand made on the impeller is the production of flow. The flow pattern produced is a function of the impeller design and the vessel design. A rotating impeller with vertical blades produces radial flow by centrifugal action. Inclining the blades at an angle to the horizontal will produce axial flow components, used for drawing flow from the liquid surface or directing flow against the bottom of the tank for solids suspension. (See Figs. 1-1 through 1-4, Chapter 1.)

A rotary pattern (swirling) has little radial or vertical flow components. Vertical wall baffles in the tank divert purely radial components, producing vertical streams for blending. The resulting discontinuities in the flow produce the highest overall shear rate in the system.

## TANK DESIGN

It is not always possible to optimize the mixer design for a given process if the tank design is fixed. This is especially true if we must use existing vessels. Tank geometry plays a very important role in the final mixer design. In some cases it might be less costly in the long run to start with the proper tank design, rather than designing a mixer around a fixed tank geometry.

In general, the best tank design results in a $Z/T$ range of 1.0 to 1.2. This refers to the ratio of the static liquid depth, $Z$, to the tank diameter, $T$. This $Z/T$ range allows an optimum distribution of power and optimum power input in most cases. However, if conditions do dictate another $Z/T$ ratio, the *minimum* normal operating level, $Z$, should be about one-third of the tank diameter. When $Z/T > 1.0$-$1.2$, multiple mixing impellers will usually be required.

Tank bottoms should either be flat or dished but conical or spherical bottoms may also be used. Normally a dish-bottom tank is preferred, but flat-bottom tanks can be used for many processes without any difficulties. We should also consider the use of a small fillet in the flat-bottom tank if we are concerned about solids suspension in large vessels (Fig. 2-9).

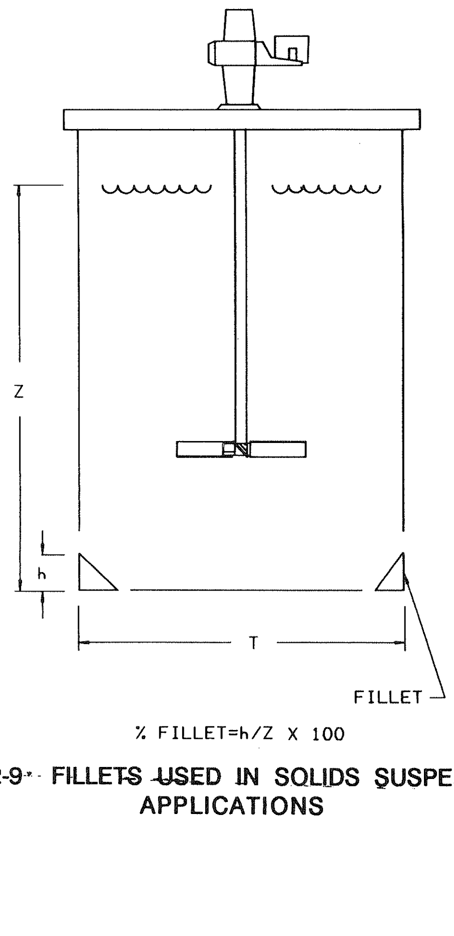



Fig. 2-9 FILLETS USED IN SOLIDS SUSPENSION APPLICATIONS

# three

## FLOW-CONTROLLED OPERATIONS

### INTRODUCTION

Flow-controlled operations involve physical contacting and circulating the process components. Typical examples include blending, suspension of solids and heat transfer. All mixing processes require some degree of flow (impeller pumping capacity) and shear stress. In a flow-controlled operation, process result is optimized by maximizing the pumping capacity of the impeller while minimizing the fluid shear rate. Shear *rate* is the velocity gradient, $\Delta V/\Delta Y$, measured from the velocity profile of the mixing impeller (Fig. 3-1). Only because of fluid shear rate can fluid particle A "catch up" with fluid particle B. At this point shear stress dictates the interaction between A and B. Shear *stress* is the product of shear rate and viscosity, $\mu\Delta V/\Delta Y$. Shear stress is the agent responsible for producing the small scale fluid interchange and intermixing which *is* the mixing process.

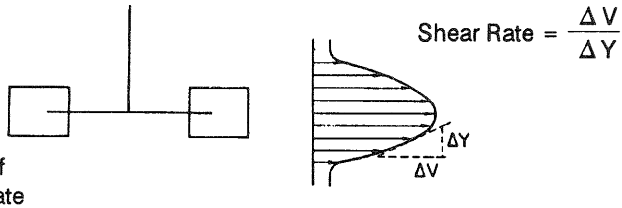



Fig. 3-1
Illustration of
fluid shear rate

## Blending and Solids Suspension

Although blending and solids suspension are separate topics, they are related. Both are flow-controlled applications; generally the more flow, the better the process result. Materials are blended *because* of flow, and solids are suspended for the same reason:

$$\textbf{PROCESS RESULT} \propto \textbf{FLOW} \qquad (3\text{-}1)$$

A mixer is essentially a pump (although not a very efficient one) and has many of the same characteristics as a pump. The most fundamental of these is that the impeller power consumption is proportional to flow and head,

$$P \propto QH \qquad (2\text{-}9)$$

where,

P = Power
Q = Flow
H = Velocity head or fluid shear rate

In other words, all of the power that a mixer supplies to a fluid produces flow, velocity head or shear rate. (In fluid mixing velocity head and fluid shear rate are used interchangeably.) At constant power, we can change the relative amount of flow or shear applied to a fluid by changing the impeller design. There is a wide range of flow-to-head ratios which can be achieved by using this approach (refer to Fig. 2-8, Impeller Spectrum, Chapter 2).

Since blending and solids suspension are both flow-controlled applications, it is necessary to choose an impeller that provides relatively more flow than shear. For these applications, we normally use *axial flow impellers* rather than radial flow impellers because axial flow impellers are more efficient pumping devices, generating more flow per unit of power expended (Fig. 3-2).

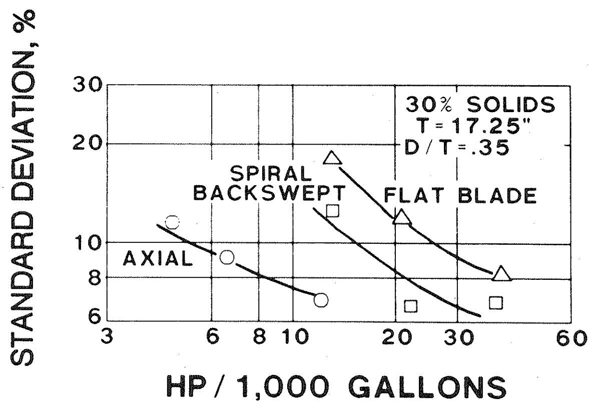



Fig. 3-2 The standard deviation of solids uniformity expressed as the standard deviation from the mean of samples taken at 25 locations.

## Flow Definition

The term primary flow refers only to the flow produced directly by the impeller; it can be measured in the discharge area of any mixing impeller. *Total* flow includes primary flow and entrained flow (flow attributed to the *circulation pattern* of the impeller). Fig. 3-3 shows that total flow is severely throttled, at constant power, when D/T ratios are greater than 0.6. In general when we speak of impeller flow rates, we mean *primary* flow unless otherwise specified.

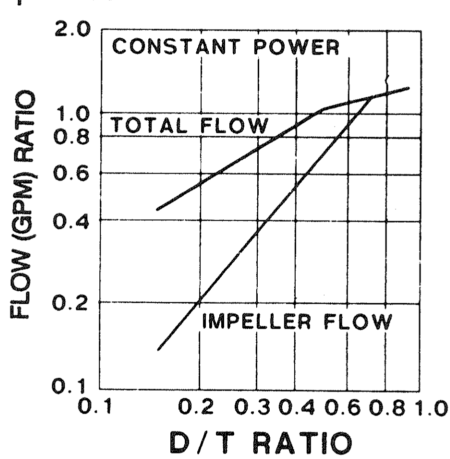



Fig. 3-3 Total flow and primary flow at constant power. Total flow is severly throttled at D/T ratios >0.6.

There are many types of axial flow impellers, including the pitched-blade turbine (designated the A200) marine-type propeller (A100) and the high-efficiency axial flow impeller (A310). Fig. 3-4 compares two of these widely-used impellers.

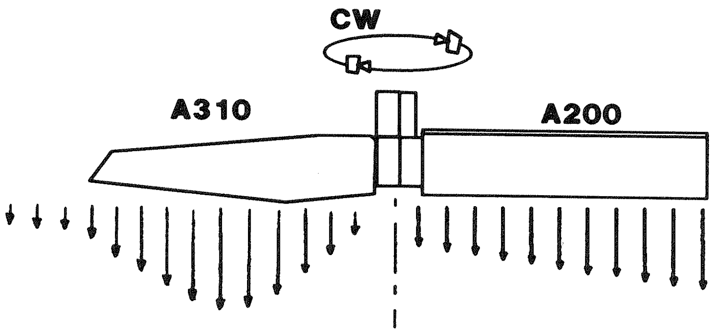



Fig. 3-4 The A310 impeller is specifically designed to produce high pumping capacity at a much lower power level with uniform shear rates across the diameter of the impeller.

NOTE: At the present time there is no universally accepted standard for identifying impeller types. For example, we use the designation "A200" to identify the pitched-blade turbine. Others may use PBT, AFT or other identifiers for the same impeller. As a result, we use the terminology with which we are most familiar in identifying impellers throughout this book. In each case a description or diagram accompanies each impeller type.

From equations (2-11) and 2-12) we can derive equations to express Q and N *at constant impeller power* in terms of D:

$$Q_{(P=k)} \propto D^{4/3} \qquad (3\text{-}2)$$

$$N_{(P=k)} \propto D^{-5/3} \qquad (3\text{-}3)$$

As a result, at constant power (P=k), as impeller diameter increases flow increases, but impeller speed decreases rapidly. This implies that more process results can be obtained at constant power by using a larger impeller diameter at a lower speed. There are, of course, practical limits on both speed and diameter. Furthermore, at constant power, as impeller diameter increases and speed decreases the resulting torque increases:

$$\text{Torque} \propto P/N \quad (3\text{-}4)$$

Increased torque necessarily means a larger gearbox and a larger gearbox means higher equipment cost. As illustrated in Fig. 3-5, there are many combinations of power and torque which give the same process result for flow-controlled applications and, in the final analysis, the correct mixer is strongly a matter of cost. Capital cost is proportional to torque, and operating cost is proportional to power consumption, therefore both of these factors must be considered in optimizing the mixer design.

System requirements must be studied and analyzed in order to select the correct kind of agitation and the correct type of mixer. Mixer optimization must also be tied to overall *process* optimization while considering the economic objectives as well.

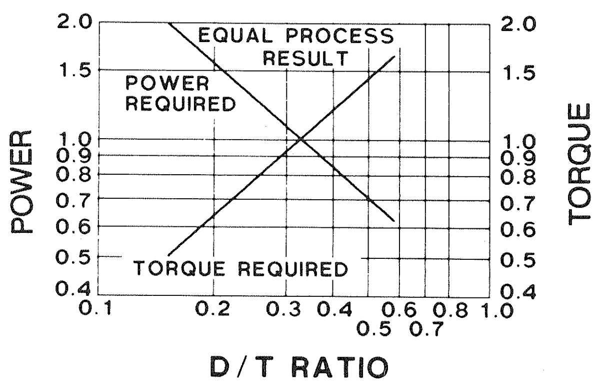



Fig. 3-5 For equal results there are many possible mixer selections.

## Blending

Blending may be described as the operation of combining components of different viscosities and/or densities to produce a material with uniform properties such that it will not change with time.

In combining solutions or suspensions to form a continuous uniform material, the viscosity of the incoming streams and the final viscosity of the mixture will determine the character of the mixing to be applied. Flow throughout the system is the primary requirement which calls for the use of axial flow impellers in low viscosity systems. With increased viscosity, the impeller size (D/T) is increased for an economical and process-effective selection.

Blending operations may be divided into two broad categories:

Low viscosity blending <50,000 cP (<50 Pa· s)
High viscosity blending >50,000 cP (>50 Pa· s)

This differentiation is used only as a guide in determining the types of impellers required. Blending controls the same variables in either low viscosity or high viscosity applications:

1) Temperature uniformity
2) Reactant/catalyst uniformity
3) Product uniformity
4) Fluctuations

### Low Viscosity Blending

In low viscosity blending applications these factors determine the agitator selection:

1. Starting conditions - this refers to whether the mixer will be running during the addition of components. If the process fluids are allowed to stratify before mixing, it could easily take 5 to 10 times as long to reach uniformity compared with running the mixer while components are added.

2. Characteristics of the materials to be blended - this includes the viscosity and density of each component. The greater the difference in viscosities and/or densities, the more power required. Also, the higher the final blended viscosity, the more power required.

3. Blend time - this refers to the time it takes from when all the components have been added with the mixer running until the tank contents are blended to the degree required by the process.

4. Tank size and shape - the larger the tank volume, the more power required to obtain a desired blend time, and as the Z/T ratio increases (the ratio of fluid depth to tank diameter) multiple impellers are often required. For very large volume tanks with low Z/T ratios (e.g., petroleum storage tanks) side-entering mixers are normally much more cost-effective than top-entering mixers.

### High Viscosity Blending

Only about 2% of all blending applications fall into this category. Nevertheless, high viscosity blending is of extreme importance in some industrial processes.

Most high viscosity fluids are non-Newtonian pseudoplastic, i.e., their viscosities decrease rapidly with increasing shear rate (refer to Fig. 3-10).

This might suggest using a high-shear mixing impeller to reduce the fluid viscosity for easier mixing. Unfortunately the resulting viscosity reduction, when using a high-shear impeller, is seen mostly in the immediate vicinity of the impeller; where shear rates are very high. Elsewhere in the fluid (away from the impeller) shear rates and flow rates would quickly diminish to zero and the fluid would remain stagnant. It is true that given enough power and fluid shear, non-Newtonian blending could be accomplished with high-shear impellers, but the resulting mixer design would be grossly inefficient and could create mechanical design problems as well.

Each of the three basic classes of impeller, the open type*, contour type (e.g., the anchor) and helical type are effective blending devices for specified viscosity ranges. Each class is suitable for a wide range of viscosities with the ranges often overlapping. Thus, there can be more than one impeller type used for a given fluid viscosity. However, there is always a best choice based on the criteria governing the process. For example, if low power consumption is essential or a minimum of heat input is desired, then a helical impeller would satisfy those conditions. If initial cost is a main criterion then an open impeller would be appropriate because the resulting mixer selection would require a smaller, less expensive gearbox compared with a helical impeller.

* An open impeller is one which is not restricted by its immediate environment. It is an impeller located in a body of fluid in which there is ample space between the impeller and physical boundaries. Conversely pump impellers or impellers located in a casing, (e.g., the rotor in a rotor-stator homogenizer and an impeller in a draft tube) are <u>not</u> open impellers because their hydraulic characteristics are controlled by their immediate physical boundaries. Anchors and helices are called close-clearance impellers.

In most cases impellers used for very high viscosity blending are of the close-clearance variety. These include anchor type impellers and double helices, each of which has its own blending characteristics. The anchor impeller (Fig. 3-6) provides most of its mixing through eddies formed in the wake of the blade. It produces little top-to-bottom motion. However, a helical impeller (Fig. 3-7) does provide good top-to-bottom motion and more complete blending than an anchor impeller at the same power level.

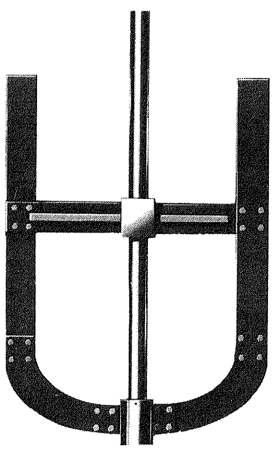


Fig. 3-6 ANCHOR IMPELLER

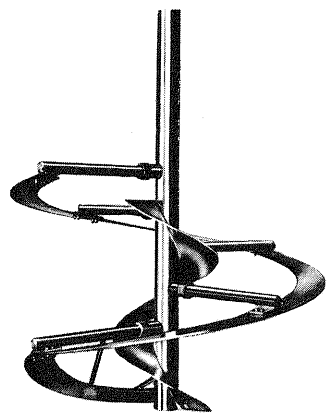


Fig. 3-7 HELICAL IMPELLER

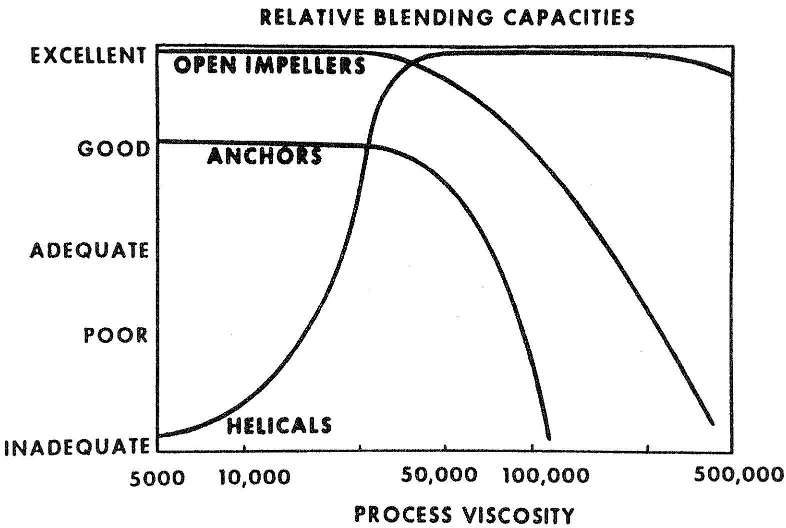



Fig. 3-8 Relative blending capacities of different impellers as a function of process viscosity

In blending high viscosity materials, especially pseudoplastic fluids, mixer design is usually based on achieving a *minimum fluid motion* in every part of the vessel. At the condition of minimum fluid motion, even though the fluid may be moving at high velocities in the vicinity of the impeller, the fluid is just barely in motion in remote parts of the vessel. Any reduction in mixing intensity would produce incomplete motion throughout the batch, i.e., a portion of the vessel might be blended quite rapidly but complete motion (therefore, complete blending) would not occur.

## HIGH VISCOSITY BLENDING WITH OPEN IMPELLERS

### Impeller Viscosity and Process Viscosity

Impeller viscosity is the viscosity the impeller sees at the impeller shear rate. For a geometrically similar series of impellers the shear rate is independent of impeller diameter; it is solely a function of rotational speed. For example, for a given open impeller type, a 20" (508mm), 50" (1270mm) or 100" (2540mm) diameter impeller operating at constant speed all see the same viscosity. Impeller viscosity is used only to determine impeller power consumption. Process viscosity is the viscosity seen by the process at the process shear rate. The process shear rate is a function of impeller speed, diameter and D/T ratio. Process viscosity is used in determining the process result.

Impeller shear rate and process shear rate are average rates and they are the controlling factors in design. For different systems there are different maximum and minimum shear rates and shear rate distributions.

Impeller shear rates are always higher than process shear rates and, for pseudoplastic fluids, the impeller viscosity is always lower than the process viscosity. These two viscosities are found by obtaining data from a pseudoplastic fluid and relating those data to a known Newtonian fluid correlation.

## Impeller Viscosity Derivation

Figure 3-9 shows a power number-Reynolds number relationship for a specific impeller, the data having been obtained from Newtonian fluids.

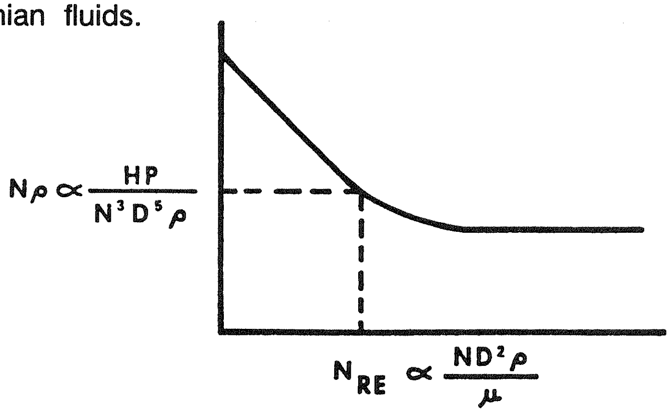



Fig. 3-9 Power Number VS Reynolds Number data obtained from Newtonian Fluids

To find the impeller viscosity for a given speed, we would make power measurements for the impeller in the specified fluid. We would measure speed, impeller diameter, horsepower and specific gravity then calculate the power number.

Using Fig. 3-9 we would then determine the Reynolds number at the calculated power number and back-calculate the impeller viscosity from the Reynolds number equation. This is the viscosity the impeller had to "see" to draw the amount of power which was measured. This is also known as the "apparent viscosity".

By taking power measurements over a range of speeds, a relationship of speed and back-calculated viscosity could be found. If an impeller of different diameter was used, and was geometrically similar to the first, we would find the same viscosity when back-calculated for the same rotational speed, demonstrating that the impeller viscosity is independent of diameter and only depends on rotational speed.

## Impeller Fluid Shear Rate Derivation

To find the impeller shear rate, we would use a relationship similar to Figure 3-10. This is a relationship of shear rate and viscosity for the pseudoplastic fluid used in back-calculating the viscosities.

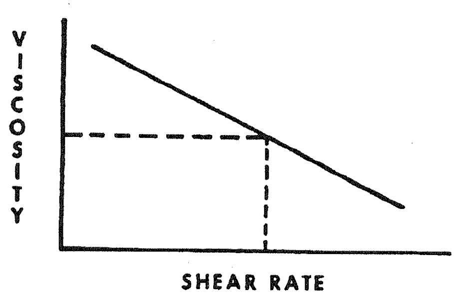



Fig. 3-10 Typical viscosity VS shear rate plot for Pseudoplastic fluids

The viscosities found at each speed relate to specific shear rates, as we can see from Fig. 3-10. Therefore, a given speed for this series of geometrically similar impellers has a specific shear rate. When repeating this for other viscosities we find that the rotational speed and the shear rate are related by a simple constant. This constant remains the same for a geometric series of impellers. Thus we can determine the average shear rate of the impeller by simply knowing its operating speed.

Any rotating device, mixing impeller, or viscometer spindle has a shear rate constant. We can relate viscosities measured or seen with a viscometer to the viscosity an impeller sees by knowing the ratio of the constants for both devices.

## Process Viscosity Derivation

Process viscosity is analogous to impeller viscosity. It is found by applying data obtained from non-Newtonian pseudoplastic fluids to a known correlation developed from Newtonian materials for a specific process result, such as blending.

Figure 3-11 shows a typical correlation between power input and Newtonian viscosity for a constant blend time at a specified D/T.

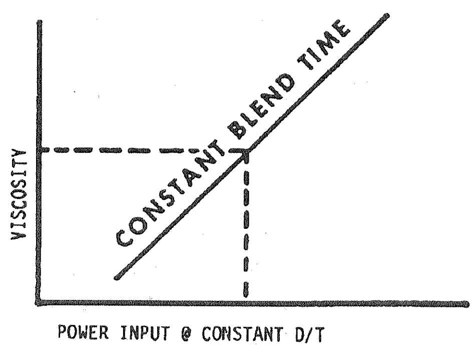



Fig. 3-11 Viscosity VS power input at constant D/T and constant blend time

When relating this to a specific tank size, we vary the power input for the constant D/T by varying the rotational speed of the impeller. In effect, this is a relationship of rotational speed and viscosity, once the impeller diameter is specified. This is similar to the power number-Reynolds number relationship shown in Figure 3-9.

To find the process viscosity for a given fluid and D/T, we can adjust the power input (by adjusting the speed) to obtain the blend time of the correlation. Then, using the power input required, the process viscosity or the viscosity the process sees, can be read from Figure 3-11.

By relating the process viscosity to a known viscosity vs shear rate curve (as in Fig. 3-10) we can determine the process shear rate to which the process viscosity corresponds. The process result (in this case, the blend time) strongly depends on the process shear rate.

At a given D/T the process shear rate is related to the impeller speed by a constant. Except for the additional dependence of D/T, this is similar to the impeller shear rate but with a different constant. Since both constants (one for the impeller and one for the process) relate to the same impeller speed, they relate to each other. By knowing one, the other can be calculated if the relationship between viscosity and shear rate is known. This relationship can be derived from viscometer data.

## BLENDING WITH CLOSE-CLEARANCE IMPELLERS

### Anchor Impellers

Anchor impellers ( Fig. 3-6) were developed as an extension of large D/T open impellers to improve fluid motion at the wall of the vessel and to improve heat transfer. However, anchors in general do not provide the axial or radial flow components which are necessary to produce a uniform blend within a reasonable period of time. This fact led to the development of gate impellers which put blades where mixing is poorest, in the center of the vessel, while still maintaining the close-clearance of the anchor impeller (Fig. 3-12).

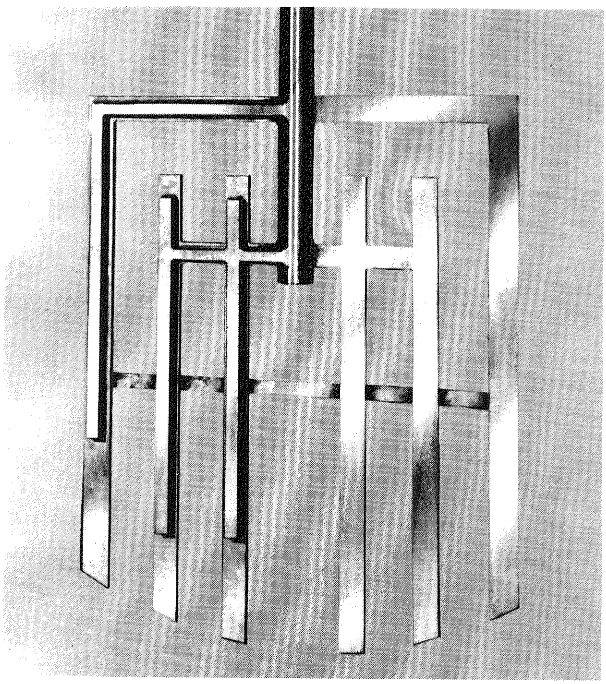


Fig. 3-12 Typical Gate Impeller

Since anchor impellers or modified anchor impellers are not efficient blending devices, their use is limited to applications which can tolerate large gradients and non-uniformities. For example, at viscosities of approximately 20,000 cP (20 Pa· s) or greater, anchor impellers are suitable for heat transfer if fluid temperature uniformity is not required.

## Helical Impellers

Unlike anchor impellers, helical impellers produce positive axial flow in high viscosity environments. As discussed earlier, open impellers can also mix highly viscous materials in excess of 100,000 cP (100 Pa · s) but only at excessively high power input. Therefore, large volume, high viscosity applications cannot normally be processed economically using open impellers. Furthermore, if heat removal is a process consideration, the additional heat input from the high power consumption of open impellers may be intolerable. In addition, the high fluid shear rates associated with open impellers is often unacceptable. There are several types of helical impellers (Fig.3-13) each of which must be evaluated for blending efficiency, bulk turnover rate, surface renewal, dispersion of liquids and/or solids and heat transfer to determine which is best suited for the application.

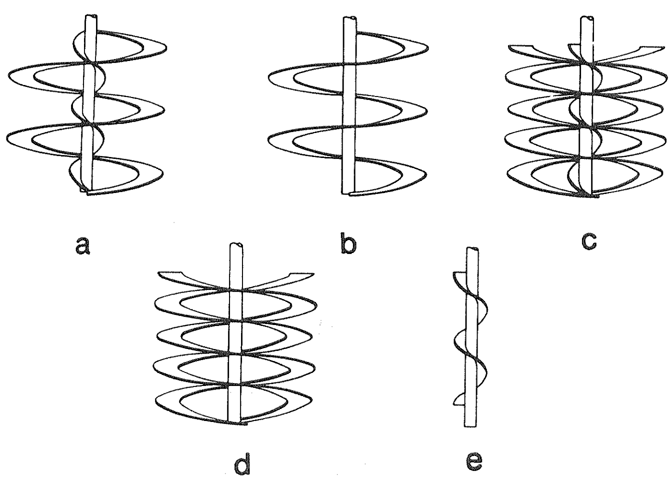



Fig. 3-13 Helical Impeller design variations

TABLE 3-1 shows comparative speed, power and torque ratios for various configurations and fluid properties at constant blend time, using the double helix (Fig. 3-13a) as a base.

TABLE 3-1 Speed, Power and Torque Ratios of Helices

| | NEWTONIAN FLUIDS | | | PSEUDOPLASTIC FLUIDS | | |
|---|---|---|---|---|---|---|
| | SPEED RATIO | POWER RATIO | TORQUE RATIO | SPEED RATIO | POWER RATIO | TORQUE RATIO |
| DOUBLE HELIX (Fig.3-13a) | 1.0 | 1.0 | 1.0 | 1.0 | 1.0 | 1.0 |
| SINGLE HELIX (Fig.3-13b) | 1.0 | .95 | .95 | 1.33 | 1.78 | 1.33 |
| DOUBLE HELIX with twin outer FLIGHT (Fig.3-13c) | .75 | 1.05 | 1.4 | .75 | 1.05 | 1.40 |
| SINGLE HELIX with twin outer FLIGHT (Fig.3-13d) | .75 | 1.01 | 1.35 | 1.0 | 1.80 | 1.80 |
| FULL HELIX (Fig.3-13e) | 2.0 | 2.0 | 1.0 | 2.5* | 3.1 | 1.25 |

## SOLIDS SUSPENSION

Solids suspension is also a flow-controlled process. However, in order for solids to be suspended in a fluid we need, in addition to flow, to impart an adequate fluid velocity to the solid particles. This velocity must be greater than the terminal settling velocity of the particles, otherwise, solids suspension would not occur. Axial flow impellers and the flow patterns, volumetric flow rates and velocities they produce are best suited for solids suspension applications.

The important considerations for solids suspension include:

1) Physical properties - including solids particle density, mesh size, percent solids composition, fluid density and viscosity. These factors are used in calculating the settling velocity of the solid particles in the fluid. As the particles become larger and/or more dense, the settling velocity *increases*. As the fluid density and/or viscosity increase, the settling velocity *decreases*.

Solids settle freely when they are permitted to settle with their natural terminal velocities. Settling becomes hindered if the presence of other particles inhibits that velocity. In general, particles with a settling velocity greater than 1 ft/min (300mm/sec.) are called *free settling solids* and particles with a settling velocity less than 1 ft/min (300mm/sec.) are called *hindered settling solids*. To suspend free-settling solids it is only necessary to produce vertical fluid velocity components large enough to offset the terminal velocity of the solids. Most applications involve free settling solids. Hindered settling solids are usually the result of concentrations of very small, finely divided particles which exhibit the same properties as viscous fluids. As the size of the particles decreases and their numbers increase, the viscosity of the suspension increases. Thus, hindered settling solids applications require both suspension and viscous mixing. Free settling or hindered settling can be determined by comparing solids concentration with the specific gravity ratio shown in Fig. 3-14.

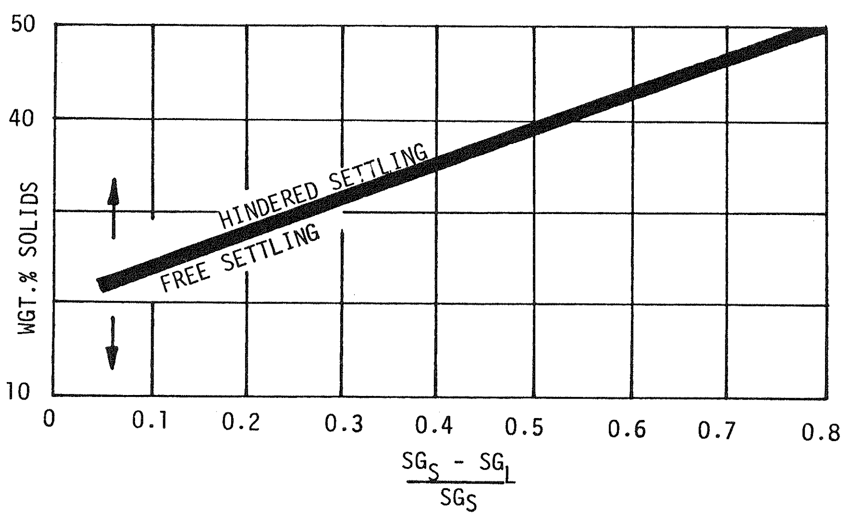



Fig. 3-14
Hindered and free-settling zones

2) Degree of suspension - solids suspension applications are usually categorized as follows:

a. Complete On-Bottom Motion

In this condition all of the solid particles are in motion regardless of particle size. Smaller or lighter particles may be suspended upward but the governing criterion is that all particles achieve a velocity, either upward or along the bottom of the tank.

b. Complete Off-Bottom Suspension

All particles are moving with a vertical velocity at some time, though not necessarily simultaneously. While larger particles may be moving upward, but only at small distances from the bottom, smaller or lighter particles may be suspended to the upper surface of the fluid.

c. Complete Uniformity

All particles are suspended uniformly throughout the vessel. Complete Uniformity is a relative term and is defined as the condition for which a further increase in mixer speed or power will not appreciably affect the solids concentration profile throughout the vessel.

The relative power required to achieve each solids suspension category is given in Table 3-2.

Table 3-2
RELATIVE POWER & SPEED REQUIRED FOR SOLIDS SUSPENSION WITH SOLIDS SETTLING VELOCITIES FROM 16 TO 60 FT/MIN

| CRITERION | SPEED RATIO | POWER RATIO |
|---|---|---|
| ON-BOTTOM SUSPENSION | 1.0 | 1.0 |
| OFF-BOTTOM SUSPENSION | 1.7 | 5.0 |
| COMPLETE UNIFORMITY | 2.9 | 25.0 |

As implied, applying more power than is necessary would mean wasting power and would increase capital and operating costs. Off-bottom suspension is usually adequate for most solids suspension applications.

3) Tank Size and Shape

This is much more important in solids suspension than in blending. As the Z/T ratio increases, tanks get taller and the solids must be suspended at higher levels in the vessel, requiring more power. At Z/T > 1.25, dual impellers are required. See Figures 3-15 and 3-16.

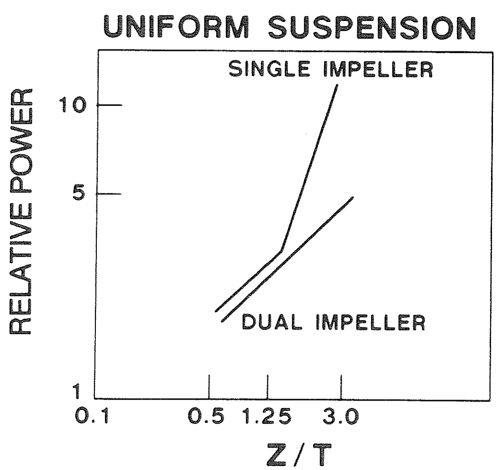



Fig. 3-15 Relative Power vs Z/T

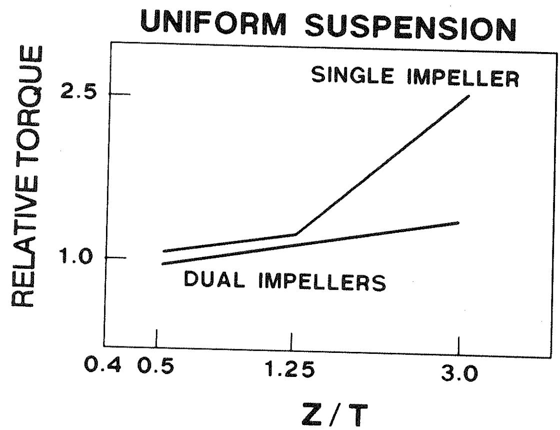



Fig. 3-16 Relative Torque vs Z/T
for uniform suspension

Dish bottom tanks are optimum for solids suspension because the bottom shape aids in directing the flow. Steep cone bottoms are shapes to be avoided as solids settle in the heel of the cone and can be very difficult to suspend.

In solids suspension, not only the cost of the agitator, but the total tank/agitator package cost must be evaluated. Each process must be considered individually to determine the optimum tank and agitator configurations.

## HEAT TRANSFER

### The Importance of Heat Transfer

Control of the flow of heat in a desired manner is a most important consideration in chemical engineering. Most chemical and biological processes are either endothermic or exothermic, usually requiring some kind of heat transfer control. Approximately 85% of all chemical processing mixer applications involve heat transfer of some nature.

Typical systems in which heat transfer is an important process parameter are:

1. Chemical Reactions
2. Polymerizations
3. Fermentations
4. Esterifications
5. Hydrogenations

Heat Transfer Equation

The quantity of heat added or removed in a system is expressed by the following equation:

$$Q = U_o \; A_o \; \Delta t \qquad (3\text{-}5)$$

where:

$Q$ = quantity of heat per unit of time, BTU/hr
$U_o$ = rate factor, overall heat transfer coefficient, BTU/hr/ft$^2$/°F
$A_o$ = heat transfer surface area, ft$^2$
$\Delta t$ = driving force or mean temperature differential, °F

The rate factor, $U_o$, is a combination of several factors and can be expressed as:

$$\frac{1}{U_o} = \frac{1}{h_o} + \frac{1}{h_i} + \frac{\ell}{k} + \text{f.f.} \qquad (3\text{-}6)$$

where:

$\frac{1}{h_o}$ = resistance of outside or mixer-side film coefficient
$\frac{1}{h_i}$ = resistance of inside film coefficient (normally water or steam)
$\frac{\ell}{k}$ = transfer tube wall resistance
f.f. = dirt film fouling factor

In many fluid mixing applications, the controlling process parameters may be something other than heat transfer. Solids suspension, mass transfer, dissolving, or blending requirements (or a combination of these) may ultimately control the process. When a process *is* heat transfer controlled, those variables *not controlled by mixing* may have a *much greater* effect on heat transfer (e.g., increasing the mean temperature differential or increasing the heat transfer surface area). Mixing can only affect the mixer-side film coefficient, $h_o$, (Eq. 3-7) which represents only a part of the overall heat transfer equation (Eq. 3-8):

$$h_o \propto \mu^{-0.3} D^{1.44} N^{0.67} T^{-0.6} k^{0.63} d^{-0.6} \rho^{0.67} C_p^{0.37} \tag{3-7}$$

$$N_{NU} = \frac{h_o d}{k} = 0.17\, N_{Re}^{0.67} N_{Pr}^{0.37} \left(\frac{D}{T}\right)^{0.1} \left(\frac{d}{T}\right)^{0.5} \left(\frac{\mu}{\mu_s}\right)^{m} \tag{3-8}$$

(For helical coils and R100 impellers only. The constant in equation 3-8 varies with impeller type.)

where:

| | |
|---|---|
| $h_o$ | = mean mixer-side film coefficient, at tank temperature |
| $k$ | = thermal conductivity of fluid, BTU/hr/ft²/°F/ft |
| $d$ | = outside diameter of heat transfer tube or coil |
| $D$ | = impeller diameter |
| $T$ | = tank diameter |
| $\mu$ | = bulk viscosity of fluid at tank temperature |
| $\mu_s$ | = viscosity of fluid at film temperature at heat transfer surface |
| $N_{Pr}$ | = Prandtl No., $\frac{C_p \mu}{k}$ |
| $C_p$ | = fluid specific heat |
| $N_{Re}$ | = Reynolds No., $\frac{ND^2 \rho}{\mu}$ |
| $\rho$ | = fluid density |
| $N$ | = impeller RPM |
| $m$ | = experimentally determined exponent, depending on bulk viscosity |
| $N_{NU}$ | = Nusselt number |

## Heat Transfer vs Fluid Regime

*Forced convection* is the heat transfer condition in which the fluid being heated is in *turbulent flow* at the heat transfer surface. Increasing the mixing intensity will not economically provide a significant change in $h_o$:

$$h_o \propto P^{0.22} \tag{3-9}$$

(Turbulent Flow, Forced Convection)

(e.g., applying *10 times* the power would only increase $h_o$ by a factor of 1.66).

The term, $h_o$, can be easily calculated from equation (3-6). This is the only part of the overall heat transfer coefficient that is significantly affected *by the mixer:*

$$h_o = 0.17\, N_{Re}^{0.67}\, N_{Pr}^{0.37} \left(\frac{D}{T}\right)^{0.1} \left(\frac{d}{T}\right)^{0.5} \left(\frac{\mu}{\mu_s}\right)^{m} \frac{k}{d} \tag{3-10}$$

A plot of the exponent, m, vs bulk viscosity for a typical fluid is shown in Fig. 3-17.

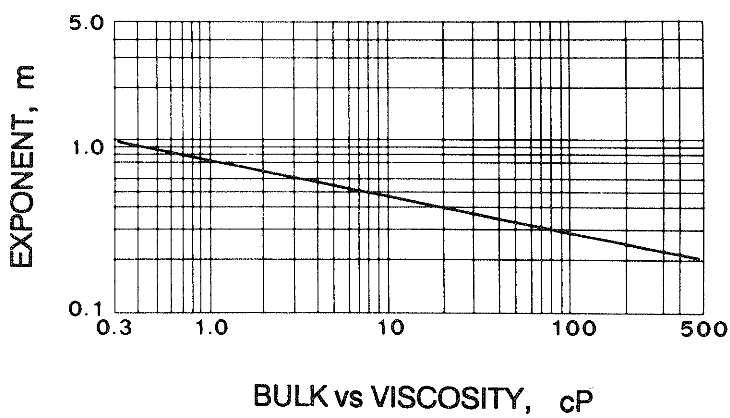



Fig. 3-17 Exponent, m, vs Bulk Viscosity

If the fluid regime is streamlined (laminar) the heat transfer condition is *natural convection*. This is where the fluid being heated or cooled is in laminar flow at the heat transfer surface. Increasing the mixing intensity produces a corresponding increase in the mixer-side film coefficient, $h_o$:

(3-11) $$h_o \propto hp$$

(e.g., applying twice the horsepower would effectively double $h_o$ in the laminar range)

If, for a given system, power is varied and the corresponding $h_o$ values are calculated, a plot of $h_o$ vs power will determine the heat transfer condition. Fig. 3-17 illustrates this relationship:

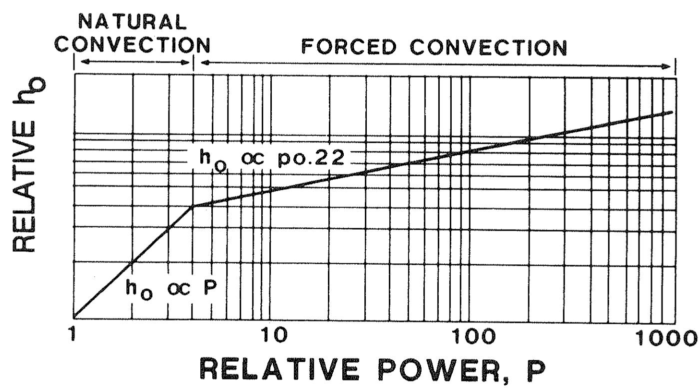



Fig. 3-18 $h_o$ vs Power (for determining heat transfer condition)

## Thermal Effectiveness of the Heat Transfer Area

There are many different methods of installing heat transfer areas in mixing vessels. Most of the research in this field has been done with helical coils. As more heat transfer area is installed in a vessel with banks of coils, the thermal effectiveness decreases as shown in Table 3-3.

TABLE 3.3 Relative Effectiveness of Heat Transfer Areas

| HEAT TRANSFER AREA | RELATIVE THERMAL EFFECTIVENESS |
|---|---|
| Tank Jacket | 1.0 |
| Vertical Tubes | 1.54 |
| Helical Coils (1st bank) | 1.54 @P=k |
| Helical Coils (2nd bank) | 1.31 |
| Helical Coils (3rd bank) | 1.08 |
| Plate or Panel Coils | 1.12 |

Figure 3-19 shows the effect of viscosity on mixer-side film coefficients for organic fluids.

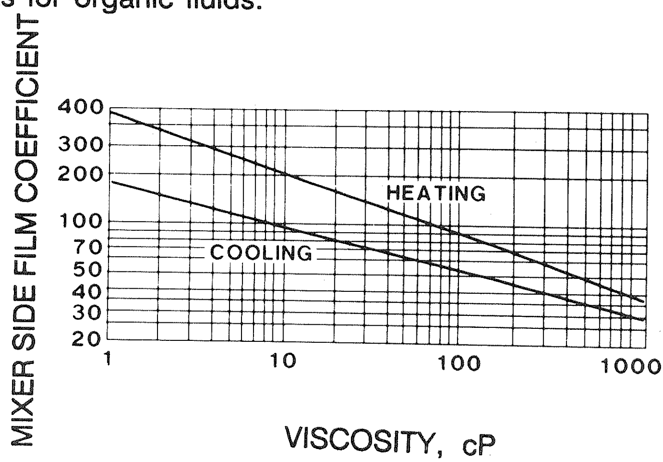



FIG. 3-19 Typical Mixer-side film coefficients for heating & cooling organic chemicals (Thermal Effectiveness = 1.54)

### Tank Baffle Requirements for Heat Transfer

Baffles may or may not be required depending on the helical coil arrangement, the fluid viscosity and the number of coils. When baffles are required they are installed inside the coil to prevent swirling, especially if the coil is tightly wound. Tube supports and the turns of the coil also act as baffles. Table 3-4 contains guidelines for baffle requirements in fluid viscosities of 100 cP (0.01 Pa ▪ s) or less. As fluid viscosity increases beyond 100 cP (0.01 Pa ▪ s), the need for baffling decreases as previously indicated in Fig. 2-7, Chapter 2.

**TABLE 3–4 Baffle Requirements For Low Viscosity Fluids**
(100 cP or less)

| COIL TYPE | DEGREE OF BAFFLING AS A FUNCTION OF NORMAL BAFFLE REQUIREMENT |
|---|---|
| SINGLE COIL | 50% OF STANDARD BAFFLES |
| DOUBLE COIL | 25% OF STANDARD BAFFLES |
| 3 OR MORE COILS | NO BAFFLES REQUIRED |

Vertical heat transfer tubes also act as baffles. Their effect on the degree of baffling and impeller power response is a function of the number of tubes, their location and size.

# four

## GAS-LIQUID OPERATIONS

### INTRODUCTION

There are two distinct areas of gas dispersion with which we are concerned:

(1) Physical Dispersion
(2) Dispersion for Mass Transfer

In both cases, radial flow impellers have normally been used because they have lower Q/H ratios than many other impellers. You may recall that flow-controlled processes require axial flow impellers because they produce higher Q/H ratios, in the axial direction, than do radial impellers (Fig. 4-1). However, for gas dispersion applications, not only is flow important, but so is fluid shear. The R100 radial flow impeller (Fig. 4-2) provides both the flow and shear required in many gas-liquid dispersions.

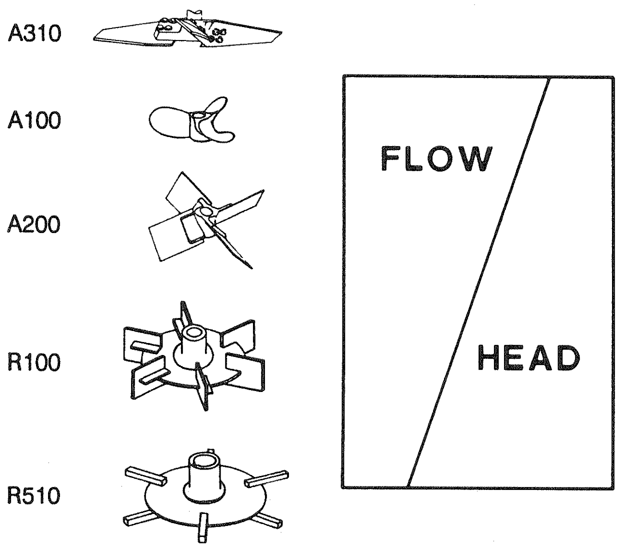



FIG.4-1 Comparison of impeller types and Q/H ratios

Flow is essential because it is the mechanism responsible for physically distributing the gas. Flow provides gas-liquid contacting when the solubility of the gas is great enough, such that the controlling step is physical dispersion and not mass transfer.

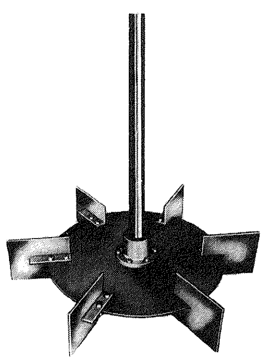


FIG 4-2 R100 Impeller which is used for many gas dispersion applications

A mixer is required in a gas-liquid system to promote:
(1) increased interfacial area
(2) longer gas residence time

Shear rate is the velocity gradient $\Delta V / \Delta Y$ of the flow produced by the impeller. Fluid shear is required in gas-liquid applications in order to produce smaller gas bubbles from larger ones, which increases the surface contact area between gas and liquid. This inherently decreases the reaction time and it simultaneously increases the system efficiency. Simply stated, when adequate shear is applied to the fluid, gas hold-up time is increased which means a longer residence time therefore a higher efficiency.

## PHYSICAL DISPERSION

In order to disperse gas at all, the applied impeller horsepower (shaft horsepower) must be equal to or greater than the gas expansion horsepower. As shaft horsepower increases the dispersion patterns are intensified. If there were not enough shaft horsepower to overcome the gas expansion horsepower, geysering would occur (Fig. 4-3).

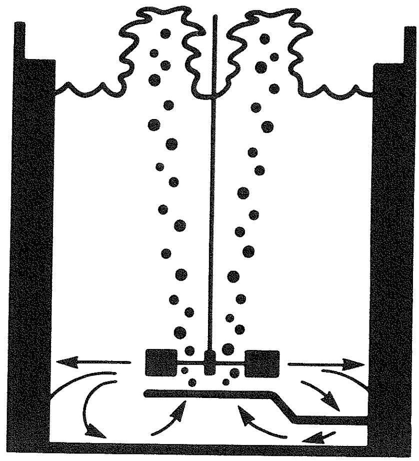


FIG 4-3 Illustration of Geysering

When shaft horsepower is greater than gas expansion horsepower, gas dispersion will occur, and the *degree* of dispersion is a function of the shaft horsepower applied. There are three basic types of physical dispersion:

(1) minimum
(2) intimate
(3) uniform

Figures 4-4, 4-5 and 4-6 illustrate these three basic dispersion patterns.

In order to obtain the proper levels of dispersion for the three conditions, the following shaft horsepower levels are required:

(1) Minimum dispersion, 1.0 times gas expansion hp

(2) Intimate dispersion, between 1.0 and 3.0 times gas expansion hp

(3) Uniform dispersion, 3.0 times gas expansion hp

To determine the gas expansion horsepower, we must consider the general equation for work, which is essentially what is accomplished by the gas:

$$W = P_1V_1 \ln \frac{P_1}{P_2} \qquad (4\text{-}1)$$

$$\frac{HP}{V} = \frac{(4.051 \times 10^{-3})\,(P_2 + 62.4Z\rho)}{Z} \ln \frac{(P_2 + 62.4Z\rho)}{(P_2)} F_1$$

where,
- $P_1$ = Pressure at tank bottom, $lb/ft^2$
- $P_2$ = Pressure at tank top, $lb/ft^2$
- $Z$ = Batch height, ft
- $V$ = Batch volume, thousands of gallons
- $\rho$ = Fluid specific gravity
- $F_1$ = Superficial gas velocity, ft/min.

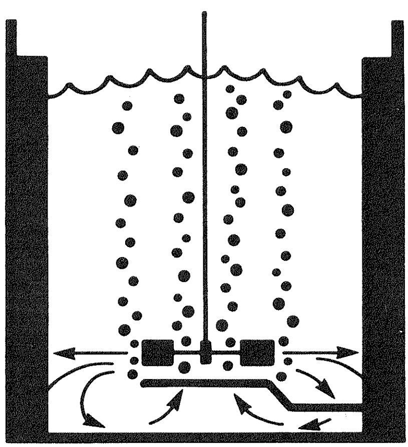


Fig. 4-4 Minimum Dispersion

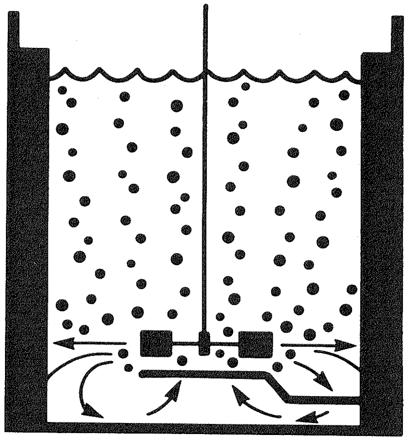


Fig. 4-5 Intimate Dispersion

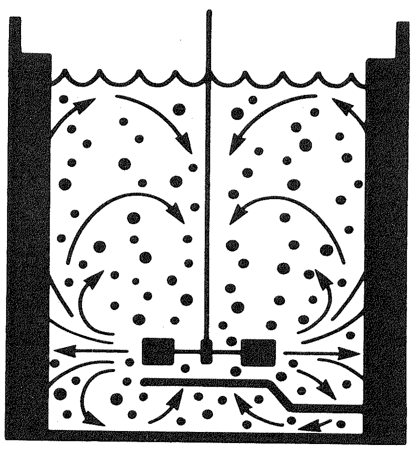


Fig. 4-6 Uniform Dispersion

Once the gas expansion horsepower is determined, the mixer can be sized for various types of dispersion. But mixer horsepower is not the only important parameter for the required dispersion. The D/T ratio is also significant. The most effective ratios to use for the horsepower and gas rate ranges are shown in Table 4-1 for radial flow impellers.

TABLE 4-1 Most Effective D/T Ratios

| GAS RATE \ HP | HIGH | LOW |
|---|---|---|
| HIGH | 0.17 | 0.33 - 0.50 |
| LOW | 0.17 - 0.50 | 0.17 |

At this point, it may appear that sizing the mixers for gas dispersion is the only thing we have to consider. However, we must also be sure that the impeller is not flooded. In an agitated gas-liquid system, the mixer and gas compete to control the flow pattern. When the mixer is controlling, gas is dispersed uniformly and gas holdup in the system can be measured (by measuring the fluid volume expansion). However, there is a maximum gas rate that a given impeller and speed can handle. If this rate is exceeded, the impeller becomes flooded.When flooding occurs, the mixer loses control and the system flow pattern becomes gas-controlled. The result is loss of dispersion, poor mixing and loss of gas-holdup or bubble retention.

## MIXER LOADING

The power consumed by an R100 impeller in a gas-liquid system at fixed diameter and speed is less than the power consumed in the ungassed liquid. The ratio $P_g / P_o$ is called the "K" factor of the impeller and is a function of the gas rate and impeller speed.

For example, assume we have determined that for a given level of dispersion, $P_g = 13.5$ and $P_o = 27$ (or the K factor = 0.50).
If we sized the mixer on the ungassed basis, we would obviously require a larger mixer than if sizing on a gassed basis but, we would be assured of operating safely with the gas either on

or off. However, if we sized on the basis of gassed horsepower, it would limit us to running only with the gas turned on unless a 2-speed motor is supplied. A 2-speed motor would allow both modes of operation using a smaller horsepower mixer and still satisfy the process. Loading a mixer based on the gassed horsepower and using a 2-speed motor offers these advantages:

1) smaller mixer, therefore, lower initial cost
2) lower operating cost
3) lower maintenance cost
4) smaller, less expensive starter box

## SPARGING DEVICES

There are many ways to introduce the gas into the liquid. The manner in which this is accomplished is *not* arbitrary however. Three of the common sparge designs are illustrated in Figures 4-7, 4-8 and 4-9. Normally a sparge ring is used because an open pipe can contribute to mechanical problems and poor mass transfer. This is because of either the large asymmetry of the impeller with respect to the pipe opening during normal shaft deflections or improper positioning of the pipe during installation. The result is non-uniform gas loading, producing larger than normal mixer shaft deflections and larger power variations. To remedy the problems of open pipe spargers, the sparge plate (Fig. 4-8) is used to evenly distribute the gas from an open pipe.

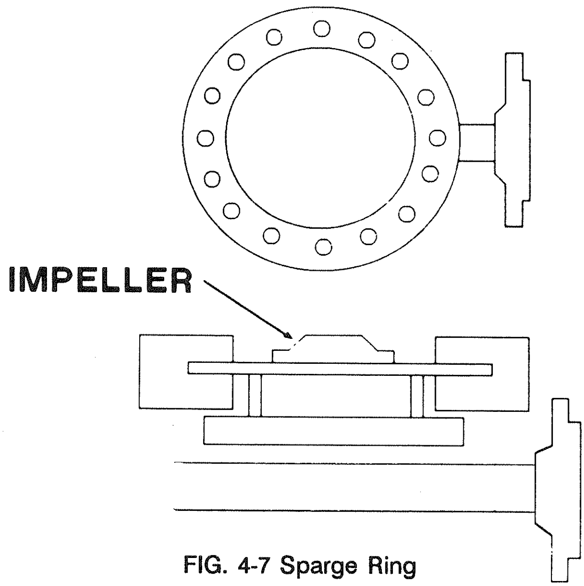



FIG. 4-7 Sparge Ring

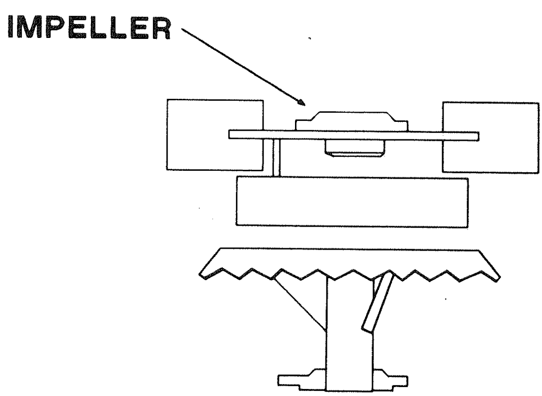



FIG. 4-8 Sparge Plate

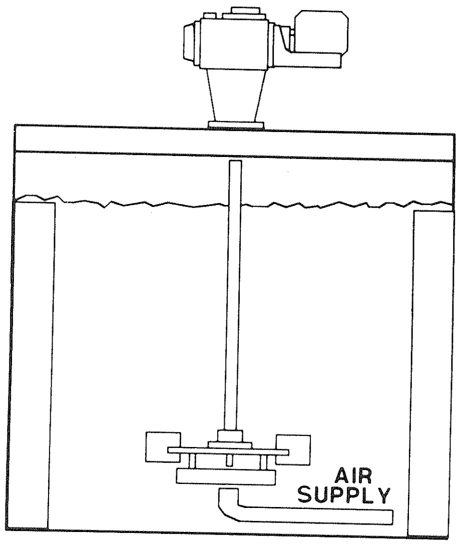



FIG. 4-9 Open Pipe Sparge System

## OTHER IMPELLERS

We have previously stated that radial flow impellers provide the best Q/H ratios for gas dispersion. Most open impellers (such as a radial flat-blade paddle type) should not be considered (see Chapter 3, page 36 for definition of open impeller). They require more power to disperse the gas and their flow patterns are not conducive to good dispersion. Furthermore, the natural tendency of the gas bubbles to rise in the liquid competes with the open impeller flow pattern.

For other gas-liquid applications (for example, submerged turbine aerators for waste and water treatment) other radial impellers and special spargers are used. An example of this is shown in Fig. 4-10.

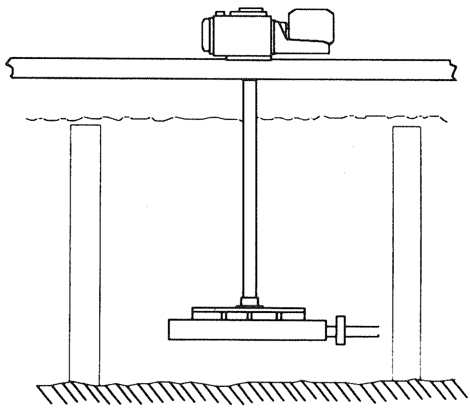


Fig.4-10 R321 Sparge and specially disigned ring

## High-Solidity Axial Flow Impellers

Historically, radial flow, Rushton-type (R-100) mixing impellers have been the choice for gas-liquid operations. These impellers provide both the fluid shear required for bubble size reduction and the necessary flow for blending, solids suspension, gas distribution and heat transfer. However, use of other impellers or combinations of impellers has been successfully employed in commercial applications. For example, Fig. 4-11 is a diagram of a 250 hp fermenter using both axial and radial flow impellers. This impeller combination was used in a particular application where power reduction was an important consideration. It is indicative of applications in which impellers are closely matched with their functional capabilities. The radial flow impeller was placed where it was needed most, at the bottom of the vessel above the sparge ring for shearing gas bubbles, while the axial flow impellers were situated in the upper regime of the batch where flow production is required but high shear is not. Although this impeller configuration was adequate for the process described, it does not represent a panacea for all gas-liquid applications. There are normally severe gas rate limitations imposed on using axial flow impellers, due to flooding. In fact, flooding is the main reason why axial flow impellers are not usually considered for gas dispersion, they have quite limited gas handling capabilities. As is true with other open impellers in gas-liquid processes, gas escapes the flow field of the axial impeller by migrating to the liquid surface along a path which is coincident with the impeller's axis. Thus gas bubbles cannot be distributed and flooding occurs.

Recently, the necessity for higher yields, lower fluid shear rates and reduced energy costs in fermentations prompted a renewed interest in axial flow impellers. Improved yield is, of course, a major goal for any process and lower energy costs are always desirable. One of the ways to improve yields in fermentations is to provide a climate conducive to the growth of specific micro-organisms, which is required for all fermentation processes. However, some micro-organisms are very sensitive to fluid shear rates, often thriving only in lower fluid shear rate environments. Axial flow impellers produce much lower macro-scale shear rates, satisfying the need of shear-sensitive micro-organisms. In flow-controlled applications, it is well documented that axial flow impellers provide the same process result as radial flow impellers while simultaneously consuming much less power than radial impellers. Thus the major requirements of improved yield and decreased energy consumption are easily satisfied by using axial flow impellers. Furthermore, other gas-liquid operations require adequate suspension of solids in promoting chemical reactions on solid particle surfaces. Once again, axial flow impellers also satisfy those needs. However, the main obstacle (flooding) which has traditionally prevented using axial flow impellers for higher gas rate applications, remained until recently. A new "high solidity" axial flow impeller (designated A315, Fig. 4-12) has been developed to solve the problem of flooding while still providing the other benefits of low power consumption, low shear rates and high fluid flowrates. Solidity ratio is the term used to express the ratio of the projected blade area of the impeller to the projected area which is swept by the impeller as it rotates. It has been determined that axial flow impellers used for gas dispersion must have a solidity ratio between 0.8 and 1.0 to prevent flooding at practical gas rates. By comparison, the solidity ratio of the A315 is 0.87, the A200 (or pitched blade turbine) 0.43 and the A310 0.22. As a result the A315 provides significantly higher mass transfer coefficients compared with an R100 (improvements of 30% have been measured). In addition, the A315 greatly extends the range of gas rates for axial flow impellers. This allows axial flow impellers to be used where lower macro-scale fluid shear rates are beneficial to shear - sensitive organisms in fermentations. The A315 axial flow impeller also minimizes the power requirements for solids suspension while simultaneously dispersing gas.

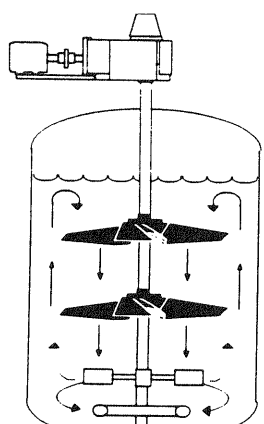


FIG. 4-11 Diagram of 250 hp fermenter using combined axial and radial flow impellers

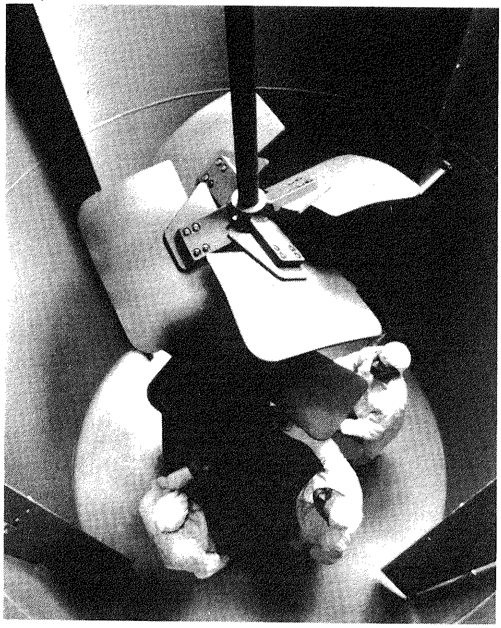


FIG. 4-12 A315, High Solidity Ratio, Axial Flow Impeller

## MASS TRANSFER

Many chemical processes involve liquid-solid, liquid-gas, or liquid-liquid mass transfer or a combination of all three. Depending upon the specific process, one or any combination of these mass transfer operations may be involved. It is important to determine the effects of mass transfer operations on the overall process objectives. Ideally, the system should be designed to first meet the reaction rate requirements, thus it should not be mass transfer rate limited. It is just as important *not* to design a system whose mass transfer capabilities *exceed* the requirements needed to satisfy a reaction rate. An understanding of the relationship of the agitation parameters to the process performance is essential to optimize the process design. One method used to establish an overall system design is as follows:

(1) Decide which step is controlling as far as the overall process results are concerned. This involves optimizing time requirements, product yields and/or quality, and reactant utilization. The effects of agitation on each of these parameters can be determined experimentally.

(2) Optimize the commercial scale design to meet controlling requirements established in step (1). This involves determining the most effective design of vessel geometry and mixer configuration to best meet process and economic objectives.

It is important to realize that uniform physical contacting may not necessarily provide an optimum mass transfer design. In some cases, the process requirements can be satisfied with a non-uniform fluid regime while others will require agitation power levels significantly higher to provide a measureable degree of fluid uniformity. Care should also be taken to assure that agitation shear levels are optimized. Higher shear levels than necessary could hinder results which might otherwise be mass transfer limited.

The absolute value of the overall mass transfer coefficient ($K_{Ga}$) cannot be calculated for a variety of processes without pilot planting since $K_{Ga}$ depends on the physical properties of the materials involved. The overall mass transfer relationship may be expressed as:

$$K_{Ga} = \text{lb-moles/hr/ft}^3\text{/atm} \qquad (4\text{-}1)$$

and

$$K_{Ga} \propto (P/V)^a (F)^b (\mu)^c \qquad (4\text{-}2)$$

Where:

$K_{Ga}$ is the gas/liquid mass transfer coefficient

P = Power
V = Volume
F = Superficial Gas Velocity
a,b, and c are functions of the process variables

Mass transfer is usually thought of in terms of the following relationship:

$$\text{Mass Transfer Rate} = K_{La} \quad (\Delta c) \qquad (4\text{-}3)$$

Where: $K_{La}$ = Mass Transfer Coefficient, $hr^{-1}$

and

$\Delta c$ = Concentration Gradient

The reaction kinetics of a specific gas-liquid mass transfer process establish the overall mass transfer rate or process requirement. A typical evaluation shows (Fig. 4-13) that the overall mass transfer coefficient is composed of a system constant, $K_s$, and a transfer area, A, which is effected by agitation. The driving force, $\Delta c$, is a system variable affected by gas partial pressure and vessel geometry.

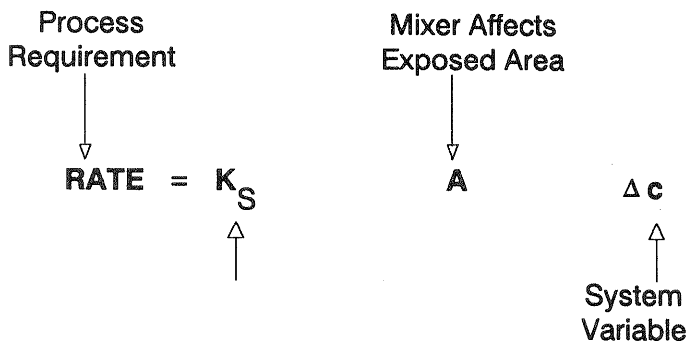



Fig. 4-13 Overall Gas-Liquid Mass Transfer coefficient is composed of a system constant, transfer area and system variable

It is always necessary to determine which process step is limiting. Referring to Fig. 4-14, the gas-liquid mass transfer capabilities of a system should be tailored to either the liquid-solid reaction requirements or the chemical reaction requirements. Insufficient gas-to-liquid transfer may limit the process results while too much would mean unnecessary capital expense and excessive operating costs.

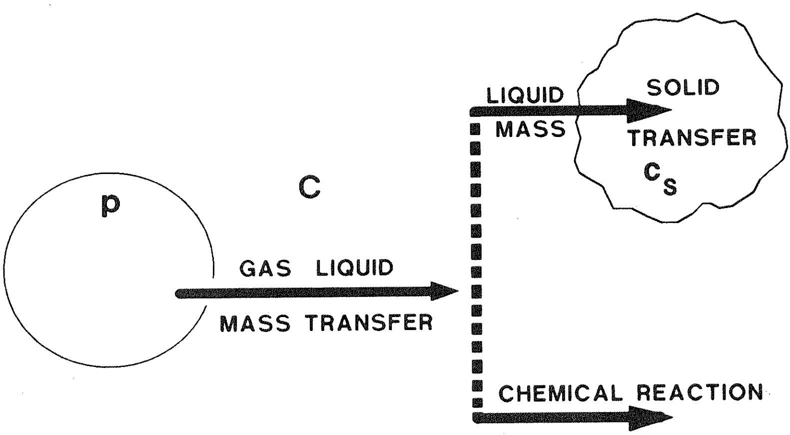



Fig. 4-14 The limiting process must be determined

The effect of agitation on the liquid-solid mass transfer coefficient, $K_S$, is quite pronounced up to the point where off-bottom suspension is achieved. Beyond that point, as suspension approaches uniformity, the effect of agitation on $K_S$ diminishes (Fig. 4-15).

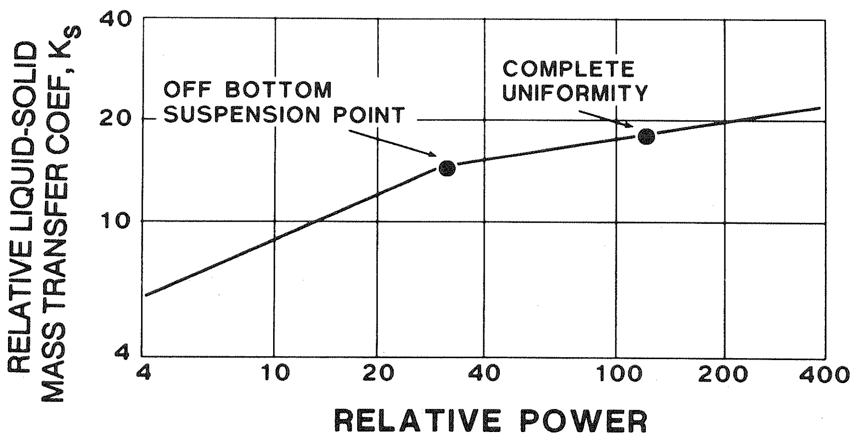



Fig. 4-15 Effect of power level on Liquid-Solid Mass Transfer coefficient for various suspension criteria

In gas-liquid mass transfer processes there is a minimum power level required for dispersion. This prevents the gas from rising straight through the fluid (Fig. 4-5). Uniform gas dispersion (Fig. 4-6) distributes gas bubbles throughout the vessel. In these examples the R100 impeller is used because it produces high fluid shear rates along with the required pumping capacity, and it produces a controlled dispersion. The system requirements for gas-liquid mass transfer determine the optimum power level and degree of dispersion.

Fig. 4-16 illustrates two cases in which processes can be either mass transfer rate limited or reaction rate limited. In both cases oxygen transfer is required by gas-liquid contacting. The figure on the left indicates a decrease in dissolved oxygen (D.O.) with time and a decrease in reaction rate because the system is oxygen deficient. The system on the right shows an increased dissolved oxygen level with time because the reaction rate is not fast enough to consume all of the available oxygen. Thus, low product yield does not necessarily result from insufficient agitation. Therefore, the system-controlling factors must be understood to obtain high yields.

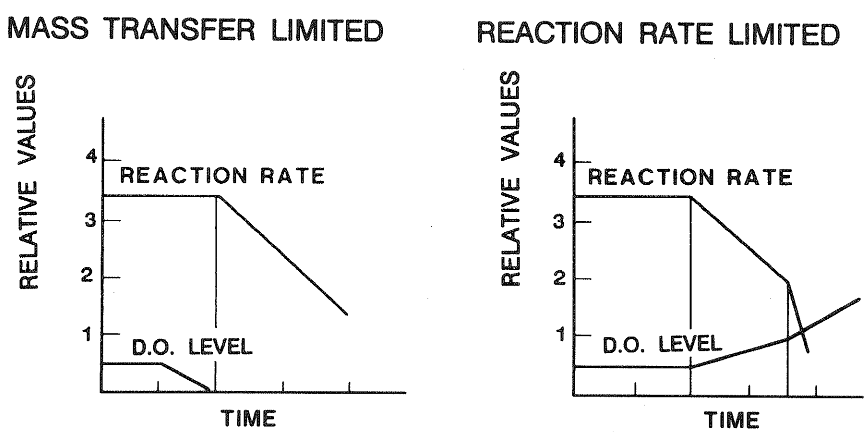



Fig. 4-16 Examples of Mass Transfer-Limited and Reaction Rate-Limited Processes

In addition, the mass transfer coefficient, $K_{Ga}$, varies with the superficial gas velocity, $Q_g/A$, and power (Fig. 4-17). Pilot studies at various superficial gas velocities and mixer power levels allow optimization. (Superficial gas velocity, F, is the ratio of the volumetric gas rate, $Q_g$, to the vessel cross sectional area, A).

$$F = \frac{Q_g}{A} \tag{4-4}$$

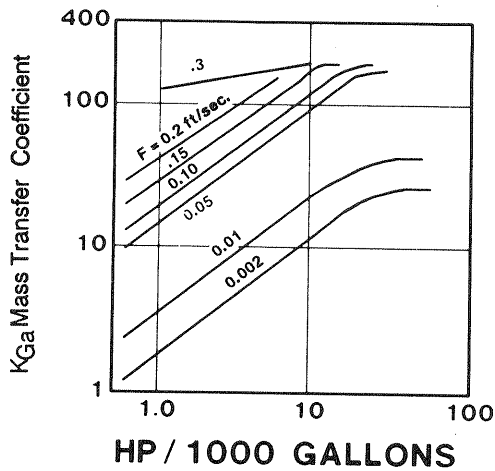



Fig. 4-17 Mass transfer coefficient, $K_{Ga}$, as a function of mixer power per unit volume and superficial gas velocity

Furthermore, the shear rate supplied by an agitator impeller influences the interfacial transfer area between two phases (Fig. 4-18). In gas-liquid or liquid-solid mass transfer, if the shear rate is too high, results may be undesirable. This is especially true for biological processes (e.g., fermentation) where some micro-organisms are very shear-sensitive. Furthermore, in liquid-liquid contacting, care must be taken to avoid dispersions which are difficult to separate. This also requires careful control of fluid shear rates.

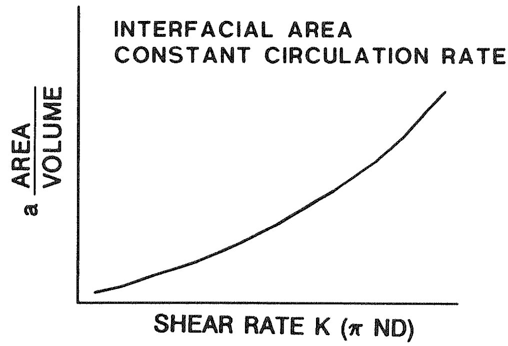



Fig. 4-18 The effect of shear rate on interfacial area

# five

## INTRODUCTION TO PILOT PLANT OPERATIONS/SCALE-UP

Some of the material presented in this chapter is a review of material presented earlier in order to facilitate the application of basic concepts to pilot planting and scale-up.

### INTRODUCTION

It is normal practice to study processes on one scale to determine the practicality and economic feasibility on another scale. This procedure is accomplished via the PILOT PLANT. Pilot plants are used to obtain data which will allow extrapolating results to the production scale, once feasibility has been established.

Conversely, for an *existing production scale process*, the proposed process design changes can be studied on another scale to determine their effects on the production scale. This operation is called MODELING. Modeling is especially useful for optimizing *existing* processes based on the latest technical innovations in fluid mixing. (Production scale processes which were designed many years ago may be designed quite differently today due to advances in fluid mixing technology. Often, the net result is a large saving in first cost, operating and maintenance costs).

Both pilot planting and modeling save much time, material, labor and expense compared to experimenting on a larger production scale. These procedures also allow much greater flexibility in selecting the design parameters to be studied.

Many fluid mixing processes no longer require either pilot planting or modeling since manufacturers of mixing equipment have obtained large data banks from which scale-up and scale-down operations can be readily determined for most processes. But for processes for which data and experience are limited (or non-existent) pilot planting or modeling is often essential.

## PILOT PLANTING AND MODELING

In order to predict the process result on a production scale, it is necessary to study the effects of mixing on a pilot scale. This provides the scale-up criteria necessary for an accurate prediction. Unfortunately it is not possible, by any scale-up technique, to maintain all of the properties of a fluid mixer constant with respect to the pilot plant. Pilot studies determine which mixing parameters are significant and quantify their effects on the process (Table 5-1, 5-2).

Table 5-1 Properties of a fluid mixer on scale-up.

| Property | Pilot Scale (20 gal) | Plant Scale (2500 gal) | | | |
|---|---|---|---|---|---|
| P | 1.0 | 125.0 | 3125.0 | 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 |
| N D | 1.0 | 1.7 | 5.0 | 1.0 | 0.2 |
| $ND2\rho/\mu$ | 1.0 | 8.5 | 25.0 | 5.0 | 1.0 |

Table 5-2 Properties of a fluid mixer on scale-down.

| Property | Pilot Scale (2500 gal) | Plant 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 |
| N D | 1.0 | 0.48 | 0.56 | 0.62 | 1.0 |
| $ND2\rho/\mu$ | 1.0 | 0.07 | 0.08 | 0.06 | 0.09 |
| D/T | 0.35 | 0.35 | 0.35 | 0.30 | 0.30 |
| $D\varpi/D$ | 1.0 | 1.0 | 1.0 | 1.0 | 0.25 |

## Similarities

In fluid mixing there are three types of similarity to be considered: geometric, kinematic and dynamic. Geometric similarity exists when the ratios of dimensions for one system are equal to the corresponding ratios of a second system. Kinematic similarity exists when two systems are geometrically similar and the ratios of velocities between corresponding points in each system are equal. Dynamic similarity exists when the systems are geometrically and kinematically similar and the ratios of *forces* between corresponding points in each system are equal.

## Force Ratios

The input from the mixer is called the *inertia force*, $F_I$. Force ratios illustrating the ratio of inertia force to other forces have been defined in Chapter 2 ($N_{Re}$, $N_{Fr}$ and $N_{We}$). Inertia force is determined by the impeller type, diameter and speed.

It can be shown that only the ratios of two of the four forces, $F_I$, $F_V$, $F_g$ and $F_\sigma$, can be constant ratios when using the *same fluid* (or another fluid with identical physical properties) on different scales.

Each dimensionless group provides a rule for scale-up, but they often conflict. Therefore, it is usually required to design pilot equipment to minimize the effects of certain groups in favor of other groups. For dynamic similarity, scale-up should depend mainly on a single dimensionless group representing the ratio of inertia force to the appropriate opposing force. In addition, the fluid regime (see Chapter 2, Fig. 2-3) should not change from one vessel size to another. For example, the opposing forces should be largely due to either gravity, surface tension or viscosity, but not to a combination of the three.

The three dimensionless groups, $N_{Re}$, $N_{Fr}$, and $N_{We}$ are respectively proportional to $ND^2$, $N^2D$ and $N^2D^3$. If one of them is used for scale-up, the scale-up rules for the other two are automatically void.

For a given system, various groups may be far less significant than others. For example, $W_e$ is only important where more than one physical phase exists. Therefore, in a single phase fluid system, the effects of $W_e$ can be eliminated. To illustrate another example, in a baffled mixing tank the baffles eliminate vortexing (the effect of gravity), therefore, $N_{Fr}$ is no longer important, etc.

## Power Consumption

As we have seen, impeller power consumption data obtained on a small scale can be easily and accurately projected to a larger scale, assuming geometric similarity and identical fluids. In the turbulent region, power is calculated by the proportionality:

$$P \propto N^3D^5 \qquad (2\text{-}7)$$

For transition and laminar regimes, impeller power is calculated from the power number/Reynolds number curve for the impeller (see Fig. 2-3, Chapter 2). If the Reynolds number is the same for both scales, or if operation is in the turbulent regime for both scales, then the same proportionality holds ($P \propto N^3D^5$). If the Reynolds numbers are different (excluding the turbulent regime) then power is also proportional to the ratio of power numbers for the respective Reynolds numbers.

## Fluid Shear Rates and Velocities

As mentioned previously, shear stress is the mechanism by which mixing occurs. Fluid shear rate is one component of shear stress and it is also an essential element of mixing. Shear rate is a velocity gradient; without it, mixing would not occur. There are various types of fluid shear rates - maximum and average impeller zone shear rates and also a range of shear rates throughout the fluid and at the fluid boundaries.

The average velocities of a radial flow turbine (R100) are shown with respect to impeller speed and sample point location in Fig. 5-1. The maximum and average impeller zone shear rates are given as a function of turbine speed in Fig. 5-2. Fig. 5-3 illustrates maximum and average impeller zone shear rate as a function of impeller diameter when speed is constant.

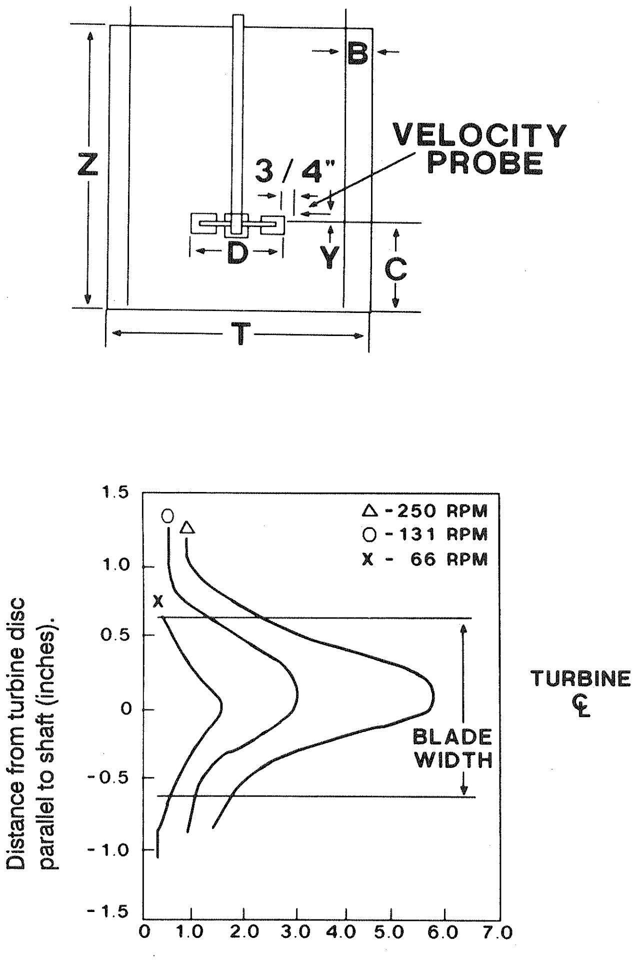



Fig. 5-1 Experimental apparatus from which average velocities are obtained for a radial flow turbine (R100) at various impeller speeds.

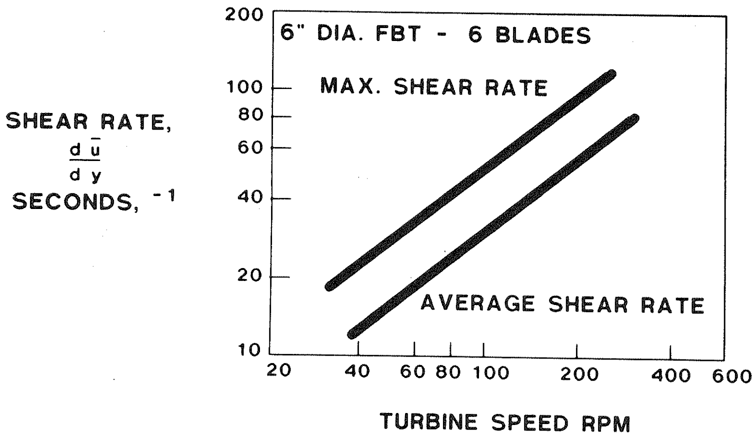



FIG. 5-2 Maximum and average impeller zone shear rates for 6" R100 as a function of speed.

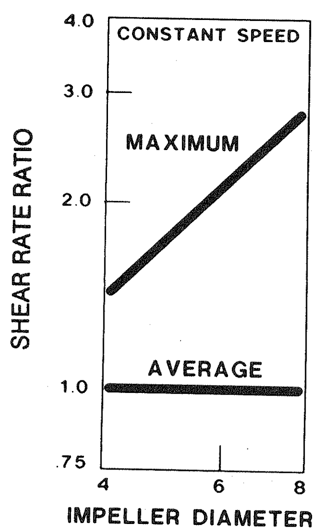



FIG. 5-3 Ratio of maximum and average shear rates at constant speed as a function of diameter.

If we examine the effect of scale-up on maximum and average impeller zone shear rates at constant power per unit volume, we have the curves shown in Fig. 5-4. Note that impeller speed normally decreases, while tip speed increases on scale-up. The net result is a larger range of shear rates in larger vessels.

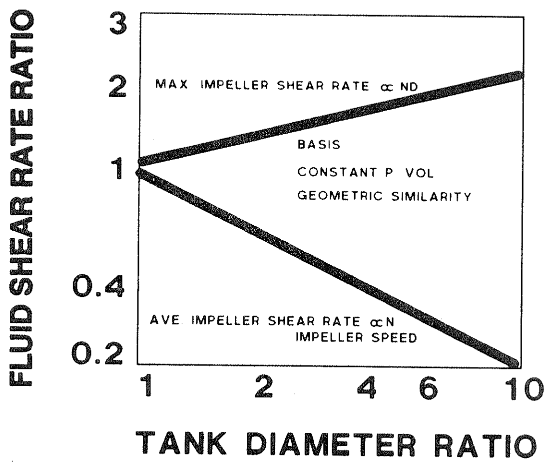



Fig. 5-4 Effect of scale-up on maximum and average shear rates.

Non-geometric scale-up techniques are required to affect the divergence of maximum and average shear rates. Consequently, pilot plant experiments must define the shear rate levels which affect the process. An example of this is shown in Table 5-2 using a standard impeller and a modified impeller. By varying speed and diameter the effect of shear rate on the process can be determined.

TABLE 5-2
Effect of impeller design and speed on shear rate at constant P/V.

| IMPELLER DIA. | RPM | RELATIVE BLADE WIDTH | SHEAR RATE MAXIMUM | SHEAR RATE MINIMUM |
|---|---|---|---|---|
| 1 | 1 | STD | 1 | 1 |
| 0.7 | 1.8 | STD | 1.3 | 1.8 |
| 1 | 1.3 | NARROW | 1.3 | 1.3 |

Power per Unit Volume

Every aspect of the full scale process requirement must be considered before designing and operating the pilot plant. Pilot tests must be run under mixing conditions conducive to scale-up for practical commercial scale equipment. Fig. 5-5 shows the relation of speed ratio (pilot plant-to-full scale) and tank diameter ratio (full scale-to-pilot plant) to power per unit volume.

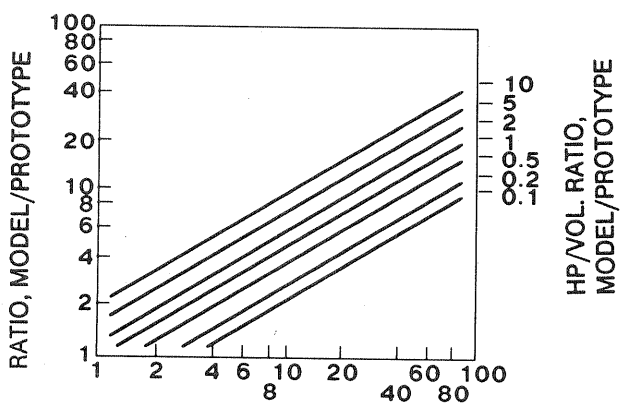



FIG. 5-5 Speed ratio (pilot plant-to-full scale) VS tank diameter ratio (full scale-to-pilot plant) as a function of P/V.

It is important to note that power per unit volume is *not* normally a constant parameter on scale-up, however, it can be a useful correlating parameter. If there is a quantitative parameter directly related to process result, and if it is compared to the power required to produce the result, the slope of the resulting curve provides a good indication of the controlling mechanism. For example, in Fig. 5-6, a steep slope indicates that increased agitation will increase process results, for example in a mass transfer- limited process. At relatively lower slopes, the process is reaction-rate limited and agitation level is less influential. As the slope further decreases, the effects of increased agitation also decrease and, finally, at a slope of zero, there is little if any effect of agitation on process result.

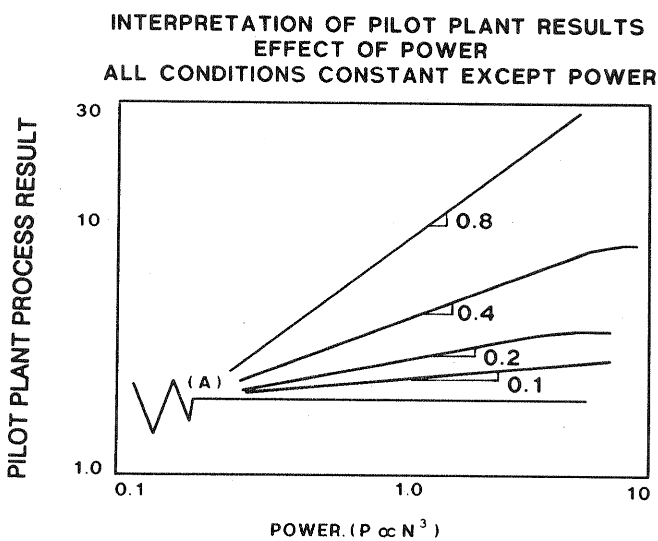



FIG. 5-6 Interpretation of pilot plant results as a function of power level (all conditions constant except power).

# six
## MECHANICAL DESIGN

### INTRODUCTION

The mechanical design of a mixer cannot be completely qualified without an understanding of the magnitude and the nature of the forces that are imposed on it. These forces are a result of the fluid dynamics produced by the impellers interaction with the vessel internals and the process involved which, in turn, produce both constant and variable stresses on the mixer assembly. Before a complete understanding of the interrelationship of the mechanical design and those forces can be completely evaluated, the following must be considered:

1)The magnitude and nature of the forces must be identified.

2)The conceptual design of the agitator must take those forces into account.

3)The engineering standards and manufacturing tolerances must be consistent with the design of the equipment.

The fluid forces imposed on a mixer can be divided into two groups. The first is related to the fluid dynamics produced by the agitator and the interaction with the vessel internals. These forces are normally associated with baffles, impeller location, and the energy density input of the mixer. The second group includes forces affected by the process. This would normally include, for example, fluids from violent chemical reactions, gas dispersion, and reactions containing high percent solids. The ability to predict these forces normally depends on the literature that has been published, the knowledge of the process involved and the laboratory testing of various mixers.

The nature of the forces applied to mixing systems normally depicts the conceptual design required to withstand them. For example, mild agitation (which produces definable fluid forces) can be achieved by directly connecting the mixer shaft to a speed reducer that has been specifically designed to withstand those loads. If the loads are of exceptional magnitude and/or difficult to predict, a design may be required which dampens potentially destruction loads and isolates the speed reducer from them. This is particularly important since the reducer often represents a large cost of the system.

Finally, consideration must be given to the engineering standards and manufacturing tolerances used and adhered to by the manufacturer. Those practices must be consistent with sound engineering practice and the assumptions used in designing the equipment.

The most important aspects of *total* mixer design are presented in the following sections. They include shaft design, mixer drive design, mechanical seal design and coupling design.

## SHAFT DESIGN

This section deals with the factors involved in mixer shaft selection and operation. It includes data and methods for determining allowable shaft lengths.

To be properly designed, a mixer shaft must satisfy the limitations imposed by:

1) Critical speed
2) Shaft stress
3) Shaft deflection

For a given length, the shaft having the minimum diameter that will satisfy these criteria is the optimum selection because a larger shaft diameter would by itself be more expensive. Furthermore, a larger-than-required shaft diameter might also mean a larger, more expensive speed reducer.

## CRITICAL SPEED

Critical speed is the most important factor in determining allowable shaft length. It is purely a mechanical consideration depending only upon materials of construction, shaft diameter, shaft length, bearing spacing, and impeller weight. Critical speed is defined as:

> that speed at which the rotational speed of the shaft is equal to its natural lateral vibration frequency measured in cycles per minute

When the operating speed of a shaft approaches the natural frequency, the shaft will vibrate with increasing amplitude, such that a failure may occur from excessive bending stress (fatigue) or by exceeding the yield point of the material.

In determining critical speed, certain assumptions must be made. The most important of these is that the mixer mounting is perfectly rigid. In actual field structures, this is often not the case and the operating limits on shaft speed must allow for some flexibility of the mounting in determining allowable shaft lengths. In extreme cases the mixer support or tank top may have to be strengthened for safe mixer shaft operation.

The Critical speed, $N_C$, can be calculated by the following:

$$N_C = K_1 \left(\frac{d}{L}\right)^2 \sqrt{\frac{E}{\left(\frac{L+a}{L}\right)\left(W + \frac{K_2 \times W_e}{L}\right)}} \qquad (6\text{-}1)$$

where:

| | | |
|---|---|---|
| $d$ | = | shaft diameter |
| $L$ | = | shaft length |
| $E$ | = | modulus of elasticity |
| $W$ | = | shaft weight per unit length |
| $W_e$ | = | equivalent impeller weight |
| $K_1$ and $K_2$ | = | constants |
| $a$ | = | bearing spacing |

From this equation, it is obvious that careful consideration must be given to the type of material used for shafts, the equivalent impeller weight, the shaft diameter, the shaft length and the bearing spacing.

Similarly, the critical speed varies inversely as the square root of the ratio of equivalent impeller weights. Thus, a heavier impeller lowers the critical speed of a shaft and reduces the allowable shaft length for operation at a given speed. Conversely, a lighter impeller increases critical speed and increases the allowable shaft length at a given speed.

Of greater importance is the effect upon critical speed of changes in shaft length and shaft diameter. A change in either factor changes the critical speed by the square of the ratio (equation 6-1). Thus, while changes in equivalent weight and type of material do have an effect on the critical speed, the magnitude is relatively small, but changes in shaft length or diameter have a larger effect.

## SHAFT STRESS AND DEFLECTION

Shaft stress results from a combination of three loads, one of which is a function of the shaft deflection. They are:

1) Torsional Loads
   Torsional loads are those caused by the torque necessary to drive the impeller.

2) Axial Loads
   Axial loads are due to the weight of the shaft and impeller, and also the thrust of the impeller. Thrust axial loads are of significance only with axial flow impellers.

3) Radial Loads
   Radial loads are due to the hydraulic fluid forces resulting from the action of the fluid on the impeller, and from centrifugal or eccentric dynamic loads. Radial loads introduce shaft deflection which in turn produces bending stresses.

Radial loads and bending stresses are especially important when the power input to the system is high and the radial loads due to fluid forces are great. Deflection is also a controlling factor when it is necessary to have low shaft run-out at the stuffing box, mechanical seal or impeller.

Deflections of the shaft are greatest during draw-off (draining the vessel), when the impeller is rotating partly in liquid and partly in air and the unbalanced loads may be large. Deflections with the impeller operating fully submerged may also be significant, though not as frequently.

Impeller stabilizing rings and fins are used to increase the allowable shaft length of a fixed diameter shaft at a fixed speed. These devices do not change the critical speed but allow operation closer to the critical speed by dampening the effect of unbalanced hydraulic loads. Such loads are always present during mixer operation and are larger through draw-off. Stabilizers are most effective when they are operating totally in the liquid with the impeller partially exposed.

## MIXER DRIVE DESIGN

A properly designed mixer drive is more than just a speed reducer, although the speed reducer is a very important part of the drive system and is most often the largest portion of the cost.

Before we can design a drive, we must first define the loads to which these components will be subjected.

The forces experienced in a mixer application are bending moment - due to the fluid forces, and torque - due to the rotational resistance. The bending moments are, in reality, fluctuating loads of various amplitudes that depend upon vessel geometry and, more importantly, process conditions. If the process requires mild agitation, these forces are clearly defined. Normal torsional loads are simply proportional to power divided by speed.

Figure 6-1 depicts the bending moment seen by the mixer shaft, thus the drive, under normal operation. The amplitude of the bending moments becomes less definable as the power/unit volume increases, and as the possibility of a process upset increases. For example, in a gas-liquid application, the amplitude of fluctuations is increased since the incoming gas produces additional vibrational forces on the agitator system. Even if the amplitude of the fluctuating loads is known, we must still decide to what limit the reducer should be designed (e.g., should it withstand 80% of the fluctuating loads, or 90% or what?)

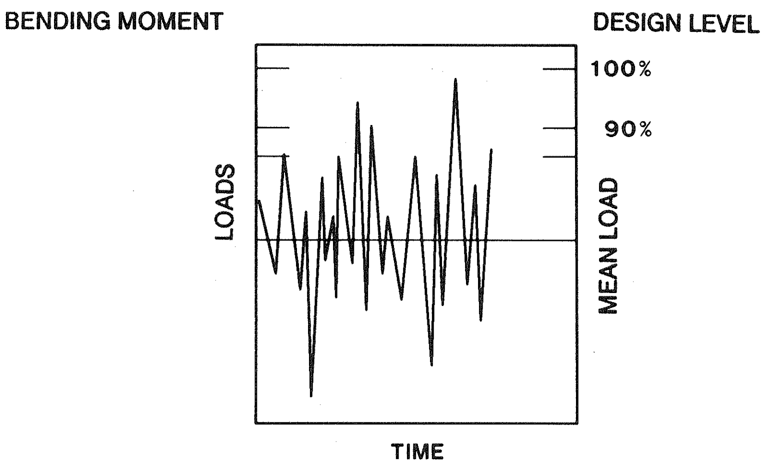



FIG. 6-1 Mixer shaft bending moment over time

Lets examine some of the mixer drive designs that are currently available. First, we have a design in which the mixer shaft is directly coupled to the speed reducers bearings.

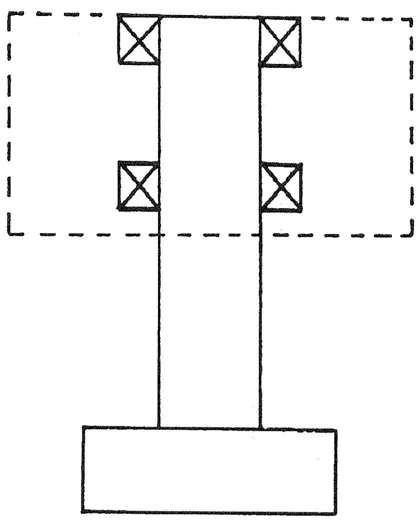
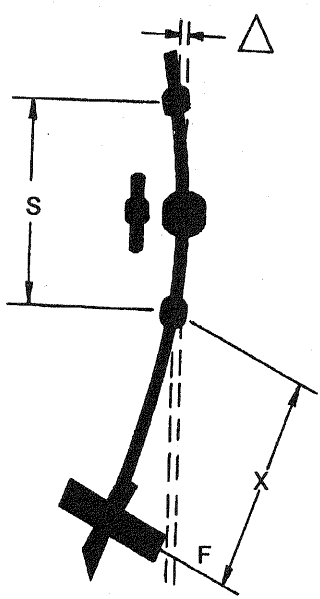


FIG. 6-2 Mixer output shaft and bearing assembly

With this type of drive design, the bending moments seen by the mixer shaft are also seen by the reducer shaft and reducer bearings. This type of design subjects the bearings to exceptionally high loads due to the small support span, overhung shaft length, and large fluid forces as shown in Fig. 6-10.

This design is also subject to gear problems when high fluid forces are present (producing shaft deflection between the bearings of the gear reducer). A deflection in this area can cause the low-speed gearing to deflect (Fig. 6-3) causing excessive gear wear due to improper meshing of the gears. An increase in the reducer shaft diameter can decrease gearing misalignment due to shaft deflection, but it does not eliminate it. Therefore, this type of design is normally used when fluid forces are definable and are within allowable limits of the engineering standards and manufacturing quality of the speed reducer.




$$\Delta \propto \frac{F(x)(S)^2}{EI}$$

FIG. 6-3 Deflection of a shaft which is directly coupled to the reducer bearings

This design also allows all of the torsional shock loads to be directly transmitted to the gear reducer, thus the gearing. Therefore, this type should only be used in operations which do not include torsional shock loading.

Reducer Protection - A second design is one which isolates bending moments from the speed reducer through the use of an independent bearing support and a flexible coupling between the mixer shaft and reducer shaft.

It has often been called the hollow quill design because the speed reducer portion of the drive has a hollow low-speed shaft (i.e., like a quill).

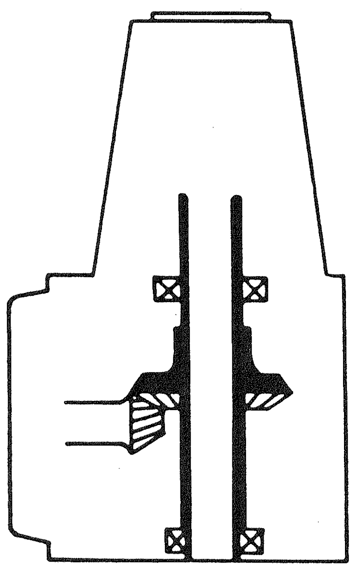


FIG. 6-4 Reducer with hollow quill shaft and gears

The speed reducer in Fig. 6-4 is shown schematically with the last reduction of the double reduction gearing illustrated. Note that the low-speed reducer shaft (thus the low-speed gear) is supported by its own double set of bearings. This is the quill portion of the design.

Next, the low-speed mixer shaft fits through the hollow quill and is supported by its own double set of bearings. This is shown in Fig. 6-5A.

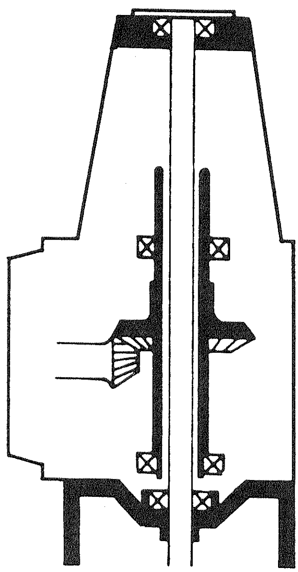


Fig. 6-5A Hollow quill design showing independent bearing supports for quill shaft & mixer shaft

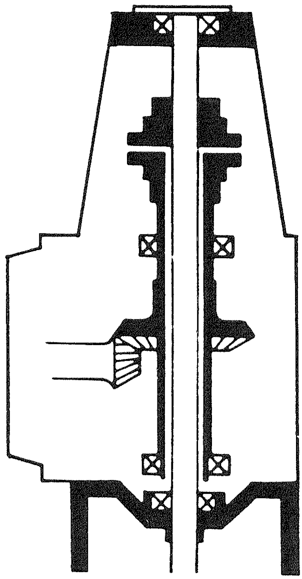


FIG. 6-5B Quill shaft and mixer shaft are joined by flexible, taper grid coupling

The mixer shaft is connected to the speed reducer (quill) shaft by means of a flexible, tapered grid coupling, as in Fig. 6-5B.

Thus bending loads on the mixer shaft, resulting from mixing in the vessel, are isolated from the speed reducer portion of the drive due to the separate bearing support for each. Also, the peak torque or power loads are dampened by the torsionally resilient coupling. Both of these concepts are illustrated in Fig. 6-6.

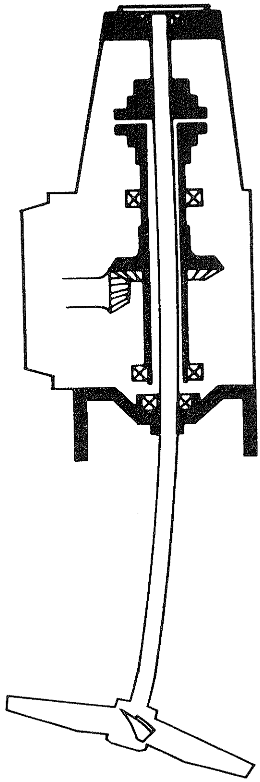


FIG. 6-6 Bending loads on mixer shaft are isolated from speed reducer shaft and gears

Flexible couplings dampen out the top 30-35% of peak shock loads (i.e., if the shaft is seeing shock loads of two times normal running torque, then the flexible coupling will only transmit loads below 1.3 to 1.4 times normal running torque). This is illustrated in Fig. 6-7.

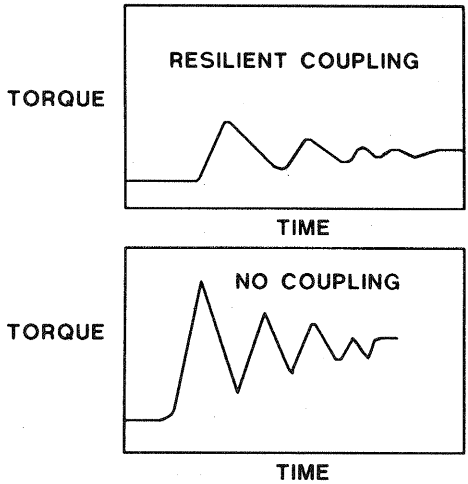



FIG. 6-7 Comparison of torque dampening with and without flexible couplings

Both of the graphs in Fig. 6-7 show the output torque variation with the same input load variation. Thus the speed reducer, which is torsionally resiliently isolated, sees much lower loads and the drive has a higher service factor, based on the loads.

For example, a reducer having an AGMA (American Gear Manufacturers Association) service factor of 1.5 would have a corresponding service factor of 2.14, based on the loads if they were dampened out by a torsionally resilient coupling.

## MECHANICAL SEAL DESIGN

The mechanical seal, and how it fits into the overall drive design is very important, as a seal will only work properly and dependably if it is a part of a properly designed drive.

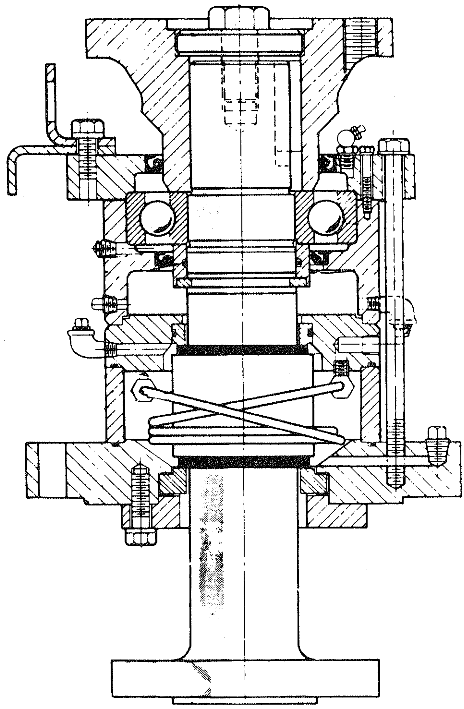


FIG. 6-8 Double Mechanical Seal

This discussion is confined to a double mechanical seal (Fig. 6-8). A single seal configuration is the lower half of a double seal and this discussion is valid for both halves. The detail of the rotating and stationary elements of the seal is shown in Fig. 6-9.

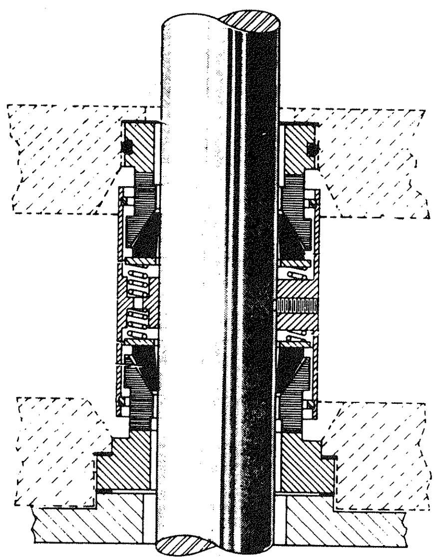


FIG. 6-9 Details of Mechanical Seal

In discussing mechanical seals, the most common concern is the deflection of the shaft in the seal area. Generally, calculated deflection versus bearing span must also be considered. In Fig. 6-10, note that the seal is located X units below the inboard shaft bearing, and a force, F, acts at the end of an overhung, rotating shaft.

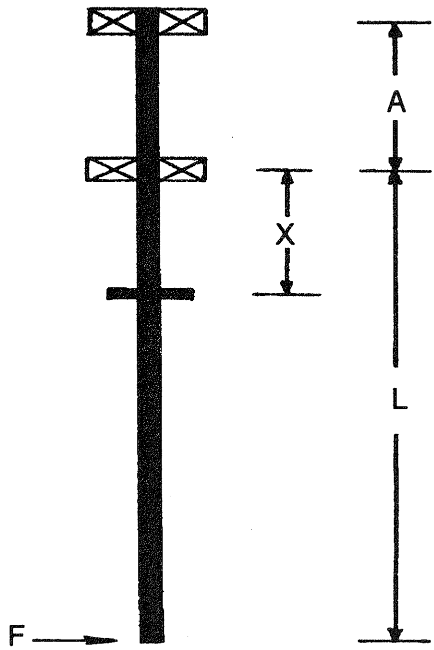



FIG. 6-10 Load and calculated deflection on mixer shaft bearings

LOAD ON BEARINGS

$$\frac{FL}{A} + F$$

CALCULATED DEFLECTION

$$\frac{FX}{6K}\left(2AL + 3LX - X^2\right)$$

Notice that the radial load on the bearings is higher for the shorter bearing span, A. For the same force, a shorter bearing span will require larger bearings (or the same bearing will have a shorter life).

On the other hand, a shorter bearing span will result in a lower *calculated* deflection. Thus, all else being equal, shorter bearing span, A, means less deflection, and normally, less seal wear.

The seal life is not affected by the calculated deflection per se, but more importantly, the calculated deflection compares with the allowable deflection of the seal faces.

## SEAL FACE DESIGN

As shown on the left of Fig. 6-11, the width of the rotary seal face (the upper face in the diagram) is the same as the width of the stationary face. This is in contrast to the relatively narrower rotary face on the right which is normally used in pumps. A narrow rotary face seal is suitable for pump service but not necessarily for mixer service.

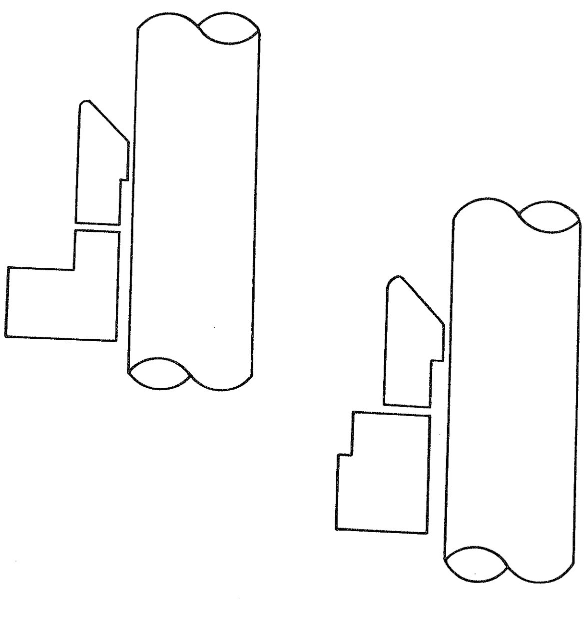


FIG. 6-11 Comparison of sealing faces required for mixer service vs pump service

When the seal faces are new, either type works well, but, as with all parts in friction, wear ultimately occurs. Although a carbon rotary face will wear faster than an alloy stationary face, both faces do wear.

With the face on face type seal, equal face widths provide leak-free sealing, even with abnormal deflections.

With the narrow face pump seal, the rotary face wears a groove in the stationary face, and as the shaft experiences normal deflections, the outer edge of this groove is rounded off, creating a path for premature seal leakage. A more severe problem results when abnormal deflections occur. As is shown in Fig. 6-12, the groove in the stationary face will attempt to hold the rotary face, resulting in the seal face being fractured at the corner, which causes massive leakage of lubricant.

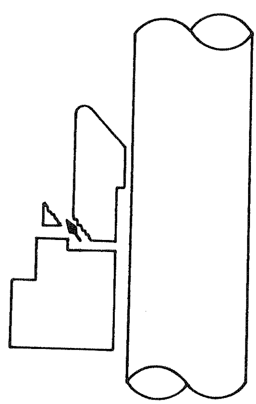


FIG. 6-12 Rotary face damage results in mixer service when seal faces are of unequal width

## PACKAGE SEALS AND CARTRIDGE SEALS

Seal life is also strongly affected by the mixer shaft deflections as a result of the seal design and machining tolerances. Short bearing-span units in particular develop problems due to machining tolerances resulting in higher *actual* deflections than would have been predicted by a theoretical calculation. Many short bearing-span mixers have mechanical seals which are supplied on a sleeve. The sleeve is sealed to the mixer shaft by O-rings which in itself may cause problems. We could pre-test the sleeve subassembly either statically or dynamically but when the seal package is installed on the mixer shaft, the O-ring seal still remains to be tested. Therefore, leakage, if any, can only be determined after the mixer is put on stream.

With a package seal there is a likelihood of compounding the effects on manufacturing tolerances. The design incorporates a bearing on a sleeve which is mounted on another sleeve then mounted on the mixer shaft (Fig. 6-13). In addition there is an additional gap required for the O-ring. The misalignment due to accumulated tolerances results in the centerline of the bearings being non-coincident with the centerline of the mixer shaft; thus, a built-in deflection occurs which could be as large as the shaft deflection caused by the force at the end of the shaft.

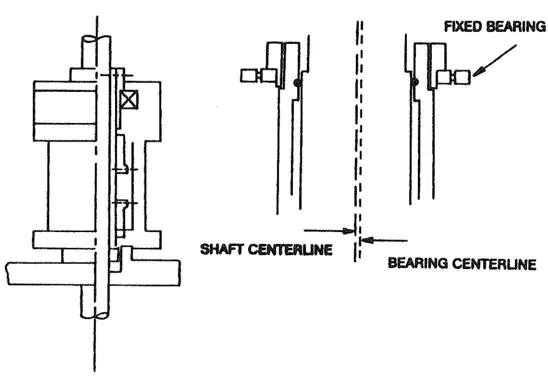



FIG. 6-13 Compound tolerance effect of a package seal

The larger bearing-span units which use a tubular shaft reducer eliminate the need for a sleeve seal package. This type of design minimizes seal deflection (Fig. 6-14). The fixed bearing is pressed onto the seal shaft directly above the rotating seats. This ensures true alignment between the bearing and the shaft. Tolerance buildup as associated with sleeve seals is non-existent.

This type of seal assembly is normally referred to as a cartridge design since the complete assembly can be pretested at the mixer operating speed and temperature before mounting on the tank nozzle. A cartridge seal should always be considered if downtime is of importance to the process or to plant economics.

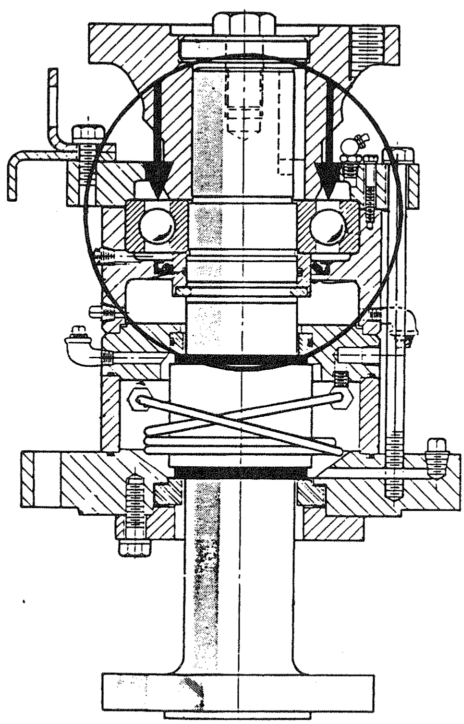


FIG. 6-14 Cartridge Seal

## COUPLING DESIGN

Another possible source of deflection is the method of attaching the removable coupling-half above the seal. The coupling shown in Fig. 6-15 depicts how the bending moment is transmitted from the shaft through the coupling. By maintaining the tubular section of the coupling in compression with bolt tension, the coupling does not rely on an interference fit between the shaft and coupling to transmit bending moments. Thus, fretting and undesirable deflection of the shaft are eliminated.

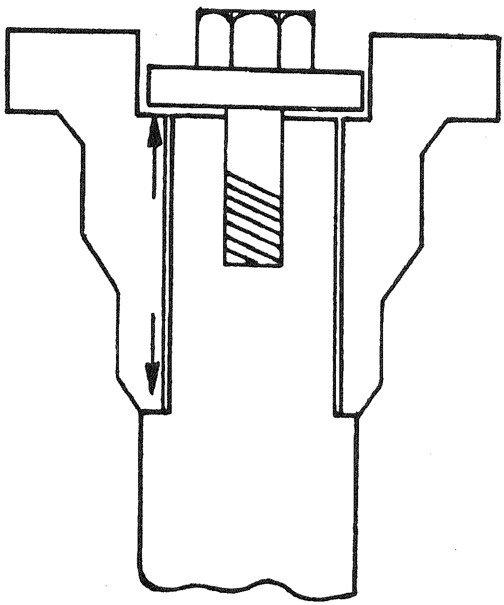


FIG. 6-15 Removable Coupling

A tapered coupling is theoretically the best way to make a coupling removable; as the tapers slide together, there is no gap. However, each taper of the matching set is not necessarily exactly the same. Fig. 6-16 illustrates what often occurs.

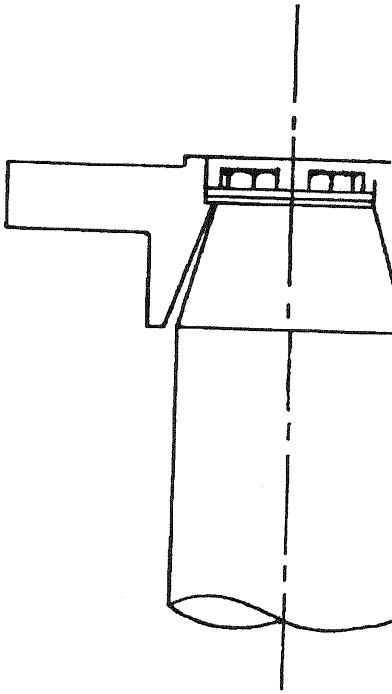


FIG. 6-16 Tapered Coupling

This will add more deflection to the system. If a burr occurs on one of the tapers during maintenance, the tapers will not seat properly. The resulting situation would not be much different from the behavior of a ball joint.

An ill-fitted coupling can cause so much movement that fretting at the seal can occur between the Teflon wedge (between the springs) and the rotary seal face. If a sleeve design is used, the contact between the shaft and sleeve may be damaged.

If either the fretting or galling is so pronounced that integrity of the metal surface is ruined, then the complete mixer drive normally has to be removed to remachine the damaged surface.

# seven

## FLUID MIXING LABORATORY MANUAL

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NOTE: The following mixing experiments assume the use of a laboratory dynamometer. If a dynamometer is not available, power can be measured using a torque cell on the mixer shaft or by using a recording wattmeter. If a recording wattmeter is used, care must be taken to ensure that the power measurement is corrected to produce mixer impeller horsepower (shaft horsepower). All of the power measurements required in the experiments are impeller horsepower values. The impellers used in these experiments are identified by the nomenclature given in Experiment #I, Fig. I-1 which is used by LIGHTNIN. Other manufacturers may use different identifiers to describe the same or similar impellers.

WARNING: SAFETY GLASSES MUST BE WORN AND SAFETY PROCEDURES FOLLOWED WHEN CONDUCTING THESE EXPERIMENTS.

# EXPERIMENT I

## POWER CONSUMPTION OF MIXING IMPELLERS

OBJECTIVES:

1. To determine the effect of impeller speed and diameter on impeller power consumption.

2. To determine the effect of fluid density on impeller power consumption.

EQUIPMENT REQUIRED:

Laboratory dynamometer
18" diameter vessel fitted with four flat wall baffles, 1 1/2" wide (see sketch below).
One 6.8" (173mm) diameter A310 impeller
One 5.2" (132mm) diameter A310 impeller
One 4.5" (115mm) diameter A310 impeller
Approximately 50 lbs. (23kg) sodium chloride
Tap water source
Lift table

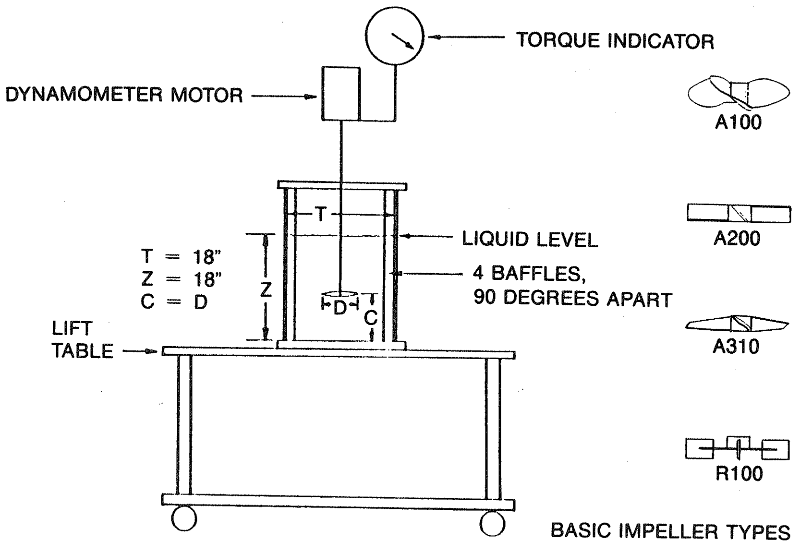



Fig . I-1

PROCEDURE I:

Mount a 6.8" (173mm) diameter A310 impeller at the end of the dynamometer shaft and locate the impeller in the center of the vessel, one impeller diameter off the bottom of the vessel.

Add water to the 18" (457mm) level (Z=18). With the mixer speed control set at zero RPM, turn on the dynamometer motor and adjust the speed to produce a stable torque reading. Record the speed and torque and calculate impeller power consumption:

$$P = N \times T \times k$$

where P = impeller horsepower
N = impeller speed, rpm
T = torque scale reading, lb.
k = dynamometer constant

Run at least six other speeds within the limits of the dynamometer's load range and calculate impeller power consumption for each speed. (Although speed selection is arbitrary, the suggested speeds for this impeller diameter are 350, 400, 500, 600, 700, 800 and 1000 RPM - this range will provide a good spread in the data.)

Repeat the experiment using the 5.2" (132mm) diameter A310 located 5.2" (132mm) off-bottom. (Suggested speeds - 400, 500, 600, 700, 800, 900, 1000, 1100 and 1200 RPM.) Record diameter, speed and horsepower in each case.

Repeat the experiment using the 4.5" (114mm) diameter A310 located 4.5" (114mm) off bottom.

ANALYSIS OF DATA:

Correlate power (P) with speed (N) for the 6.8" (173mm) diameter impeller (D). Do the same for the other two impellers. You may use statistical techniques or plot power vs speed on log-log graph paper, calculate constants, measure slopes, and write the equations. Write equations to predict power consumption as a function of speed for each impeller:

$P_{6.8\ (173mm)}$ = ______________________________

$P_{5.2\ (132mm)}$ = ______________________________

$P_{4.5\ (114mm)}$ = ______________________________

What is the relationship among the three equations? Write a general equation to predict the power consumption of any A310 impeller diameter at any speed under the same test conditions.

$P_D$ = ______________________________

PROCEDURE II:

Repeat all of the experiments in Procedure I using a salt water solution. The specific gravity required is arbitrary but it should be large enough to give measurable results. It is recommended that 50 lbs. (23kg) of sodium chloride be completely dissolved in water and the final liquid depth adjusted to 18" (457mm) in the vessel. Measure or calculate the specific gravity of the solution. Record this value and run the experiment.

Write equations for power consumption as in Procedure I.

For specific gravity ( $\rho$ ) = __________ (concentrated salt solution)

$P_{6.8\ (173mm)}$ = ______________________

$P_{5.2\ (132mm)}$ = ______________________

$P_{4.5\ (114mm)}$ = ______________________

How do these results compare to those obtained in Procedure I?

PROCEDURE III:

Remove about half of the salt solution and readjust liquid level to 18" (457mm) with tap water and mix thoroughly to produce about 50% of the salt concentration in Procedure II. Measure and record the specific gravity. Rerun the experiment for each impeller and write the following equations:

For $\rho$ = ______________ (diluted salt solution)

$P_{6.8\ (173mm)}$ = ____________________

$P_{5.2\ (132mm)}$ = ____________________

$P_{4.5\ (114mm)}$ = ____________________

Now compare the results from Procedures I, II and III. (NOTE: in Procedure I, $\rho$ = 1.0.)

Write a general equation for the power consumed by any A310 impeller diameter at any speed and any fluid density, assuming the fluid viscosity is 1.0 cP:

$P_{D,\ \rho}$ = ____________________________

## EXPERIMENT II

### Power Number/Reynolds Number Determination

OBJECTIVE:

To determine the relationship between power number and Reynolds number for a given mixing impeller design.

EQUIPMENT REQUIRED:

Laboratory Dynamometer
18" diameter vessel fitted with four flat wall baffles,
1 1/2" (38mm) wide (See Fig.I-1, Experiment I)
One 6" (152mm) R100 impeller
One 6" (152mm) A200 impeller
One 6" (152mm) A100 impeller
Water
Lift table
Approximately 20 gals. (76 litres) of corn syrup
Brookfield viscosimeter (or equivalent)

DISCUSSION:

Reynolds number, $N_{Re}$, and power number, $N_P$, are dimensionless groups which are used to express relative power consumption for different impeller designs. They are defined as:

$$N_{Re} \propto \frac{ND^2\rho}{\mu} \text{ (ratio of inertia force to viscous force)}$$

and

$$N_P \propto \frac{P}{N^3D^5\rho}$$

where $P$ = impeller power consumption
$N$ = impeller speed
$D$ = impeller diameter
$\rho$ = fluid density
$\mu$ = fluid viscosity

The proportionalities can be written as equations, and in the English system:

$$N_P = \frac{1.53 \times 10^{13} P}{\rho N^3D^5}$$

$$N_{Re} = \frac{10.7\, ND^2\rho}{\mu}$$

where $P$ = impeller horsepower
$N$ = impeller speed, rev/min
$D$ = impeller diameter, inches
$\rho$ = fluid specific gravity, dimensionless
$\mu$ = fluid viscosity, cP

In metric units:

$$N_P = \frac{2.158 \times 10^{17} P_1}{N^3 d^5 \rho}$$

and $$N_{Re} = \frac{1.667 \times 10^{-5} N d^2 \rho}{\mu_o}$$

where $P_1$ = impeller power, W
$d$ = impeller diameter, mm
$\mu_o$ = dynamic viscosity, Pa· s
$N$ = impeller speed, rev/min
$\rho$ = fluid specific gravity, dimensionless

For a given impeller geometry, a single $N_P$ vs $N_{Re}$ curve (Fig.II-2) can be generated which holds true for any impeller diameter, for that specific geometry - e.g. when $N_{Re} = k_o$:

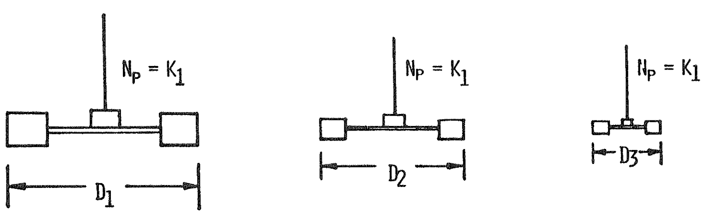



Fig. II-1

or, at $N_{Re} = k_o$, $N_{p(D1)} = N_{p(D2)} = N_{p(D3)} = N_{p(Dn)} = K_1$

Therefore, at a given Reynolds number, power number is constant for a given impeller design for any diameter and geometrically similar systems.*

A typical power number vs Reynolds Number curve looks like this (Fig. II-2) when plotted on log-log paper:

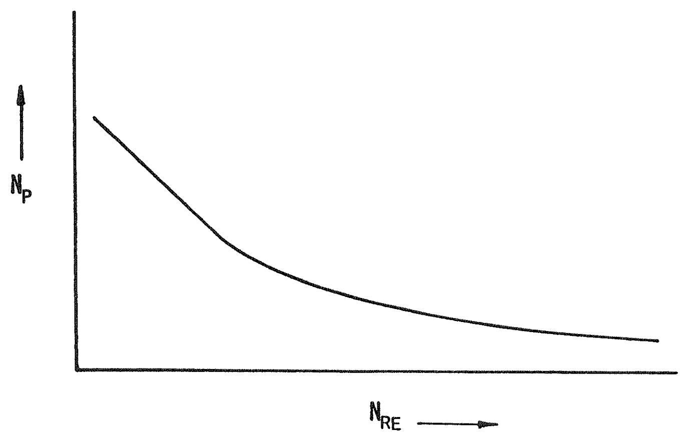



Fig. II-2

* where the ratios D/T, C/D, Z/T are constant (refer to Fig.I-1, Experiment #1).

This curve can be separated into three distinct fluid regimes - laminar, transition and turbulent, Fig. II-3 (similar to the friction factor curves for flow in pipes).

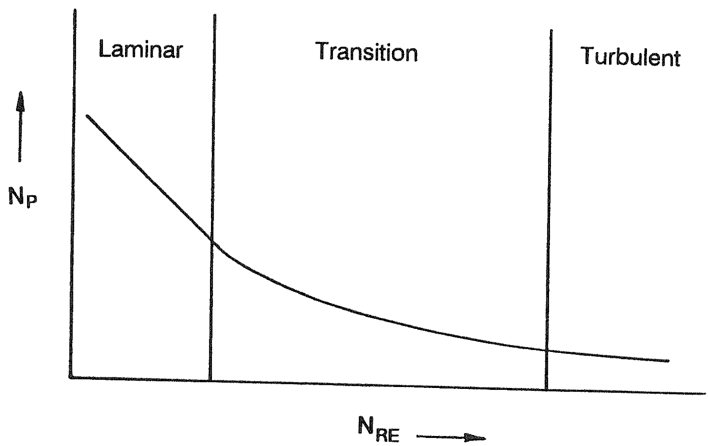



Fig. II-3

Note that in the LAMINAR region, power number varies inversely as Reynolds number. In the TRANISTION region, slope varies continuously and in the TURBULENT region the slope is constant (zero) therefore, power number is constant.

If we had three completely different impeller designs, three separate curves could be generated as symbolized in Fig. II-4.

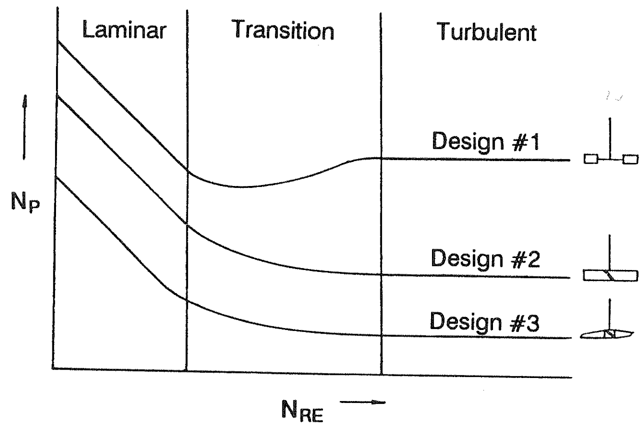



Fig. II-4

PROCEDURE I:

From the 6.8" (173mm) A310 impeller data obtained in Experiment #1 (where D = 6.8, (173mm) $\rho$ = 1.0, $\mu$ = 1.0) calculate power number and Reynolds number for each data point.

What did you observe about power number range?

What did you observe about Reynolds number range?

In which fluid regime was the impeller operating? Why?

Assume that the points of inflection for the $N_P$ vs $N_{Re}$ curve for an A310 are (10, 2.60) and (2,500, 0.30) as shown in Fig. II-5.

A310 IMPELLER

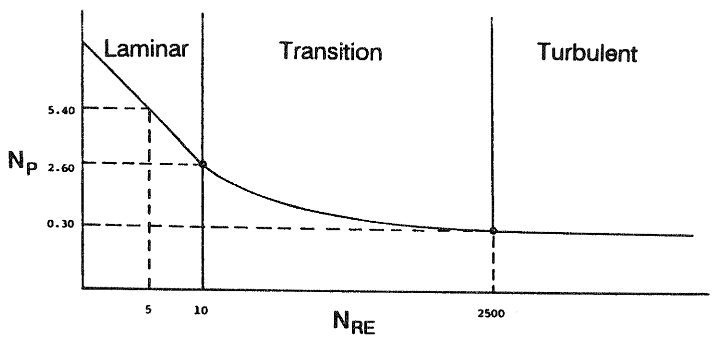



Fig. II-5

If the viscosity of the fluid were changed to 1000 cP (1.0 Pa · s) assume $\rho$ = 1.0) what happens to Reynolds Number for each of the speeds used in the previous exercise? What is the fluid regime for each speed?

| N | $N_{Re}$ ( =1,000) (1.0 Pa·s) | FLUID REGIME |
|---|---|---|
| ___ | ___ | ___________ |
| ___ | ___ | ___________ |
| ___ | ___ | ___________ |
| ___ | ___ | ___________ |
| ___ | ___ | ___________ |

At 110 RPM and 10,000 cP (10.0 Pa·s), in which fluid regime is the 6.8" (173mm) A310 impeller operating? ________________ Using < , > or =, express the resulting power number __________. If we obtained a Reynolds number of 5.0 by adjusting the impeller speed or viscosity, what would the power number be if we then reduced the speed by 20%? __________ Why? ______________________________.

PROCEDURE II: For each of the following impellers, obtain power data in an 18" (457mm) vessel with four 1-1/2" (38mm) wide baffles and an 18" (457mm) liquid (water) depth. In each case, position the impeller one impeller diameter off the vessel bottom. Calculate the power numbers (assume $\rho$ = 1.0).

$N_p$ (6"R100) (152mm) =

$N_p$ (6"A200) (152mm) = Turbulent Regime

$N_p$ (6"A100) (152mm) =

Based on the results in the turbulent regime,

What is the power number for a 9" (229mm) R100? ____________

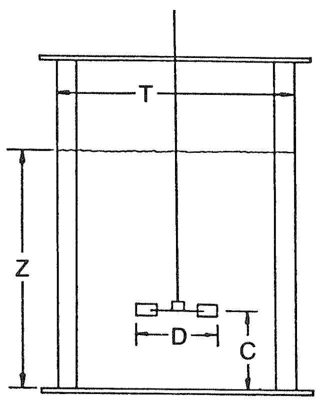
What is the power number for a 7" (178mm) A200? ____________

What is the power number for a 2" (51mm) A100? ____________

Turbulent Regime

Why is this true?

In Procedure II we have maintained the following geometry:




$Z = T$
$D = T/3$
$C = D$

Fig. II-6

If we tripled each dimension, then took power data using the same fluid, one of the following statements would be correct:

1. $N_P$ would be greater than that obtained for a 6" (152mm) impeller.

2. $N_P$ would be equal to that obtained for a 6" (152mm) impeller.

3. $N_P$ would be less than that obtained for a 6" (152mm) impeller.

Which statement is correct? _____ Why?

PROCEDURE III: Select one of the 6" (152mm) impellers used in Procedure II. Using the same system geometry (T=18 (457mm), Z=18 (457mm), C=D) obtain power data with successive dilutions of corn syrup and generate the power number vs Reynolds number curve for the impeller selected.

Suggestions: Some commercial grade corn syrups have viscosities as high as 170,000 cP (170 Pa · s) and as low as 90,000 cP (90 Pa · s), undiluted, at room temperature. It is important to predetermine the effects of dilution and temperature on viscosity before running power tests since only a small amount of dilution and/or a small temperature change will produce a dramatic effect on viscosity. By checking viscosity for successive dilutions and/or temperature changes in a beaker-size sample, you can predetermine the approximate desired dilution to produce the desired viscosity. Calibration curves of viscosity vs temperature can be used to determine the fluid viscosity at each impeller speed. Since the power applied by mixing the fluid will tend to increase the fluid temperature, viscosity will tend to decrease as mixing continues, therefore, it is necessary to know the temperature of the fluid each time a data point is recorded. Furthermore, the specific gravity of the fluid will decrease upon dilution. This value must also be determined for each data point. Also, for a given fluid viscosity and specific gravity a 100 fold change in Reynolds number can be measured by a 100 fold change in speed because

$N_{Re} \propto N$. Therefore, a single dilution would cover two orders of magnitude in Reynolds number if speed is also varied by two orders of magnitude.

Plot the results on log-log paper and draw the $N_P$ vs $N_{Re}$ curve for the impeller.

# EXPERIMENT III

## BAFFLE DESIGN

OBJECTIVE: To determine the effect of baffle design on power response and mixing intensity.

EQUIPMENT REQUIRED:
18" (457mm) diameter vessel any mixing impeller, where 0.2 - D/T - 0.5
6 flat baffles, T/36 inches wide
6 flat baffles, T/18 inches wide
6 flat baffles, T/12 inches wide
Water
Laboratory dynamometer and lift table

DISCUSSION: It is obviously desirable to minimize the power required to produce the desired mixing result. For given fluid properties and a given impeller type, diameter and speed, the baffle design and number of baffles controls the power applied to the fluid. Baffles also influence flow pattern and mixing intensity. Thus a given mixing system may be underbaffled, overbaffled or adequately baffled to produce optimum results.

In a vertical/cylindrical tank with mixer located on the tank centerline, proper baffle design produces these effects:

1. assures consistent, stable, power draw
2. helps dampen radial shaft loads
3. promotes optimum mixing flow pattern by reducing or eliminating rotary (relatively static) motion

4. prevents vortexing, swirling and air induction
5. directs flow from the impeller to produce the necessary vertical currents
6. improves loading accuracy

Baffles are not normally required for portable mixers or side-entering mixers (or for top-entering mixers where viscous fluids are encountered - i.e. the laminar flow regime). In other cases, with rare exception, baffles are usually required. An unbaffled system would produce wider variations, swirling, vortexing and very little real mixing.

PROCEDURE I: Select any impeller and tank diameter such that 0.2 - D/T - 0.5. Fill the tank with water to Z/T = 1.0. Use tracer particles (e.g. plastic beads or other insoluble solid particles) to observe the flow pattern and mixing intensity.

At an arbitrary impeller speed, using four standard flat baffles (baffle width = T/12) located at the tank wall, 90 degrees apart, note the flow pattern and impeller horsepower. Remove the baffles and run the impeller at the same speed. Note the flow pattern, horsepower and mixing intensity (record results after power response has stabilized).

Compare the power, flow pattern, mixing intensity and particle motion in a baffled tank to the same characteristics in an unbaffled tank.

PROCEDURE II: Impeller power consumption and, to some extent, mixing results can be correlated with baffle width and number of baffles. For each of the conditions given in Table III-1, run the same impeller at various speeds, record the corresponding horsepower and note mixing characteristics (intensity, flow pattern, swirling or non-swirling, vortexing or non-vortexing, tracer particle behavior, etc.)

Correlate relative power consumption as a function of baffle width and number of baffles using power for four standard baffles (T/12 in width) as the base.

Record observations relative to apparent mixing intensity, uniformity of particle distribution, etc. for each baffle arrangement.

TABLE III-1

| BAFFLE WIDTH | NO. OF BAFFLES |
|---|---|
| T/36 | 6,4,2,1 |
| T/18 | 6,4,2,1 |
| T/12 | 6,4,2,1 |
| 0 | 0 |

# EXPERIMENT IV

## SOLIDS SUSPENSION WITH MIXING IMPELLERS

OBJECTIVE: To study the effects of impeller diameter, speed (power) impeller position and impeller type on the ability to suspend solids.

EQUIPMENT REQUIRED:
Laboratory dynamometer and lift table
18" (457mm) diameter vessel fitted with four flat wall blaffles, 1 1/2" wide
4" (102mm) and 6" (152mm) diameter A200 impellers
6" (152mm) diameter R100 impeller
4.5" (114mm), 5.2" (132mm) and 6.8" (173mm) A310 impellers
Plastic beads or other insoluble tracer particles
Water

DISCUSSION: Suspending solids is one of the most common applications in fluid mixing. The degree of suspension is extremely important because the mixer power requirement increases rapidly as we go from simple particle movement along the bottom of the vessel to a completely uniform suspension.

In many processes, the process result is satisfied by supplying enough mixer power to impart complete particle motion along the bottom of the vessel, with no need for particle suspension (on-bottom motion). In other processes enough power must be available to suspend all of the particles, but not necessarily uniformly (off-bottom suspension). Still another process criterion is the power required for uniformly suspending all of the particles (complete uniformity).

The ease or difficulty of suspending solids depends on particle density, mesh size, concentration and the fluid density and viscosity. As the solid particle becomes larger or denser, the particle settling velocity increases. As the fluid density or viscosity increases, the particle settling velocity decreases. Therefore, meeting any suspension requirement depends on overcoming the particle settling velocity which is a function of particle size, density and fluid properties.

The relative impeller speeds and power requirements for a range of settling velocities are given in Table IV-1, where N is impeller speed and P impeller horsepower. This illustrates the wide range of power requirements as a function of settling velocity range, for the same process criterion.

TABLE IV-1

| Settling Velocity ft/min (m/min) 16-60 (4.9-18.3) | | Settling Velocity ft/min (m/min) 4-8 (1.2-2.4) 0.1-0.6 (0.03-0.18) | | |
|---|---|---|---|---|
| N | P | | P | P |
| 2.9 | 25 | Complete Uniformity | 9 | 2 |
| 1.7 | 5 | Off-Bottom Suspension | 3 | 2 |
| 1.0 | 1 | On-Bottom Motion | 1 | 1 |

Another important factor in determining process power requirement for solids suspension is the use of fillets in the tank. In tanks that have flat bottoms, material settling at the intersection of the tank wall and tank bottom (a fillet) is often very difficult to suspend. The additional power required to suspend this material is significant. As a result, it is often necessary to allow a non-progressive fillet to form in order to minimize the required mixer power (Fig. IV-1).

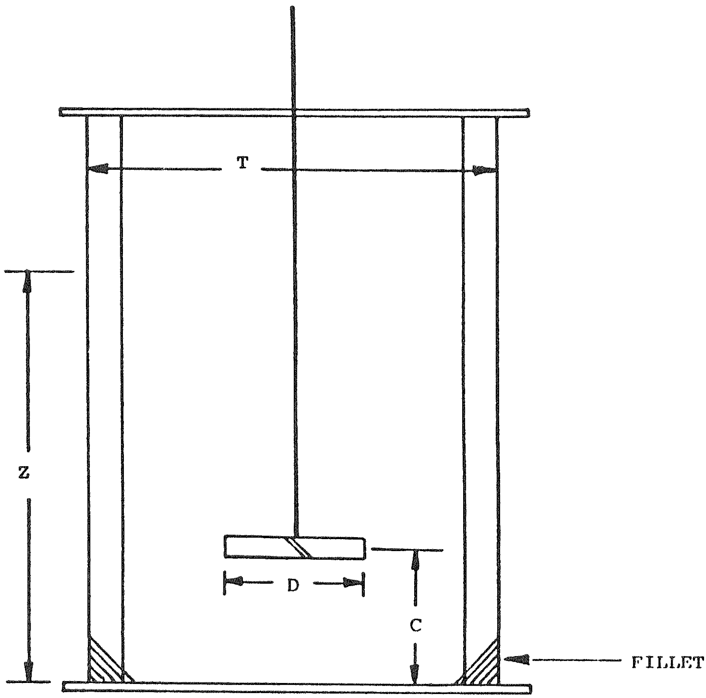



Fig. IV-1

For a given solids concentration, as the allowable fillet size increases, the power requirement decreases for a given process requirement.

Other factors which affect mixer power requirement are: the ratio of fluid depth to tank diameter, Z/T, the ratio of impeller diameter to tank diameter, D/T, location of the impeller with respect to the tank bottom, C/D.

PROCEDURE I: Add water and 5 lb. (2.25kg) of plastic beads (or other insoluble tracer particles) which have a specific gravity greater than 1.0 to the 18" (457mm) tank and adjust the liquid level to Z/T=2/3. Using a 4" (102mm) A200 impeller at C/D=1.0, determine the impeller power required for each process requirement (on-bottom motion, off-bottom suspension, complete uniformity) without fillet formation. Adjust the impeller position to C/D=0.5 and rerun the experiments. Run the same experiment using a 6" A200 impeller. Next, adjust the liquid level to Z/T=1.0 and add enough beads to produce the same solids concentration as in the previous experiments. Run the experiments again for each impeller at C/D=1.0 and 0.5. Correlate the power requirements with respect to D/T, Z/T and C/D for each process condition.

PROCEDURE II: Using 4.5" (114mm), 5.2" (132mm) and 6.8" (173mm) A310 impellers run the same experiments as in Procedure I. Correlate the data and compare the results for the A310 with the results for the A200. (See sample data sheet.) What are your conclusions?

Additional experiments can be run for various solids concentrations and the data can be correlated to reflect the effect of % solids.

NOTE: In determining the impeller power that satisfies the process result, visual observation of particle behavior is required. Care must be exercised to use the same judgement criterion for a given process result. The following descriptions may be helpful:

ON-BOTTOM MOTION: This condition is satisfied when every particle is in motion at some time. It is not necessary for particles to rise off the bottom of the tank although some particle suspension will occur. The minimum power applied to meet these conditions is recorded as the power required for ON-BOTTOM MOTION.

OFF-BOTTOM SUSPENSION: This condition is satisfied when every particle has vertical velocity at some time. All of the particles must be suspended off the bottom of the tank, not necessarily continuously. As long as the particles achieve an upward velocity at some time during the test, the process requirement will be satisfied. The minimum power to produce this result is the power required for OFF-BOTTOM SUSPENSION.

COMPLETE UNIFORMITY: This condition is satisfied when the solids appear to be uniformly distributed throughout the tank.The minimum power to achieve this result is the power required for COMPLETE UNIFORMITY.

SAMPLE SOLIDS SUSPENSION DATA SHEET
ON-BOTTOM MOTION

| IMPELLER TYPE | D | D/T | Z/T | C/D | % SOLIDS | N | hp |
|---|---|---|---|---|---|---|---|
| A200 | 4 | 0.222 | 0.667 | 1.0 | | | |
| A200 | 4 | 0.222 | 0.667 | 0.5 | | | |
| A200 | 4 | 0.222 | 1.000 | 1.0 | | | |
| A200 | 4 | 0.222 | 1.000 | 0.5 | | | |
| A200 | 6 | 0.333 | 0.667 | 1.0 | | | |
| A200 | 6 | 0.333 | 0.667 | 0.5 | | | |
| A200 | 6 | 0.333 | 1.000 | 1.0 | | | |
| A200 | 6 | 0.333 | 1.000 | 0.5 | | | |

(Use same format for OFF-BOTTOM SUSPENSION and COMPLETE UNIFORMITY.)

## EXPERIMENT V

### LOW VISCOSITY BLENDING

OBJECTIVE: To determine blend time as a function of impeller design, power speed and diameter.

EQUIPMENT REQUIRED:
Laboratory dynamometer
$N_aOH$ solution
HCl solution
18" (457mm) tank with baffles
Timer or stopwatch
Recording pH meter (optional)

DISCUSSION: It is fairly obvious that given a system for blending low viscosity materials the time required to achieve uniformity varies inversely with applied mixer power. But the blending capabilities of various impeller types also vary. Measuring blend time as a function of power, speed and impeller diameter will provide relative blending correlations for each impeller type.

One of the simplest ways to measure blend time for low viscosity and low specific gravity fluids is to find the time required to reach thermal or physical equilibrium. Methods such as blending fluids of different temperatures or different concentrations of soluble constituents have been used by many investigators. Another method is measuring pH change and recording the time to reach equilibrium. This is probably the most practical method for most laboratories.

PROCEDURE I: Fill an 18" (457mm) diameter baffled tank to the 18" (457mm) level with water. Add phenolphthalein indicator. Select any impeller type and diameter and mount it one impeller diameter off the bottom of the tank. Add enough sodium hydroxide to produce a color change. Select an operating speed or horsepower. With the mixer running, add enough hydrochloric acid to neutralize the sodium hydroxide. Record the time for the first disappearance of color. This is blend time. By alternately adding acid or base, obtain several blend times for various power levels. Use another impeller diameter (of the same type) and repeat the experiment. Determine the relation for power level, speed, D/T and blend time. Run the same tests for a different impeller type and compare the results. (This experiment is somewhat subjective since color change must be judged by eye. An alternate is to use a recording pH meter and record the time to reach equilibrium.) In this experiment it is assumed that acid-base reaction time is always small when compared with blend time.

PROCEDURE II: An alternate method fordetermining the blending characteristics of mixing impellers is to measure the time required to uniformly blend miscible layers of materials. This is called stratified blending. Normally, on a commercial scale, stratified blending is accomplished in large tanks (e.g. petroleum blending) using side-entering mixers. However, the effect of variations in power, speed impeller geometry, impeller diameter and impeller location in the vessel can also be observed using top-entering mixer.

If a two-layer system of miscible fluids is prepared, blending to uniformity can be measured in a number of ways.

These include recording the time to achieve a physical uniformity such as a uniform temperature, viscosity, concentration, specific gravity, etc.

For example given a two-layer system of fresh water and salt water, the time required to blend the system to uniformity can be determined as a measure of conductivity, specific gravity, or colorimetrically, by dyeing one layer with a water-soluble food dye (Fig. V-1).

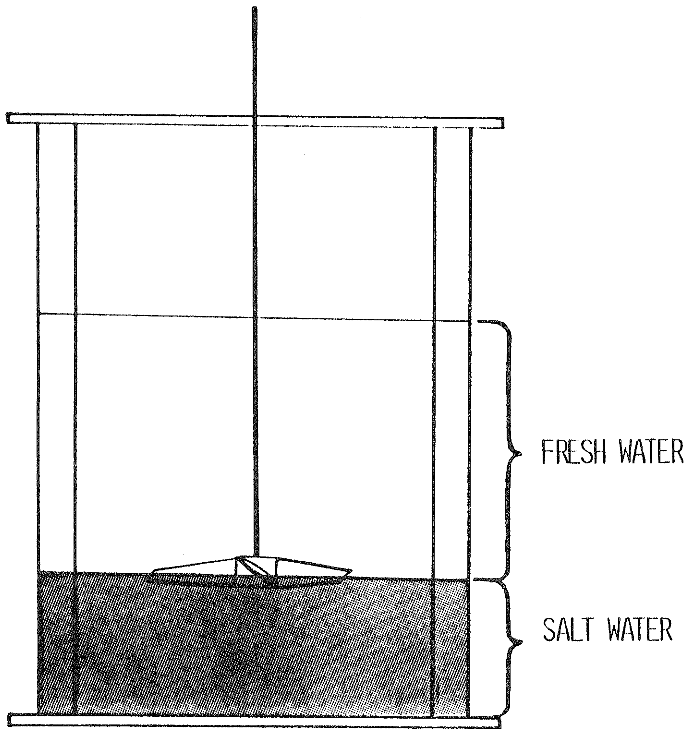



Fig. V-1

# EXPERIMENT VI

## GAS-LIQUID MASS TRANSFER

OBJECTIVE: To study the effects of impeller type, diameter, speed (power) and gas rate on the mass transfer coefficient in a gas/liquid system.

EQUIPMENT REQUIRED:
Laboratory dynamometer and lift table
Tap water
18" (457mm) diameter vessel with four 1-1/2" (38mm) wide baffles
6" (152mm) R100 impeller, 6" (152mm) A200 impeller
Clean air supply
Rotameter (to meter airflow)
Sparge ring or tube (see Figs. VI-1, and VI-2)
Sodium sulfite, $Na_2SO_3$
Cobaltous chloride, $CoCl_2.6H_2O$
Reagent chemicals used in dissolved oxygen analysis (the Azide Modification of the Winkler Method - consult STANDARD METHODS FOR THE EXAMINATION OF WATER AND WASTEWATER - Sec. 421B, 15th Edition).

Dissolved oxygen analyzer (alternate method for measuring dissolved oxygen - if this method is used, reagent chemicals are not required for calibrating the dissolved oxygen meter). 60 ml dissolved oxygen sample bottles (8 or 10 required) for chemical method of dissolved oxygen analysis - not required if dissolved oxygen meter is used.

DISCUSSION: One way to illustrate the mass transfer coefficient as a function of mixing variables is to measure the dissolved oxygen transfer rate in tap water for various conditions.

By first removing dissolved oxygen (DO) from water, then reaerating and sampling over time until the water is saturated with DO, the oxygen transfer rate and mass transfer coefficient can be determined.

PROCEDURE:

Equipment should be as shown in Figs. VI-1 and VI-2.

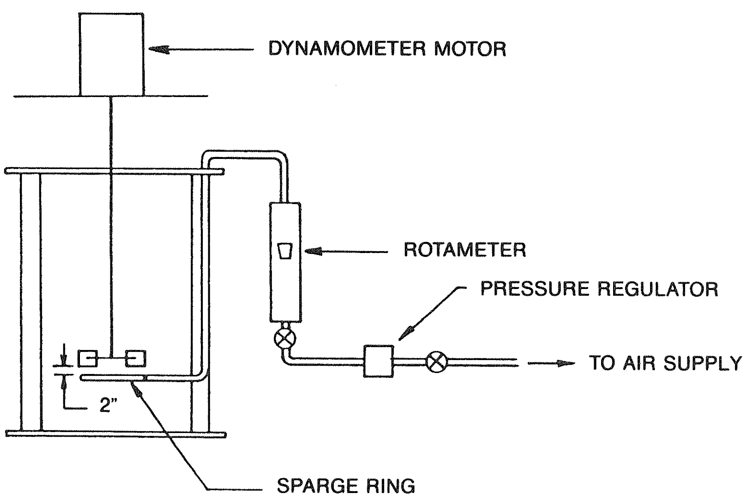



Fig. VI-1

Sparger may be a ring or open tube:

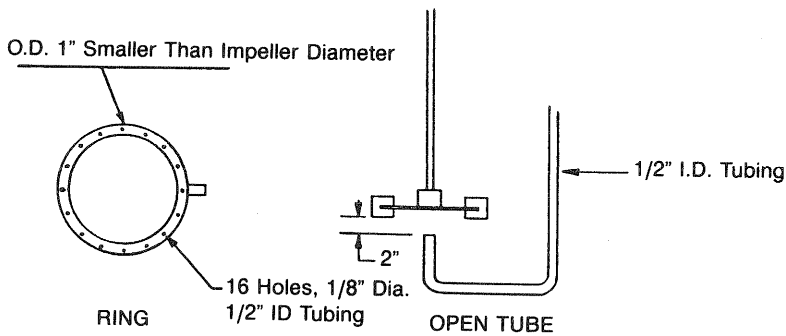



Fig. VI-2

The mass transfer coefficient, $K_G a$, is often expressed as a function of gassed impeller power consumption and gas rate. Furthermore, gassed impeller power consumption is usually expressed as power per unit volume, Pg/V, and gas rate as superficial gas velocity, f :

$$K_G a = \frac{\text{lb. mol}}{(\text{HR})(\text{ft}^3)(\text{atm})} = F(Pg/V, f)$$

$$f = \frac{Q}{A}$$

where f = superficial gas velocity, ft/sec

Q = gas rate, $ft^3$/sec
A = tank cross sectional area, $ft^2$

and

Pg/V = gassed impeller horsepower/1000 gal.

Recommended test run variables for each impeller to be tested are given in Table VI-1.

TABLE VI-1

| Impeller Power (Gassed), Pg/V (hp/1000 gal.) | Superficial Gas Velocity (ft/sec) | | | |
|---|---|---|---|---|
| 1.0 | 0.001 | 0.003 | 0.01 | 0.02 |
| 2.0 | " | " | " | " |
| 3.0 | " | " | " | " |
| 4.0 | " | " | " | " |

PROCEDURE II: Add tap water to the 18" (457mm) liquid level in the 18" (457mm) diameter test tank. Add 2.0 ppm $Co^{++}$. This is required to catalyze the reaction between $SO_3^{=}$ and D.O. Measure the water temperature and determine the equilibrium dissolved oxygen concentration at saturation. (This can be done by saturating the water with air, sampling and analyzing for D.O., or by referring to DO solubility tables in Standard Methods For The Examination of Water and Wastewater, 15th edition.

Set the gas rate at f = 0.001 and the gassed mixer hp at 1.0/1000 gal.

Calculate the stoichiometric quantity of sodium sulfite required to remove the D.O. from the water in the test tank. Multiply this value by 1.2 to 1.5* and dissolve that amount of $Na_2SO_3$ in the tank. Begin sampling at the first indication of uptake when using a D.O. meter (or within a few seconds of sulfite addition when using the Winkler method). Continue sampling, and record the time and D.O. for each sample, until saturation occurs. Calculate $K_Ga$ for the test conditions. (Note: an excess of sulfite beyond the stoichiometric amount necessary for complete reaction is required since the system will be aerating before the first sample is drawn. Without an excess of sulfite, especially at high uptake rates, D.O. saturation could occur before samples are collected. This is especially true if the Winkler method is used because there is no quick way to determine when uptake begins. When using a D.O. meter, this problem is minimized since the meter will immediately indicate uptake. Nevertheless, whether the Winkler method is used or a D.O. meter is used, an excess of sulfite is required because some of the sulfite is oxidized before sampling begins).

Repeat the procedure for the remaining superficial gas velocities and impeller gassed horsepowers given in Table VI-1. Correlate $K_Ga$ as a function of Pg/V and f.

Using a 6" (152mm) A200 impeller, repeat the tests and correlate the data. Compare the results for the two impellers.

NOTE: For an explanation of the theory of oxygen transfer in clean water (as opposed to wastewater) consult Standard Methods For The Examination of Water and Wastewater, 15th edition.

*Use 1.2 for low gas rates and low power levels and 1.5 for high gas rates and high power levels.

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