

# Mixing in Pipelines

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## 7-1 INTRODUCTION

Most industrial mixing processes take place in tanks or vessels. They are ubiquitous in the process industries. However, mixing can and often does take place in the pipes connecting these process vessels, and when this is the case, the pipelines themselves actually serve as process vessels. In many cases the pipe, especially when equipped with static mixing internals, is a better place to mix and more economical than a vessel. This is often true when fast blending is required or when long hold-ups associated with vessels are not desirable: for instance, when dealing with molten polymers that degrade with time. There are a number of pipeline devices in mixing related service throughout the chemical and hydrocarbon processing industries:

- Static mixer
- Tee mixer
- Impinging jet mixer
- Spray nozzle
- Empty pipe or duct, elbows, etc.
- In-line mechanical mixer

Pipeline mixing is most useful when:

- Process is continuous versus batch

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- Component feed rates are uniform
- Plug flow is preferred to backmixing
- Short residence time is desirable (long residence times are special consideration involving slow reactions)
- With solids of consistent concentration and usually small particle size
- Gas phase continuous (agitated tanks not applicable)
- High pressure (seal concerns)
- Limited space available, limited access—low maintenance desirable

See Myers et al. (1997) for additional selection criteria.

Pipeline or in-line mixing has evolved to play a well-established role in process engineering. Equipment design and use in both the turbulent and laminar flow regimes are well documented. Application reports by industrial users, academic papers, patents, and the literature of equipment manufacturers present a wealth of technology. It is very clear that the value of pipeline mixing technology in the process industry far exceeds the equipment capital cost. Investment in pipeline equipment is small compared to that of in-tank dynamic agitators and other mechanical mixing devices, but is increasing. This growth results largely from static mixers having proven their capability, not only in bulk blending and mixing, but also in applications involving the dispersion of immiscible fluids, heat transfer, interphase mass transfer, and establishing plug flow in tubular reactors.

Static in-line mixers are continuous radial mixing devices, characterized by an effective degree of plug flow, depending on the specific design and application. Unlike dynamic mix tanks, large recirculation flows are not required to achieve desired results. Since there is little backmixing, residence times can be very short, and consequently, many commercial scale static mixers are compact relative to the scale of fluid flow being processed. Since they have short residence times and little backmixing, proper dosing of the feed components with no fluctuation in time is a prerequisite for good performance. When backmixing is required, static mixers are incorporated into pump-around loops. Like agitated vessels they provide a single stage of contacting for interphase mass transfer. Since static mixers have no moving parts, they are low maintenance and sealing problems are nonexistent.

Static mixers range in size from a few millimeters in diameter to units with equivalent diameters exceeding 3 m and volumes exceeding 100 m<sup>3</sup>. They exist in both round and nonround cross-section. Small static mixers, for example, are found in laboratory scale processes for mixing and fast reaction, used in meter-mix-dispense systems to combine reactive components at the point of application, and in synthetic fiber production for flow homogenization prior to spinning. Large static mixers result from requirements to handle large flow rates, to hold a process stream for a long residence time, or to provide a large surface area for heat transfer. Mixer applications involving very high volumetric flow include gas mixing in utility scale power/incineration plants and additives blending in municipal water treatment facilities. Mixers required to provide large fluid

hold-up typically involve slow reactions, as, for example polymerization, hydrolysis, and the catalysis of soluble enzyme reactions, or are mass transfer limited, as in some absorption and extraction processes. Heat transfer applications involve both heating and cooling of viscous and/or heat-sensitive materials. Examples are the rapid preheating of a polymer solution prior to devolatilization and the cooling of viscous products prior to packaging. Applications, device selection criteria, and design fundamentals are presented in the chapter. It is our intent to provide the reader with both a general overview of pipeline mixing technology and an ability to select and size equipment for a variety of applications.

## 7-2 FLUID DYNAMIC MODES: FLOW REGIMES

The flow regime, laminar or turbulent, sets the mechanisms and the relations used in the selection and detailed design of in-line mixing equipment. An early first step in the understanding of a pipeline mixing application is the identification of the fluid dynamic mode or flow regime in which the process operates. The determinates are the fluid flow rate and physical properties. Flow regime can vary with processing rate. Additionally, it must be recognized that fluid properties can change with time during the mixing process, which in the case of pipeline mixing devices means fluid properties and possibly flow regime can vary along the length of the mixer. An understanding of fluid dynamic flow regime is required to calculate degree of mixing, energy expenditure (pressure drop), heat transfer, and drop size in the case of multiphase processes. The design correlations are often valid in only one flow regime. Different correlations are required as flow changes from laminar to turbulent, and vice versa.

### 7-2.1 Reynolds Experiments in Pipeline Flow

The Reynolds number characterizes turbulence in any given pipeline flow or mixing device. It is instructive to consider first the empty or unpacked pipe and look at the classic experiment by Osborne Reynolds (1883). His demonstration consisted of flowing water through a clear glass tube with capability to vary the water flow rate to achieve a broad range of fluid velocity. At the center of the tube a fine jet of water-soluble dye is introduced through a capillary tube so that a thin filament of dye injected coaxially into the stream of water has a velocity equal to that of the water at the point of introduction. Figure 7-1*a* shows that at low water velocity the dye filament retains its identity in the water stream, tending to widen very slightly during the downstream passage because of molecular diffusion of the dye into the water. At a slightly higher mean velocity as shown in Figure 7-1*b*, the dye filament breaks up into finite large eddies. Further downstream the eddies break up further, and the dye that has been introduced tends to become homogeneously dispersed or mixed with the water. At much higher mean velocity (Figure 7-1*c*) the eddy activity becomes extremely violent, and the region of homogeneous dye color approaches the point of dye entry. From visual observation it is evident that the eddies in normal pipe

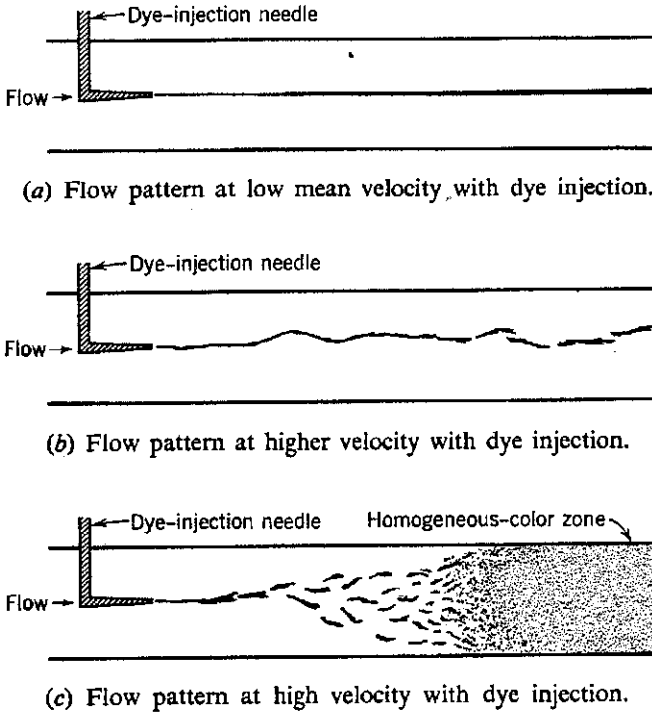


Figure 7-1 Reynolds experiments.

flow were on the order of one-tenth the pipe diameter and move in completely random patterns. Subsequent experiments showed further that eddy formation was influenced by system factors such as pipe wall finish, vibration, dissolved gases, and other factors. Abnormal or metastable flow aside, it was shown that an upper limit of viscous flow and a lower limit of turbulent flow seemed to exist and that the limits were separated by a transition region.

The conclusions that can be drawn from the Reynolds experiment are:

1. Above a certain mean velocity for a given system, relatively large eddies form that flow cross stream in some random behavior.
2. These eddies are larger and more abundant at the center of the tube.
3. An increase in mean velocity of the fluid widens the turbulent core until the tube is essentially filled with the core of eddy activity.

## 7-2.2 Reynolds Number and Friction Factor

The *Reynolds number*, defined as

$$Re = \frac{\rho DV}{\mu} \quad (7-1)$$

was proposed to delineate the flow regime in ducts and pipes. Various forms were later proposed and used for systems other than circular pipes and for pipes containing structures such as static mixers. All are the dimensionless ratio of momentum transferred by eddy mechanisms to momentum transferred by molecular transport. Two systems that operate at the same Reynolds number are dynamically similar with respect to forces associated with momentum transfer.

A friction factor,  $f$ , was also derived to express the ratio of total momentum transferred to momentum transferred by turbulent mechanisms. Friction factor is a function of Reynolds number,  $f = \Phi(\text{Re})$ .

The *Fanning friction factor*, the form used in this chapter, is defined as follows:

$$f = \frac{\Delta P}{\rho V^2} \frac{D}{2L} \quad (7-2)$$

Another definition of friction factor is that used in the Darcy equation for pressure drop. The *Darcy friction factor*,  $f'$ , is defined as

$$f' = \frac{2\Delta P}{\rho V^2} \frac{D}{L} \quad (7-3)$$

and thus

$$f = \frac{f'}{4} \quad (7-4)$$

The European literature often uses the term *Newton number* for the friction factor. The Newton number is defined as

$$\text{Ne} = \frac{\Delta P}{\rho V^2} \frac{D}{L} \quad (7-5)$$

and is one-half the Darcy friction factor or twice the Fanning friction factor.

The definition of friction factor should be noted carefully when using friction factor–Reynolds number plots contained in academic and vendor literature. Also, the influence of pipe roughness should be taken into account when designing for turbulent flow. Publications dealing with pipeline mixing are typically based on measurements taken on clean commercial pipe and plate surfaces. If you contemplate using something else, a roughness factor should be considered.

A plot of the Darcy friction factor versus Reynolds number for flow in open circular pipe of various roughnesses from Moody (1944) is shown in Figure 7-2. The plot shows three distinct flow regimes:

- $\text{Re} < 2100$                       Laminar
- $2100 < \text{Re} < 10\,000$       Transition (the range from 2100 to 3500 is especially unstable)
- $\text{Re} > 10\,000$                     Fully turbulent

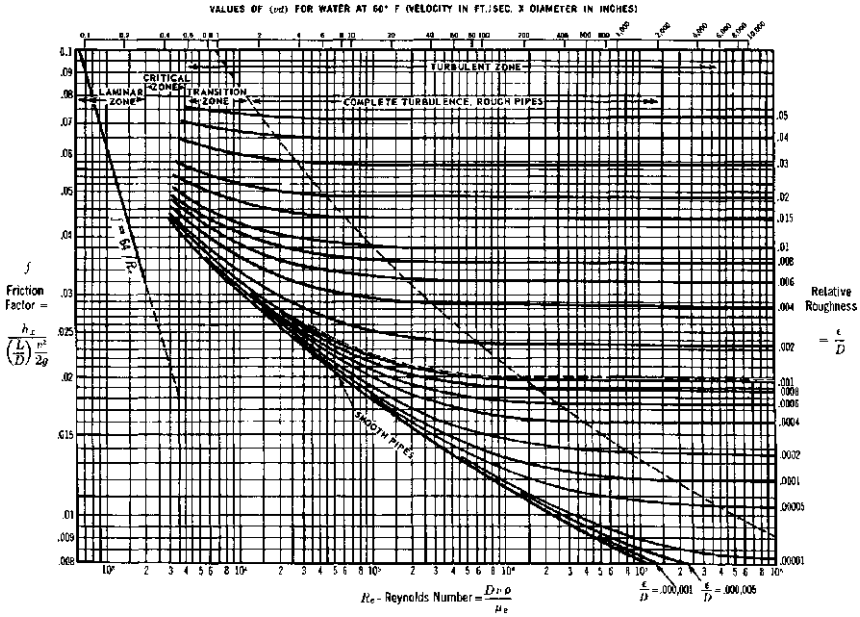


Figure 7-2 Friction factor (Darcy) versus Reynolds number.

These regimes change when inserts such as static mixers are installed, due to their shape, which changes the hydraulic diameter of the mixing channel. Vendor literature should always be consulted.

Later studies have supplemented the early work of Reynolds, providing a better understanding of the nature of turbulence, and more important for those in the process industries, how to apply the information to fluid processing. With an understanding of the fundamentals, equipment can be engineered to meet specific process requirements efficiently. Pipeline mixing is applied over the entire range of fluid flow regimes. It is an attractive processing option for bulk blending, additive/reactant mixing, multiphase dispersion and contacting, tubular reactors, and heat transfer.

In each of these applications, turbulence strongly influences the mixing process and the ability to achieve desired results. These applications are further discussed later in the chapter. Fundamentals of turbulence and the influence on mixing are developed further elsewhere in the book. For more information, see Chapter 2.

### 7-3 OVERVIEW OF PIPELINE DEVICE OPTIONS BY FLOW REGIME

There are a broad variety of method and equipment options for the continuous processing of fluids in pipelines to achieve objectives in mixing, dispersion, heat transfer, and reaction. The fluid flow regime is a main determinate for

equipment selection. Additionally, the available pressure in both the main stream and additive stream are important in the selection criteria.

Fluid flow in the pipe itself may generate adequate turbulence to accomplish simple mixing and dispersion processes. This option is often used successfully in highly turbulent flow where mixing length and time are not important.

Tee mixers, impingement jet mixers, and spray nozzles (for liquid into gas) are also often used, especially when adequate pressure energy is available or can be made available in the additive stream. These sidestream additive injection devices are sometimes used in combination with static mixers to optimize design and performance. The design and application of tee mixers are described in detail in the following section.

Static mixers are the dominant design choice for motionless pipeline mixing. They are essential in the laminar flow regime. They are well established in turbulent processes, both single and multiphase, due to their simplicity, compactness, and energy efficiency. Properly designed static mixers offer predictable performance and operate over a broad range of flow conditions with high reliability. Static mixer design options and basic design principles are described in the following sections.

In-line mechanical mixers are primarily the rotor–stator type and extruder design. They are mentioned because of their importance in pipeline mixing processes but are really beyond the scope of this chapter. They are relatively high energy devices capable of applying high shear stresses to fluids. A typical application is the dispersion of solids into a liquid matrix. High shear stress can prevent the formation of solid agglomerates, and break them up if formed in the initial stage of the process. Extruders are also used where the heat generated by mechanical shear stress is used for melting materials and for providing heat in devolatilization processes. High-speed rotor–stator mixers are very attractive where extremely short contact time is required.

Table 7-1 provides a rough method for initially selecting among the various pipeline equipment options. Subsequent detailed design is used to select the optimum device where several options are possible.

In summary:

- Energy is required for achieving the desired result.
- In laminar flow, static mixers are typically required. The energy for mixing must be made available as pressure that will be dissipated in the process. Line pressure drop will be a design criterion.
- In turbulent flow, if there are no time or length restrictions, the simple pipeline uses the minimum energy and is often the best choice for blending applications. If there are energy limitations, the energy for mixing must be supplied by either the main stream or the side stream.
- Simple mixing or blending can typically be done at low energy expenditure if time is available. With miscible components, diffusion will eventually result in a homogeneous mix. The in-line mixer device merely accelerates the process by bringing the components into more intimate contact.

**Table 7-1** Pipeline Equipment Options

Flow Regime	Pipe	Tee Mixer	Impingement Jet Mixer	Spray Nozzle	Static Mixer	Inline Mechanical
Laminar regime						
Mixing/blending					×	×
Dispersion					×	×
Heat transfer					×	×
Reaction					×	×
Plug flow					×	
Turbulent regime						
Mixing/blending	×	×	×	×	×	
Dispersion	×	×	×	×	×	
Heat transfer	×			×		
Reaction			×	×	×	×

- Dispersion involves the creation of surface area and is more energy demanding. With immiscible components the end result with time is phase separation. The in-line mixer must overcome this to create a dispersion. Significant energy input may be required, depending on the fluid properties.

### 7-3.1 Turbulent Single-Phase Flow

When the flow is highly turbulent single phase, there are many design options, including:

- Empty pipe or pipe works without any special internals
- Valves, nozzles, and orifice plates
- Tee mixers and jet mixers
- Static or motionless mixers

Special features of these design options are described below.

**7-3.1.1 Empty Pipe or Pipe Works without Special Internals.** Blending or simple mixing is achieved as a result of naturally occurring turbulent eddies in straight pipe and additional turbulence resulting from changes in flow direction, as, for example, in 90° elbows, pipe restrictions and expansions, and so on. Typically, a very long flow path length is required to achieve homogeneity, especially in large pipes and flow conduits. The main advantage is that cost is low and pressure drop increase is minimal or nothing if the pipe works already exists or is being built for fluid transport. Empty pipe is used when residence time and residence time distribution are not design issues.

**7-3.1.2 Valves, Nozzles, and Orifice Plates.** These devices provide some degree of control of the mixing process versus empty pipe. Mixing length can be



reduced, but only at the expense of increased pressure drop. They are sometimes also used in multiphase flow to create dispersions and mists, but it is difficult to predict the drop/bubble size and size distribution. The drop size distribution is typically very broad.

**7-3.1.3 Tee Mixers and Jet Mixers.** Both of these pipeline devices can be designed to achieve rapid turbulent mixing in a short length of pipe. Since the energy required to achieve mixing is in the additive or side stream, the process side pressure drop required to achieve homogeneity is very low. These designs are most workable when the additive can be supplied at a pressure significantly above the mainline pressure and can be injected into the main stream at a single point. This design option becomes less attractive when dealing with either very small or very large volumetric flows. Small additive flows necessitate small injector size and are subject to plugging. Large flows in large diameter pipe necessitate multiple injection points, increasing complexity and concerns about feed distribution to the various injection points.

**7-3.1.4 Static or Motionless Mixers.** These mixing devices are readily available and highly engineered for continuous operation. Static mixers achieve predictable mixing performance through a definable pressure drop. A high degree of homogeneity can be achieved in a very short length of pipe. The most attractive designs for single-phase flow components at Reynolds numbers greater than 10 000 are based on vortex generating principles—large scale vortex flow is initiated at a mixer blade or blades with bulk flow mixing immediately downstream. Mixers of about five pipe diameters of total length, including the built-in empty pipe sections, are capable of achieving variation coefficients below 0.05 for moderate amounts of additive. A value of 0.05 (5%) is considered thoroughly mixed in most industrial applications (see Section 7-5.2). Shorter mixing lengths are possible with mixers built of structured plate or bars and well-designed inlet injectors. These designs more aggressively direct the flow, using the increased turbulent energy to achieve mixing. Mixers of corrugated plate design are often used in large pipes and ducts, where length is limited. In many cases involving large flow conduits (both round, square, and rectangular cross-section), the mixer internals represent significantly lower installed cost versus empty conduit based on achieving equivalent mixing. Unlike jet mixers and tee mixers, most of the energy cost is with the main process steam, and higher than the empty pipe option.

### 7-3.2 Turbulent Multiphase Flow

When the flow is highly turbulent multiphase, there are only two practical design options:

- Static or motionless mixers
- Valves, nozzles, and orifice plates

**7-3.2.1 Static or Motionless Mixers.** Static mixers are well established in multiphase turbulent flow and meet industrial requirements for absorption, reaction, extraction, and heat transfer/phase change. Designs are engineered to achieve specific results at minimum cost and energy expenditure. The mixers are very compact, making them very attractive for single-stage contacting applications versus countercurrent flow options such as packed, tray, or mechanically driven towers.

Static mixers are recommended for multiphase flow applications with a continuous liquid phase and a dispersed gas or immiscible liquid phase. Turbulent shear is applied efficiently to the additive liquid or gas to create a dispersion or droplets or bubbles. The mean drop size depends on the energy expenditure. Also important is the drop or bubble size distribution. Static mixers are designed specifically for the application to create uniform drop size distributions with the interfacial surface area required for reaction or extraction. Uniform size distribution also facilitates downstream separation of the phases in some type of gravity or inertial separator. In addition to creating interfacial surface area, the static mixer performs bulk homogenization, ensuring that all flow components are distributed uniformly in the cross-section and exposed to similar levels of turbulent energy dissipation in the fluid surrounding the droplet or bubble. The required mixer pressure drop or energy dissipation depends on the amount of interfacial surface area required for mass transfer-limited and reaction rate-limited applications as well as the required residence time when reaction rate is limiting. Surface area generation varies with power input per unit mass, and consequently, there are turndown limitations that must be considered when designing static mixer processes for multiphase applications. Scale-up criteria are well established for the static mixer designs that are used in turbulent multiphase flows. This is a very important consideration since many processes are lab scale or pilot scale tested prior to commercialization.

Static mixers in multiphase applications where the gas is continuous are typically highly structured designs, providing large surface area per unit volume. Surface area is needed for absorption of gas phase components, stripping of components from the liquid, condensation, or vaporization. The properly selected mixer is a compact, highly efficient phase contactor. Turbulent flow energy is used to break up the liquid feed, achieving some equilibrium droplet size and corresponding total surface area. Flow turbulence is maintained uniformly over the pipe cross-section in individual interconnected flow channels. Liquid that wets the mixer surfaces is continuously stripped off and redispersed in the gas stream. Flow stability is maintained over a greater range of gas-liquid flows versus what would occur in an empty pipe, an important factor considering that liquid- and gas phase mass and volume flow rates could change significantly during the process as a result of phase change of all or part of the streams. As with all multiphase processes the initial drop (or bubble in the continuous liquid analog) size is an important design factor. Spray nozzles (with or without an atomizing fluid) are often used to create the initial drop size distribution utilizing additive stream energy and designing for relatively low mixer pressure drop.

**7-3.2.2 Valves, Nozzles, and Orifice Plates.** These devices are used in continuous liquid-phase processes, but less and less so as static mixing technology has evolved to dominate multiphase continuous liquid processes. In continuous liquid applications these devices are less efficient and often require a higher energy expenditure than that required by a properly designed static mixer. Also, as a result of highly concentrated energy dissipation, the drop size distribution is less controlled and typically much broader than that which can be achieved in a static mixer. Broad drop size distributions have a negative impact on equipment size and performance of downstream phase separation equipment as described above for static mixing. Valve designs do have the benefit of providing a degree of adjustment to tune the process and deal with turndown, but have proved difficult with respect to continuously achieving design performance. They are rarely used in continuous gas processes, except for two fluid nozzles, where the gas component can be used to atomize the liquid stream. Design uncertainty and potential for flow instability are typical concerns.

### 7-3.3 Laminar Flow

When the flow is laminar, either single or multiphase, there is only one design class option: static or motionless mixers. Other pipeline mixing devices described for turbulent flow are not usable for even the simplest mixing applications in the laminar regime. All rely on turbulence and cannot function at low Reynolds numbers. The only alternative technology is in-line dynamic mixers, which include extruders, rotor–stator mixers, and a variety of rotating screw devices. None of these has the benefits of simplicity and the little or no maintenance characteristic of static mixers. In-line mechanical mixers are discussed briefly later in the chapter.

**7-3.3.1 Static or Motionless Mixers.** Static mixers are proven in a broad range of laminar flow processes involving both Newtonian and shear thinning fluids. Some processes are more complicated than others. Very often, commercial installations follow laboratory or pilot scale evaluations, and success is dependent on proper scale-up. Scale-up methodology is well established for the predominant static mixer designs used in the laminar flow regime. In addition to mixing applications there is value in the use of static mixer packings to enhance laminar flow heat transfer and for creating plug flow in laminar tubular reactors.

**7-3.3.2 Blending of Fluids with Similar Viscosity.** There exist a broad range of static mixer design options for blending or distributive mixing. Selected static or motionless mixers are designed specifically for operation in laminar flow. At a minimum they operate by cutting and dividing the feed into substreams, distributing the substreams across the pipe diameter and recombining them in a continuous manner creating fluid layers. The layers are stretched, reducing layer thickness, and recombined as they are folded into each other. The number of

layers is increased exponentially as fluids flow through the mixer. Layer thickness is decreased until thickness is so small that differences in composition or temperature are indistinguishable on a macroscopic level. Final micromixing occurs by diffusion or conduction made possible in the laminar region at very small distances between layers. The mixer is designed to achieve the desired degree of homogeneity. The time required for diffusion may influence the mixer design, but in most cases it is not an issue since the time normally required to create the striations is quite significant, at least relative to what can be accomplished in short time when operating in the turbulent regime. This process of division and recombination of the process stream as fluids flow through the static mixer represents what is called *simple* or *distributive mixing*. Many static mixer designs can achieve the desired result, but there are significant differences in mixer residence time, length, and pressure drop. The optimum static mixer from the perspective of equipment design would be the one most compact and operating at the lowest pressure drop. The optimum mixer design would be the one that is the most efficient generator of fluid surface, or a high division rate device.

**7-3.3.3 Blending of Fluids with High Viscosity Ratio.** When laminar distributive mixing is complicated by viscosity differences, static mixer design options are limited. Elongational flow static mixers are required. The description of the simple distributive mixing process above applies to cases where the component fluids are not only miscible, but in addition can be infinitely divided into each other. In other words, resistance at the interface between flow streams is minimal or nonexistent. There is, however, a phenomenon in viscous laminar flow called the *miscible interface* which serves to block the mixing process. It comes into play, for example, when processing polymeric material, where there are significant differences in molecular weights and/or crystallinity, and most important, when there is a significant viscosity difference between the two components. Mixer design options are now more limited than for simple blending applications.

When there is a resistance at the interface, the static mixer must be one that operates at uniform shear stress. Additionally, when there is some degree of immiscibility and a significant difference in viscosity, elongational shear has been demonstrated to be more effective than rotational or simple shear (Grace, 1982). Miscible materials actually behave like they are immiscible when there is a large difference in viscosity, and this must be taken into account when designing the mixer. Channeling of the low viscosity component at the wall must also be prevented for a successful mixer operation. The structured X or cross-bar design has many of the characteristics necessary for these difficult applications and at this time is the only proven performer in many industrial processes.

**7-3.3.4 Liquid-Liquid and Gas-Liquid Dispersion.** When the flow is laminar and multiphase, elongational flow static mixers are required to mix and disperse additives into viscous bulk streams. The dispersion of a low viscosity immiscible additive into a viscous mainstream is a common and very difficult

static mixing application in laminar flow. As mentioned above, the uniform application of shear is required to prevent channeling of the low viscous additive. Mixing efficiency is strongly dependent on elongational flows within the mixer structure. Mixing of higher viscosity material into a lower viscosity but laminar stream is not as common but equally difficult, requiring controlled elongational flows. Shear stressing of the additive gas or liquid results in it being extended to the point where it becomes unstable and breaks to smaller size. This process continues until the droplet or bubble is reduced to a size that is stable under mixer flow conditions. In addition to creating this dispersion, the mixer must also distribute the additive phase uniformly over the pipe cross-section. The structured X or cross-bar design is at this point in time, the only significant commercially available design for this very difficult application.

**7-3.3.5 Heat Transfer Enhancement.** When the application is laminar flow with heat transfer, static mixers can enhance the heat transfer and provide a more gentle product treatment than what can be achieved in empty or unpacked tubes. Heat transfer enhancement is achieved by the use of static mixer heat transfer packings (mixing elements) in jacketed pipes and the tubes of multitube heat exchangers. The internals mix the fluid during the heat transfer process, continuously exchanging material at the wall with material at the core. Disruption of the laminar boundary layer at the wall results in increased heat transfer capability along the length of the pipe. In laminar flow this is usually the controlling resistance. Thermal stratification is continually reestablished as material flows through the heat exchanger. In addition to improving the process side heat transfer coefficient, the static mixer packing provides a more gentle thermal processing. Baking on the wall is reduced in heating, and skinning or precipitation minimized in cooling. The fluid flowing through a static mixer packed heat exchanger pipe has a more uniform thermal profile, uniform shear history, and narrower residence time distribution compared to fluid flowing through an empty (unfilled) pipe.

In addition to designs with static elements serving as inserts in pipe and tubes, there exists commercially available designs where the static mixing element itself is made out of hollow tube through which heat transfer fluid flows. The entire mixing element is an active heat transfer surface with process fluid flowing externally. Large diameter monotube designs are possible, and very attractive in many processes, especially when the process fluid viscosity increases dramatically, as in polymerization reactions and some cooling applications.

It is important to note that in most cases the value static mixing brings to heat transfer does not translate to turbulent flow, where cost and pressure drop cannot be justified by the heat transfer enhancement. Turbulent flow heat transfer processes are best handled by empty tubes and tubes with spiral wrapped cores or tubes containing twisted tapes. These other pipeline devices, though not discussed here, are nevertheless important in industry. The reader is encouraged to seek other literature if interested (see Burmeister, 1983a,b).

**7-3.3.6 Tubular Plug Flow Reaction.** When the application is a laminar flow tubular reactor, static mixing internals can provide great benefits in terms of

performance, control, and reliability. Static mixer reactor packings are specially designed to create plug flow in tubular reactors. The parabolic velocity profile of laminar flow in empty pipe is flattened as material flows through the static mixer. Reactors can be designed to achieve near-plug-flow conditions even in pipes with large diameters and short lengths. Static mixers are applied in laminar flow reactors processing both low and high viscosity materials. Although the designs are significantly different, depending on viscosity, plug flow is readily achieved in both cases. Density-driven internal recirculations are eliminated by proper mixer selection and detailed design. When a heat load exists simultaneously with a reaction, a static mixer with temperature-controlled surfaces may be required. Lab scale and/or pilot testing is most often required to establish exact conditions.

## 7-4 APPLICATIONS

Very many commercial scale applications are efficiently handled with in-line static mixing equipment. Other in-line devices are also used, but the range of applications is less broad. For the purpose of the applications discussed in this section, we are referring primarily to static mixing applications, but recognizing that in some specific cases, other in-line devices may also be employed.

### 7-4.1 Process Results

Processes carried out with in-line mixing equipment are very similar to those with stirred tanks. They include the following three main classes:

- Blending of miscible fluids or distribution
  - Same physical properties
  - Different physical properties
  - For homogeneous chemical reactions
- Area generation or dispersion
  - Liquid in a liquid
  - Gas in a liquid
  - Solid in a liquid
- Heat transfer

Additionally, there are applications where there is no analogy with stirred tanks, for example:

- Blending of gases
- Contacting of gases with liquids (scrubbing, vaporization, desuperheating)
- Creating plug flow in a pipe

**Table 7-2** Applications of Pipeline Devices

Application	Laminar Flow	Turbulent Flow	Measurement Criteria
Blending of components	×	×	Variation coefficient
Temperature or thermal homogenization	×	×	Variation coefficient
Liquid–liquid dispersion extraction/reaction	×	×	Drop size, drop size distribution, mass transfer
Gas–liquid dispersion with gas continuous	–	×	Drop size, total surface area
stripping/vaporization absorption	–	×	Mass transfer, reaction
Gas–liquid dispersion with liquid continuous			
Reaction	×	×	Bubble size, bubble size distribution
Absorption	×	×	Mass transfer, reaction
Heat transfer enhancement	×	–	Heat transfer coefficient, heat duty
Fast reaction	–	×	Mix time, variation coefficient
Slow reaction (plug flow)	×	–	Plug flow characteristics
Gas–solid fluid bed	–	×	Contacting efficiency

### 7-4.2 Pipeline Mixing Applications

Pipeline devices are used in a broad spectrum of applications (Table 7-2). Performance criteria are well established.

### 7-4.3 Applications Engineering

Static mixers differ widely in their construction and performance characteristics. Technical criteria should be used to determine the best design for each specific application. The process requirements should dictate the static mixer design or design options. There are three fundamental steps in the thought process to select the correct mixing design for a given application:

1. Determine if pipeline mixing is applicable.
2. If applicable, determine what type of pipeline equipment is best for the application.
3. Complete a detailed design after selection of equipment type.

A nine-step design procedure is recommended:

1. Clearly identify the application (blending, dispersion, heat transfer, reaction).
2. Fully define process flow conditions (stream flow rates, densities, viscosity, etc.).

3. Identify constraints (e.g., space limitation, available pressure drop).
4. Specify desired process results and measurement criteria.
5. Pick candidate designs (several will probably achieve the desired process result).
6. Identify secondary requirements.
7. Evaluate candidates on secondary requirements (includes cost, length, etc.).
8. Select an optimum design.
9. Design the mixer.

Note that the mixer is chosen on secondary requirements. All candidate mixers must achieve the process result. Thus “better” or optimum design will be based on secondary considerations. Typical secondary considerations are cost, length, pressure drop, delivery, and past experience.

Design of the static mixer is possible only after a thorough specification of the application. Numerous product brochures and technical publications are available to assist the designer. Additionally, the major static mixer manufacturers offer application and design assistance as part of their proposal process. Secondary, but also important factors that need to be considered are startup, turndown, upset conditions, and mechanical requirements. Most static mixer designs are available in a broad selection of materials of construction, both metals and plastics, to meet plant/process requirements.

Many applications have become routine, and equipment can be sized following well-established design procedures. However, as the technology evolves, so does interest in more difficult applications, many of which involve large commercial scale production rates, complicated chemistry and physical interaction between components, unique products not already characterized at flow conditions, and continuous tubular systems involving simultaneous mixing, heat transfer, and reaction. These should be laboratory and/or pilot tested prior to attempting full scale design. Design fundamentals and the ability to scale-up a design then become very important. Development of new static mixing processes should start with analysis of the full scale system. After final requirements are identified, laboratory and pilot scale mixers can be selected based on their ability to perform and their suitability for scale-up.

It is interesting to note that it is sometimes advantageous to change the mixing design along its length to optimize results at any given point in the mixing process. For example:

- Follow a high intensity jet mixer, utilizing side-stream energy and operating at high local turbulence, with a low pressure drop static mixer to distribute or bulk mix the mixer products throughout the pipe cross-section.
- Provide a length of empty pipe downstream of a vortex-generating device to achieve bulk mixing.
- Change pipe diameter to maintain turbulence as volumetric rate changes, as, for example, during gas absorption (rate decreasing) or conversely during



vaporization of a liquid or gas being generated as a result of a reaction (rate increasing).

- Arrange mixers of different design (diameter) in parallel to handle extreme process changes and turndown.
- Change mixer design or diameter to account for a change in fluid viscosity, as, for example, during cooling of a viscous product, or reactions building molecular weight, both of which lead to increasing viscosity and reduction in turbulence. This is analogous to polymerization reactions conducted in a series of agitated tanks, each designed with an agitator best suited for process viscosity, mixing, and heat transfer requirements.

#### 7-4.4 Sample of Industrial Applications

##### *Chemicals*

- Mixing miscible/dispersing immiscible reactants
- Dissolving gases (e.g., chlorination processes)
- Providing plug flow and controlled-reaction conditions in tubular reactors with low or high viscous fluids
- Dispersing liquids in extraction and washing processes
- Mixing gases in front of catalytic reactors (e.g., the production of styrene, nitric acid, maleic anhydride)
- Vaporizing liquids in front of oxidation reactors (e.g., xylene in phthalic anhydride plants)
- Co-current scrubbing acid process gas components
- Homogenizing process and product streams for representative sampling
- Controlled heating and cooling of slurries in catalyst production
- Neutralizing or pH adjustment/control of process streams with caustic or acid

##### *Cosmetics and Detergents*

- Saponifying greases with caustic soda
- Sulfonating fatty alcohols with oleum
- Mixing components of toothpaste, lotions, shampoo, soaps, or detergents
- Diluting surfactants

##### *Energy*

- Mixing blast furnace and coke oven gas
- Reheating flue gas in desulfurization plants
- Blending emulsifier for water into fuel, dispersing/emulsifying water, and fuel
- Blending fuel gases with air before combustion

*Environmental Protection*

- Scrubbing H<sub>2</sub>S from exhaust gas with caustic
- Oxidizing sulfite-laden scrubber blow over with air
- Vaporizing ammonia solution and mixing it with exhaust gas before the catalyst bed in a SCR DeNox installation

*Foods*

- Dissolving CO<sub>2</sub> into beer, fruit juice, or wine
- Heating coffee extract before spray drying of flash evaporation
- Heating and cooling chocolate mixtures
- Heating and cooling starch slurries under plug flow conditions
- Mixing enzymes and chemicals into starch suspensions
- Diluting concentrated juices and admixing flavorings
- Mixing fruits and flavors into yogurt and ice cream
- Diluting molasses and sugars
- Mixing color and flavor into pet food

*Natural Gas*

- Scrubbing H<sub>2</sub>S from natural gas with caustic or specialty chemicals
- Dehydrating natural gas with glycols
- Adjusting the Btu content of natural gas with propane
- Cooling natural gas in the compressor loop of LNG terminals by injection and vaporizing LNG

*Polymers, Plastics, and Textile Fibers*

- Mixing additives, catalyst, and inhibitors into polymer melts and solutions
- Providing plug flow in polymerization reactors (e.g., polystyrene, PA6, silicone, and many others)
- Dispersing a low viscosity stripping agent into polymer solutions prior to devolatilization (e.g., water into polystyrene)
- Rapid uniform heating of polymers prior to flash devolatilization
- Mixing additives (e.g., mineral oil, pigments, ultraviolet stabilizer, antioxidants) into polymer prior to pelletization)
- Homogenization of temperature and colorants in polymer melts in extruder and injection molding machines
- Cooling polymer melts before processing and removing heat of polymerization

*Petrochemicals*

- Chlorinating hydrocarbons (e.g., ethylene to EDC)

- Mixing ethyl benzene with stream before the first dehydrogenation reactor in styrene plants
- Scrubbing acid components with caustic from exhaust gas during catalyst regeneration

### *Pulp and Paper*

- Mixing bleaching chemicals with pulp stock
- Blending stocks and diluting stock with water for consistency control
- Mixing retention agents in front of the head box
- Admixing caustic or acid for pH control
- Admixing chlorine and chlorine dioxide to stocks for bleaching
- Steam injection for heating

### *Refining*

- Homogenizing crude for representative BS&W measurement, custody transfer
- Desalting crude with water
- Dispersing sulfuric acid or HF in hydrocarbons in alkylation processes
- Neutralizing/washing hydrocarbon streams with caustic and water
- Sweetening kerosene and gasoline with caustic
- Blending different hydrocarbon streams and additives into gasoline
- Scrubbing acidic components with water from hydrogen gas
- Establishing vapor–liquid equilibrium in hydrocarbon streams
- Adjusting the viscosity of heavy oil with gas oil and other additives
- Contacting steam with catalyst in the FCC catalyst stripper

### *Water and Wastewater*

- Aerating water (e.g., for improvement of oxygen level or oxidation of iron and/or manganese)
- Neutralization, adjusting pH/conditioning of water with acid, caustic, lime solution, or by dissolving CO<sub>2</sub>
- Diluting flocculants (e.g., polyacrylamide)
- Mixing flocculants such as FeCl<sub>3</sub> or Al<sub>2</sub>(SO<sub>4</sub>)<sub>3</sub> into water, wastewater, or sludge
- Stripping excess CO<sub>2</sub> with air for deacidification
- Dissolving ozone

## **7-5 BLENDING AND RADIAL MIXING IN PIPELINE FLOW**

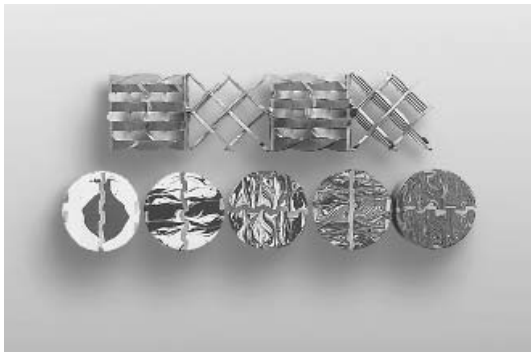
Blending in a pipeline can be radial or axial. The best designs create a high degree of plug flow, achieving radial mixing while minimizing backmixing. This

is an important characteristic of in-line devices, especially static mixers. Back-mixing, however, often occurs at the point of additive injection, something that must be considered in the overall system design. The amount of mixing that can be achieved in any given pipeline mixer and energy expenditure is strongly dependent on the flow regime, laminar or turbulent. If a high degree of back-mixing is required, as, for example, to even out time fluctuations in the feed, an agitated tank may be a better design option.

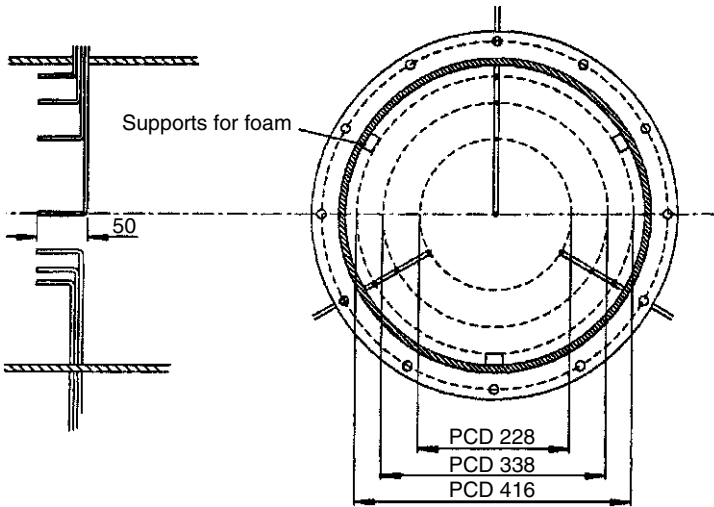
### 7-5.1 Definition of Desired Process Result

The process result of heat transfer is a heat transfer coefficient. For dispersion it is a drop or particle size and size distribution. For blending in tanks it is blend time to achieve a certain degree of mixing. The equivalent for mixing in pipeline flow is not as clear. Alloca and Streiff (1980) proposed using a radial coefficient of variation, and this concept is now widely accepted. Since it is unique in the process industries to pipeline flow, it merits some extended discussion.

**7-5.1.1 Coefficient of Variation as a Measure of Homogeneity.** Consider the cross-section of a pipe to which a small amount of material has been added. Initially, it is completely segregated into two areas, one occupied by each fluid. As mixing occurs, the areas intermingle. Figure 7-3, showing the mixing of two fluids in laminar flow in a motionless mixer (type SMX), presents the process graphically. How can the differences in this set of pictures be described? If we now superimpose a grid of squares over the cross-section, we can describe the process by estimating how much of each color is in each individual square. The overall average concentration will stay the same, but the individual boxes will start very segregated and approach the average with continued mixing length. Alternatively, we could sample at many points in the cross-section at axial positions downstream of the mixer and use these point values as a measure of segregation. Of course, the sampler should not interfere with or promote



**Figure 7-3** Homogeneity with laminar mixing of two fluids (shown with an SMX mixer). (Courtesy of Koch-Glitsch, LP.)



**Figure 7-4** Sampling array arrangement: 10 points. (Courtesy of BHRG.)

the mixing process. This can be done with temperature probes or by extracting individual samples at points located in the cross-section. Other methods can be worked out. The results are a set of numbers at defined spacing which with time approach the average. Given this set of numbers, statistics can be used to produce a measure of uniformity or mixedness. Table 7-3 gives some typical experimental data obtained by BHRG for a mixer study for DuPont. A mixture of air and air with  $\text{CO}_2$  was used for the two streams. Ten points were sampled using a gas chromatograph. The sampling array is shown in Figure 7-4. Measurements were made with time and with position. The time variation was due to fluctuations in the feed system. This sets the minimum degree of uniformity that can be obtained. From the 10 discrete readings, an average and standard deviation are calculated. The average should not change, of course. This is just an internal check of the data quality. The standard deviation is normalized by dividing it by the average, giving the function called the *coefficient of variation* ( $\text{CoV} = \text{standard deviation of concentration measurements} / \text{mean concentration}$ ). This is a useful concept, as coefficient of variation (often reported as a percent) is easy for laypeople to comprehend. This is often also called the *intensity of mixing* or *degree of segregation*. From statistics for a normal or Gaussian distribution of data, two-thirds of the data will lie within  $\pm 1$  standard deviation, 95% within  $\pm 2$  standard deviations and  $>99\%$  within  $\pm 3$  standard deviations. Thus, in practice, one can say that  $\pm 2$  or 2.5 CoV is the spread of all the data. One can talk about everything within 10, 5, or 1% of the average. Often, the process will indicate what is an acceptable coefficient of variation. For example, in a typical industrial mixing process, an additive might be considered well mixed at 5% CoV, while in a more critical application such as the addition of color to an extruded sheet, product might require 0.5% CoV in order to escape the strong discriminating



ability of the human eye. Note that the final CoV is usually independent of the amount to be mixed. The mixer length required to achieve a given CoV depends on the amount to be mixed because of the initial state of unmixedness. It is of interest to look at the original state of mixing in similar terms. The coefficient of variation for an unmixed sample ( $\text{CoV}_0$ ) is given by statistical theory to be based on the unmixed volume fraction,  $C_v$ :

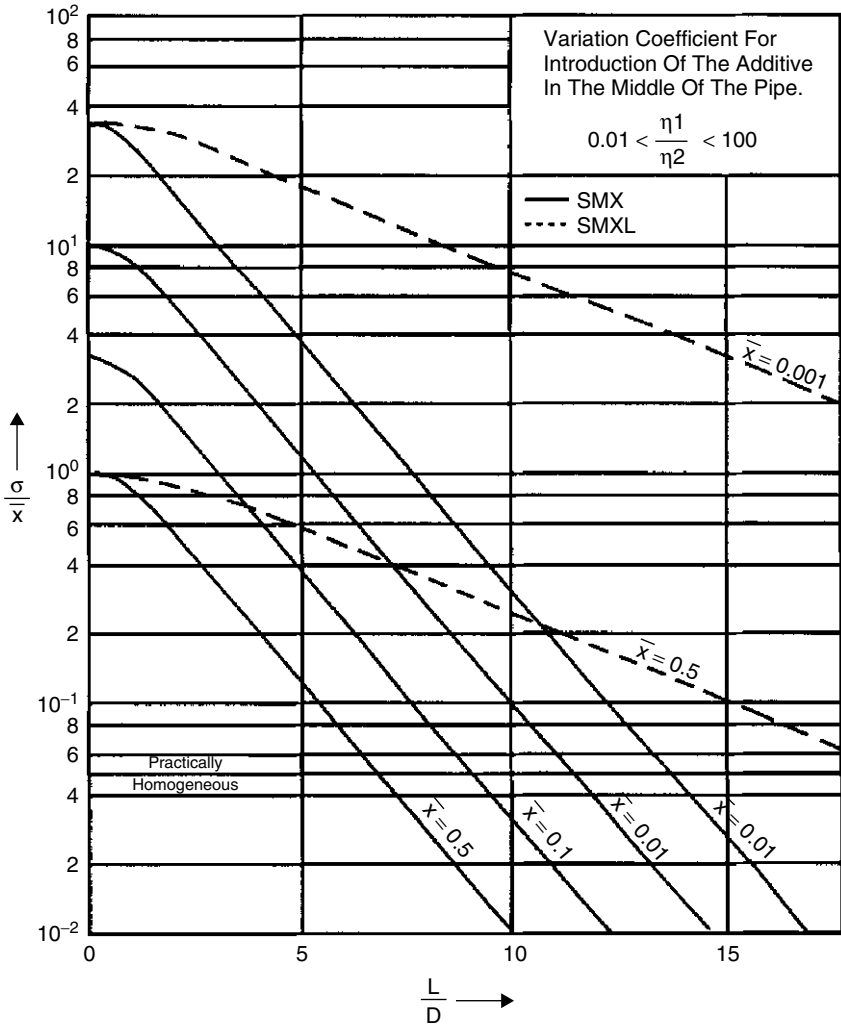
$$\text{CoV}_0 = \left( \frac{1 - C_v}{C_v} \right)^{0.5} \quad (7-6)$$

Thus, the initial degree of “unmixedness,”  $\text{CoV}_0$ , depends on how much needs to be mixed. The following table is illustrative.

Initial Additive (Volume Fraction), $C_v$	$\text{CoV}_0$
0.5	1.0
0.1	3
0.01	10
0.001	33

Thus, the smaller the amount added, the greater the initial CoV or state of unmixedness. We define a mixing task as reducing from an initial coefficient of variation ( $\text{CoV}_0$ ) to a final chosen CoV which is independent of the initial coefficient of variation ( $\text{CoV}_0$ ). It is believed that the performance of a motionless mixer is set by this mixing task and is independent of the magnitude of the initial or final values. Some recent unpublished data challenge this, but until better data are obtained, we will presume that the mixer performs as a reducer of variance. This is the assumption of most vendors.

This, of course, neglects molecular diffusion effects, which tend to shorten the process. The motionless mixer can then be thought of as a transfer function which reduces the CoV from an initial value to a lower final value. Thus, CoV reduction as a function of length is a measure of quality of mixing of a motionless mixer. The smaller the volumetric flow of additive (mean concentration, or % addition), the longer the mixer. Variation coefficient versus mixer length for two mixer designs (SMX and SMXL) operating in laminar flow at 0.1, 1, 10, and 50% additive rates is shown in Figure 7-5. This is also shown in turbulent flow for a KVM vortex mixer ( $L/D = 5$ ) at 0.1, 1, and 10% additive rates in Figure 7-6 and for the SMV mixer at 0.1, 1, 10, and 50% additive rates in Figure 7-7. Note by comparison with Figure 7-5 that as one would expect, in turbulent flow homogeneity is achieved much more quickly than in laminar flow and that there is additional mixing achieved after the mixer due to the turbulence in the tailpipe section.

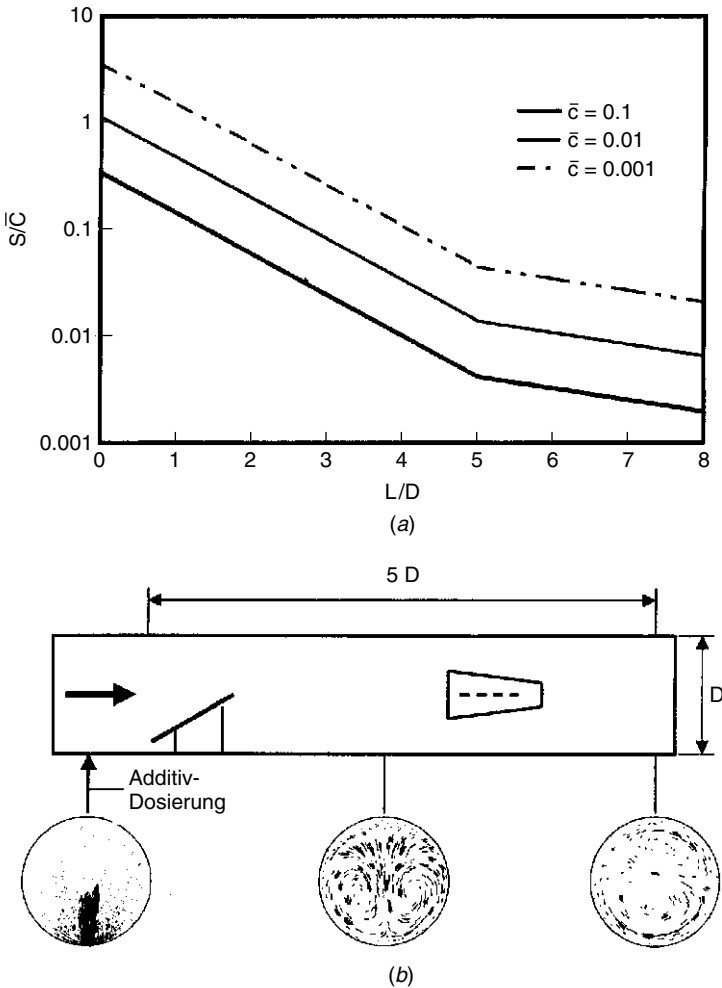


**Figure 7-5** Homogeneity expressed as variation coefficient versus mixer length for SMX and SMXL static mixers operating in laminar flow. (From Schneider, 1981.)

**7-5.1.2 Other Characterization Measures.** Many other mixing measures have been developed. Several of them have been taken from work on mixing of particulate solids. However, in miscible liquid blending and gas mixing there will always be diffusion at the lowest and final scales to finish off the process. There is also no mechanism for de-mixing as can occur with solids due to particle mass differences.

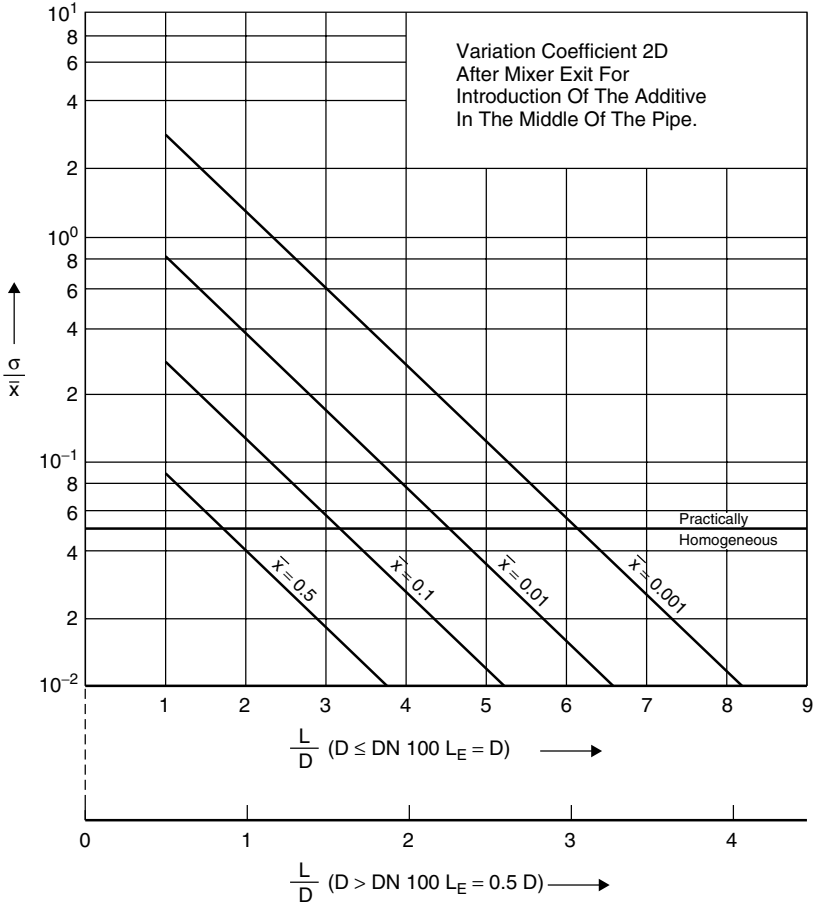
In recent years the advances in computational fluid mechanics (CFD, computational fluid dynamics) and its application to mixing (CFM, computational





**Figure 7-6** (a) Homogeneity expressed as variation coefficient versus mixer length for a vortex-type static mixer type (KVM at  $L/D = 5$ ) in turbulent flow. (b) KVM mixer layout and computational velocity vectors shown at three points. (From Streiff et al., 1999.)

fluid mixing) have allowed more detailed analysis to be performed of the state of mixing. Early computational studies showed more rapid mixing than experimental studies. This was due to a phenomena called *numerical diffusion*, where with coarse grids numerical rounding errors cause a smoothing of concentration gradients. With finer grids and the tracking of large number of massless particle tracking, the accuracy of the calculation matches well with the experimental measurements. This has been demonstrated in laminar flow for both the Koch SMX and the Kenics KMS motionless mixers by Zalc et al. (2003) and for turbulent mixing with Kenics HEVs by Bakker and LaRoche (1993).



**Figure 7-7** Homogeneity expressed as a variation coefficient versus mixer length for an SMV static mixer operating in turbulent flow. (From Schneider, 1981.)

Experimentally, Etchells et al. (1995) calculated both CoV and Danckwerts mixing length and found similar relations in laminar flow. The Danckwerts length has the dimension of distance and may be thought of as laminae or diffusional thickness in laminar flow.

In the previous examples, color and composition were used; however, concentration or temperature can be used as well. In some cases not enough data exist for calculation of a coefficient of variation. In that case, another measure, such as spread, can be used. This is most useful in dealing with temperature problems, where the spread can be estimated coming in and the desired spread specified going out. Note that the mixing concepts above are independent of flow regime. The concept of coefficient of variation to describe mixing in pipeline devices is valid in both the laminar and turbulent flow regime and is used routinely in

process design to describe or specify homogeneity for blending applications. In Chapter 2, several other concepts are discussed in detail, using the motionless mixer as an example.

## 7-5.2 Importance of Physical Properties

The two key properties in single-phase flow are the fluid density and the viscosity. The density is quite straightforward; it is the mass per unit volume. In turbulent flow, pressure drop is directly proportional to density, so that the accuracy of the density is the accuracy of the pressure drop prediction. It is easy to get better than 1% accuracy on such values. Viscosity, on the other hand, is a more complex measurement. Low viscosity systems usually run in turbulent flow, where the viscosity has little or no effect on mixing or pressure drop. For low viscosity material the prime use of the viscosity is in calculating a Reynolds number to determine if the flow is laminar or turbulent. If turbulent, little accuracy is needed. An error in viscosity of a factor of 2 will have negligible effect. In laminar flow, however, the viscosity becomes all important and pressure drop is directly proportional to it, so that an accuracy of 10% or less is often required. For laminar processing a complete relation of stress versus strain or shear rate versus shear stress is required. See Chapter 4 for the means and type of data required.

**7-5.2.1 Laminar Flow Regime.** In laminar pipeline flow the velocity vectors are parallel and there is no radial mixing. Because of the parabolic velocity distribution the velocity across the pipe is nonuniform. This results in a residence time distribution that is not plug flow (see Figure 7-33). The presence of this residence time distribution does not indicate backmixing, but rather, de-mixing, in that it works against radial mixing. The age distribution of fluid elements is being increased and this may be undesirable. For all practical purposes there is no radial mixing in laminar pipeline flow. Pipeline laminar flow often introduces gradients in age and temperature that must be removed by mixing. This mixing task is often referred to as *simple blending* or *low homogenization* since there are no additives being introduced to the bulk stream. Contents of the flow stream are merely being blended with themselves to eliminate gradients.

Static or motionless mixers are the only effective pipeline devices in the laminar flow regime. Flow inverters are related devices since they take annular fluid flowing at the pipe wall and interchange it with material at the center of the pipe. The purpose of this flow inversion is to equalize residence time to prevent degradation effects in sensitive processes such as melt fiber spinning. Static mixers are good flow inverters and, in addition, accomplish mixing. Flow inverters are, however, not good mixers—in principle, a second ideal flow inverter device in series with another merely undoes any mixing achieved by the first.

**7-5.2.2 Turbulent Flow Regime.** With turbulent flow there is mass interchange in both the radial and axial directions, due to turbulent eddies. Radial mixing has been summarized extensively by Gray (1986). The study by Ger and

Holley (1976) is most useful. They looked at single-point coaxial addition with equal main and injection velocities. They determined a radial diffusion coefficient in terms of a friction factor and an average velocity. They showed how an initial centerline injection spread out to the walls of the pipe as the fluid traveled along. Because of the linear relation between time and diffusivity, an increase in velocity increases turbulent diffusivity but reduces contact time so that the critical parameter become the length/diameter ratio ( $L/D$ ) and friction factor, which was relatively insensitive to velocity. Thus, the empirical observation by many early workers that radial mixing in an empty pipe took between 50 and 100 diameters was verified by this work. In addition, the work showed that side injection took about twice as long as centerline, due to the longer diffusion path. Ger and Holleys's relation is

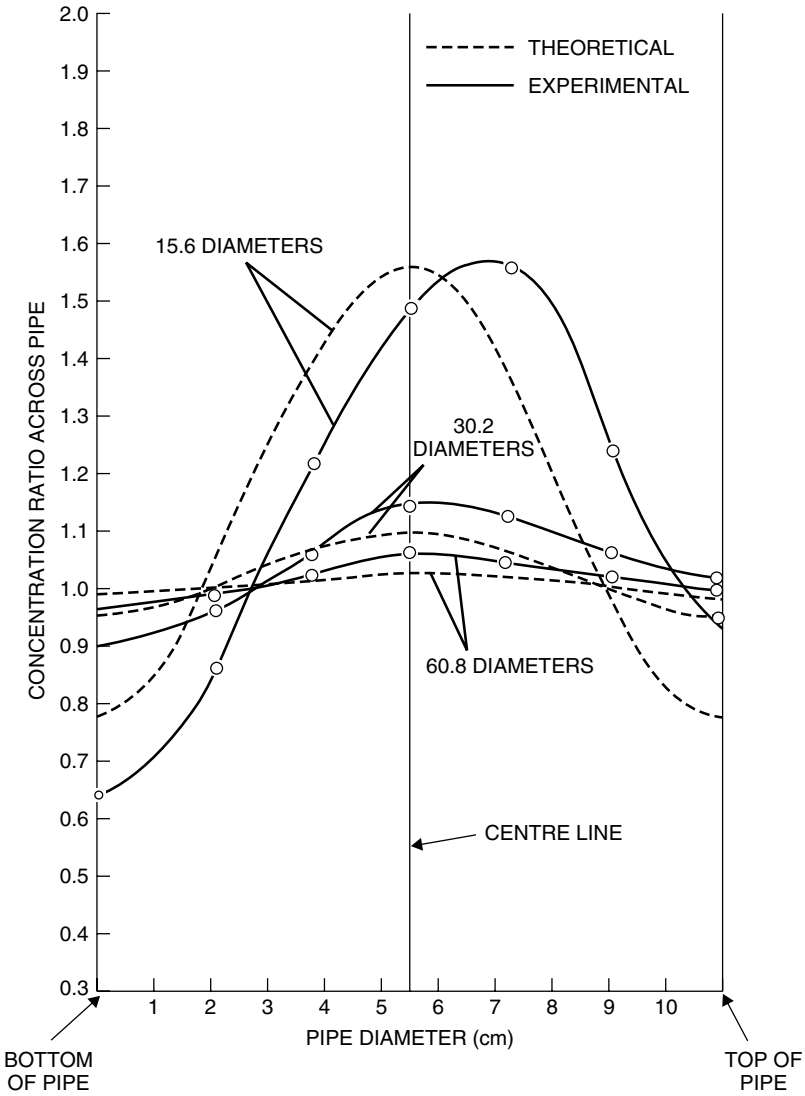
$$\frac{L}{D_{\text{mix}}} = 20.5 \text{Re}^{0.10} \left( \frac{f_s}{f} \right)^{0.5} \log(2.40 \text{CoV}) \quad (7-7)$$

Unfortunately, Ger and Holley do not include the initial degree of unmixedness ( $\text{CoV}_0$ ) in their correlation. It is interesting to note that the length required for mixing increases with increasing Reynolds number. See Figure 7-8 for measurements across a 4 in. diameter pipe at various distances downstream of a centerline injection of radioactive tracer into water at 1% concentration with a Reynolds number of 77 000. The publications also present results achieved with additive injection at the wall.

There is also axial dispersion. Again this is not mixing but a mechanism that introduces a residence time distribution. If a pulse is added to a turbulent pipeline, it will gradually lengthen with time. The best discussion of this is in Levenspiel's book (1967), where the work of Levenspiel and Bischoff is discussed. Mixing time, an important concept for reactive mixing, is given by

$$t_{\text{mix}} = 50 \frac{D}{V} \quad (7-8)$$

Thus, small pipes had short mixing times, and larger pipes take proportionally longer. This could present a scale-up problem from semiworks to full scale plant. In many plant cases, involving large pipe, the mixing length and time are too long. In many control and measurement situations, shorter time and distances are needed or used. This requirement to reduce mixing length and time has led to the concept of the multicoaxial mixer. There has been much academic work (Toor, 1975) on using parallel arrays of tubes for coaxial injection. This adjusts the initial scale of turbulence to a lower value (higher turbulence) than an empty pipe. While of continuing academic interest, they are not commonly used in industry; they are too complex and relatively ineffective. Tee mixers and static mixers are the two designs routinely employed to accomplish distributive mixing in the turbulent flow regime. Occasionally, they are used together. They are described in detail in subsequent sections of this chapter.



**Figure 7-8** Variation in concentration across a round pipe at different lengths downstream of a central injection with  $RE = 77\ 000$ . (From Clayton, 1979.)

### 7-6 TEE MIXERS

One of the two most popular approaches to pipeline mixing in turbulent flow involves the use of side injection tees. The other is the use of motionless static mixers (which are discussed later). The most complete work on side tees is by Forney and Lee (1982). They and others have found that the momentum of

the side stream must be high enough to mix across the pipe with the bulk stream for shortest mixing length. When the momentum is low, the side stream will be deflected and become a sidewall injection, and the mixing length will be about 50 to 100 diameters. Therefore, much of the work has been centered around finding the optimum relation between main and side flow rate and velocity or diameter ratios. For low side stream momentum, the side stream is plastered along the near wall and mixing rates are similar to that with a pipeline. For too high a side-stream momentum, the incoming jet plasters against the far wall and backmixing occurs, which can be undesirable in the case of reactions. Figure 7-9 provides a diagram of the flow variables and shows the turbulent jet created by the properly designed tee. Tee mixing is the prime technology used in many reaction injection molding (RIM) systems for making polyurethanes. Both side tee and opposed tees are used. Angled jets that cause swirl reduce the amount of mixing and lengthen mixing length. The following rules from Forney and Lee (1982) give a set of optimum relations:

$$\frac{v}{V} = \frac{1.0}{(d/D)^{0.5}} \quad (7-9)$$

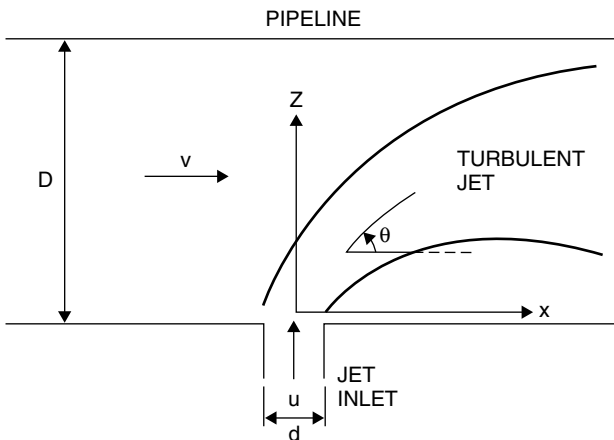
or

$$\frac{q}{Q} = \left(\frac{d}{D}\right)^{1.5} \quad (7-10)$$

since

$$\frac{q}{Q} = \frac{v}{V} \left(\frac{d}{D}\right)^2 \quad (7-11)$$

Slight adjustments would be required to account for differences in densities between the two streams.



**Figure 7-9** Turbulent jet created by Tee mixer. (Modified from Forney, 1986.)

The energy for mixing comes mainly from the higher flow velocity in the side tee. Because there is a higher flow velocity, there must be a higher pressure to drive the side stream. This pressure loss is typically defined in terms of velocity head loss and depends on the design of the nozzle, which sets the head loss coefficient.

$$\Delta P = \frac{Kv^2}{2} \quad (7-12)$$

where  $\Delta P$  is the side-stream pressure loss,  $K$  the head loss coefficient, and  $v$  the side-stream velocity.  $K$  can vary between 1.0 for a very smooth inlet to 2.5 for an orifice, with 1.5 being very typical.

For many practical situations the side stream has to come in between two and five times the velocity of the main stream. As the volumetric amount of side stream becomes smaller, this ratio grows higher, the inlet hole gets very small, and the pressure drop gets very large. Tee mixers become less attractive for adding a small amount of one material into a large amount of a main stream.

Multijets have been studied, and four arranged circumferentially around the pipe wall is somewhat better than one. For an optimum design, the quality of mixing in terms of coefficient of variation is given by Forney et al. (2001) as

$$CoV^2 \left( \frac{x}{D} \right)^E = \frac{0.32}{B^{0.86}} \quad (7-13)$$

where

$$B = n^2 R^2 \left( \frac{d}{D} \right)^2 \quad (7-14)$$

Here  $n$  is the number of jets,  $R$  the velocity ratio (jet over main),  $d$  the side-stream diameter, and  $D$  the main stream diameter,  $x$  is the tailpipe length up to  $5D$  and  $E$  is an empirically determined constant depending on  $B$  and geometry:

$$E = 1.33 \quad \text{for } B < 0.7 \text{ and } n = 1$$

$$E = 1/33 + 0.95 \ln \frac{B}{0.7} \quad \text{for } B > 0.7 \text{ and } n = 1$$

$$E = 1.97 \quad \text{for } B < 2.0 \text{ and } n = 4$$

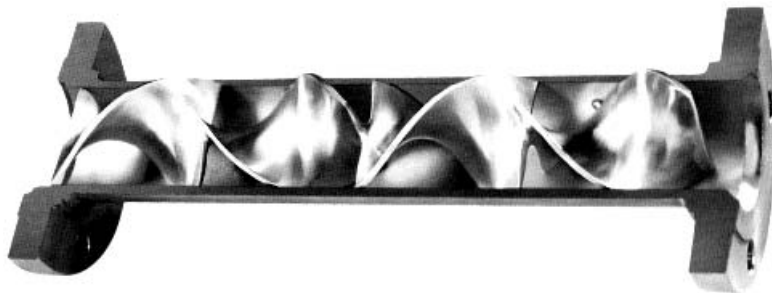
$$E = 1.97 + 0.95 \ln \left( \frac{B}{2} \right)^{2.0} \quad \text{for } B > 2 \text{ and } n = 4$$

Equations (7-13) and (7-14) is applicable only to a mixing length of five pipe diameters for systems turbulent in the tailpipe. Equations (7-13) and (7-14) have been developed for low viscosity fluids. As viscosity increases, the flow approaches transitional. Due to the high energy dissipation in the tee, the flow can be turbulent even at Reynolds numbers down to about 1000, depending on design and flows. However, once outside the mixer, the flow rapidly becomes laminar in the tailpipe. Equations (7-13) and (7-14) should be used with great care with such systems.

## 7-7 STATIC OR MOTIONLESS MIXING EQUIPMENT

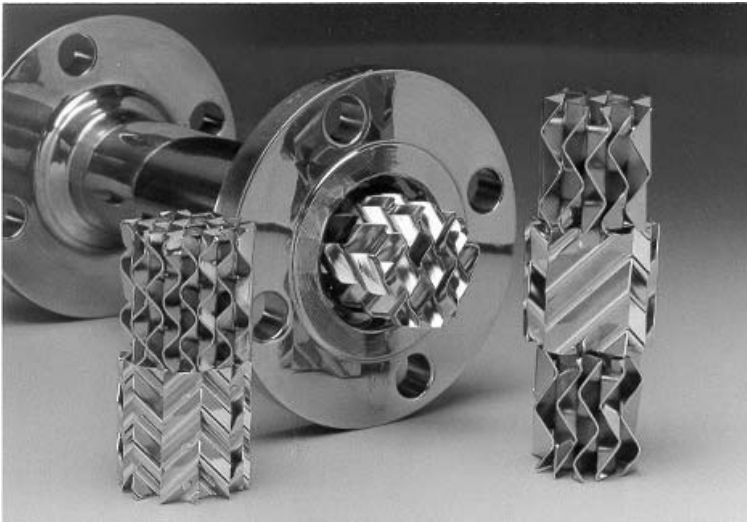
In the early 1950s a number of devices were developed in industry to handle thermal nonhomogeneous regions in polymer piping (transfer lines). Molten polymers usually are in laminar flow and have no radial exchange when flowing. Thus temperature gradients can form and be propagated. Heating and cooling of these materials through the wall is very difficult because of the residence time distribution associated with the laminar velocity profile. Material at the center moves much faster than the material at the wall, and in addition, has less contact time because of the poor conductive heat transfer at the center. These devices would reduce radial thermal gradients that occur in polymer processing. They were called *thermal homogenizers* and *flow inverters*. One of the first commercial units was the Kenics device. In the Kenics, a set of twisted elements with left- and right-hand twists caused the material to move from the wall to the center and from the center to the wall (see Visual Mixing CD). After traveling through a number of these elements, the fluid is homogenized with respect to age, composition, and temperature. These devices were called *motionless mixers* or *static mixers* because the mixer did not move, although the liquid did. The term *static mixer* was originally copyrighted by Kenics Corporation, but the term is now commonly used for all such in-line motionless mixers.

Over the years a large number of companies have produced motionless mixers all based on the principle of moving the streams radially by a series of metal baffles. These baffles may consist of twists of metal, corrugated sheets, parallel bars, small-diameter passages, or tabs sticking out from the wall. They are essentially plug flow devices with some small degree of backmixing, depending on the exact design. Two common types are the twisted-ribbon mixer (Kenics KMS; see Figure 7-10) and the structured-packing mixer, one of which makes use of layers of crisscrossed corrugations (Koch-Sulzer SMV; see Figure 7-11). Another structured packing static mixer is the overlapping lattice type (Koch-Sulzer SMX and SMXL; see Figure 7-12). The lattice members in the SMX are all oriented at  $45^\circ$  to the direction of flow and at  $30^\circ$  for the SMXL. A simplification of the lattice type which generates a mixing flow somewhat similar to the twisted ribbon is made of crossed elliptical plates whose flat surfaces are at  $45^\circ$  to the



**Figure 7-10** Spiral static mixer. (Courtesy of Chemineer, Inc.)





**Figure 7-11** Corrugated plate static mixer (SMV). (Courtesy of Koch-Glitsch, LP.)

direction of flow (Koch SMXL-B). A version of this has triangular plates connecting the straight sides of the ellipses, making each element resemble a crude twisted ribbon. Another incorporates a flat to divide the ellipse at the centerline (Komax mixer). An example of a radial mixer with tabs extending from the pipe wall is the Koch-Sulzer SMF, a low intensity mixer with wide-open structure used in highly plugging service.

A recent static mixer innovation for application exclusively in highly turbulent flow is the use of small tabs projecting from the wall of the pipe into the core region of a turbulent flow (Kenics HEV; see Visual Mixing CD). Another design (Koch KVM; Figure 7-13) utilizes a single larger tab mounted off the tube wall to create large counterrotating vortices for mixing. The industry found a number of years later that even in turbulent flow where there is radial turbulent mixing, this mixing can be enhanced by using motionless mixers. The mechanism was different in detail, but the effect was the same. Some of the improved radial mixing came from increased radial turbulent diffusion. In the Ger and Holley formulation for empty pipe there is a friction factor. In motionless mixers the friction factor is many times larger than for empty pipe (i.e., the pressure drop is higher). This in itself would increase mixing and reduce mixing length. Also, there is in some motionless mixers a bulk radial flow. Etchells and Short (1988) took some limited data on SMV motionless mixers and showed that the improved mixing rate over an empty pipe was due almost entirely to the increased friction factor. Subsequent data on the HEV, however, do not fit that model. It is now believed that only a portion of the pressure drop energy expended goes into radial mixing and that the rest is lost in skin friction. Thus, the newer motionless mixers for application in turbulent flow rely on vortex generation away from surfaces to mix and take less pressure drop to get equivalent blending results.

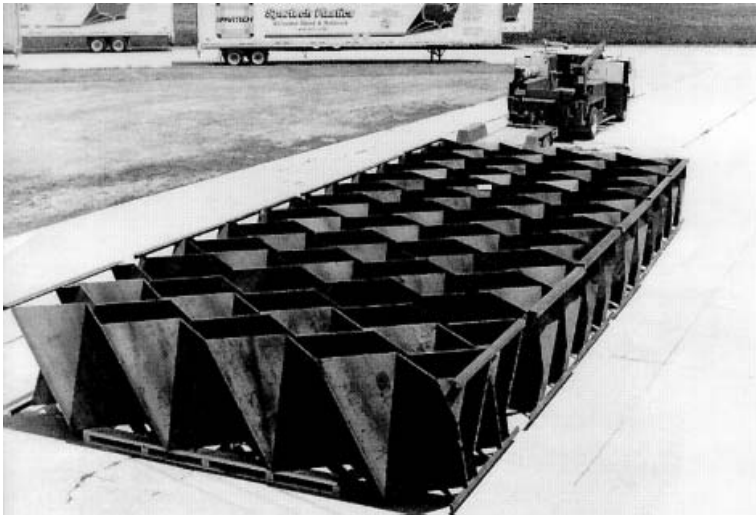


**Figure 7-12** SMX static mixer shown with two mixing elements. (Courtesy of Koch-Glitsch, LP.)

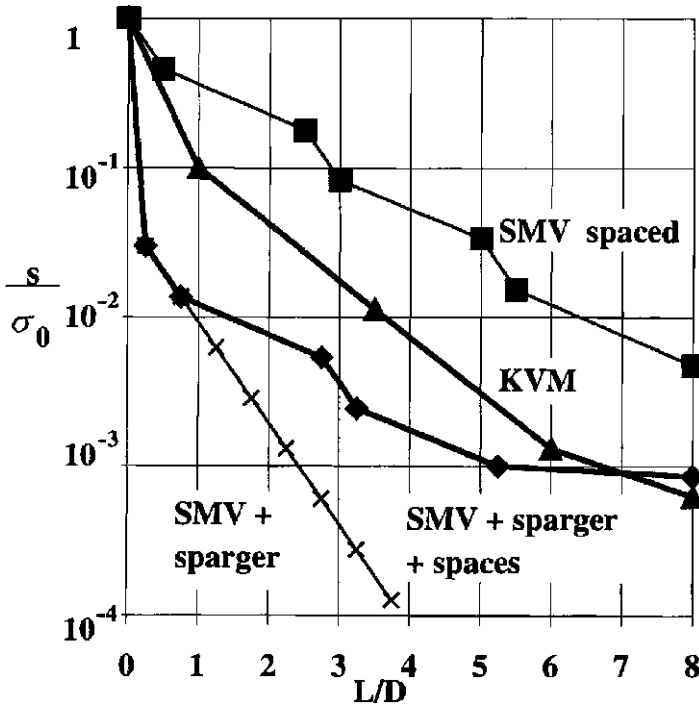
Plate-type mixers (SMV) are, however, still very attractive design options for turbulent flow applications in large diameter ducts and pipe where mixing length is limited. The method for introduction of the additive stream becomes a very important part of design optimization. An example of this is found in the selective catalytic reduction process for the removal of nitrogen oxides from combustion flue gas (DeNox), where a small amount of ammonia is added and mixed with the flue gas prior to the catalyst bed. A large SMV static mixer element for this application is shown in Figure 7-14. A high degree of mixing is achieved in only two pipe diameters using sparger designs that introduce the additive to each mixing cell of an SMV mixing element (Fleischli and Streiff, 1995). Pressure drop is low since some of the mixing is accomplished downstream of the individual mixing elements by utilizing the swirl flow, which is induced in the wake of the mixer hardware. Figure 7-15 compares the mixing achieved using a plate mixer (SMV) with that of a vortex design (KVM). Note that spacers are included between the mixing elements to take advantage of mixing in the tailpipe



**Figure 7-13** KVM vortex static mixer shown in flanged housing. (Courtesy of Koch-Glitsch, LP.)



**Figure 7-14** Large duct static mixer (SMV) for application in mixing ammonia with hot flue gas selective catalytic reduction DeNox process. (Courtesy of Koch-Glitsch, LP.)



**Figure 7-15** Homogeneity for plate (SMV) and vortex (KVM) static mixers in turbulent flow at additive concentration = 0.01 (1%). (Modified from Streiff et al., 1999.)

downstream of each element. Also note the contribution of a properly designed sparger to the total mixing achieved. Modern mixer design for turbulent flow involves optimization of the sparger or injector, the static mixing element, and properly located empty pipe spaces.

### 7-7.1 Types of Static Mixers

Examples of the most commercially significant static mixers are, by manufacturer:

*Chemineer, Inc. (Kenics)*

- *KMS*: twisted ribbon or bowtie type, with alternating left- and right-hand twists. An element is 1.5 or 1.0 diameter in length. The KME variation is edge sealed to the tube wall. See Visual Mixing CD.
- *KMX*: a series of inclined retreat curve rods forming an X lattice; alternating in direction every diameter an element is one diameter in length.
- *HEV*: a series of four tabs spaced around the pipe. An element consists of four tabs symmetrically placed. Axially, the tabs are about 1.5 diameters apart. See Visual Mixing CD.

*Koch-Glitsch, LP*

- *SMV*: several stacked sheets of corrugated metal running at 30 or 45° to the pipe axis. Each element is 0.5 to 1.0 diameter in length and adjacent elements are rotated 90° relative to each other. Mixer hydraulic diameter is determined by the height of the corrugation or the number of stacked corrugated sheets.
- *SMX*: guide vanes are intersecting bars at 45° to the pipe axis. Each mixing element is 1.0 diameter length. Adjacent elements are rotated 90°. See Visual Mixing CD.
- *SMXL*: similar to the SMX but with intersecting bars at 30° to the pipe axis. Typically, fewer bars per element, and the element length is variable, depending on application.
- *SMR*: guide vanes are hollow tubes through which heat transfer fluid circulates. The tubular bundle is arranged similar to the shape of the SMX design.
- *KVM*: single inclined tab mounted off the tube wall. Axially, tabs are about 2.5 diameters apart.
- *KHT*: twisted ribbon with alternating right- and left-handed twists.
- *SMF*: three guide vanes project from the tube wall so as not to contact each other. This is a special design for high plugging applications. Element length is approximately 1.0 diameter.
- *KFBE*: special version of the SMX/SMXL design with guide bars for exclusive application in gas fluidization of solid particles.

*Komax Systems, Inc.*

- *Komax mixer*: crossed elliptical plates with a flat at the centerline. Adjacent mixing elements are rotated 90°.

*Charles Ross & Son Company*

- *ISG*: solid tube inserts with shaped ends so that adjacent elements form a tetrahedral chamber, each with four holes drilled at oblique angles.

*Sulzer Chemtech*

- *SMV, SMX, SMXL, SMF, and SMR*: as described above. These products, initially developed jointly by Koch and Sulzer under a licensing agreement, now expired.

See Figure 7-16 for an illustration of some of these static mixers. There are many more types, and new ones are being developed constantly. Most of the above are characterized by good vendor technical information. Omission of a type from the list above, however, does not necessarily indicate an inferior product, just a lack of quantitative information. Many manufacturers have copied the basic



**Figure 7-16** Static mixer design options. From left: vortex mixer (type KVM), corrugated plate (type SMV), wall-mounted vanes (type SMF), cross-bar (type SMX), helical twist (type KHT), cross-bar (type SMXL). (Courtesy of Koch-Glitsch, LP.)

**Table 7-4** Rough Guidelines for Applications in the Laminar and Turbulent Flow Regimes<sup>a</sup>

Flow Regime	Static Mixer Design									
	KMS	KMX	HEV	SMV	SMX	SMXL	SMR	KVM	SMF	ISG
Laminar										
Mixing/blending	c	a			c	c			a	a
High–low viscosity		a			c	a				a
Dispersion	a	a			c	a				a
Heat transfer	c				b	c	c			
Plug flow	b				c	b	c*			
Turbulent										
Mixing/blending										
High turbulence	a		c	c <sup>†</sup>				c		
Low turbulence	c			c	a	a			a	
Dispersion										
Liquid–liquid	c			c	a	a	c*		a	
Gas in liquid	c			c	a	a	a*		a	
Liquid in gas	a			c	a					
Fluidized beds					c <sup>‡</sup>					

<sup>a</sup> a, Applicable; b, typically applied; c, best design choice. \*, Where temperature control is required; <sup>†</sup>, especially for very large diameters and nonround cross-sections; <sup>‡</sup>, gas fluidized solid particles, specialized design (Koch-type KFBE).

helical mixing element, Kenics type KMS/KME, and have competing products on the market.

### 7-7.2 Static Mixer Design Options by Flow Regime and Application

Table 7-4 provides rough guidelines for applications in the laminar and turbulent flow regimes. Equipment selection and sizing should be based on application engineering to meet specific process requirements.

### 7-7.3 Selecting the Correct Static Mixer Design

See Figure 7-17.

## 7-8 STATIC MIXER DESIGN FUNDAMENTALS

It is appropriate to start this section with a word of caution. The equations and design constants given here and in the tables for the various types of commercial motionless mixers come from the open literature and the vendors' literature. All calculations based on them should be confirmed by the vendors. There are differences in design and construction in different pipe sizes which can significantly affect pressure drop and energy dissipation. For example, high pressure drop designs tend to be made of heavier construction (thicker sheets or bars) so that the actual velocities are somewhat higher than the superficial or empty pipe velocity usually calculated. Additionally, new mixers are constantly being developed, and the parameters of those are not given.

### 7-8.1 Pressure Drop

In both laminar and turbulent cases, the addition of the baffles of motionless mixers increases the pressure drop, and therefore extra energy is required to get the additional mixing effect. Twisted-ribbon and structured packing static mixers will increase pressure drop per unit pipe length over standard open pipe by as much as a factor of 7 to several hundred, depending on the Reynolds number. Vortex-generating designs operate with less flow resistance, but the pressure drop is still significantly higher than it is for the same pipe size without elements. A fundamental fact is that pressure energy is required to mix in pipeline flow. Reduced mix time (shorter mixer length) requires higher-energy dissipation rates.

The pressure drop in a static mixer of fixed geometry is expressed as the ratio of the pressure drop through the mixer to the pressure drop through the same diameter and length of open pipe  $K_L$  for laminar and  $K_T$  for turbulent flow.

$$\Delta P_{sm} = \begin{cases} K_L \Delta P_{pipe} & \text{when laminar} \\ K_T \Delta P_{pipe} & \text{when turbulent} \end{cases} \quad (7-15)$$

$$(7-16)$$

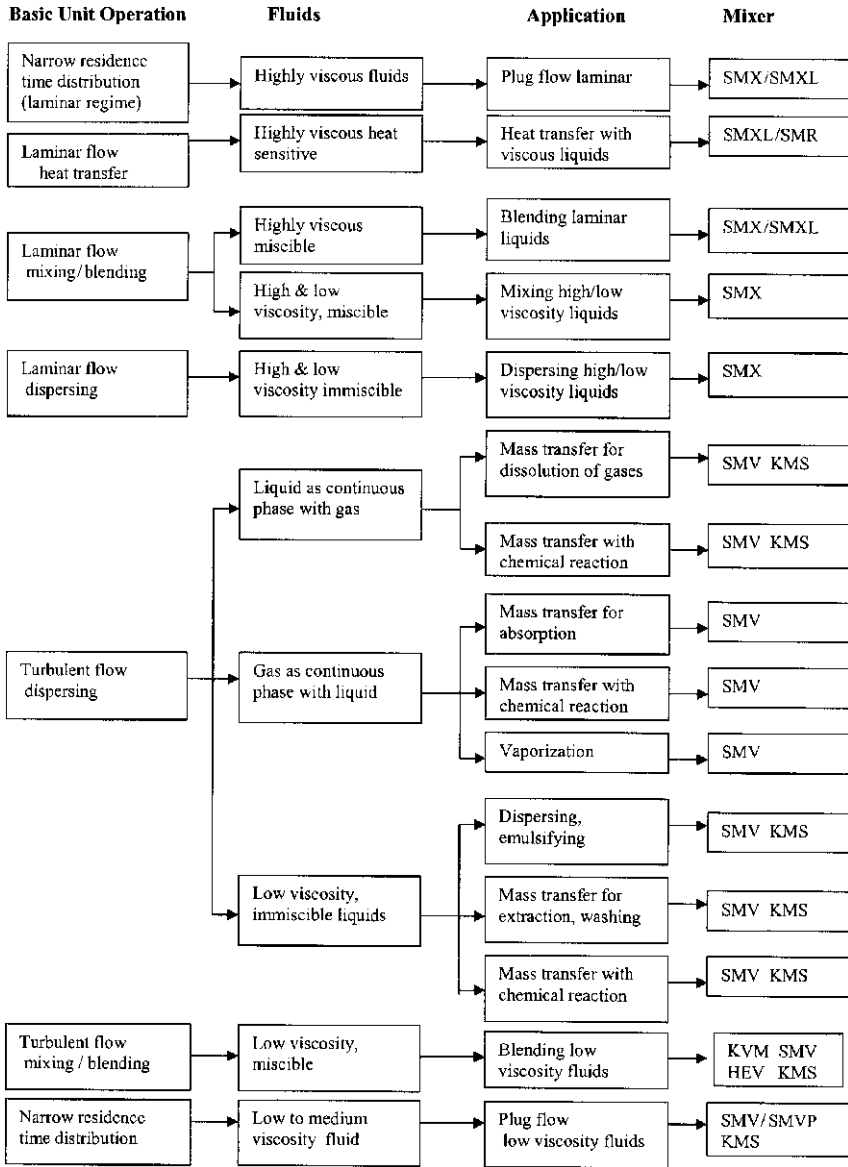


Figure 7-17 Correct static mixer design: applications.

In Tables 7-5 and 7-6, values of  $K_L$  and  $K_T$  are given. These values are considered good to about 15%.

For completeness the standard pressure-drop equation for open pipe is

$$\Delta P = 4f \frac{L}{D} \rho \frac{V^2}{2} \tag{7-17}$$



**Table 7-5** Laminar Blending and Pressure Drop Parameters for Motionless Mixers

Device	KL	KiL
Empty pipe	1	—
KMS	6.9	0.87
SMX	37.5	0.63
SMXL	7.8	0.85
SMF	5.6	0.83
SMR	46.9	0.81

Source: Streiff et al. (1999).

**Table 7-6** Turbulent Blending and Pressure Drop Parameters for Motionless Mixers

Device	Ne <sup>a</sup>	KT	KiT
Empty pipe	0.01	1	0.95
KMS	1.5	150	0.50
KVM	0.24	24	0.42
SMX	5	500	0.46
SMXL	1	100	0.87
SMV	1–2	100–200	0.21–0.46
SMF	1.3	130	0.40

<sup>a</sup>Ne is the Newton number, equivalent to 2f, twice the Fanning friction factor.

Source: Streiff et al. (1999).

where  $f$  is the Fanning friction factor introduced in Section 7-2.2. It is correlated empirically for turbulent flow in smooth pipes by the Blasius equation, given by

$$f = \frac{0.079}{\text{Re}^{0.25}} \quad (7-18)$$

for Reynolds numbers between 4000 and 100 000, and in laminar flow by

$$f = \frac{16}{\text{Re}} \quad (7-19)$$

for Reynolds numbers below 2000.

Most vendors have more accurate correlations that take into account a slight Reynolds number effect in transitional and turbulent flow, and the volume fraction occupied by the mixer, which varies with mixer diameter and pressure rating. A more detailed approach is necessary for some designs that have the option for variable but similar geometry. For the most accurate pressure drop predictions, the manufacturer should always be consulted.

### 7-8.2 Blending Correlations for Laminar and Turbulent Flow

The results for blending for motionless mixers can be correlated by plotting coefficient of variation reduction  $CoV_r$  versus  $L/D$ . In laminar flow there is no effect of viscosity, flow rate or initial  $CoV$  on these correlations.  $CoV_r$  is usually found to correlate with the  $L/D$  in an exponential form,

$$CoV_r = K_i^{L/D} \quad (7-20)$$

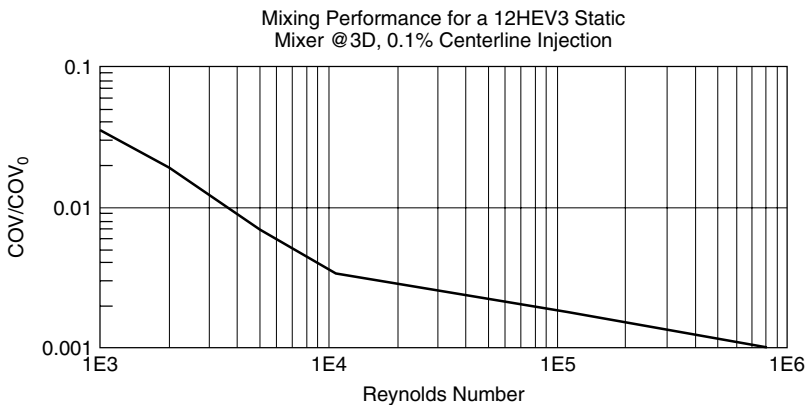
where  $K_i$  depends on the mixer type. Tables 7-5 and 7-6 give typical values for both the blending coefficient ( $K_iL$  for laminar,  $K_iT$  for turbulent) and pressure drop coefficient ( $KL$  for laminar,  $KT$  for turbulent).

At low  $L/D$  there is some deviation as the flow develops, but this is usually neglected because there is also little mixing. The Kenics HEV shows a weak Reynolds number dependence (Figure 7-18), along with a length/number of element dependence. This vortex-generating mixer design is typically applied at a Reynolds numbers above 10 000.

**7-8.2.1 Laminar Flow: Effective Shear Rate.** In laminar flow, the fluids are often shear thinning (i.e., the viscosity decreases with increasing shear rate). The apparent or effective shear rate in an empty pipe with Newtonian fluids is expressed as

$$G' = 8 \frac{V}{D} \quad (7-21)$$

For motionless mixers in laminar flow, the shear rate is higher. This and the extra surface area are what contribute to the higher pressure drop. Table 7-7 gives some



**Figure 7-18** Coefficient of variation reduction versus Reynolds number for the HEV mixer. (Courtesy of Chemineer, Inc.)

**Table 7-7** Effective Shear Rate in Motionless Mixers  
 $KG = G'/(V/D)$

Device	KG
Empty pipe	8
KMS	28
SMX	64
SMXL	30
SMV	50
SMF	25
SMR	60

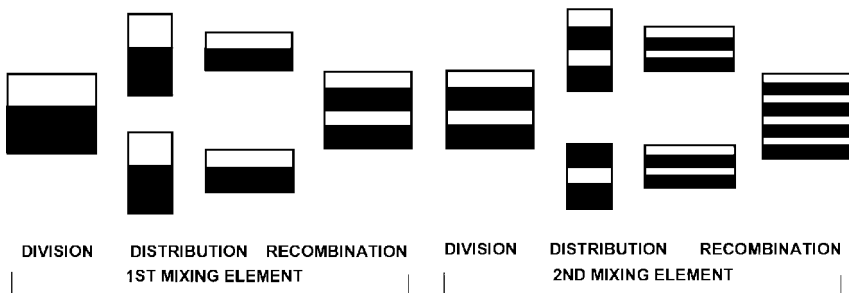
Source: Streiff et al. (1999).

estimated effective shear rates in a variety of mixers, based on

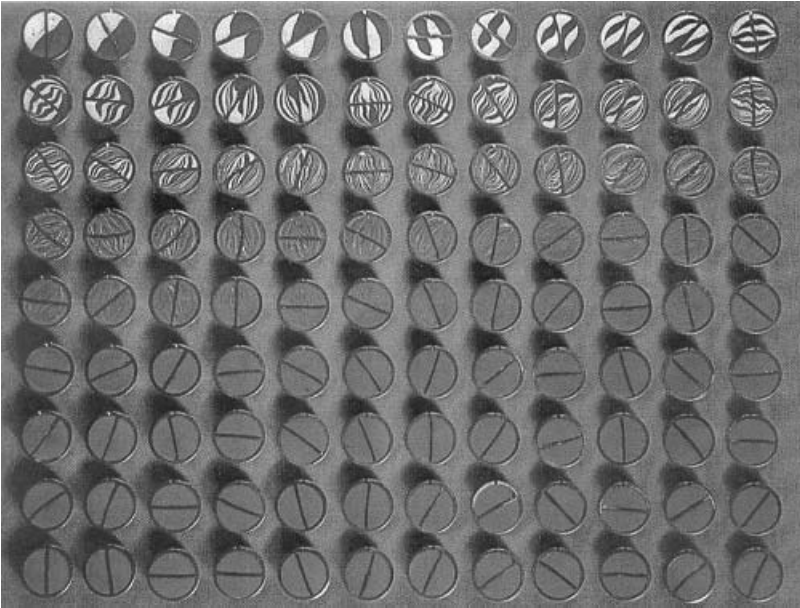
$$K_G = \frac{G'}{V/D} \tag{7-22}$$

With this apparent or effective shear rate a rheogram relating effective viscosity to shear rate can be used to calculate an effective viscosity for pressure drop calculations. For another experimental approach, see Jaffer and Wood (1998).

**7-8.2.2 Laminar Flow: Layer Generation.** In laminar flow, mixing of miscible components with similar viscosity and nonelastic behavior is achieved by the formation of layers as the materials are stretched and deformed into each other. All static mixers employ the principle of dividing the flow into substreams, distributing the substreams radially, and recombining them in a reordered sequence. The number of layers is increased and layer thickness is reduced by each successive mixing element. The process is represented schematically in Figure 7-19 and shown for real mixing processes in Figure 7-20 (KMS cross-cuts, cutting flow into two substreams) and Figure 7-3 (SMX cross-cuts, cutting flow into eight



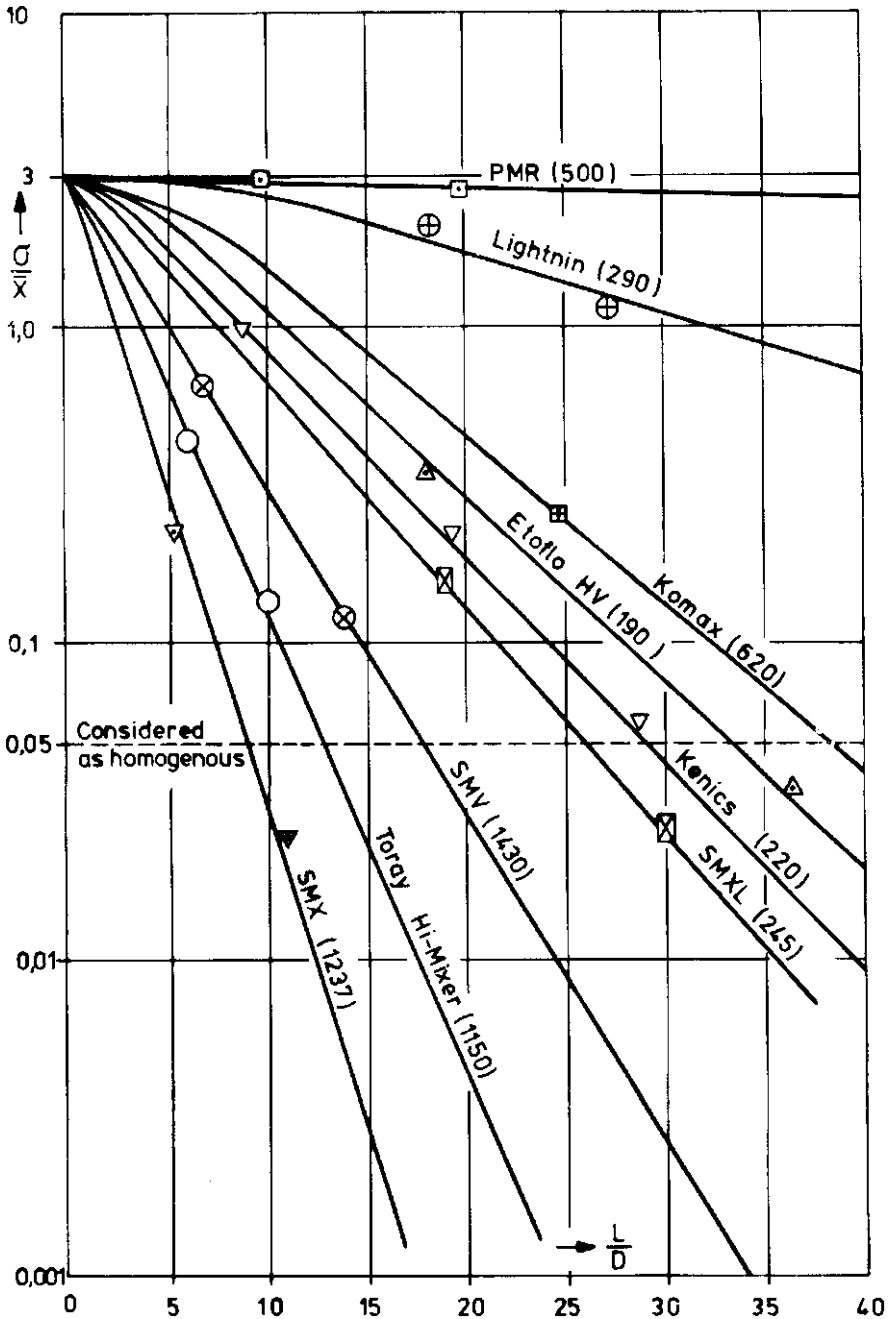
**Figure 7-19** Generalized portrayal of simple mixing in a static mixing device (each element dividing the fluid flow into two substreams).



**Figure 7-20** Laminar mixing of fluids by division and recombination (KMS mixer). Cross-sections of the mixer are shown in sequence from left to right, top to bottom. (Courtesy of Chemineer, Inc.) See insert for a color representation of this figure; for the animation, see the Visual Mixing CD affixed to the back cover of the book.

substreams). It is interesting to note that the rate of layer generation differs with different designs.

**7-8.2.3 Comparison of Several Static Mixers in Laminar Flow.** Several static mixer designs were studied (Alloca and Streiff, 1980) for the purpose of comparing their performance as distributive mixers operating in laminar flow. The conductivity tracer method was used to measure the degree of mixing achieved at 10% additive fraction. Testing was performed using concentrated glucose solutions with a viscosity in the range 7 to 9 Pa · s. The additive and bulk stream were equal in composition except for a small amount of tracer in the additive stream. Additionally, they were of equal viscosity to eliminate the effect of viscosity differences at the interface. Variation coefficient as a function of relative mixer length ( $L/D$ ) is plotted in Figure 7-21. The difference in the degree of mixing per unit mixing length is shown clearly. Table 7-8 provides a comparison of mixer diameter, length and volume, fluid hold-up, and pressure drop to achieve the same degree of mixing, in this case a variation coefficient of 0.05 (5%). There is a significant degree of design flexibility in laminar distributive mixing. Mixers can be optimized to minimize pressure drop, length requirement, or residence time, or some combination. Of the mixers shown in Figure 7-21, the PMR design is of unknown manufacture and the Lightnin mixer tested most often used in turbulent flow.



**Figure 7-21** Coefficient of variation (CoV) versus relative mixer length for several static mixer designs operating in laminar regime at 10% additive and equal viscosity. (From Alloca and Streiff, 1980.)

**Table 7-8** Comparison of Static Mixers for Equivalent Homogeneity in Laminar Flow

Mixing Unit	Measured Values		Comparisons				Pressure Drop <sup>b</sup>	
	L/D for $\sigma/\bar{x} = 0.05$	$Ne \cdot Re_D$	Volume <sup>a</sup>	Holdup <sup>a</sup>	Diameter <sup>a</sup>	Length <sup>a</sup>		
SMX	9	1237	1	1	1	1	1	
SMXL	26	245	1.8	1.8	0.84	2.4	0.6	
SMV	18	1430	4.6	4.5	1.3	2.7	2.3	
Kenics	29	220	1.9	1.8	0.84	2.7	0.6	
Etoflo HV	32	190	2	2	0.84	2.7	0.6	
Komax	38	620	8.9	8.2	1.3	5.4	2.1	
Lightnin	100	290	29	27	1.4	15.3	2.6	
PMR	320	500	511	460	2.4	86	14.5	
Cunningham			No mixing					
Toray	13	1150	1.94	0.88	1.1	1.6	1.35	
N-Form	29	544	4.5	3.8	1.1	3.6	1.40	
Ross ISG	10	9600	9.6	3.4	2.1	2.3	8.6	

<sup>a</sup>Multiple of volume, fluid holdup, diameter and length as compared to the SMX design for equal volumetric flow, viscosity, pressure drop, and variation coefficient (CoV = 0.05).

<sup>b</sup>Multiple of pressure drop as compared to the SMX design for equal volumetric flow, viscosity, pipe diameter, and variation coefficient (CoV = 0.05).

**7-8.2.4 Effect of Physical Property Differences in Blending.** Two fluids may be miscible but not have the same physical properties. For example, a high viscosity soluble dispersant may be added to water. A small amount of solvent may be added to a highly viscous polymer stream. Two polymers of different molecular weight and therefore different viscosity may need to be blended. The correlations given above are for materials that are miscible and of the same physical properties. The effect of density is usually not large but can be significant (see Section 7-11.1 on orientation of motionless mixers for more on density effects). On the other hand, viscosity can differ by orders of magnitude, and the materials are still miscible.

In turbulent mixing systems the trick is to keep the more viscous materials (almost always the additive phase) from getting into a low-turbulence area. There is interesting work on stirred tanks by Smith and Schoenmakers (1988) which is equally applicable to static mixers. They found that if the high viscosity additive is allowed to touch the wall of the vessel, it takes a long time to dissolve. If added into the turbulent zone, mixing time is the same as with low viscosity material. For motionless mixers this suggests that the additive should be added not at the inlet but between the elements, where high levels of turbulence occur. When added at the inlet, the additive stream can drift into a low-turbulence area.

In viscous systems the additive viscosity (dispersed phase,  $\mu_d$ ) is usually of lower viscosity than the bulk stream (continuous phase,  $\mu_c$ ). With such systems the low viscosity additive slips between the areas of high shear rate and shear stress and the flow is segregated. Mixing length is much greater to reach a desired CoV.

The following empirical relation seems to describe the situation (Streiff et al., 1988):

$$\left(\frac{L}{D}\right)_{\text{unequal}} = \left(\frac{L}{D}\right)_{\text{equal}} + K \log \frac{\mu_c}{\mu_d} \quad (7-23)$$

This equation applies for long mixers where the outlet CoV is low. The number of values for  $K$  is limited. For the SMX design it has been determined experimentally to be 1.0. For other designs, values between 2 and 10 are probably realistic. Note that for large viscosity ratios  $\mu_c/\mu_d$  such as 10 000 : 1, which are not uncommon in the polymer industry, the mixer length can be 1.5 to 3 times longer than it would be if the streams were of equal viscosity. Design selection is limited when mixing low viscosity additives into viscous bulk streams.

For miscible additives of higher viscosity with a  $\mu_c/\mu_d$  ratio of 1 : 10 000, Streiff (1999) claims that no mixing is possible using static mixers operating in laminar flow. Diffusion will still occur and this suggests that multiple inlets will be helpful.

A special case is one where materials are miscible but because of molecular differences have different surface tensions, and so when initially contacted there is a finite interfacial tension. Such materials act initially as immiscible, but as mass transfer takes place the interfacial forces disappear and the system acts totally miscible. Extra mixer length is required while this transformation takes place.

### 7-8.3 Which In-line Mixer to Use

**7-8.3.1 Turbulent Blending.** In all cases energy is required to mix. For in-line mixers that energy comes from pressure drop. Motionless mixers usually do not take very high pressure drops compared to the total for liquid systems. For gas systems, however, the pressure drop, while low, is often significant compared with that of the whole system. Special high-efficiency designs have been developed which are most useful for gas systems.

If there are no time or length limitations, the simple pipeline uses the minimum energy. If there are limitations, then:

- If the main stream has sufficient pressure, a static or motionless mixer should be used.
- If the main stream does not have sufficient pressure but the additive side stream does, a Tee mixer (or spray nozzle for liquids into gas) should be used.
- If neither the main stream nor side stream has adequate pressure for mixing, an in-line mechanical device where the power can be supplied externally should be considered.

**7-8.3.2 Laminar Flow.** In laminar flow there is no radial mixing without a motionless mixer, mechanical in-line mixer, or a stirred tank. The choice is

between various motionless mixers. On a simple pressure drop basis it is a trade-off between length needed to get to a certain quality of mixing versus the required pressure drop. This balance can be determined by using the data in the earlier tables. Usually, the device with the smallest  $KL$  is the one with the lowest pressure drop but the longest length. Then the investment must be looked at. Longer mixers cost more. In some cases the extra length required to obtain a certain  $CoV$  is not possible due to space limitations, and shorter mixers operating with higher pressure drop are preferred. When the viscosity ratio of the main stream to the additive stream is large (greater than 100), only the SMX and KMX have been demonstrated to be effective.

### 7-8.4 Examples

**Example 7-1: Gas–Gas Blending—Turbulent Blending.** Two gases are to be mixed prior to entering a reactor with a catalytic bed. Prior to contacting the catalyst, the gases are inert. Since there is little radial mixing in the catalyst bed, a high degree of uniformity of stoichiometry is required in the feed gas. To this end a mixer has been proposed. Three types will be evaluated: a tee mixer, an SMV motionless mixer, and a HEV motionless mixer.

The main pipe is 762 mm in diameter. The main stream flow is  $11.7 \text{ m}^3/\text{s}$  and the side stream  $2.74 \text{ m}^3/\text{s}$ . Densities are  $1.79$  and  $1.77 \text{ kg/m}^3$ , respectively; viscosities are  $0.014$  and  $0.020 \text{ mPa} \cdot \text{s}$ .

*Physical Properties:*

	Major Flow	Minor Flow	Total
Density ( $\text{kg/m}^3$ )	1.79	1.77	1.79
Viscosity ( $\text{Pa}\cdot\text{s}$ )	$1.4 \times 10^{-5}$	$2.0 \times 10^{-5}$	$1.4 \times 10^{-5}$
Flow ( $\text{m}^3/\text{s}$ )	11.7	2.74	14.44

*Dimensions:*

$$D = 0.762 \text{ m} \quad \text{area} = \frac{\pi}{4}(0.762)^2 = 0.46 \text{ m}^2$$

*Velocity:*

$$14.44/0.46 = 31.4 \text{ m/s}$$

*Reynolds number:*

$$\frac{\rho DV}{\mu} = 0.762 \times 31.4 \times \frac{1.79}{1.4 \times 10^{-5}} = 3 \times 10^6$$

The flow is turbulent. The density difference is negligible.



The initial degree of unmixedness, from eq. (7-6):

$$C_v = \frac{2.74}{14.44} = 0.19$$

$$\text{CoV}_0 = \left( \frac{1 - 0.19}{0.19} \right)^{0.5} = 2.06$$

Specify the side tee:

$$V = \frac{11.7}{0.46} = 25.43 \text{ m/s}$$

$$\frac{q}{Q} = \frac{2.74}{11.7} = 0.23$$

From eq. (7-10),

$$\frac{q}{Q} = 0.23 = \left( \frac{d}{D} \right)^{1.5}; \quad \text{then } \frac{D}{d} = 2.66, \quad v = 45.67 \text{ m/s}$$

The side tee nozzle diameter is 0.29 m with a velocity of 45.67 m/s.

This is the optimum design. The degree of mixing is given by eqs. (7-13) and (7-14):

$$B = n^2 R^2 \left( \frac{d}{D} \right)^2 = (1) \left( \frac{45.67}{25.43} \right)^2 \left( \frac{0.29}{0.762} \right)^2 = 0.47$$

for  $B = 0.47$ ,  $E = 1.33$ . Use the maximum  $x/D$  of 5.0 for the mix length:

$$x = 5 \times 0.762 = 3.81 \text{ m}$$

$$\text{CoV}^2 \times 5^{1.33} = \frac{0.32}{0.47^{0.86}} = 0.62$$

$$\text{CoV} = \left( \frac{0.62}{5^{1.33}} \right)^{0.5} = 0.27$$

Thus, at the end of 3.81 m, the CoV has been reduced from 2.06 to 0.27. That is still a high variability and probably not good enough for this application. A long tailpipe would still be required.

For the HEV, use Figure 7-18 at a Reynolds number of  $3 \times 10^6$  the coefficient of variation reduction,  $\text{CoV}_r = \text{CoV}/\text{CoV}_0 = 0.001$  for three sets of HEV tabs as measured three diameters downstream. This means a CoV of 0.002 at 2.2 m downstream for the last tab. The three sets of HEV tabs will take up about another three diameters, for a total length of 4.4 m. The pressure drop must be estimated by the vendor.

For the SMV motionless mixer, use eq. (7-20) and Table 7-6.

For the SMV the friction factor is 1-2 and  $K_f T$  is between 0.21 and 0.46, depending on exact design. We will use an average 0.33.

From eq. (7-20),  $\text{Co}V_r = 0.33^{L/D}$

To get the same  $\text{Co}V_r$  as with the HEV,

$$\frac{L}{D} = \frac{\ln 0.0010}{\ln 0.33} = 6.23$$

The mixer length would have to be 4.74 m.

An alternative would be to use Figure 7-7; however, it does not go out to such a low  $\text{Co}V_r$  value. The vendor would probably optimize this further for pressure drop and length by employing a multipoint sparger and spacers between the elements, both subjects beyond the scope of this chapter.

**7-8.4.2 Thermal Homogenization—Laminar Blending.** A polymer solution exiting a simple shell-and-tube heat exchange shows erratic behavior in downstream processing. Contact temperature measurements around the exit pipe gave temperatures of 113, 93, 115, and 126°C. We will design a motionless mixer to get the temperature spread down to  $\pm 0.5^\circ\text{C}$ . The polymer solution has a density of 1100 kg/m<sup>3</sup> and a viscosity of 650 mPa·s at process conditions. The flow rate is 15 060 lb/hr (6845 kg/h). The exit pipe is 50 mm inner diameter.

*Physical properties:*

$$\rho = 1100 \text{ kg/m}^3 \quad \mu = 0.650 \text{ Pa} \cdot \text{s}$$

$$\text{flow} = 6845 \text{ kg/h} = 1.6 \times 10^{-3} \text{ m}^3/\text{s}$$

$$\text{pipe size} = 2 \text{ in.} = 50 \text{ mm} = 0.05 \text{ m} \quad \text{area} = 0.002 \text{ m}^2$$

$$\text{velocity} = 0.88 \text{ m/s} = V$$

$$\text{Re} = \frac{\rho DV}{\mu} = 0.05 \times 0.88 \times \frac{1100}{0.650} = 74.5$$

Flow is laminar.

$$\text{maximum temperature difference} = 126 - 93 = 33^\circ\text{C}$$

$$\text{desired temperature difference} = 1^\circ\text{C}$$

$$\text{desired reduction } 1/33 = \text{Co}V_r = 0.033$$

From Table 7-2, for the Kenics KMS,

$$K_i = 0.87 \quad K_L = 6.9$$

$$\frac{L}{D} = \frac{\ln 0.033}{\ln 0.87} = \frac{3.41}{0.14} = 24.36$$

$$L = 1.22 \text{ m}$$

For the SMX,

$$K_i = 0.63 \quad K_L = 37.5$$

$$\frac{L}{D} = \frac{\ln(0.033)}{\ln(0.63)} = \frac{3.41}{0.46} = 7.41$$

$$L = 0.37 \text{ m}$$

For pressure drop in an empty pipe, use eq. (7-17):

$$\Delta P = 4f \frac{L}{D} \rho \frac{V^2}{2}$$

For 1 m of pipe without mixers,

$$f = \frac{16}{\text{Re}} = \frac{16}{74} = 0.215$$

$$\Delta P = 4 \times 0.215 \times \frac{1.0}{0.05} \times 1100 \times \frac{0.2^2}{2} = 7324 \text{ Pa or } 0.07 \text{ bar}$$

For the KMS, the length is 1.22 m:

$$\Delta P = 1.2 \times 6.9 \times 7324 = 63\,444 \text{ Pa or } 0.63 \text{ bar}$$

For the SMX, the length is 0.37 m:

$$\Delta P = 0.37 \times 37.5 \times 7324 = 103\,928 \text{ Pa or } 1.04 \text{ bar}$$

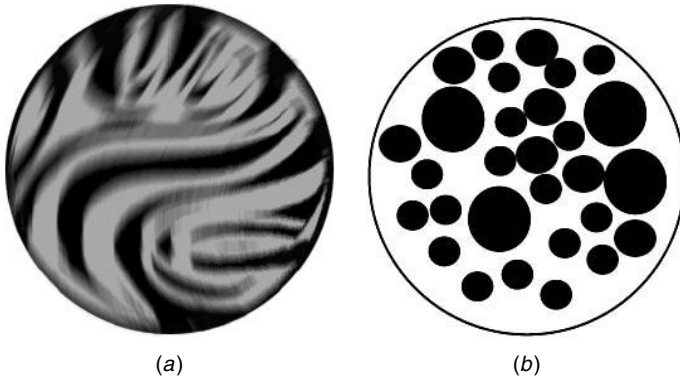
Thus, the SMX takes more pressure drop but is shorter and has less volume, while the KMS takes less pressure drop but has more volume and length. The final best choice of mixer for this application would probably depend on what the secondary criteria were, since the pressure drops are not high in either case.

## 7-9 MULTIPHASE FLOW IN MOTIONLESS MIXERS AND PIPES

Dispersive multiphase mixing is distinguished from simple blending or distributive mixing in that the additive phase breaks up into discrete drops or bubbles which are surrounded by the other phase. A basic comparison is shown in Figure 7-22.

### 7-9.1 Physical Properties and Drop Size

In the process result of area generation or dispersion, the most important physical property for low viscosity fluids is the interfacial tension. This is the force at an interface between immiscible fluids: liquids with liquid or liquids with gas



**Figure 7-22** Blending (simple distributive mixing) and dispersive mixing: (a) blending: flow streams are interleaved; (b) dispersive mixing: one phase is discontinuous.

that resists elongation or deformation. For single liquids this is called the *surface tension*. There is no surface tension for gases, as they form no interfaces. However, for liquids it is a physical property that can and often is measured. For two totally immiscible fluids, the force at the surface is called the *interfacial tension*. It is the difference in their individual surface tensions. However, if there is mutual solubility of the components in each other, the interfacial tension is less than this ideal defined value. Interfacial tensions can be lowered further by the presence of surface-active chemicals in even trace amounts in the liquids. These surface-active agents reside at the interface and change the surface energy of the drops. They affect not only the interfacial tension but also the tendency for interfaces to break or coalesce.

With two immiscible fluids it is also necessary to know which is the continuous phase and which is the dispersed phase. For liquid and gases, all drop size/bubble correlations are for gas dispersed in liquid. In theory, at high enough gas rates it should be possible to make a spray in a motionless mixer, but such a phenomenon has not been reported. Even at inlet volume ratios of eight parts of gas to one of liquid, the liquid phase is continuous, with gas holdups well over 50%. In a two-phase gas–liquid flow system there are several possible flow regimes. These include bubble flow, wavy flow, slug flow, annular flow, dispersed flow (also referred to as spray or mist flow), and a few other variations. A good description of these regimes for horizontal flow in empty pipe is provided by Lockhart–Martinelli (Govier and Aziz, 1972). When it is desirable to operate in the dispersed flow regime, liquid volumetric flow should be kept below 10% of the gas volumetric flow. Baker and Rogers (1989) report that the mixer/contacter geometry has a strong impact on the point at which dispersed flow is achieved. The SMV mixer will induce dispersed flow at much lower gas velocity than that required in an empty pipe. Flow regime (and operating pressure drop) at any gas and liquid flow rate will be influenced by mixer orientation and flow direction (horizontal, vertical up or down).

When dealing with the contacting of liquid in continuous gas flow in static mixers, one thinks in terms of liquid film flow on the mixer surfaces and droplets being formed as the liquid film is sheared off the mixer blades. The total interfacial surface is a combination of film plus droplet surface. Film flow is well defined and equal to the mixer plate surface area and is often used as the contact surface area for calculation of mass and heat transfer capacity. Mixers with high surface/volume ratios are most often the preferred design.

For two liquids the continuous phase is determined by physical and surface properties in a way not thoroughly understood in either agitated tanks or pipeline mixers. A rough guideline is that the liquid with volume fraction below 30% is dispersed. Between 30 and 70% is an ambiguous region where there are no hard rules, although it should be noted that water tends to be the continuous phase even above 50% organic (see Pacek et al., 1994). With static mixers providing a large surface area, the phase that best wets the mixer surface will tend to be the continuous phase, even if the minor component is within limits in the ambiguous region. Thus, water is the continuous phase in metal and glass equipment and would tend to be the dispersed phase in polyethylene, PVC, and polytetrafluoroethylene. This has little effect on the drop size, which is most strongly affected by interfacial tension. In turbulent flow, which phase is dispersed has little effect. However, it may be important for downstream operation such as decanters or at high volume fractions where coalescence is important. In laminar flow, where viscosity of the continuous phase is important, the determination of which phase is continuous is important even in dispersion.

**7-9.1.1 Turbulent Flow: Dispersed Phase Drop or Bubble Size.** The breakup of drops and bubbles in turbulent fields in agitated tanks is discussed elsewhere in the book. Similar theories apply to motionless mixers as uniform turbulence generators and similar equations can be developed. In pipeline flow the equations for dispersed gas and dispersed liquid are identical. The physical property differences are handled in the physical property corrections. Little or no coalescence is assumed. The equations apply in vertical flow and in horizontal flow when the criterion in Chapter 12 is met. Unlike stirred tanks, the motionless mixer system reaches equilibrium droplet size very quickly, in a few pipe diameters of mixer length. The most popular equation form (Middleman, 1974; Streiff et al., 1999) is

$$d_{\max} = k_1 \left( \frac{\sigma}{\rho_c} \right)^{0.6} \left( \frac{\rho_c}{\rho_d} \right)^{0.2} \varepsilon^{-0.4} \quad (7-24)$$

$k_1$  is on the order 1.0, and  $d_{\max}$  is about 1.5 times  $d_{32}$ .

An alternative is the Weber number form,

$$\frac{d_{32}}{D} = \frac{K}{We^{0.6}} \quad (7-25)$$

Here  $K$  is different for various mixers and the form does not include all the density, viscosity, and concentration effects. For fluids with a low density

ratio and low viscosity ratio in turbulent flows,  $K = 0.49$  for the Kenics KMS mixer (Calabrese and Berkman, 1988). For other mixers the drop size is inversely proportional to the friction factor to the 0.4 power.

Additional terms are often included to take into account dispersed phase viscosity and coalescence due to holdup. They usually depend on the volume fraction of dispersed phase:

$$1 + kC_v \quad (7-26)$$

where  $k$  is 3 to 5 and  $C_v$  is the volume fraction of dispersed phase. Streiff et al. (1997) report that this  $k$  depends on flow orientation and varies from 1.7 to 3.4 for upflow versus downflow for air in demineralized water. This effect of dispersed phase volume fraction on drop size is probably a coalescence rather than a turbulence dampening effect as suggested by some. Several authors also include a density ratio to get both gas and liquid dispersed phase data in the same correlation. This ratio is continuous over dispersed and the exponent varies from 0.1 to 0.5.

Notice that this relation predicts that gas bubbles will be larger than liquid droplets at the same energy dissipation. This is observed experimentally. A similar correlation is used for gas-driven sprays where the gas is the continuous phase, and this predicts that spray drops would be finer than drops in a motionless mixer. This is also observed.

For a viscous dispersed phase the derivation of the drop size equations are modified (Calabrese and Berkman, 1988; Streiff et al., 1997) with an extra term representing the viscous resistance to drop breakup. This adds a new term:

$$d_{\max} = K_1 \left( \frac{\sigma}{\rho_c} \right)^{0.6} \left( \frac{\rho_c}{\rho_d} \right)^{0.2} \varepsilon^{-0.4} (1 + Vi) \quad (7-27)$$

where

$$Vi = \frac{\mu_d (\varepsilon d_{\max})^{0.333}}{\sigma (\rho_c / \rho_d)^{0.5}} \quad (7-28)$$

or

$$Vi = \mu_d \frac{V}{\sigma} \left( \frac{\rho_c}{\rho_d} \right)^{0.5} \quad (7-29)$$

Sometimes the Ohnesorge number ( $Oh$ ) is used to show the breakpoint between viscosity-controlled and surface tension-controlled breakup. Unfortunately, it is depends on the drop size and is of limited usefulness.

It is, however, used extensively in the spray literature (Lefebvre, 1989):

$$Oh = \frac{(We)^{0.5}}{Re} = \frac{\mu_d}{(\rho_c \sigma d)^{0.5}} \quad (7-30)$$

It thus is the ratio of the resisting viscous force to the surface force.

Though seldom explicitly mentioned, the dispersed phase viscosity in the drop breakup equations should be the elongational viscosity. For simple Newtonian fluids it is three times the shear viscosity. For elastic fluids the elongational viscosity can be much larger than the shear viscosity. Few data exist for the breakup of elastic drops.

Streiff et al. (1999) propose an empirical equation for low viscosity dispersed phase, similar to eq. (7-16) but allowing some of the dimensionless constants to float:

$$d_{\max} = 0.93 \left( \frac{\sigma}{\rho_c} \right)^{0.6} \left( \frac{\rho_c}{\rho_d} \right)^{0.1} \varepsilon^{-0.4} \quad (7-31)$$

and for all the data (Streiff et al., 1997),

$$d_{\max} = 1(1 + 1.7C_v) \left( \frac{\sigma}{\rho_c} \right)^{0.6} \left( \frac{\rho_c}{\rho_d} \right)^{0.1} \varepsilon^{-0.4} 0.9^{0.6} (1 + 0.3Vi)^{0.6} \quad (7-32)$$

Figure 7-23 is a comparison of the measured drop size for different mixer types with the drop size calculated according to this equation [see Streiff et al. (1997) for more information].

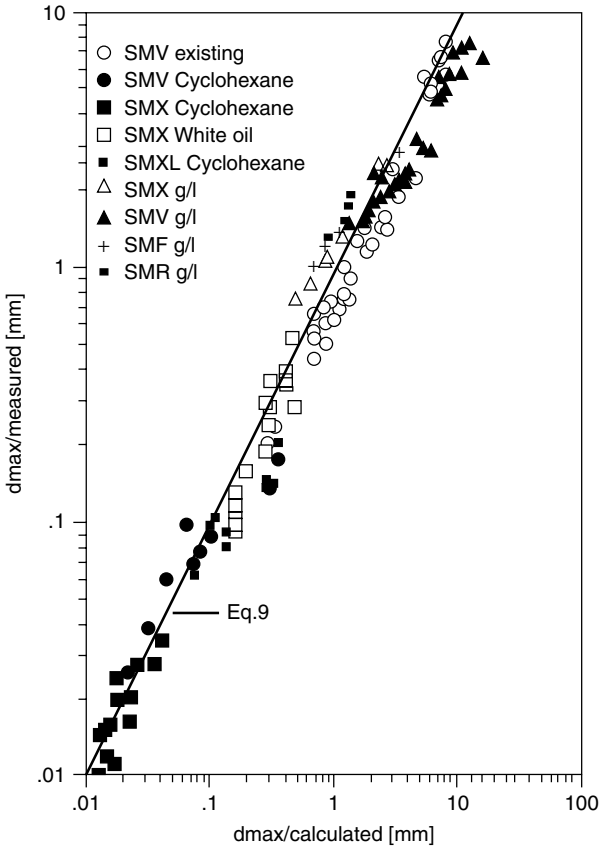
In addition, Chandavimol et al. (1991a,b) have estimated the kinetic rate at which the bubbles go from initial size to the maximum equilibrium size as a function of energy dissipation. The rate of dispersion was found to be approximately proportional to energy dissipation rate. [See Figure 7-24 for a comparison of bubble breakup rate between vortex (HEV) and spiral (KMS type) static mixers.] In general, the equilibrium drop size is reached in a few pipe diameters. However, the drop size distribution is narrowed as the simultaneous processes of drop breakup and coalescence are continued, depending on the mixer design and fluid properties. See also Hesketh et al. (1987, 1991).

In the correlations above,  $\varepsilon$  for a motionless mixer is given by

$$\varepsilon = \frac{\Delta PQ}{\rho AL} = \frac{\Delta P}{\rho L} V \quad (7-33)$$

Most theories and analysis of the dispersion process neglect the coexisting coalescence process. This is probably valid at very low concentration or when antioalescing (stabilizing) chemicals are present. For example, in agitated mix tanks, if a dispersion is made at one speed and then agitated at a lower speed, the drops will grow via coalescence. Similarly, with static mixers there is an equilibrium drop size that depends on fluid properties and specific energy dissipation within the device. If the energy input is reduced, as for example in the downstream pipeline device to provide residence time, the drop will grow to a new equilibrium value.

Droplet breakup and coalescence occur in parallel in most industrial processes. An example is found in the mining industry, where metals are extracted from leach solutions by contacting with an organic phase. In both agitated vessels and static mixers, there are two stages of contacting. The first is with high specific



**Figure 7-23** Comparison of measured to calculated drop size for different static mixers. (From Streiff et al., 1997.)

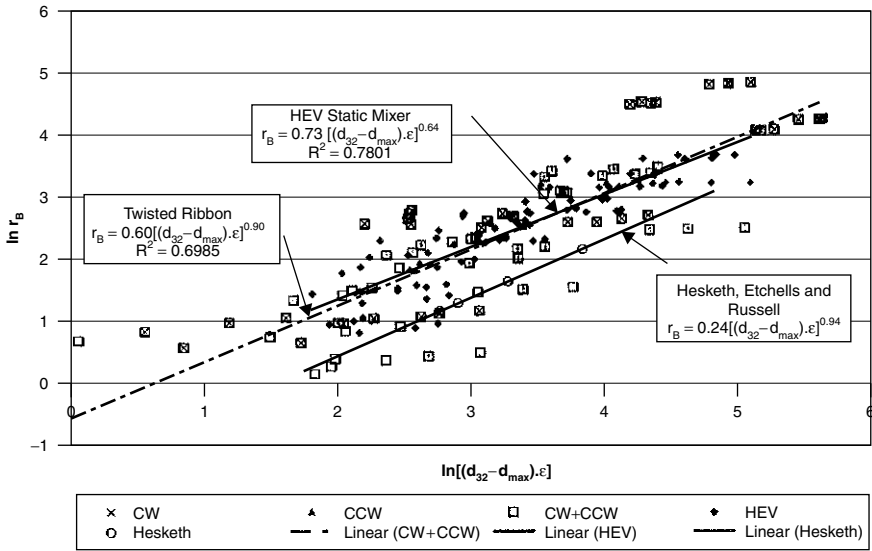
energy dissipation, creating small droplets and a high interfacial surface area for mass transfer and reaction. This is followed by a second stage, operating at a lower energy input, providing residence time and maintaining a dispersion with increasing drop size due to coalescence. It is also interesting to note that in these processes, which operate at a high dispersed phase fraction, the mass transfer rate between phases is dependent on which phase is continuous and which is dispersed.

There have been a few studies of drop size distributions, and they appear to be similar. Calabrese et al. (1988) give

$$f_v(x) = 0.5 \left[ 1 + \operatorname{erf} \left( \frac{X - X_{\text{ave}}}{1.414\sigma} \right) \right] \quad (7-34)$$

where  $f_v(X)$  is the volume frequency distribution, erf the error function,  $X_{\text{ave}}$  is the mean of the drop size distribution, and  $\sigma$  is the volume-weighted standard





**Figure 7-24** Comparison of overall correlation of bubble breakup rate for vortex (HEV) and spiral (KMS) static mixers. (Data from Chandavimol et al., 1991b.)

deviation.  $X = d/d_{32}$ ,  $X_{ave}$  and  $\sigma$  are the two constants that define the distribution. Calabrese and Berkman (1988) found that  $X_{ave} = 1.12$  and  $\sigma = 0.31$  fit both tank and motionless mixer data. Middleman (1974) got slightly narrower parameters, of  $X_{ave} = 1.06$  and  $\sigma = 0.25$ . Figure 7-25 shows the distribution obtained by Calabrese and Berkman (1988).

Also of use is the table from Streiff et al. (1997) for calculating the characteristic drop size of interest:  $d_{min}$ ,  $d_{10}$ ,  $d_{50}$ ,  $d_{32}$ ,  $d_{90}$ , and so on. For example,  $d_{50}$  means that 50% of the drop swarm volume is in drops below this diameter. The drop diameter  $d_{50}$  shown below is 60% of  $d_{max}$ . For mass transfer the surface/volume mean diameter,  $d_{32}$ , is used. It is the drop diameter that will give the same mass transfer surface area as the swarm.

$$\begin{aligned}
 d_{min} &= 0.2d_{max} & d_{10} &= 0.35d_{max} & d_{50} &= 0.6d_{max} \\
 d_{32} &= 0.65d_{max} & d_{90} &= 0.85d_{max}
 \end{aligned}$$

**7-9.1.2 Laminar Flow: Dispersed Phase Drop or Bubble Size.** The laminar mechanism for dispersed phase breakup is discussed in detail elsewhere. Again there is the balance between the forces holding the dispersed phase together and those generated by the flow through the mixer. Figure 7-26 illustrates how dispersive mixing occurs in a static mixer. The drop will break up until the force holding the drop together is larger than the continuous phase force. Thus, at

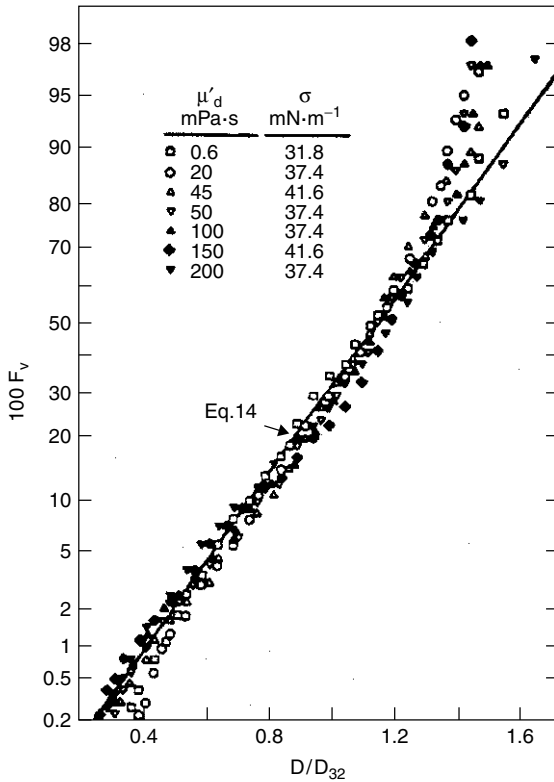


Figure 7-25 Normalized volume distribution for constant conditions of agitation at  $Re = 18\ 000$ . (From Calabrese and Berkman, 1988.)

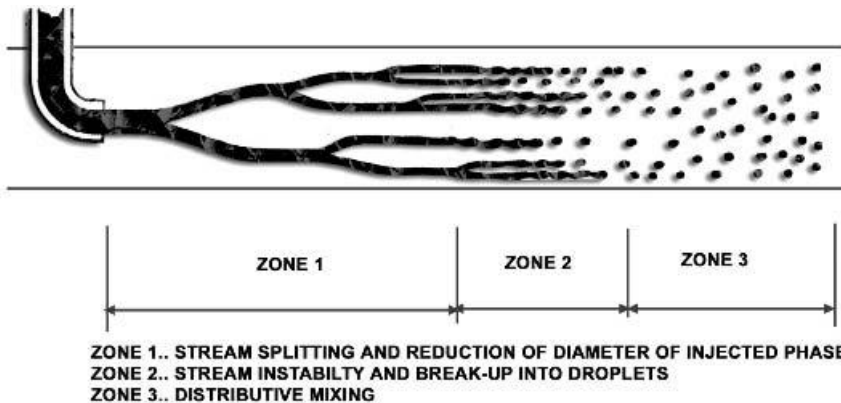


Figure 7-26 Simplified portrayal of dispersive mixing in a static mixing device. A video clip of this process is provided on the Visual Mixing CD affixed to the back cover of the book.

equilibrium the forces are about equal.

$$\frac{\sigma}{d} = \mu_c G' \tag{7-35}$$

This leads to the relation (sometimes called a *capillary number*)

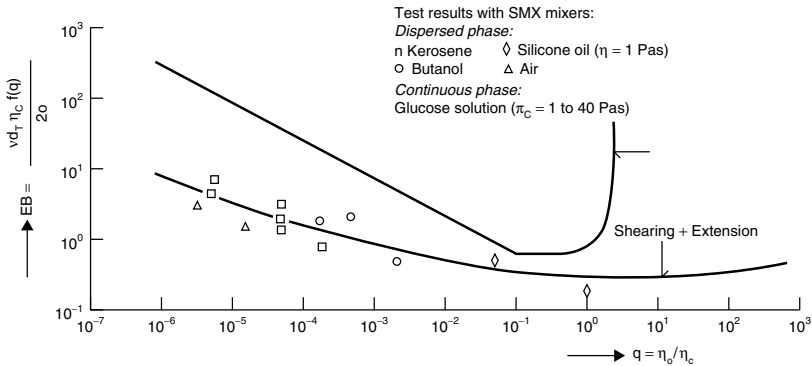
$$Ca = \frac{d\mu G'}{\sigma} \tag{7-36}$$

which can be considered as a drop size. One would expect this to depend only on the flow regime (elongational or shearing) and the viscosity ratio. Grace (1982) showed that such a relationship exists for drops broken up in laminar flow in a four-roll mill. Mutsakis et al. (1986) showed that a similar relation could be obtained for a motionless mixer. This relation and work are shown in Figure 7-27. Note that the breakup of drops only seems to occur at dispersed/continuous viscosity ratios of less than about 4. At this and higher ratios the droplets do not seem to break in shear flow. At low viscosity ratios the exponent on the viscosity ratio is between 0.33 and 0.5. Some authors give a theoretically derived viscosity ratio effect,

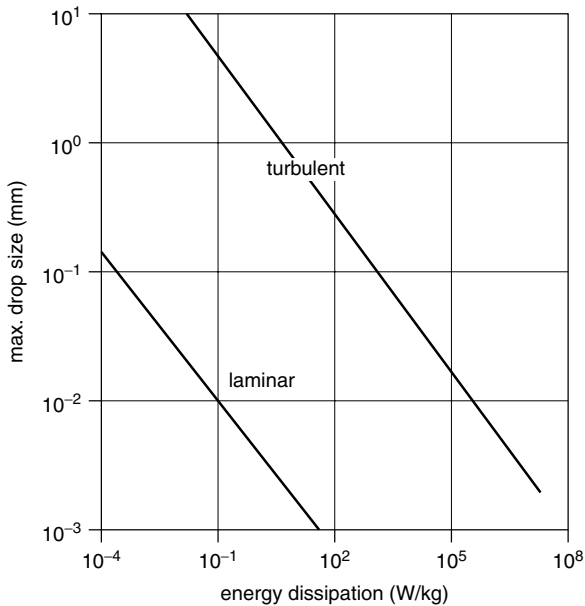
$$\frac{q + 16}{q + 19} \tag{7-37}$$

where  $q$  is the viscosity ratio, dispersed to continuous. For a range of  $q$  from 1 to 0.0, this value varies from 16/19 to 17/20. This is a very minor effect and does not accurately predict the effects measured.

From the above the maximum stable drop size can be estimated. There will be smaller drops, but in theory no drops larger than this. No data on distribution as yet exist for laminar breakup. Figure 7-28 compares drop size by laminar mechanisms with those calculated for turbulent flow. Smaller droplets are expected for laminar versus turbulent flow at the same energy dissipation rate.



**Figure 7-27** Shear stressing, causing drop disintegration. (From Mutsakis et al., 1986.)



**Figure 7-28** Maximum drop size versus energy dissipation for laminar and turbulent flow. (From Streiff et al., 1999.)

Data from Grace indicate that elongational flow is more efficient at breakage than is simple shearing flow. The flow in most motionless mixers is a combination of shear and elongation. An interesting comparison of several commercially significant static mixers utilizing flow numerical analysis is provided by Rauline et al. (1998). The investigation compares mixers using criteria of extensional efficiency, stretching, mean shear rate, and intensity of segregation. Highest efficiency-creating dispersions with static mixers are achieved by designs with high extensional efficiency throughout the mixer volume. If the region of high extensional efficiency is segregated within the mixer volume, there is no guarantee that the drop to be dispersed will pass through it, and the mixer efficiency is reduced. Two designs of commercial significance in this area of application are the SMX and ISG. Use of the ISG is limited due to its high pressure drop versus that of the SMX and other designs.

Note that after the dispersion is created, distributive mixing of the discrete drops is often required to achieve a uniform mix. This can be done at a lower energy expenditure than that required to create the dispersion initially.

### 7-9.2 Dispersion of Particulate Solids: Laminar Flow

By analogy to liquid–liquid drop breakup, we can determine the mechanism for the breakup of agglomerates of fine particles. This is a common phenomenon in polymer processing, where various solid pigments and additives need to be

deagglomerated and then dispersed into a viscous fluid. Tadmor and Manas-Zloczower (1994) give for the cohesive strength of an agglomerate,

$$\tau = \frac{9}{8} \left( \frac{1 - \varepsilon}{\varepsilon} \right) \frac{F}{d^2} \quad (7-38)$$

where

$$F = \frac{Ad}{24z^2} \quad (7-39)$$

$\varepsilon$  is the porosity,  $F$  the inter particle force,  $d$  the primary particle size,  $A$  the Hamaker constant, and  $z$  the physical adsorption distance.

From this we see that densely packed agglomerates of fine particles will be quite strong. Breakup will be by attrition, and it will take many breakups to get to all primary particles. For example, titanium dioxide has a primary particle size of 0.25  $\mu\text{m}$  but exists as a powder as agglomerates of 10 to 100  $\mu\text{m}$ . Similarly, carbon black has a primary particle size of 0.03  $\mu\text{m}$ , some aggregates of 0.15  $\mu\text{m}$ , and agglomerates of 100  $\mu\text{m}$ . Again there is a tendency for primary particles broken off agglomerates to re-form unless there are chemicals added to prevent reagglomeration (see Section 7-11.2.3).

Results of tests with the SMX mixing element in high-concentration slurries are given by Furling et al. (2000). The tests demonstrate that it is possible to prepare high solid content slurries with in-line static mixers. The slurry product is reported to have the same quality as those prepared with in-tank agitators operating at significantly higher power input. Unfortunately, a method to wet out particles using a motionless mixer has not been developed, and a tank and agitator to make the initial slurry are usually required.

### 7-9.3 Pressure Drop in Multiphase Flow

In both laminar and turbulent flow it is assumed that the mixture is pseudohomogeneous with respect to density; that is, a volume average density is used. The viscosity of the continuous phase is used. Note that for very small dispersed phase drops or bubbles (under 10  $\mu\text{m}$ ), the viscosity may even be higher and non-Newtonian, such as in foams and emulsions. In such cases direct measurements are required of a well-dispersed sample. Avoid correlations that average viscosities. For gas-liquid systems, the method of Lockhart and Martinelli (see Govier and Aziz, 1972) for turbulent flow is very successful and more accurate than the pseudohomogeneous method. The pressure drop for each phase flowing alone is calculated. The liquid/gas pressure drop ratio is made. This is used with an empirical correlation to get an enhancement factor for the liquid-alone pressure drop,

$$\left( \frac{\Delta P_{\text{liquid}}}{\Delta P_{\text{gas}}} \right)^{0.5} = X \quad (7-40)$$

and the two-phase pressure drop is given by

$$\Delta P_{tp} = \left( 1 + \frac{20}{X} + \frac{1}{X^2} \right) \Delta P_{liquid} \quad (7-41a)$$

or

$$\Delta P_{tp} = (1 + 20X + X^2) \Delta P_{gas} \quad (7-41b)$$

### 7-9.4 Dispersion versus Blending

Note that the mechanisms for blending and dispersion are very different and lead to very different relations. In simple blending (without extreme viscosity differences) the degree of mixing depends on the length of the mixer and is independent of the flow rate, shear rate, and continuous phase fluid properties. Turndown does not change the blending CoV. Extra length improves mixing. The cost of mixing (i.e., pressure drop) does depend on the flow and physical properties. In dispersion the physical properties have a strong influence. Flow rate sets the drop size, a function of specific energy input. The length to achieve equilibrium drop size is short, and more length beyond that does not affect the process result of droplet size.

### 7-9.5 Examples

**Example 7-3: Liquid–Liquid Contacting—Turbulent Dispersion.** A stream from a reactor is to be contacted with an immiscible solvent to extract the product. A motionless mixer is planned. After the mixer the two streams will enter a decanter (Table 7-9). The cut size of the decanter is 125  $\mu\text{m}$ , so a goal drop size for the mixer is 500  $\mu\text{m}$ . Choose a Kenics KMS mixer based on past experience. No line size is given.

Typical line velocities are 2 to 3 m/s. Use 2 m/s as a first guess. The line size initial estimate:  $0.0075/2 = 0.0037 \text{ m}^2 \Rightarrow 0.069 \text{ m} = 69 \text{ mm}$ ; use 3 in. pipe with an inside diameter of 76.2 mm. The velocity is then 1.63 m/s.

**Table 7-9** Process Stream Data

	Main Stream	Secondary Stream
Density ( $\text{kg}/\text{m}^3$ )	1154	982
Viscosity ( $\text{mPa} \cdot \text{s}$ )	1.07	0.475
Flow ( $\text{kg}/\text{h}$ )	21 755	8149
Flow ( $\text{m}^3/\text{s}$ )	0.0052	0.0023
Combined Streams		
Total flow	0.0075 $\text{m}^3/\text{s}$	
Interfacial tension	12 $\text{mN} \cdot \text{m}$	
Volume fraction	0.30	

Do a rough estimate of drop size using the simpler but less exact Weber number from eq. (7-25):

$$\text{We} = D \frac{\rho V^2}{\sigma} = \frac{0.0762 \times 1154 \times 1.63^2}{12 \times 10^{-3}} = 19\,283$$

$$d_{32} = \frac{KD}{\text{We}^{0.6}} = 0.49 \times \frac{0.076}{(19\,283)^{0.6}} = 100 \mu\text{m}$$

Then the average drop size from eq. (7-25) is 100  $\mu\text{m}$ .

This is too small; we must go to larger pipe—use 4 in. pipe with a diameter of 0.100 m and a velocity of 0.95 m/s:

$$\text{We} = \frac{0.100 \times 0.95^2 \times 1154}{12 \times 10^{-3}} = 8596$$

$$d_{32} = 0.49 \times \frac{0.100}{(8596)^{0.6}} = 213 \mu\text{m} \quad \text{looks good}$$

Check the Reynolds number:

$$\text{Re} = \frac{\rho DV}{\mu} = \frac{0.1 \times 0.85 \times 1154}{1.07 \times 10^{-3}} = 91\,672 \quad \text{turbulent}$$

Check for orientation using eq. (7-42) to calculate the Froude number:

$$\text{Fr} = \frac{\rho V^2}{\Delta \rho Dg}$$

$$= \frac{1154 \times 0.95^2}{(1154 - 982) \times 0.1 \times 9.8} = 6.2$$

A vertical installation is required since the Froude number is less than 20.

For the exact droplet size calculation, the energy dissipation is obtained from eqs. (7-33) and (7-17):

$$\varepsilon = \frac{\Delta P V}{L \rho} = \frac{4f V^3}{D 2} = 2f \frac{V^3}{D}$$

The friction factor for the KMS is given in Table 7-6 as  $0.5 \times 1.5 = 0.75$ :

$$\varepsilon = \frac{2 \times 0.75 \times 0.95^3}{0.1}$$

$$\frac{\sigma}{\rho} = \frac{15 \times 10^{-3}}{1154} = 1.3 \times 10^{-5}$$

Using eq. (7-31) yields

$$d_{\max} = \frac{0.93 \times (1.3 \times 10^{-5})^{0.6}}{(1154/982)^{0.1} (12.86)^{0.4}}$$

To check the viscosity effect, use eq. (7-28):

$$V_i = \mu_d \frac{(\varepsilon d_{\max})^{0.33}}{\sigma} \left( \frac{\rho_c}{\rho_d} \right)^{0.5}$$

$$= 0.475 \times 10^{-3} \frac{(12.86 \times 5 \times 10^{-4})^{0.33}}{15 \times 10^{-3}} \left( \frac{1154}{982} \right)^{0.5} = 0.0058$$

so the effect of viscosity is to increase the droplet size by 1.056, which is negligible.

To check the density effect:

$$\frac{\rho_c}{\rho_d} = \frac{1154}{982} = 1.1752^{0.1} = 1.01 \quad \text{negligible}$$

To check the concentration effect, use eq. (7-26):

$$1 + 0.3 \times 5 = 2.5 \quad \text{significant}$$

We can use a 4 in. pipe with a KMS mixer located vertically. Use about 10 diameters of mixer for a residence time of about 1 s.

Using Figure 7-25, and eq. (7-34) with  $d/d_{32} = 125/500$  gives a volume fraction less than 0.2% under 125  $\mu\text{m}$ , and this would be the approximate carryover in the decanter with a cut size of 125  $\mu\text{m}$ .

**Example 7.4: Blending and Dispersion—Large Viscosity Ratios.** A molten polymer is to have several materials added. One is a soluble antioxidant and the other is an immiscible silicone oil slip agent (Table 7-10). The pipe size is 50 mm. We will work out a motionless mixer design.

$$\text{flow} = \frac{1500/3600}{900} = 4.62 \times 10^{-4} \text{ m}^3/\text{s}$$

$$\text{area} = \frac{3.1416}{4 \times 0.05^2} = 1.96 \times 10^{-3} \text{ m}^2$$

$$\text{miscible volume fraction } C_v = \frac{15/600}{15/600 + 1500/900} = 0.015$$

**Table 7-10**

	Polymer	Miscible	Immiscible
Flow (kg/h)	1500	15	10
Density (kg/m <sup>3</sup> )	900	600	1010
Viscosity (Pa · s)	Shear-thinning index of 0.7 1200 at 1 s <sup>-1</sup>	100	0.5
Interfacial tension (m · N/m)		12	



$$\text{velocity} = 0.2357 \text{ m/s}$$

$$\text{empty pipe shear rate } G' = 8 \frac{V}{D} = 8 \times \frac{0.2357}{0.05} = 37 \text{ s}^{-1}$$

$$\text{viscosity ratio miscible} = \frac{0.5}{1200 \times 1000} = 4 \times 10^{-7}$$

use an SMX because of the viscosity ratio.

$$\text{shear rate in mixer (Table 7-7) } G' = 64 \times \frac{0.2357}{0.05} = 301 \text{ s}^{-1}$$

$$\text{effective viscosity } \mu = \frac{1200}{G'^{(1-0.07)}} = 216 \text{ Pa} \cdot \text{s}$$

$$\text{viscosity ratio in mixer } \frac{0.5/216}{1000} = 2.3 \times 10^{-6}$$

$$\text{Reynolds number } Re = \frac{\rho DV}{\mu} = \frac{0.05 \times 0.2357 \times 900}{216} = 0.0491$$

$$\text{friction factor } f = \frac{16}{Re} = 325$$

$$\begin{aligned} \text{pressure drop empty pipe per meter} &= \frac{\Delta P}{L} = \frac{4f \rho V^2}{D} \\ &= \frac{4(325)}{0.050} \times 900 \times \frac{0.2357^2}{2} = 649\,987 \text{ Pa/m or } 6.5 \text{ bar/m} \end{aligned}$$

which is rather high.

From Table 7-5, the pressure drop for the SMX is 37.5 times larger. What length is required to get to an outlet coefficient of 5%? The starting CoV from the volume fraction is, from eq. (7-6),

$$\text{CoV} = \left( \frac{0.985}{0.015} \right)^{0.5} = 8.1$$

$$\text{CoV}_r = \frac{0.05}{8.1} = 0.0062$$

For equal viscosities, using Table 7-5 we have

$$\text{CoV}_r = 0.63^{L/D}$$

$$\ln 0.0062 = \frac{L}{D} + \ln 0.63 = 4.62$$

Add a length for the viscosity ratio, eq. (7-23):

$$1.0 \log(2.3 \times 10^6) = 6.36 \text{ additional length}$$

$$\text{total length} = 4.62 + 6.36 \text{ for } 11 \times 50 \text{ mm} = 0.549 \text{ m}$$

For the immiscible system, from Figure 7-27 and eq. (7-36):

$$\frac{\mu_d}{\mu_c} = \frac{100}{260 \times 1000} = 4 \times 10^4$$

The capillary number is

$$\begin{aligned} 1.0 &= d \frac{\mu_c G'}{\sigma} = d \times \frac{260 \times 301}{12 \times 10^{-3}} \\ &= 0.15 \text{ } \mu\text{m} \end{aligned}$$

Because of the length needed for blending the system should reach this value. The final pressure drop is  $6.5 \times 0.549 \times 37.5 = 133$  bar. We may decide that the pressure drop is too high and go to larger pipe, say 75 mm.

The mixing length for blending will not change in terms of  $L/D$  because there is no effect on blending of flow rate. The actual length will now be

$$11 \times 75 \text{ mm} = 0.825 \text{ m}$$

The new velocity, Reynolds number, and shear rate will be lower:

$$\begin{aligned} V &= 0.1053 \text{ m/s} \\ G' &= 64 \times \frac{0.1053}{0.075} = 89 \text{ s}^{-1} \\ \mu &= \frac{1200}{89^{0.3}} = 312 \text{ Pa} \cdot \text{s} \\ \text{Re} &= 0.075 \times 1053 \times \frac{900}{312} = 0.0228 \end{aligned}$$

The new pressure drop is

$$\begin{aligned} f &= \frac{16}{\text{Re}} = 702 \\ \frac{\Delta P}{L} &= \frac{4f \rho V^2}{D} = 4 \times \frac{702}{0.075} \times 900 \times \frac{0.1053^2}{2} \\ &= 1.87 \text{ bar per meter of empty pipe} \end{aligned}$$

For the static mixer,  $\Delta P = 0.825 \times 1.87 \times 37.5 = 58$  bar about one-half the pressure drop.

**Table 7-11** Process Stream Data

	Main Steam: Sour Gas	Side Stream: Dilute Caustic
Flow rate	280 000 scfh	44 gpm
Density (kg/m <sup>3</sup> )	4.97	1057
Viscosity (mPa · s)	0.02	1.8
Pressure (psig)	73	as required
Temperature (°F)	105	100

**Example 7-5: Gas Continuous Multiphase—Turbulent Dispersion.** A refinery sour gas stream containing 1.21 mol % hydrogen sulfide is to be scrubbed to reduce H<sub>2</sub>S to below 50 yppm. Since the flow rates are relatively low, it is decided that a dilute caustic solution is the most economical chemistry (Table 7-11). With fast chemical reaction, a single stage of contacting is all that is required to achieve desired contaminate removal. This is provided by spray nozzle feeding the caustic solution into a static mixer where the contacting/absorption and chemical reaction take place. Downstream of the static mixer a mist eliminator separates entrained liquid from the cleaned product gas. A 6 in. diameter line size and horizontal installation are preferred to best fit the available space. A pressure drop of 10 psi is available for the static mixer. A SMV design mixer consisting of 6 diameters packed length is determined to be required for the process. We are to determine the suitability of the 6 in. diameter mixer and the pressure drop.

The Froude number will be evaluated to determine the suitability of horizontal flow. The pressure drop will then be calculated for both the gas phase and liquid phase flowing separately in the empty pipe. The pressure drop for the combined flow streams will then be determined using the method of Lockhart and Martinelli. Finally, the mixer pressure drop will be determined using the multiplier of the empty pipe pressure drop reported for the SMV mixer specified.

Convert from standard conditions to flow at actual operating temperature and pressure using the ideal gas law. Standard conditions are 60°F and 1 atm in the gas process industry.

$$\begin{aligned}
 Q_{\text{actual}} &= Q_{\text{standard}} \frac{P_{\text{standard}}}{P_{\text{actual}}} \frac{T_{\text{actual}}}{T_{\text{standard}}} \\
 &= 280\,000 \times \frac{14.7}{14.7 + 73} \times \frac{460 + 105}{520} = 50\,992 \text{ acfm}
 \end{aligned}$$

or in metric, the actual gas flow rate is 0.4012 m<sup>3</sup>/s.

Six-inch schedule 40 pipe specified: ID = 6.06 in. or 0.154 m, and an open cross-section of 0.0186 m<sup>2</sup>. The gas velocity in open pipe flow is (0.4012 m<sup>3</sup>/s) / 0.0186 m<sup>2</sup> = 21.57 m/s. Calculate the Reynolds number using eq. (7-1) to confirm the flow regime:

$$\text{Re} = \frac{\rho DV}{\mu} = \frac{0.154 \times 21.57 \times 4.97}{2.0 \times 10^{-5}} = 825\,462 \quad \text{the flow is turbulent}$$

Check the Froude number using eq. (7-42) to see if stratification will be a problem in horizontal flow:

$$\begin{aligned} Fr &= \frac{\rho V^2}{\Delta\rho Dg} \\ &= \frac{4.97 \times 21.57^2}{(1150 - 4.97) \times 0.154 \times 9.81} \\ &= 1.34 \quad (<20 \text{ and too low for horizontal flow}) \end{aligned}$$

Since the Froude number is below 20, the mixer should be set in vertical orientation with flow down. The horizontal installation requested would be risky, especially at startup and turndown operation.

The friction factor is required to calculate the pressure drop. Use eq. (7-18) for turbulent flow:

$$f = \frac{0.079}{Re^{0.25}} = \frac{0.079}{825\,462^{0.25}} = 0.00259$$

The pressure drop for gas flow only in the empty pipe is then determined using eq. (7-17):

$$\begin{aligned} \Delta P_{\text{gas}} &= 4f \frac{\rho V^2 L}{2 D} \\ &= \frac{4(0.00259)(4.97)(21.57)^2(0.924)}{2 \times 0.154} = 71.87 \text{ Pa or } 0.00072 \text{ bar} \end{aligned}$$

Now in similar fashion, calculate the pressure drop for the liquid phase only. At a feed rate of 44 gal/min, or 0.00278 m<sup>3</sup>/s:

$$\text{velocity, } V = \frac{0.00278}{0.0186} = 0.149 \text{ m/s}$$

$$\text{Reynolds number, } Re = \frac{1150 \times 0.154 \times 0.149}{0.0018} = 14\,660$$

$$\text{friction factor, } f = \frac{0.079}{14\,660^{0.25}} = 0.0072$$

$$\begin{aligned} \text{pressure drop, } \Delta P_{\text{liquid}} &= \frac{4(0.0072)(1150)(0.149)^2(0.924)}{2 \times 0.154} \\ &= 2.2 \text{ Pa or } 0.000022 \text{ bar} \end{aligned}$$

For the combined stream pressure drop in empty pipe, we use eqs. (7-40) and (7-41):

$$X = \left( \frac{\Delta P_{\text{liq}}}{\Delta P_{\text{gas}}} \right)^{0.5} = \left( \frac{2.2}{71.87} \right)^{0.5} = 0.175$$

$$\begin{aligned} \Delta P_{\text{total}} &= \left( 1 + \frac{20}{X} + \frac{1}{X^2} \right) \times \Delta P_{\text{liq}} \quad [\text{using eq. (7-41a), liquid predominant}] \\ &= 1 + \frac{20}{0.175} + \frac{1}{0.175^2} \times 0.000022 = 0.0032 \text{ bar in empty pipe} \end{aligned}$$

or

$$\begin{aligned} \Delta P_{\text{total}} &= (1 + 20X + X^2) \times \Delta P_{\text{gas}} \quad [\text{using eq. (7-41b), gas predominant}] \\ &= (1 + 20 \times 0.175 + 0.175^2) \times 0.00072 = 0.0033 \text{ bar in empty pipe} \end{aligned}$$

In this case there is good agreement between the values calculated for total pressure drop. Since gas flow is predominant, the total empty pipe pressure drop calculated from the gas-only pressure drop using eq. (7-41b) should be used. It should be noted that this Lockhart–Martinelli correlation is considered to be conservative when used in vertical downward flow. The original work was all in horizontal flow.

Now for the SMV mixer in turbulent flow, we use eq. (7-16) and Table 7-6:

$$\Delta P_{\text{sm}} = K_T \Delta P_{\text{pipe}} = K_T(0.0033)$$

With  $K_T$  given as ranging from 100 to 200, the pressure drop is expected to be 0.33 to 0.66 bar, or in English units, 4.8 to 9.6 psi. This is less than the maximum 10 psi allowed, and the design is acceptable based on this preliminary calculation. The static mixer vendor should be consulted for more exact determination of pressure drop based on the specific SMV mixing element being used.

There are other comments to be made about this type of application. The liquid spray nozzle should supply liquid to the face of the mixing elements without appreciably wetting the vessel wall. A 30° full cone nozzle is typically used. Ideally, the spray should consist of droplets in the range 1000 to 2000  $\mu\text{m}$  range. A quick review of spray nozzle literature indicates that the appropriate spray nozzle for 44 gal/min of alkaline water would operate at about 100 psi pressure drop. Multiple nozzles could be used if a lower liquid-side pressure drop is desired. Fine atomization spray nozzles should be avoided since fine spray drops are difficult to separate in downstream mist eliminator equipment. A filter or strainer should be installed on the liquid feed to prevent plugging the feed nozzle, especially if the nozzle orifice size is small.

## 7-10 TRANSITIONAL FLOW

The previous discussion was devoted to processes that were either laminar or turbulent. The transition between these flow regimes is set by the Reynolds

number based on the pipe diameter. For a simple pipe the traditional number is 2100, but there is a large transition range. For motionless mixers the Reynolds number is usually much lower based on pressure drop. This is due to the much higher rates of energy dissipation due to the internals. The exact value depends on the mixer design but is in the 500 range for many. However, this transition is based on the change measured by pressure drop. The eddy structure starts to change at higher Reynolds numbers. There is very limited test work showing that the quality of turbulent mixing is poorer at low Reynolds number, due to these turbulence changes. Care should be taken in this region.

The Kenics HEV mixer, which consists of tabs, shows a transition in mixing performance at a very high Reynolds number. This is believed due to the change in vortex structure off the tabs at a specific tab Reynolds number rather than a pipe Reynolds number. Since the tab/diameter ratio is kept constant, this occurs at a higher pipe Reynolds number.

## 7-11 MOTIONLESS MIXERS: OTHER CONSIDERATIONS

### 7-11.1 Mixer Orientation

**7-11.1.1 Density Ratio Effects in Blending.** If the mixer is located vertically, there is little or no effect of density ratio between the added and main streams. If the mixer is horizontal, a density ratio could cause separation before mixing. This is most important in plain pipe, where a 10% density difference is reported to increase the mixing length tenfold. Vertical orientation is recommended by some manufacturers when the densimetric Froude number is less than 20 in turbulent flow situations, and the ratio of Froude to Reynolds number should be less than 1.0 in laminar systems. Criteria for orientation in the turbulent flow regime:

$$Fr' = \frac{\rho V^2}{\Delta \rho D_h g} < 20 \quad (7-42)$$

Criteria for orientation in the laminar flow regime:

$$\frac{Fr'}{Re} = \frac{\mu V}{\Delta \rho g D_h^2} < 1 \quad (7-43)$$

**7-11.1.2 Density Ratio Effects in Dispersion.** Again, the manufacturers recommend vertical arrangements. Horizontal arrangements will cause separation and increase coalescence, which will cut down on the interfacial area. The same criteria as used above in blending are therefore recommended.

### 7-11.2 Tailpipe/Downstream Effects

**7-11.2.1 Turbulent Blending.** The length downstream of the mixer is often referred to as the *tailpipe*. In laminar flow, no further mixing occurs in this region.

In turbulent flow, however, mixing continues as the extra turbulence generated by the mixing elements dies out. This effect lasts one or two diameters. For the HEV an extra factor of 2 in CoV reduction has been observed for SMVs; the range is from 1.5 to 2, depending on the number of plates in the mixer (Tauscher and Streiff, 1979). This observation suggests that the overall pressure drop can be minimized for a given mixing task by spacing mixers and empty pipe.

**7-11.2.2 Turbulent Dispersion: Coalescence.** After the dispersed phase leaves the motionless mixer, it will tend to coalesce to an equilibrium drop or bubble size characteristic of the shear field in the downstream piece of pipe. This coalescence is not just a phenomenon of the downstream tailpipe but is a process happening in parallel with dispersion. It is not as well understood. We do know that just like dispersion, coalescence is affected by volume concentration and is promoted by turbulence. Coalescence is strongly affected by surface chemistry effects. The role of many chemicals added to stabilize dispersions is to slow down the coalescence rate.

**7-11.2.3 Laminar Dispersion: De-mixing.** Often, motionless mixers are used to mix particulate solids into fluids flowing in the laminar flow regime. A high concentration slurry is mixed into a main stream and the mixers are to distribute the material across the diameter of the pipe (e.g., a pigment concentrate is added to a polymer line). This is a common and successful application of a mixer. However, because of the nature of laminar flow, discrete particles may move to agglomerate due to the velocity gradients in laminar flow. There is a radial diffusivity that causes particles to move away from the wall. This is usually not a problem [but for more information see the work of Acrivos and Leighton (1987)]. The parabolic profile has been a problem in some cases. Because of the differential velocity, particles on one streamline can catch up with those on another. This can lead to agglomerates, which can adversely affect downstream processing. This is often termed *Smolachowski agglomeration*. Agarwal et al. (1998) and Chimmili et al. (1998) have shown that in laminar flow the key variable can be combined as

$$\frac{d}{d_0} = \exp\left(\frac{-8fC_v G't}{3\pi}\right) \quad (7-44)$$

or

$$\frac{d}{d_0} = \exp\left(\frac{-8fC_v L}{3\pi D}\right) \quad (7-45)$$

where  $f$  is the efficiency of collision. Thus, this is a problem at high concentrations and with long pipes in laminar flow. It has been suggested that this value be kept lower than 1 (see Agarwal et al., 1998). To prevent such de-mixing agglomeration, extra motionless mixers are installed along long transfer lines.

### 7-11.3 Effect of Inlet Position

In laminar flow, vendors usually recommend coaxial centerline injection, often at the edge of an element. An interesting study was done by Hobbs et al. (1997, 1998) on the effect of injection position on degree of blending in laminar flow. It found that the initial injection position affected quality of mixing and that this effect was equivalent to several lengths to diameters of mixing, depending on mixer type. For a twisted mixer (KMS), nonoptimum addition could add the equivalent of four diameters of mixer to get the same CoV. For a cross-member mixer (SMX), nonoptimum costs about two diameters of length. For short mixers, both of these are significant reductions in mixing capability. Staged addition along the mixer axis may be required when adding large volumetric rates (>25%) of low viscosity fluids to high viscosity base materials in laminar flow. If the low viscosity phase should become continuous, the shear stress will be reduced and mixing will be reduced. When adding immiscible additives, consideration must be made to ensure that the mixer hydraulic diameter is of adequate size to prevent *flooding*, a term used to describe the low viscosity additive phase becoming continuous. Staged mixers with decreasing hydraulic diameter may be required to avoid flooding and also to achieve desired drop size. In turbulent flow evidence exists that injection is also very important, as the number of mixing elements are very low. Off-center injection and bends in front of turbulent mixers can drastically reduce the effect of the first element and thereby significantly reduce overall performance. Vendors' guidelines should be followed. If they cannot be followed, extra diameters of mixers should be added.

### 7-11.4 Scale-up for Motionless Mixers

In most cases motionless mixers can be designed based on fluid physical properties and process understanding without the need to run experiments on a small scale. The equations given above and vendor correlations will allow a large percentage of mixers to be designed without any scale-up or scale-down testing. However, in some cases data exist only on a small scale and the desire is to scale up to a larger processing capacity, achieving the same results as demonstrated on the smaller scale. An example of where small scale laboratory or pilot testing is required is when the fluid physical properties or the rate constants are unknown. Another is when the exact process result is unclear for example attrition or agglomeration or dispersion. Scale-up in heat transfer applications is a special case discussed in Chapter 14. In such cases the basic understanding discussed above gives guidance for scale-up.

We will use the flow rate ratio between big and small scale,  $R$ , as a scaling parameter and small and capital letters for the various variables:

$$R = \frac{Q}{q} = \frac{V}{v} \left( \frac{D}{d} \right)^2 \quad (7-46)$$



For operation in a laminar flow regime, we get the same time effects when we keep the residence time the same:

$$\frac{L}{V} = \frac{\ell}{v} \quad (7-47)$$

If we keep the shear rate the same in both scales, we also keep the shear stresses the same:

$$\frac{V}{D} = \frac{v}{d} \quad (7-48)$$

This leads immediately to keeping the length/diameter ratio the same, and

$$D = dR^{0.33} \quad (7-49)$$

and the pressure drops will be

$$\Delta P_{\text{large}} = \Delta P_{\text{small}} \quad (7-50)$$

since the pressure drop is proportional to velocity and length and inversely to diameter squared.

For the turbulent flow regime, the residence time is again held the same, and now the energy dissipation is held constant for any reactive or two-phase effects. This leads to keeping

$$\frac{L}{V} \text{ and } \frac{V^3}{D} \text{ constant} \quad (7-51)$$

Then

$$D = d^{3/7} \quad (7-52)$$

The length/diameter ratio is no longer constant.

In some cases there may be a flow regime change so that the Reynolds number, which always increases on scale-up, must always be compared on both scales. Scale-up is most reliable when both large and small scale systems are operating in the same flow regime. As mentioned in Section 7-10, care should be taken when operating in the transition regime.

## 7-12 IN-LINE MECHANICAL MIXERS

There are a number of cases in which mechanical mixers are put in line to promote mixing. Many years ago, small tanks (often made of pipe) and mechanical agitators were manufactured to be put in line. The residence times were low and the agitators of modest power with high speed. These devices could give very short blend times or high local energy dissipations. The internal flow is high compared to the through flow to avoid bypassing. Sometimes, staged vessels were used with horizontal baffles. In most cases these devices have been replaced by

motionless mixers when the pressure drop is available. The chief incentive was removing a rotational piece of equipment with its shaft and seals from a process. Maintenance is reduced and leakage eliminated with static mixers. This movement away from in-line mechanical mixers to static mixers is given momentum by the movement toward process chemistries operating at very high pressure, 1000 psi and higher.

### 7-12.1 Rotor–Stator

Still very popular for in-line dispersion are the class of rotor–stator mixers. These devices look more like pumps than like stirred tanks. Volumes are small but rotational speeds and powers are high, giving high local energy dissipation. They are often staged with several rotors separated by stators that reduce bypassing.

The literature is small on rotor–stator devices, as discussed in Chapter 8. [They are also discussed in Chapter 12, since this is the process result (liquid–liquid and solid–liquid) that these devices are most used for.] It should be mentioned that because of their very high speed they can produce turbulent motion in some rather high viscosity fluids. Also see Cohen (1998), Dietsche (1998), and Myers et al. (1999) for general information about the industrial application of rotor–stator mixers versus other mixing options.

### 7-12.2 Extruders

Extruders are sometimes used as mixers. But the primary purpose of most extruders is to melt polymer pellets and to increase the pressure in the melt (i.e., to pump). Neither of these steps requires a lot of radial mixing. What radial mixing may occur is secondary. In particular, single-screw extruders generate little radial mixing. Some radial mixing occurs in the melting zone, but this is very hard to predict. If material is injected into the single-screw flights, very little radial mixing is found. Twin-screw extruders, however, can show some radial mixing, due to their configuration, which promotes radial mixing. Attempts to quantify this type of mixing have not been as successful as with motionless mixers. For more information, consult a good polymer processing text such as Tadmor and Manas-Zloczower (1994). Specially designed twin-screw extruders (both co-rotating and counterrotating) are used for compounding (mixing), cooling (heat transfer), reaction (mixing and plug flow), and devolatilization (heating). See Biesenberger (1983) for more on the subject of devolatilization. In all cases, static mixers are attractive alternatives. For the topic of *compounding*, a fancy name for mixing and dispersion in polymers, see Tadmor and Manas-Zloczower (1994). Many mechanical dispersion processes are in commercial operation, and others are continually being developed. Manufacturers of static mixing equipment have varying degrees of experience in these areas of application in polymer processing and should be interviewed if there is interest in this growing area of activity.

## 7-13 OTHER PROCESS RESULTS

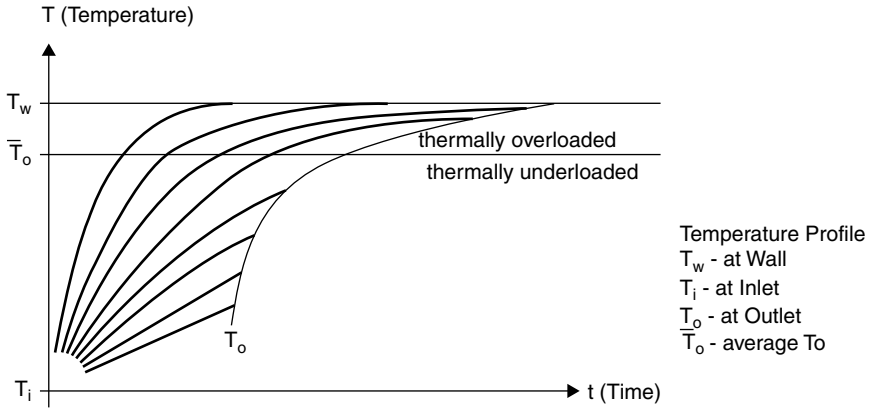
### 7-13.1 Heat Transfer

**7-13.1.1 Turbulent Flow Heat Transfer in Pipes.** In turbulent flows in pipes, the relations for heat transfer to or from the fluid and the wall are well known and are the basis of many heat exchanger designs. The overall heat transfer resistance consists of a resistance on the cooling or heating side, a wall resistance, and a fluid resistance inside the pipe. The latter depends on the nature of the fluid and the flow regime and is the one of primary interest in this section. As is common practice we will use the individual heat transfer concept to describe resistances. For a detailed discussion, see any heat transfer text (e.g., Burmeister, 1983a,b). Static mixers and other pipe inserts are not typically justified in turbulent flow, since the cost and added pressure drop are high relative to the benefit achieved. However, in some cases, such as condensing, flow turbulence promoters in the form of long, spiral internals are added to promote film forming (Burmeister, 1983a,b). A general form of the equations for predicting inside or process heat transfer coefficient is given by

$$\text{Nu} = A \cdot \text{Re}^a \cdot \text{Pr}^b \left( \frac{\mu}{\mu_w} \right)^c \quad (7-53)$$

where, typically,  $A = 1$  to  $2$ ,  $a = 0.66$ ,  $b = 0.33$ , and  $c = 0.14$ . Note that even though the flow is turbulent, there is still a significant viscosity effect. The addition of the viscosity ratio is empirical and based on the observation that when cooling, the heat transfer coefficients are lower than when heating. This adjustment is needed because heat transfer takes place in a wall film and the conditions of this film are better characterized by local properties at the local temperature. In general, the viscosity is much more sensitive to the temperature than are any of the other physical properties.

**7-13.1.2 Laminar Flow Heat Transfer in Pipes.** In laminar flow the heat transfer rates are greatly reduced. There is no radial flow and temperature gradients build up, reducing the heat transfer rate. In addition, the parabolic velocity distribution causes the center to spend little time in the heating zone and the fluid near the walls to spend more time. The overall effect is that the local heat exchange coefficient becomes a function of length and can actually approach zero at long lengths. In other words, heating in laminar flow is very poor and uniform. For example, when heating in an empty pipe, the flow can be viewed as a series of streamlines, as shown in Figure 7-29. Material near the center of the pipe flows fast and heats up slowly compared to material flowing near the pipe wall. At the outlet the product is a mixture of material with drastically different time, temperature, and shear histories. There exists a thermally overloaded zone near the wall and a thermally underloaded zone near the center.



**Figure 7-29** Thermal loading profile for laminar flow heat transfer in empty pipe. (From Heierle, 1989.)

Performance problems related to maldistribution also exist in cooling applications, especially where viscosity increases as temperature is lowered. At worst, case equipment could become inoperable, due to plugging of all but the center of the flow channel. This condition can be eliminated by the use of static mixer internals, discussed later. Heat transfer coefficients for laminar flow in empty pipe are correlated by

$$Nu = A \cdot Re^a \cdot Pr^b \left( \frac{\mu}{\mu_w} \right)^c \left( \frac{D}{L} \right)^d \tag{7-54}$$

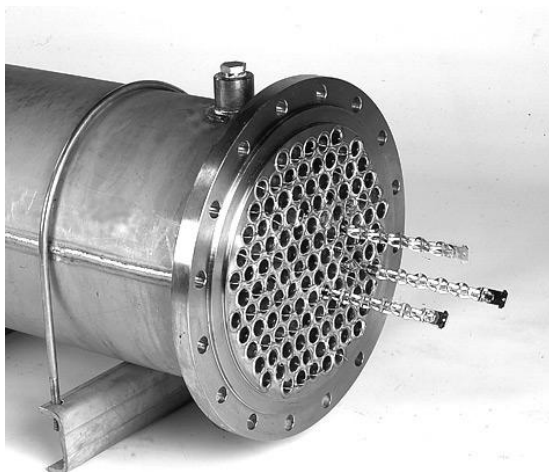
where  $A = 1.61$ ,  $a = 0.33$ ,  $b = 0.33$ ,  $c = 0$  to  $0.14$ , and  $d = 0.33$ . This equation has a limit at a Nusselt number of about 3.7. Higher Nusselt numbers are achieved only at short lengths, providing only limited heat transfer surface.

In laminar flow where pressure drops are high, there is viscous heating, which is on the order of  $1^\circ\text{C}$  per 10 bar pressure drop for most polymers. This temperature rise is quite nonuniform, causing radial temperature gradients of several degrees to appear. In addition, extruders and pumps also generate thermal nonuniformity and radial temperature gradients. Even multifold heat exchangers show large radial temperature gradients. Radial temperature gradients up to  $10^\circ$  or more are not uncommon.

Since the static or motionless mixer promotes radial flow of both momentum and heat, it will significantly enhance the heat transfer rate. Figure 7-30 shows static mixer heat transfer packing installed in the tubes of a multitube heat exchanger.

The same form of the heat transfer equation is used:

$$Nu = A(Re \cdot Pr)^a \left( \frac{\mu}{\mu_w} \right)^b \left( \frac{D}{L} \right)^c \tag{7-55}$$



**Figure 7-30** Static mixer for heat transfer enhancement in a multitube heat exchanger (SMXL). (Courtesy of Koch-Glitsch, LP.)

where  $A$  depends on vendor correlation,  $a =$  depends on vendor correlation,  $b = 0.14$ , and  $c =$  depends on vendor correlation. Note that there is now no effect of viscosity. The mechanism is one of surface renewal and differs from purely convective flow in that it is unaffected by film effects. It is also debatable whether there is an actual  $D/L$  effect, as there is continual surface renewal. There is few experimental data, and it appears that the laminar form was used by some vendors with a fixed  $L/D$ . Most vendors do not include length in their correlation for laminar flow heat transfer. Table 7-12 gives some typical values for constants  $A$  and  $a$  used in eq. (7-55) (Streiff, 1986). Note also that sometimes the constant  $a$  is set to 0.33 and  $A$  is restated for comparison of data from different sources.

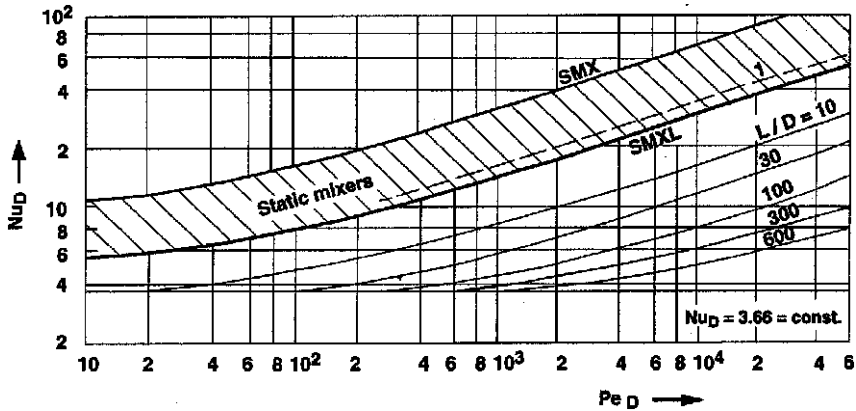
The increase in heat transfer resulting from surface renewal related to radial mixing is significant. Figure 7-31 compares the laminar flow heat transfer coefficient achieved with the type SMX and SMXL mixing element with that of an empty pipe of the same diameter (Heierle, 1988). A four to tenfold increase in

**Table 7-12** Heat Transfer Coefficients for Established Flow

	$A$	$a^a$
Kenics KMS		
Edge seal	1.5	0.33
No edge seal	2.25	0.33
SMX	2.6	0.35
SMXL	0.98	0.38

<sup>a</sup>See eq. (7-55).

Source: Streiff (1986).



**Figure 7-31** Heat transfer to the inner tube wall with laminar flow in empty pipe and static mixers. (From Heierle, 1989.)

heat transfer is shown. A slight added benefit could be obtained by conduction of heat through the element blade (called the *fin effect*), but only if the mixing element is securely sealed to the pipe wall. Also note that for SMX and SMXL mixing elements, unlike empty pipe and other heat transfer enhancements, the heat transfer coefficient is not a function of the diameter/length ratio.

An interesting motionless mixer heat exchanger is the SMR, where the cross members that provide the radial mixing are actually heat exchanger tubes with heat transfer medium flowing in them. This increases the heat exchange area per unit length five to tenfold. Heat transfer surface per unit volume can be maintained as equipment volume increases, making the design very attractive for scale-up of reactor processes requiring precise temperature control. See Figure 7-32 for a picture of the SMR mixer-heat exchanger-reactor. The radial mixing is good, giving high heat transfer coefficients (see Table 7-2). SMR mixers do not produce radial temperature uniformity on a fine scale. They are often followed by a short length of smaller conventional mixer to give fine scale radial temperature uniformity.

### 7-13.1.3 Notes Regarding the Scale-up of Heat Exchangers Containing Static Mixer Heat Transfer Packing.

Conventional static mixers such as the SMX, SMXL, and KMS types accomplish heating and cooling by the transfer of heat through the vessel wall. This limits scale-up as a single tube design. The heat transfer coefficient decreases in inverse proportion, while product volume and total heat load increase as the second power with increasing pipe diameter. Therefore, the volume-related heat transfer capacity is very small with larger diameters. Pipe diameters can be increased only up to about a nominal 100 mm (4 in.) maximum. With high throughputs and/or heat load, the required pipe length, and consequently pressure drop, would become excessive.

Multitube designs offer a partial solution to the limitation above. Tube diameters can be maintained small and high volume related heat transfer capacity



**Figure 7-32** Static mixer with internal heat transfer surface (SMR). (Courtesy of Koch-Glitsch, LP.)

achieved at high throughput when the flow is divided into a large number of parallel flow paths. High capacity and uniform product treatment are achieved, but only when flow is divided equally to all tubes. It is important to recognize that the mixing elements achieve their effect only within the individual tubes treating partial flows. With isolated flow paths, there is no radial mixing effect over the entire product stream as with the monotube design. This is proved not to be a problem in heating applications where the viscosity decreases as temperature is increased. At a moderate pressure drop, a properly designed multitube viscous heat exchanger is self-correcting under conditions of decreasing viscosity. Size (number of parallel tubes) can be increased to handle whatever maximum heat load and throughput are required.

Heat exchanger scale-up is much more difficult for cases where the viscosity increases during the process. This is the case in many viscous cooling applications and for polymerization reactions where viscosity can increase by several orders of magnitude. The *maldistribution effect* in multitube designs occurs when one or several of the parallel tubes behave a little differently than the others. There are numerous causes for this. Maybe there is a maldistribution in the feed stream; or resistance to flow may be minimally different due to slightly different surface roughness or slightly different diameter within tolerance limits; or heat transfer may be slightly different due to minor differences in coolant flow around the tube. If for some reason an individual tube cools a little faster than the others, its effective viscosity will increase at a greater rate, causing higher resistance to flow. Since the pressure drop must be the same over all parallel tubes, the flow through the offending tube will decrease. This results in a longer residence time, an even lower temperature, and still higher viscosity. At greatly diminished flow, the tube could eventually reach the temperature of the cooling fluid, and if

viscosity at this temperature is high enough, the tube is essentially plugged. Meanwhile, at constant overall volumetric throughput flow has been diverted to the unaffected tubes, resulting in higher velocities, less heat transfer, and shorter residence time. Multiple steady states are established. The situation always results in reduced cooling performance, a very wide residence time spectrum, and in many cases, a higher pressure drop than calculated originally. The design of multitube equipment for viscous cooling or polymerization reaction involving increasing viscosity requires thorough analysis of the rate of viscosity increase versus the flow rate. In many cases, very high pressure drop becomes necessary to avoid creation of a multiple steady-state situation. A single flow channel is highly desirable for heat transfer/reaction processes with steeply increasing viscosity. Isolated flow paths are avoided and radial exchange of material ensures uniform product treatment. This is achieved in the monotube SMX, SMXL, or KMS heat exchanger, but only at small throughput, due to limited availability of the heat transfer surface. At high throughput, special features such as those provided by the SMR design are required. The SMR eliminates the requirement to transfer heat through the wall of the containing vessel. Heat transfer occurs throughout the structure of the mixing element, thus providing a very high volume-related heat transfer capacity (Streiff, 1986), as shown in Figure 7-33. The SMR structure achieves mixing, boosts the viscous heat transfer coefficient, and achieves plug flow similar to that of SMX and SMXL mixing elements. A very low temperature driving force and low pressure drop operation are possible. The SMR is recommended for scale-up when viscosities are steeply increasing and when extremely uniform product treatment is required.

### 7-13.2 Mass Transfer

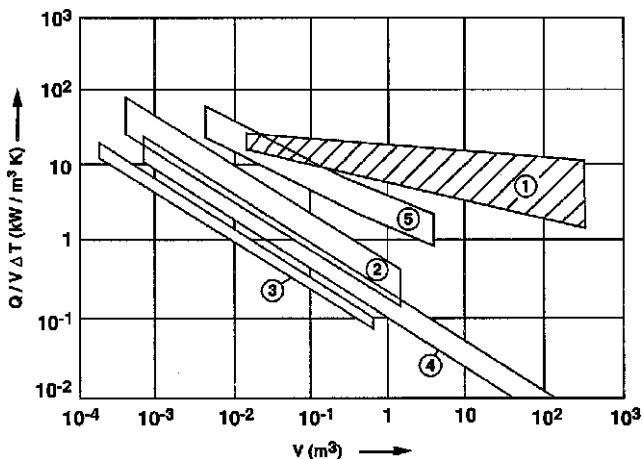
Given the drop or bubble size in laminar or turbulent flow, the mass transfer coefficient can be estimated. For turbulent flow, the techniques in Chapter 11 will be applicable. Experience shows that in such cases mass transfer is very fast and that the equivalent of an equilibrium stage (90 to 99% of equilibrium) is reached in less than 10 diameters. Static mixers are beneficial not only in creating the drop dispersion but also in maintaining the drop size distribution and a high degree of turbulence, promoting exchange at the interface.

**7-13.2.1 Chemical Reactor.** Motionless mixers and pipelines are often used as chemical reactors, particularly for fast reactions. Chapter 13 is recommended for a discussion of these aspects. Of particular note is the work of Baldyga et al. (1997), in which different mixers were found to give different energy efficiencies for reactive mixing.

**7-13.2.2 Plug Flow Characteristics.** In turbulent flow the residence time distribution (RTD) is close to plug, but in laminar flow the RTD results from the parabolic velocity distribution are quite skewed. Middleman (1977) gives for the RTD in laminar flow

$$F(t) = 1 - 0.25 \left( \frac{t}{t_{\text{ave}}} \right)^{-2} \quad (7-56)$$





**Figure 7-33** Heat transfer capacity per unit volume of SMR and other chemical reactors: (1) SMR; (2) static mixer; (3) empty pipe; (4) stirred tank; (5) extruder.

where  $F(t)$  is the fraction of material leaving the pipe before a given time  $t$ . Notice that no fluid leaves a pipe before one-half the residence time and that at the average residence time half the material has left.  $F(t)$  can also be thought of as the well-mixed normalized concentration exiting the pipe. In actuality the flow leaving a pipe is radially distributed, but  $F(t)$  as a normalized concentration is the concentration it would be if there was instantaneous mixing at the end of the pipe. This distribution can cause many problems in treatment and reactions in laminar flow. In most cases, plug flow where all parts of the fluid spend the same time in the pipe is preferred.

Static mixing technology is very applicable to slow chemical reactions requiring plug flow for long duration (see Streiff and Rogers, 1994). Plug flow is achievable with SMX and SMXL static mixers, and with the SMR when precise control of temperature during reaction is also important. Mixing and heat transfer criteria have already been discussed. Reaction presents additional requirements and constraints on the design. First laboratory, and then pilot testing is almost always necessary to establish full scale reactor design.

The static mixer reactor is preferably a monotube design. Multitube configuration presents high potential for reactor instability and nonperformance due to tube-tube variation, as described previously. This is especially true when attempting to control high viscosity exothermic polymerization reactions (Nguyen et al., 1984b; Cusack, 1999).

Commercial scale tubular plug flow reactors utilizing SMX and SMXL static mixer internals have now been operating for more than 20 years. These static mixer designs have demonstrated their ability to mix and achieve plug flow in large scale equipment. The residence time distribution and shear stress-temperature history are very uniform. The Bodenstein number (also sometimes known

as a Péclet number) is a measure of the width of the RTD:

$$Bo = \frac{LV}{D_{ax}} \quad (7-57)$$

The Bodenstein number ( $Bo$ ) is equal to zero for an ideal continuously stirred tank reactor (CSTR, backmixed reactor) and infinity for perfect plug flow as shown in Figure 7-34. For a cascade of  $j$  continuously stirred tanks in series,  $Bo = 2j$  (Levenspiel, 1967). Measurements on the SMX and SMXL static mixers in nominal 100 mm (4 in.) diameter pipe at a Reynolds number of 0.1 gave values of the Bodenstein number ranging from 50 to 100 per meter length, corresponding to a very high degree of plug flow. According to Streiff et al. (1999), well-designed static plug flow tubular reactors have residence times corresponding to greater than 30 CSTRs ( $Bo > 60$ ). Tracer studies conducted during the development of a tubular reactor design for the continuous bulk polymerization of polystyrene showed flow through an SMX packed reactor to be equivalent to a large number of continuously stirred tanks in series (Nguyen et al., 1984a).

Additionally, plug flow requires the reduction or elimination of both localized and large scale recirculation. These phenomena can originate within an improperly designed reactor as a result of temperature and density gradients occurring in the feed or generated during the reaction. Reactor performance is affected by both fluid viscosity and velocity. Parametric testing during pilot scale evaluation should be designed to identify a range of performance and points of instability. Design symmetry with regard to both the shape and size of the hydraulic flow channel should be maintained as best possible on scale-up. With large units, inlet and outlet conditions are also important. Flow simulation models are now starting to be used to evaluate these requirements.

When heat transfer is required for initiation and/or control of a viscous reaction, the SMR mixer-reactor is used. It offers, for example, a unique solution to the difficult problem of laminar flow heat transfer during exothermic polymerization reactions. As described previously, it provides a large area of heat

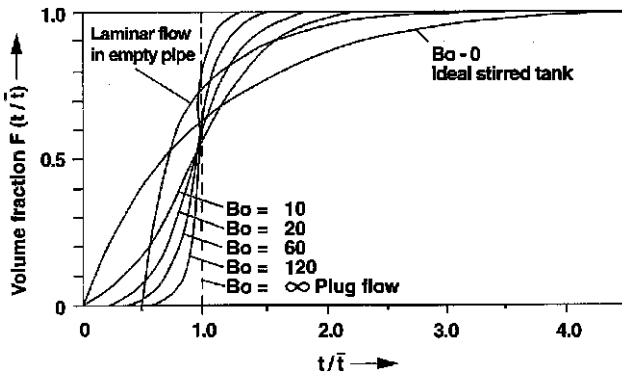


Figure 7-34 Residence time distribution.

transfer surface throughout the reacting mass as well as accomplishing mixing and maintaining plug flow. Reaction temperatures can be controlled with very small transverse temperature gradients. A near-isothermal reaction process is possible (Craig, 1987).

## 7-14 SUMMARY AND FUTURE DEVELOPMENTS

From the information presented in this chapter it can be seen that for a limited but significant number of cases, an in-line mixer and particularly a static mixer offers a way to achieve many process results. In addition, there are a large number of theory-based empirical relations based on academic and vendor correlation for calculating designs to achieve blending, dispersion, and heat transfer process results. In many cases the methods are more accurate than the available basic data. Future vendor designs will target new applications and be aimed at achieving a given process result with less pressure drop or length.

Uncertainties in the current correlations for in-line mixers are similar to those in other fields of mixing. They involve systems where the physical properties of the streams to be mixed are very different and the understanding of the physics is poor. For example:

- The role of coalescence in gas–liquid and liquid–liquid drop size prediction
- Rate processes in making dispersions
- Blending of miscible systems with large viscosity ratio

Recent unpublished work suggests that the effect of feed ratio in turbulent flow is not as strong as predicted by the coefficient of variation method. It is hoped that computational fluid mechanics will increase our knowledge in this and other areas of pipeline mixing.

As more of the fundamental physics and physical chemistry of such processes becomes understood it will be incorporated into the rather large existing knowledge base. Work is just starting to appear on the effect of non-Newtonian fluids on pressure drop and blending. Because of the simple geometry and the uniformity of the shear stresses, both laminar and turbulent and the controlled limitation of bypassing and tight RTD, the motionless static mixer may be a better device to study some of these phenomena than the conventional stirred tank with its wide variability of shear stress time history.

## ACKNOWLEDGMENTS

The authors gratefully acknowledge Chris Wolfe for her work on the examples.

## NOMENCLATURE

- |   |   |
|---|---|
| a | exponent in heat transfer correlations            |
| A | preexponential term in heat transfer calculations |

b	exponent in heat transfer correlations
Bo	Bodenstein number, $Bo = LV/D_{ax}$
c	exponent in heat transfer correlations
Ca	capillary number, $Ca = d\mu G'/\sigma$
CoV	coefficient of variation
CoV <sub>0</sub>	initial coefficient of variation
CoV <sub>r</sub>	relative CoV reduction
C <sub>p</sub>	specific heat
C <sub>v</sub>	volumetric concentration of dispersed phase
d	exponent in heat transfer correlations
d	drop or bubble diameter
d <sub>max</sub>	maximum drop or bubble diameter
D, d	pipe diameter
D <sub>ax</sub>	axial dispersion coefficient
f, f'	friction factor
f <sub>s</sub>	friction factor for smooth pipe
g	gravitational acceleration
G'	shear rate
h	heat transfer coefficient
k	thermal conductivity
K, k	arbitrary constants
K <sub>L</sub> , K <sub>T</sub>	pressure drop ratios for motionless mixers
KiT, KiL	mixing rate coefficient for blending
K <sub>G</sub>	shear rate constant
L, l	length variable
Ne	Newton number, $Ne = (\Delta P/\rho V^2)(D/L)$
Nu	Nusselt number, $Nu = hD/k$
Oh	Ohnesorge number, $Oh = (We)/Re^{0.5} = \mu_d/(\rho_c \sigma d)^{0.5}$
$\Delta P$	pressure drop
$\Delta P_{tp}$	pressure drop two-phase (gas–liquid) flow
Pr	Prandtl number, $Pr = C_p \mu/k$
q	viscosity ratio—dispersed to continuous
Q	volumetric flow rate
Re	Reynolds number, $Re = \rho DV/\mu$
V, v	velocity
Vi	viscosity number, $Vi = \mu_d(V/\sigma)(\rho_c/\rho_d)^{0.5}$
We	Weber number, $We = D(\rho V^2/\sigma)$

### Greek Symbols

$\varepsilon$	energy intensity or power per unit mass
$\mu$	viscosity
$\mu_c$	continuous phase viscosity
$\mu_d$	dispersed phase viscosity
$\mu_w$	viscosity calculated at wall temperature
$\sigma$	interfacial or surface tension

$\rho$	density of the fluid
$\rho_c$	continuous phase density
$\rho_d$	dispersed phase density
$\Delta\rho$	density difference

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# Rotor–Stator Mixing Devices

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## 8-1 INTRODUCTION

### 8-1.1 Characteristics of Rotor–Stator Mixers

The distinguishing feature of a rotor–stator mixer is a high-speed rotor (the driven mixing element) in close proximity to a stator (the fixed mixing element). Typical rotor tip speeds range from 10 to 50 m/s. They are also called *high-shear devices* because the local energy dissipation and shear rates generated in these devices are much higher than in a conventional mechanically stirred vessel. In a rotor–stator mixer the shear rate ranges from 20 000 to 100 000 s<sup>-1</sup>. The local energy dissipation may be three orders of magnitude greater than in a conventional mechanically agitated vessel. High speed, high shear, and higher power are the main characteristics of rotor–stator mixers.

The action of the rotor and stator together generates the mixing energy, shear and elongational stresses, turbulence, and cavitation (in various proportions depending on the speed, viscosity, and other fluid flow parameters), which provide the mixing or size reduction. Thus, the rotor–stator assembly is often called a *generator*.

Commercially available rotor–stator mixers range in size from small lab units to large production units capable of flow rates of 1000 gal/min or more, driven by drives with power greater than 100 hp. There are many geometric variations in rotor–stator generator design. They may, however, be classified into three main geometric groups whose features are described in Section 8-2.



Rotor–stator generators may be assembled or configured for either batch, semibatch, or continuous operation. They may also be combined in various ways and with conventional mechanical agitation for applications requiring high shear and high local energy dissipation. A few of these configurations are also described in Section 8-2.

### 8-1.2 Applications of Rotor–Stator Mixers

Rotor–stator mixers are used in the chemical, biochemical, agricultural, cosmetics, and food-processing industries. They are employed in many process operations that involve:

- Homogenization
- Dispersion
- Emulsification
- Grinding
- Dissolving
- Chemical reaction
- Cell disruption
- Coagulation (due to shear)

Their major use includes the production of latexes, adhesives, personal care and cleaning products, dispersion and microdispersions of chemicals, and agricultural pesticide formulations. These dispersions, in general, have a viscosity less than  $150 \text{ Pa} \cdot \text{s}$  ( $150\,000 \text{ cP}$ ). When the viscosity of the fluids or dispersion is much greater than this, extruders are employed instead of rotor–stator mixers.

A single rotor–stator mixer can be used to accomplish several of the above operations within a single process. This makes the rotor–stator an invaluable piece of processing equipment. An example is the grinding and subsequent dissolution of “whole bales of rubber into a solvent” using the Silverson BE2500 Disintegrating/Dissolver Plant (Silverson Machines, 2002). Here, the grinding and mixing needed to enhance dissolution occur simultaneously. These mixers can save time due to the highly localized energy and high shear rate that they input to the system. Although the ability to accomplish multiple operations is a benefit, it also makes rotor–stator mixers difficult to understand. Each task may be governed by a different hydrodynamic or operational variable that may be competitive with respect to scale-up and operation.

### 8-1.3 Summary of Current Knowledge

Despite their widespread use, the current understanding of rotor–stator devices has almost no fundamental basis. There are few theories by which to predict, or systematic experimental protocols by which to assess, the performance of these mixers. In fact, there are very few archival publications on rotor–stator

processing. Furthermore, the equipment is developed and manufactured by small, highly competitive and highly specialized companies in an environment that is not conducive to the development of a generalized knowledge base. Consequently, process development, scale-up, and operation are done mostly through engineering judgment and trial and error instead of through sound engineering principles. This leads to a multitude of problems during the startup of a process, which inevitably result in lost time to market as well as increased costs. Often, extensive qualification programs are needed to ensure that pilot and manufacturing scale processes make the same product. In recent years there has been some activity of a fundamental nature, some of which is discussed below.

**8-1.3.1 Key Issues and Current Research Efforts.** Key questions relating to the use of rotor–stator mixers include:

- When should one consider a rotor–stator mixer?
- What design(s) and/or configuration(s) should be selected, and why?
- What criteria should be used for scale-up and/or design?

There are no straightforward answers to these questions at the moment. Users currently depend mostly on manufacturers or suppliers and their publications for information on the performance characteristics, scale-up, and design of these devices. Determining what is an appropriate rotor–stator design and operating conditions for a particular process application requires extensive lab, pilot, or plant trials. Information gathered may then be used for scale-up and design. For some processes, especially in the food and cosmetic industries, prior experience may point to a specific rotor–stator design, but the relative efficiency to other designs may still be unknown.

The need for a fundamental characterization of the performance of these devices and the criteria for scale-up and design are well recognized. In response to this need, two consortia, the British Hydromechanics Research Group (BHRG) and the High Shear Mixing Research Program at the University of Maryland, have embarked upon systematic study and characterization of rotor–stator mixers.

BHRG began work in this area as part of the HILINE program in the late 1980s. Most of the research has focused on power-draw measurements, residence time distribution tests, and reactive mixing studies in a rotor–stator device using low viscosity fluids. Very little work has been done using multiphase and high viscosity fluids. Work done so far has shown that:

1. Only a small fraction of the total power input is effective in mixing.
2. Both the shaft power and power losses increase with increasing rotor speed, larger flow rate, higher viscosity, and smaller shear gap.
3. The effect of the width of the shear gap on power draw is weak.
4. A preliminary test with a set of fast competitive reactions in a single-stage toothed rotor–stator mixer revealed significant fluid bypassing of the

region of high turbulence intensity, making the mixer ineffective as a high intensity reactor. However, the extent to which the use of multiple rows of rotor and stator teeth would change the situation was not considered.

Most of the BHRG results are contained in proprietary reports that are available only to consortium members.

The High Shear Mixing Research Program (HSMRP) at the University of Maryland has focused on two fundamental aspects. The first is developing a mechanistic understanding of the governing fluid dynamics occurring in a continuous in-line rotor–stator mixer with a single row of rotor and stator teeth. Both computational fluid dynamics simulations and velocity field measurements via laser Doppler anemometry have been performed. The second involves monitoring dilute liquid–liquid dispersions in batch mixers for insights into the physics of drop breakup in this type of rotor–stator design. These studies have examined the effect of rotor speed, gap width, power draw, and the geometry of the openings in the stator head on the drop size distribution (DSD). Since one of the coauthors of this chapter heads the HSMRP consortium, some of the results are reported below and in Chapter 12.

## 8-2 GEOMETRY AND DESIGN CONFIGURATIONS

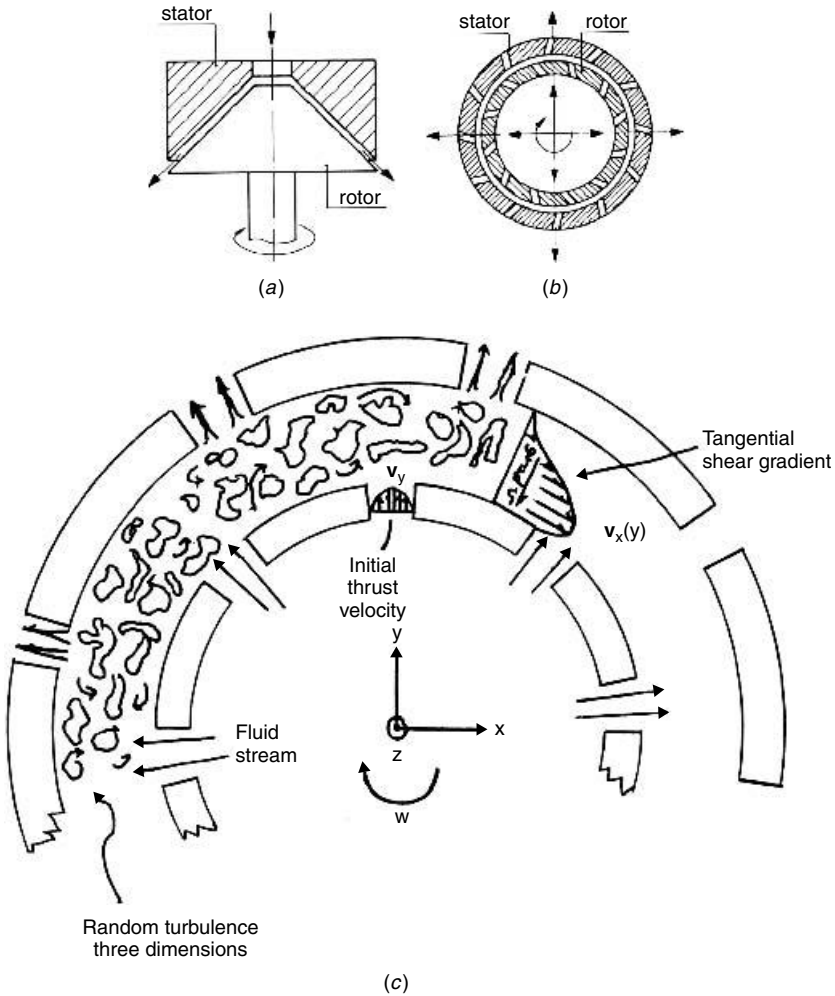
Numerous companies design and supply rotor–stator mixers. Many of the available designs often differ only slightly in geometry, although their suppliers make vastly different claims of performance. In this section the main rotor–stator geometries are described briefly and illustrated with figures. Several configurations for batch or continuous operation are also described and illustrated.

### 8-2.1 Colloid Mills and Toothed Devices

Figure 8-1*a* represents the simplest rotor–stator geometry. It is a conical couette device, called a *colloid mill*. There are many design variations of this basic couette geometry. Figure 8-1*b* represents a slightly more complex variation, with slots or teeth built into both the rotor and stator pieces to provide multiple channels for the flow of the fluid through the device. A schematic representation of the hydrodynamics is shown in Figure 8-1*c*. A commercial colloid mill with groves in the rotor and stator is shown in Figure 8-2*a*. Commercial toothed designs often have multistage assemblies of two or more sets of rotor–stator teeth in a mixing head, as shown in Figure 8-2*b*.

### 8-2.2 Radial Discharge Impeller

Another popular rotor–stator design is the Silverson Machines or Ross type. The rotor is a radial impeller that rotates inside a stationary housing with slots as shown in Figure 8-3*a*. The rotor moves the fluid radially out of the mixer head through the slots or holes in the stator. Superimposed on the radial flow is a



**Figure 8-1** Couette rotor–stator geometry, single-stage design: (a) simple Couette; (b) toothed rotor–stator; (c) schematic representation of hydrodynamics; (d) commercial example of a toothed rotor–stator device.

tangential shear flow inside the stator. Both Silverston Machines and Ross offer a variety of slot shapes and sizes, as shown in Figure 8-3b–d.

### 8-2.3 Axial Discharge Impeller

Chemineer Greenco offers a different rotor–stator geometry. In this case, as shown in Figure 8-4a and b, the rotor is an axial impeller that pushes the fluid axially through holes bored into the stator. The rotating action of the impeller, however, creates some tangential shear flow inside the stator. There is a two-stage version called the *tandem shear pipeline mixer*. This consists of a primary disperser

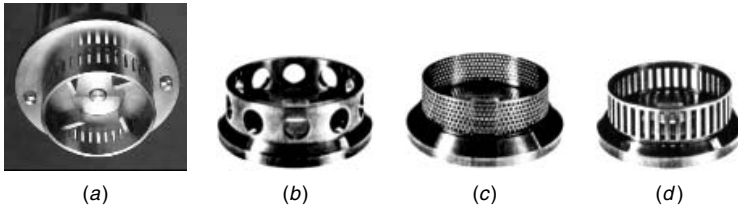


(a)

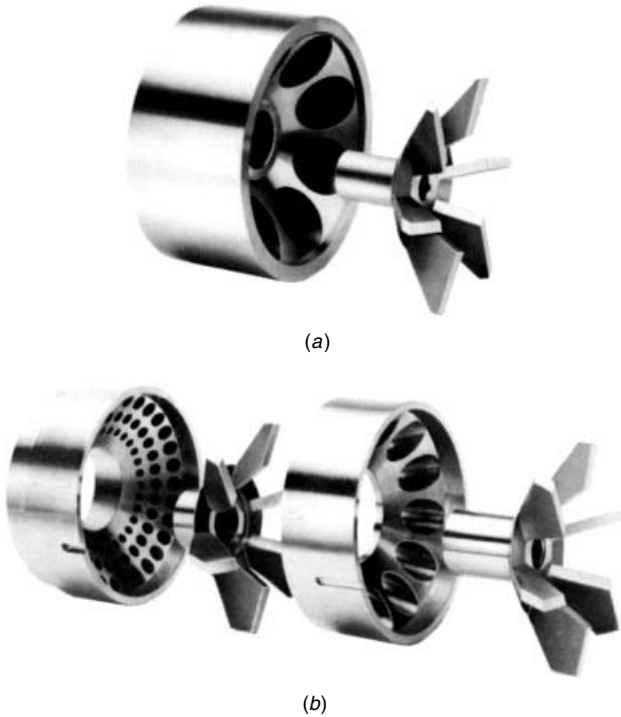


(b)

**Figure 8-2** Commercial devices: (a) grooved colloid mill; (b) IKA Works multistage rotor-stator design.



**Figure 8-3** Silverson Machines or Ross rotor–stator design: (a) head assembly; (b) general purpose disintegrating head; (c) high-shear screen; (d) slotted-screen disintegrating head.



**Figure 8-4** Chemineer Greerco rotor–stator design: (a) single-stage design; (b) two-stage design.

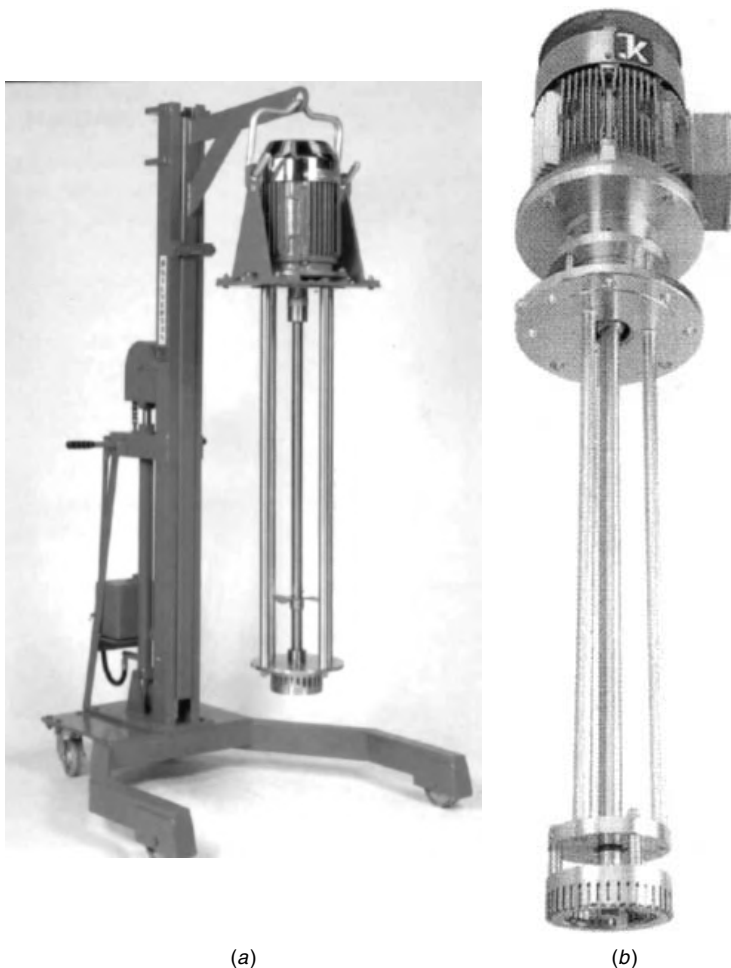
with larger holes followed by a second rotor–stator head with smaller holes in the stator. Another notable feature of this design is that the gap width between the rotor and stator may be adjusted in the field by installing washers or bushings.

**8-2.4 Mode of Operation**

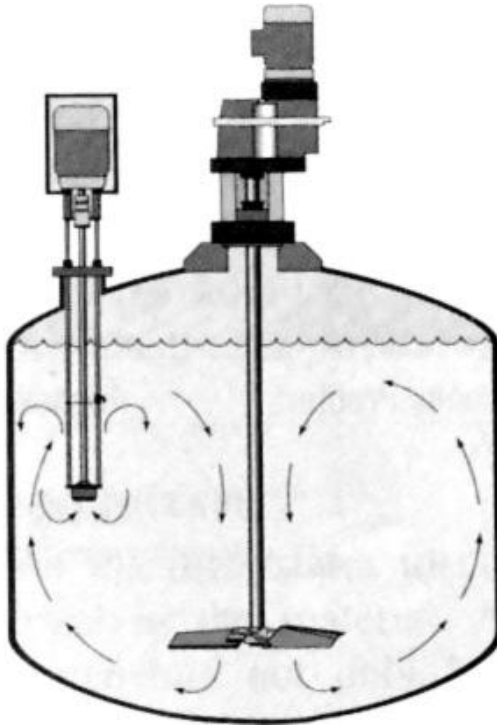
Any of the mixing heads described above can be assembled and configured for batch semibatch or continuous operation. For a batch or semibatch operation,

the mixing head is attached to a long drive shaft and supporting rods, as shown in Figure 8-5*a* and *b*. The shaft and supporting rods may also be designed with appropriate seals for mounting onto a nonatmospheric vessel, as illustrated in Figure 8-5*c*.

Because these mixers are small in diameter, they do not provide much circulation flow in batch operations. Often, especially for large vessels ( $>0.04 \text{ m}^3$  or 10 gal), the vessel may be equipped with an auxiliary impeller whose function is to create circulation flow, as illustrated in Figure 8-5*c*. For continuous operation, the head is mounted inside a casing equipped with an inlet and an outlet. Usually, the inlet is located so that the fluid enters the center or “eye” of the rotor. Figure 8-6 shows rotor-stator mixers configured for continuous operation.

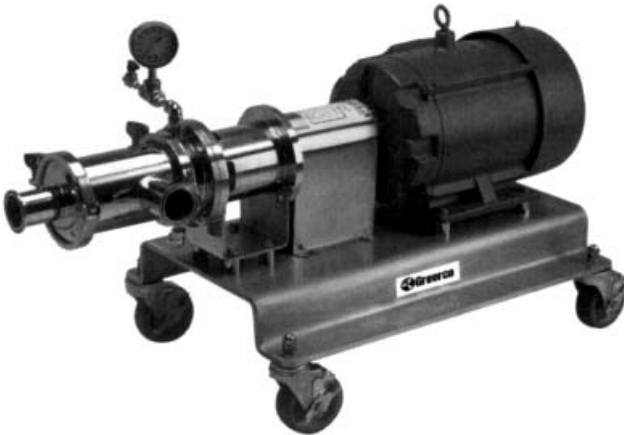


**Figure 8-5** Batch or semibatch rotor-stator assemblies: (a) Silverson Machines batch mixer on hydraulic double-lift stand; (b) IKA Works batch mixer. (Myers, 1999, reproduced with permission of AIChE © 1999.)



(c)

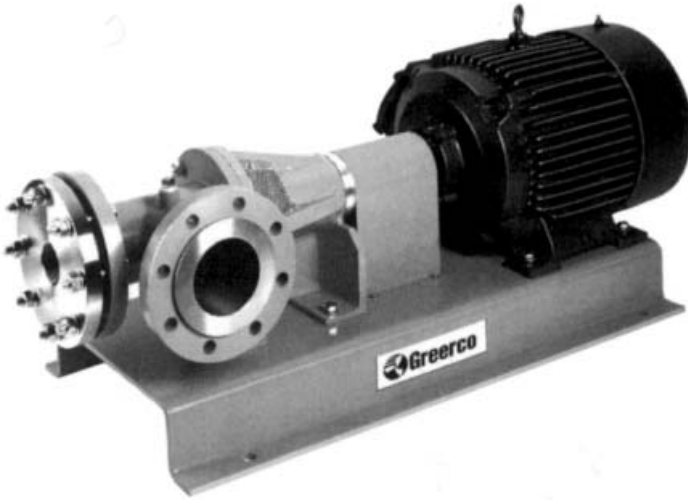
**Figure 8-5** (c) Vessel equipped with rotor–stator and impeller. (Myers, 1999, reproduced with permission of AIChE © 1999.)



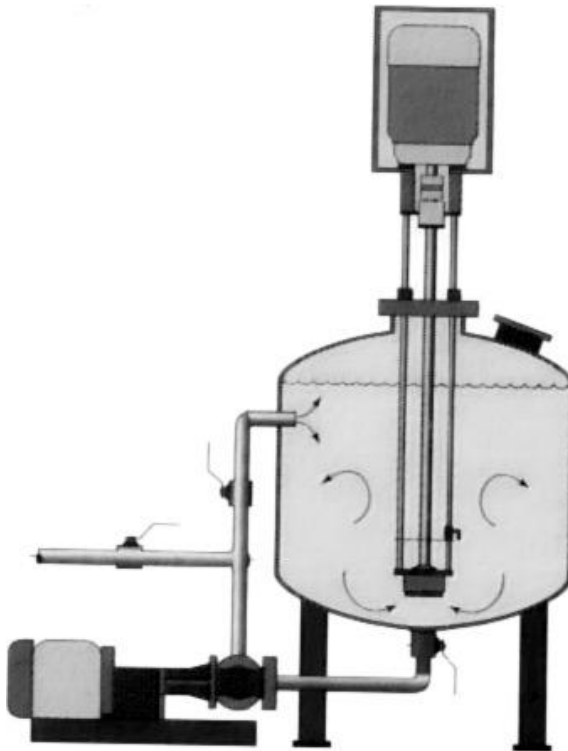
(a)

**Figure 8-6** Rotor–stator mixers for continuous operation: (a) low capacity Chemineer Greerco in-line mixer. (Myers, 1999, reproduced with permission of AIChE © 1999.)  
(Continued)





(b)



(c)

**Figure 8-6** (b) High capacity Chemineer Greerco in-line mixer; (c) Chemineer Greerco in-line mixer in a circulation loop around a batch mixer. (Myers, 1999, reproduced with permission of AIChE © 1999.)

## 8-3 HYDRODYNAMICS OF ROTOR-STATOR MIXERS

There are few published data for power draw and pumping capacity in either batch or in-line rotor-stator mixers. Even less is known about the velocity fields in these devices, so there is little hard evidence to support proposed mechanisms for dispersion and emulsification. As a result it is often necessary to rely on equipment vendors for scale-up rules. Although many vendors have facilities for customer trials, few have well-equipped laboratories for acquisition of basic data for performance characterization. In reality, it is difficult to know how many vendor data are available, since many consider the information to be proprietary. Until recently, there has been little academic interest in high-shear mixers. This work is only starting to appear in the open literature, and it is important for the practitioner to stay informed as a body of knowledge evolves.

### 8-3.1 Power Draw in Batch Mixers

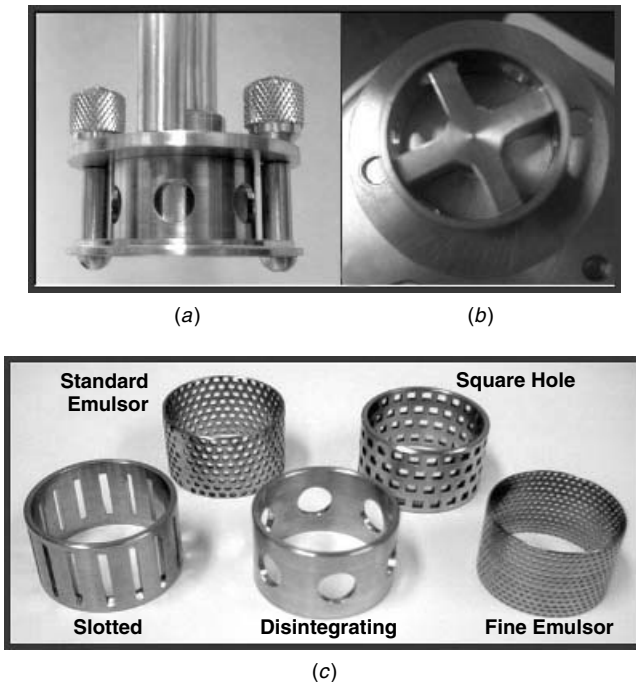
To the authors' knowledge there are no published power draw data for in-line rotor-stator mixers. The few available data sets for batch mixers are limited to devices with a single rotor with blades (not teeth) surrounded by a single stator. There are several characteristic lengths for these devices, including rotor diameter,  $D$ , and the clearance between the rotor and stator or gap width,  $\delta$ . If the stator has openings through which fluid jets exit the mixing head, an additional length scale exists. For conventional stirred tanks the primary length scale is the impeller (rotor) diameter, so the Power number and Reynolds number are defined by  $Po = P/\rho N^3 D^5$  and  $Re = \rho N D^2/\mu$ , respectively. In the laminar regime  $Po$  is inversely proportional to  $Re$ . In the turbulent regime  $Po$  is often constant and varies from about 2 to 6. The transition from laminar to turbulent flow occurs around  $Re \sim 10^4$ .

For the limited studies to date for batch rotor-stator mixers, it has been found that there was no advantage to using a different definition of  $Po$  or  $Re$ . However, for a given geometry only a single-size mixing head was studied. As a result, the  $Po$  versus  $Re$  behavior was quite similar to that of a stirred tank, displaying about the same range of  $Po$ . Myers et al. (2001) reported results for a Greerco  $1\frac{1}{2}$  HR Homomixer ( $D_{max} = 20$  cm,  $\delta = 0.25$  mm), which is an axial flow device with a conical impeller swept volume. The mixing head can be operated to pump either upward or downward and has a solid circular baffle plate above it and an annular plate below it that serve to control the flow direction. The use and position of the plates affect the power draw. The mixing head was centered in the tank, so baffling of the fluid in the vessel also played an important role. However, there was little effect of off-bottom clearance. Power numbers were similar to that for an axial impeller in a stirred tank, but the actual power draw is higher, due to increased rotational speed.

Padron (2001) acquired power draw data for a Greerco Homomixer ( $D_{max} = 4.3$  cm,  $\delta = 0.5$  mm) and found that this much smaller mixer exhibited behavior that was both similar to and different from that of the larger mixer studied by

Myers et al. (2001). The presence of baffles had little effect in upward pumping but yielded a greater power draw for downward pumping. A major difference was that this mixer could not draw power at lower speeds for low viscosity materials, resulting in an unusually large transition region not seen at the pilot scale. As a result, the onset of fully turbulent flow at constant mixer size depended strongly on fluid viscosity. The plates that are located above and below the mixer head cause flow patterns to vary widely with rotor speed and pumping direction.

Padron (2001) also acquired power draw data for the Ross ME 100LC ( $D = 3.4$  cm,  $\delta = 0.5$  mm) and the Silverson L4R ( $D = 2.8$  cm,  $\delta = 0.2$  mm) bench scale mixers. These are radial flow mixers with a four-blade rotor and replaceable stators of various geometry. The Silverson mixer head is shown in Figure 8-7. The Ross design is slightly different but has an equally large variety of stator geometry. Baffles were not used, so the mixers were placed off-center in the tanks to ensure good top-to-bottom mixing. In the laminar regime  $Po$  is somewhat independent of stator geometry, and the Silverson mixer draws slightly higher power at constant  $Re$ . This is due in part to its smaller gap width. Fully turbulent conditions occur above  $Re \sim 10^4$ , but the Ross mixer has a smaller transition region, possibly due to its larger size. In the turbulent regime the Ross mixer draws more power than the Silverson device, indicating that energy dissipation



**Figure 8-7** Rotor-stator head geometry for the Silverson L4R batch rotor-stator mixer: (a) side view; (b) bottom view showing rotor; (c) range of stator geometries.

in the shear gap is not controlling. The constant turbulent power number was dependent on stator geometry and varied from 2.4 to 3.0 for the Ross mixer head and 1.7 to 2.3 for the Silverson head. In general, for a given geometry, the power number increased with the number of openings in the stator, indicating that energy dissipation was controlled by fluid impingement on stator slot surfaces or turbulence in the jets emanating from the stator slots. For instance, for the slotted head, the power number ratio for the Ross head ( $Po = 3.0$ ) to the Silverson head ( $Po = 2.1$ ) is 1.43. The slots themselves are quite similar for both devices, but the Ross has 1.56 times as many slots, indicating that the power number per stator slot is the same.

### 8-3.2 Pumping Capacity

To obtain the pumping capacity of a batch rotor–stator mixer, it is necessary to measure the velocity field entering and/or emanating from the mixer head. Since this is often tedious and requires sophisticated instrumentation, few pumping capacity data exist for batch devices. One must usually rely on vendor information or trial-and-error experimentation to estimate batch time. Rotor–stator mixers are often operated off-center in unbaffled vessels to promote good bulk mixing. As an alternative, the vessel may be equipped with a standard axial impeller to provide bulk mixing (see Figure 8-5c). Mixer location affects both pumping capacity and blend time.

It is much easier to measure the pumping capacity of an in-line rotor–stator mixer, so vendor data are much more reliable. Units that have a rotor with blades, such as the Silverson in-line series, can simultaneously pump and emulsify/disperse material. However, many designs, such as those with multiple rows of rotor and stator teeth, have marginal pumping capacity and may even cause a pressure drop. While rotor rotation acts to create pressure and pumping, the resulting tangential velocities are redirected radially as the fluid passes through the stator slots to counteract the pumping action of the rotor. However, it is the dissipation of this pressure energy that promotes good dispersion/emulsification. This is discussed in more detail below. It is often necessary to feed the mixer with a separate pump.

### 8-3.3 Velocity Field Information

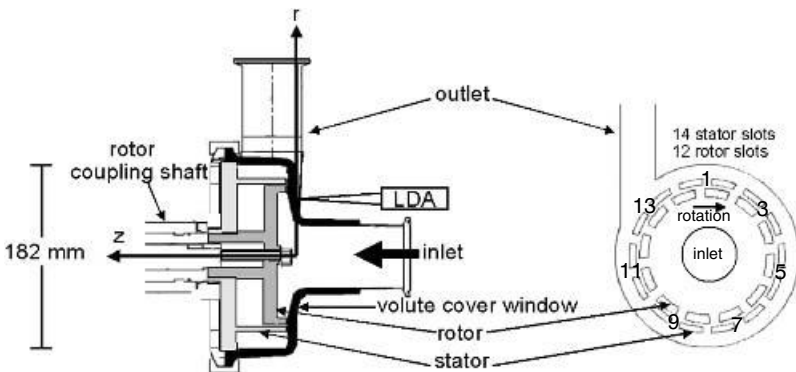
Much can be learned about the mechanisms of dispersion and emulsification in a rotor–stator mixer by studying the detailed velocity and deformation fields for a pure liquid passing through the mixer. However, this can only be accomplished using advanced computational fluid dynamics (CFD) techniques or using sophisticated measurement techniques such as laser Doppler anemometry (LDA) or particle image velocimetry (PIV). The best approach is to use a first-generation simulation to design an experimental program, to use the acquired experimental data to develop a more sophisticated model, and so on. Although still the subject of long-range academic research, rapid advancements in computational resources and instrumentation make this an attractive means to quickly increase our basic

understanding of rotor-stator devices, as well as many other little studied and geometrically complex mixers and dispersion devices.

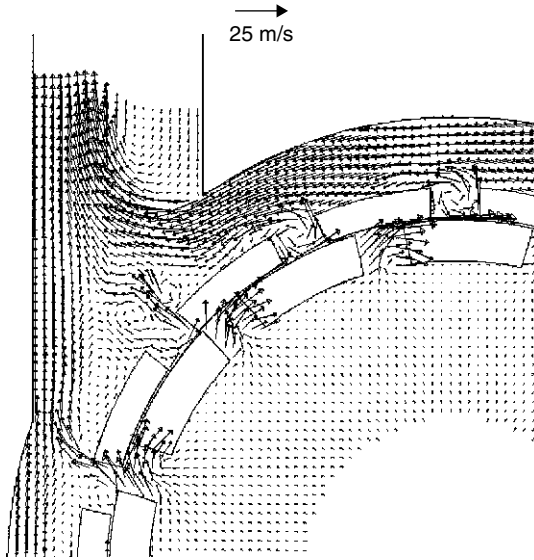
Detailed hydrodynamic studies have been performed for pump-fed in-line mixers with rotors and stators comprised of teeth with slots between them. LeClair (1995) reported an early attempt for a Kady mixer. Although the simulation was quite simplistic and considered only a small section with assumed perfect symmetry, the results revealed a complex circulation pattern in the stator slot.

Calabrese and co-workers have made a more comprehensive study, consisting of modeling and measurement. Calabrese (1999) reported preliminary results. Kevala (2001) and Calabrese et al. (2002) have reported more recent results. To allow acquisition of detailed information while maintaining realism, they considered a simplified prototype of an IKA Works, Inc. mill containing a single set of rotor and stator teeth. A schematic diagram is shown in Figure 8-8. Sliding mesh simulations with the Fluent CFD code were performed at a rotational speed of 30 rps and a throughput of 2.86 L/s (45.4 gal/min). The working fluid was water, so the flow was turbulent even in the narrow rotor-stator gap. Therefore, the RANS equations were solved with the standard  $k-\epsilon$  turbulence model. Corresponding velocity measurements were made using a two-color Dantec LDA system. The front of the volute was made of Plexiglas (or Perspex) to allow laser beam access.

Because the tooth depth was small compared to the mixer diameter, it was initially believed that the flow field was two dimensional. Figure 8-9 shows mean velocity vectors, resulting from a 2D simulation, in the quadrant closest to the exit pipe. These results are angularly resolved in that the flow field changes as the rotor passes the stator. The direction of rotor rotation is clockwise and an extremely complex flow pattern is revealed in the stator slots and volute. Circulation cells in the stator slots allow reentrainment of volute fluid back into



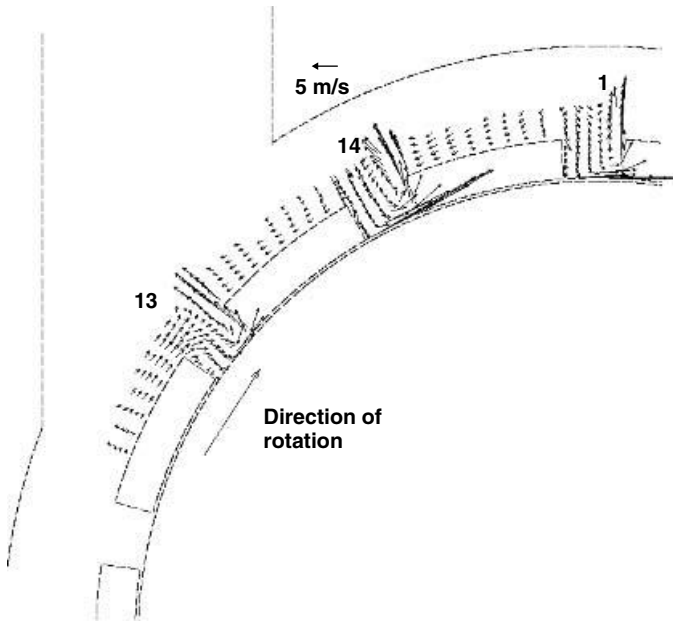
**Figure 8-8** Schematic diagram of prototype IKA mixer with 12 rotor and 14 stator teeth. Inner diameter of rotor is 11.8 cm; outer diameter of stator is 15.4 cm. The rotor tooth depth is 1 cm and the rotor-stator gap width is 0.5 mm. Approximately drawn to scale, except for gap width.



**Figure 8-9** Mean velocity vector data (m/s) acquired via LDA for IKA prototype mixer. Single-time snapshot.

the shear gap. Below the exit the flow in the volute is in the direction of rotation. To the right of the exit the volute flow is counterclockwise against the direction of rotation. As the fluid enters a stator slot it is redirected radially as it impinges on the downstream stator tooth and seeks the path of least resistance to the exit. The flow in other quadrants is equally complex. Each stator slot has a different circulation and reentrainment pattern. At the point farthest from the exit, there are recirculation zones in the volute that divide the clockwise and counterclockwise zones. Figure 8-10 shows the measured mean velocity field in the same quadrant as Figure 8-9. These data are averaged over all rotor positions. That is, LDA data are acquired in the stator slot without regard to the position of the rotor, so all angularly correlated information is lost. The angularly correlated LDA data reveal similar trends. The data were acquired at middepth in the  $z$ -direction indicated in Figure 8-8 or halfway between the front and back of the teeth. The measurements show a stronger jet emanating from the stator slots into the volute, as a result of fluid impingement on the leading edge of the downstream stator tooth, than does the 2D simulation. The recirculation zones in the stator slot are more focused and reentrainment into the shear gap is stronger. Nevertheless, it is seen that much physical insight can be gained from the more detailed and less labor-intensive simulation.

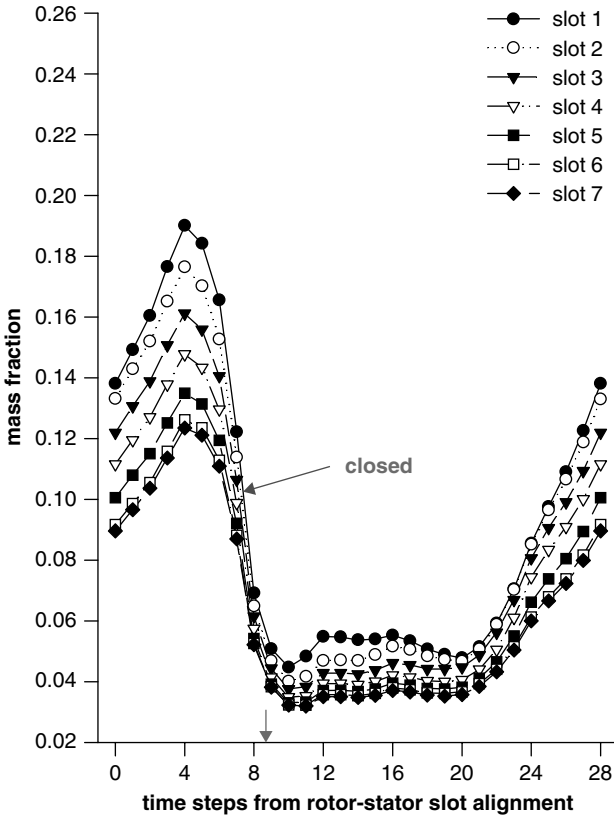
Figure 8-11 shows the total mass flow rate exiting each of stator slots 1 to 7 (refer to Figure 8-8) predicted by the 2D simulation, as a function of rotor tooth position. As the rotor tooth blocks the stator slot, the flow rate drops rapidly and does not recover until the rotor tooth passes the slot completely. The maximum



**Figure 8-10** Mean velocity vector data (m/s) acquired via LDA for IKA prototype mixer. Velocity field is angularly averaged over all rotor positions.

flow rate does not occur when the rotor and stator slots are aligned but when the stator slot is about half blocked by the rotor tooth. Stator slots closest to the exit have higher mass flow rates than those far from the exit. The LDA measurements are somewhat different. While the exit flow rate drops rapidly as the rotor tooth blocks the stator slot, it partially recovers well before the rotor tooth has passed the stator slot. Furthermore, all slots have about the same flow rate. The data reveal that the flow field is highly three dimensional and that when the rotor tooth blocks the stator slot, there is leakage flow over the top of the rotor tooth into the stator slot. As with extruder screw flights, there must be a clearance between the rotor and the front volute cover. This clearance cannot be much smaller than the shear gap itself. Preliminary 3D simulations by Calabrese and co-workers indicate that the results are much more in line with the LDA data.

Figure 8-12 shows the turbulent kinetic energy (TKE) predicted by the 2D simulation. In simple flow fields the TKE can be related to the energy dissipation rate, which is, in turn, a measure of emulsification or dispersion capacity. The LDA data are somewhat different, but the same physical insights emerge. When rotor and stator slots are aligned, the TKE is low. As the rotor tooth blocks the stator slot, the TKE builds on the leading edge of the stagnated downstream stator tooth. The most intense TKE field is created just as the stator slot is blocked and as the flow rate begins to fall. As a result, much of the fluid leaving the stator slots does not experience an intense deformation field. It is because of this “bypassing”



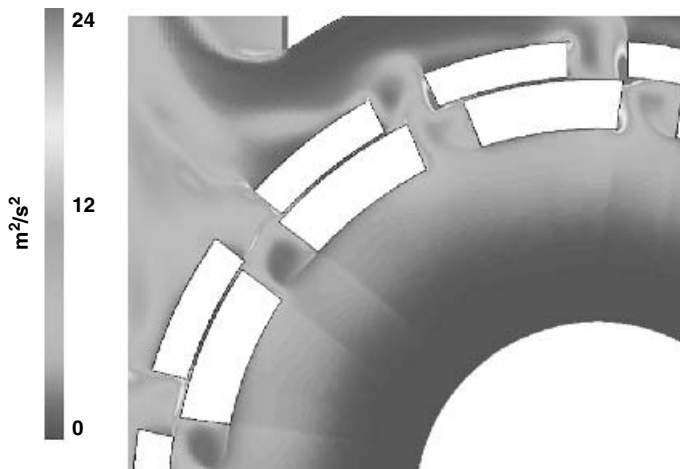
**Figure 8-11** Mass flow rate exiting stator slots 1 to 7 as a function of time or rotor position. The rotor tooth completely blocks the stator slot at time step 9 and the slot begins to open again at time step 21. Results of 2D CFD simulation. Stator slot numbers are defined in Figure 8-8.

that devices with multiple rows of rotor and stator teeth are more effective for emulsification and dispersion. It is often necessary to run multiple passes through the mixer to achieve the desired product.

For turbulent flow through rotor-stator devices with teeth, the aforementioned velocity field results indicate that flow stagnation on the leading edge of the downstream stator teeth provides a major energy field for emulsification and dispersion. It is not clear from these results what role is played by flow in the shear gap. The simulations indicate that the flow in the rotor-stator gap is not a simple shear flow but is more like a classical turbulent shear flow. Use of nominal shear rate may not be useful in scale-up.

Simulations were performed and some LDA data were acquired for a similar device that has an enlarged shear gap of 4 mm rather than the standard 0.5 mm gap. The results indicate that there is much less stagnation on the stator teeth, so





**Figure 8-12** Turbulent kinetic energy ( $\text{m}^2/\text{s}^2$ ) from 2D CFD simulation for IKA prototype mixer. Single-time snapshot. See insert for a color representation of this figure.

that high levels of TKE are not seen. When the shear gap is too large, too much fluid remains in the gap rather than being forced onto the leading edge of the stator tooth, thereby exiting the stator slot. As a result, it is necessary to have a narrow shear gap even if the shear in this gap is not a major contributor to the dispersion process.

### 8-3.4 Summary and Guidelines

As stated above, there are few fundamental data to allow design and scale-up of rotor–stator mixers available in the open literature. Academic activity is increasing and it is important to remain aware of new information. It is important to note that there is activity in this field that may be documented in non-English publications that has not been discussed above. The results discussed above were general in nature. Additional information may be available in industry-specific publications, such as those for food processing, paints and pigments, and so on.

## 8-4 PROCESS SCALE-UP AND DESIGN CONSIDERATIONS

There are many important process and mixer variables to consider in the process selection, scale-up, and operation of rotor–stator mixers. Key variables among them include:

- Chemistry and physics of the process and requirements for macro- or micro-mixing
- Density and rheological properties of the feed and product streams

- Coalescence behavior with or without surface-active agents (surfactants, dispersants, stabilizers, etc.)
- Role of the materials of construction of fluid-wetted parts on the chemistry or physics of the process
- Number of feed streams and the need for staging or premixing
- Desired throughput
- Control of the flow ratios of feed streams
- Location and order of addition of reagents to different stages of mixer heads
- Temperature of operation and the need for heating or cooling for temperature control
- Effect of shear on the degradation or coagulation, and so on, of the feed and product streams
- Operating mode: batch, semibatch, or continuous operation
- Rotor–stator design and operation
- Speed of the rotor
- Gap between rotor and stator
- Diameter of the generator or rotor–stator head or assembly
- Design or geometry of the rotor and stator, including number of teeth, tooth to tooth spacing, number of stages, percent open area, and so on.

The fluid properties at the desired operating temperature must be known. These properties determine, in large measure, the required mixing operation and how difficult it will be to achieve the desired process result. Fluids can heat up substantially while being processed in a rotor–stator mixer due to the high-energy input and small fluid volumes. As the viscosity increases, so does the viscous heating effect. Therefore, provisions for cooling, or in a few cases heating, must be considered. Jacketed units are not able to add much cooling (or heating) because of the limited surface area and short contact time. Jackets are used mainly to maintain temperature during downtime or to prevent polymerization during periods of idle operation.

For dispersion processes, the role of coalescence on the mean particle size and particle size distribution should not be overlooked. The type and amounts of surface-active agents added, as well as the material of construction of the rotor and stator, all affect coalescence, as discussed in greater detail in Chapter 12.

The appropriate operating mode (batch, semibatch, or continuous operation) will depend on the mixing process and the process result desired. Batch and semibatch operations are best suited for processes that require long time scales, greater than a few seconds.

For batch or semibatch applications, the vessel size and shape, location of the mixing head (i.e., mounted centrally or off-center), and the presence of baffles are additional items for consideration. Commercial rotor–stator mixers are matched to specific batch volumes.

For continuous applications, additional issues for consideration include:

- Flow rate through the unit and control of multiple feed streams
- Pumping capacity or need for a separate feed pump
- Premixing of feed streams and location in the mixing head where the feed is introduced
- Residence time
- Operating pressure and back pressure (for some systems), especially those that are sensitive to compression or flashing

The precise control of feed rates and ratios is vital for continuously operated rotor–stator mixers, since the extent of backmixing to even out variations in feed rates is unknown. Some processes may require premixing of the feed streams upstream of the rotor–stator mixer. For very fast reactive processes, premixing must be avoided. Others may benefit from staged addition of the streams, especially in multistaged mixers. When mixing streams of widely different viscosities, it is instructive to explore the effects of switching feed locations on the process result.

Although rotor–stator mixers can pump to some extent, it is preferred to use a pump to control the feed rate to the mixer. This way, one does not need to vary the rotor speed to control the flow rate. Instead, the rotor speed can be varied to control the energy input, turbulent kinetic energy, and shear rates in the device, independent of the flow rate. To prevent equipment failure, it is also important to ensure that the unit is “fully flooded” and not starved during operation.

Feed streams must be introduced into the mixing head so as to minimize bypassing or short-circuiting to the outlet of the mixer. An important consideration is the unit’s orientation (vertical or horizontal) for the correct delivery of the material to the shear zone. Streams must flow through the shear gaps and regions of the stator exit slots where most of the energy is dissipated. The residence time must be related to the volume in the vicinity of the rotor and stator where the mixing action occurs, not the entire volume of the device. For the device to be effective, the residence time must match the required mixing time for the processes.

The rheology of the dispersed or homogenized stream can be very different from the feed streams. This greatly influences the ability to move the fluids through the system and can affect the backpressure on the unit. Some emulsions will not form unless there is enough backpressure for the materials to stay in intimate contact with each other in the mixing zones. However, once formed, thick emulsions may need to be diluted quickly to facilitate fluid flow.

#### **8-4.1 Liquid–Liquid Dispersion**

Rotor–stator mixers are widely used in the chemical process and allied industries to produce liquid–liquid dispersions and emulsions. Although production of dispersions and emulsions in rotor–stator mixers is often highlighted in

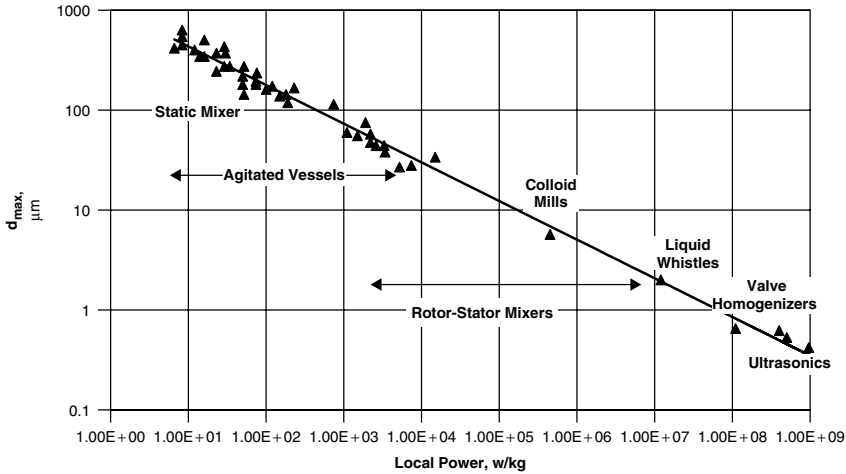
industry-specific publications, there are few studies that relate drop size to energy input or other variables related to agitation intensity. The authors are not aware of generic studies for in-line rotor–stator mixers. Such experiments are difficult and require disposal of large amounts of material, making the approach of Section 8-3, to infer mechanisms, attractive. In this section the discussion is limited to batch rotor–stator mixers, where drop size data are more easily acquired and the results more readily correlated.

Whereas power numbers for batch rotor–stator mixers are of the same magnitude as for stirred tank impellers, the power per unit volume is much higher, due to high rotational speed. Dispersions and emulsions of practical significance are usually of high dispersed phase volume fraction. As a result, surfactants or other stabilizers must be present to prevent coalescence. Unfortunately, the roles played by the surfactant and mechanical forces in determining ultimate drop size are difficult to separate and poorly understood. To circumvent this problem, Calabrese et al. (2000) made drop size measurements for dilute dispersions of low viscosity organic liquids in aqueous solutions, for which surfactants were not required, due to negligible coalescence rate. The experiments were performed in Ross ME 100 LC and Silverson L4R mixers (Figure 8-7). These are the same mixers as those used by Padron (2001) for the power number studies discussed in Section 8-3.1. The flow was turbulent.

The results showed that when the rotor–stator shear gap width was increased at constant rotor speed smaller drops were produced even though the nominal gap shear rate had decreased. Furthermore, the Ross and Silverson mixers produced the same drop size when the slotted stator head was used, since the power per stator slot was the same (Section 8-3.1). Even though the power numbers were similar, the slotted head produced smaller drops than the disintegrating head. LDA measurements in the jets emanating from the stator openings indicated that the slotted head focused the energy better and produced higher levels of TKE. These observations lead the authors to suggest that drop breakup occurred on stator surfaces or in the jets emanating from the stator rather than in the shear gap.

The smallest scale of turbulence (Kolmogorov microscale) was estimated from the power draw data and found to be slightly smaller than the drop size. That is, at high-energy input the turbulent microscale is decreased so that the resulting drop size is still governed by turbulent stresses acting on the drops. As a result, the mean drop size data could be correlated in the same way as data for stirred tank and other turbulent liquid–liquid dispersion devices. These correlations are discussed in Chapter 12. Figure 8-13 shows that for low viscosity drops of constant physical properties, the maximum drop size correlates with local energy dissipation rate (power per unit mass). The rotor–stator data fall below those for stirred tanks and above those for liquid whistles.

**8-4.1.1 Mean Drop Size and Drop Size Distribution.** Estimates of drop size that are achievable by various mixing devices for dispersing immiscible liquid–liquid systems are also shown in Table 8-1. In the table, the specific



**Figure 8-13** Maximum drop size  $d_{\max}$  versus local power draw for dilute oil-in-water dispersions. The local power per mass of fluid is the total power input divided by the mass of fluid in the high intensity dispersion region of the mixer. (After Davies, 1987.)

**Table 8-1** Performance Features of Various Dispersion Devices

Type of Device	Energy Dissipation		Comments
	Range (W/kg) or ( $\text{m}^2/\text{s}^3$ )	Typical Size Range ( $\mu\text{m}$ )	
Static mixers	10–1000	50–1000	Narrower DSD than agitated vessel
Agitated vessel	0.1–100	20–500	With Rushton turbine in a fully baffled vessel; usually, broad DSD
High speed rotor–stator	1000–100 000	0.5–100	Can be smaller with the correct chemistry
Valve homogenizer	$\sim 10^8$	0.5–1	Requires high pressure, 5000 to 10 000 psi
Ultrasonics	$\sim 10^9$	0.2–0.5	Sonification devices

energy input increases as one moves down the “Type of Device” column. In general, as local energy input is increased, a smaller droplet size is produced, as illustrated in Figure 8-13. This does not mean that the dispersion will be stable, as coalescence can occur outside the device. Surfactants are used to stabilize against coalescence. Batch systems inherently produce a wider drop size distribution than continuous systems.

When choosing a disperser or homogenizer for evaluation, important considerations include the turn-down ratio (maximum flow/minimum flow) and the ease with which equipment components can be changed to control the DSD.

### 8-4.2 Solids and Powder Dispersion Operations

Rotor–stator mixers are an important technological option for wet mixing and for dispersion of hard-to-wet powders into a liquid. They find use in the preparation of “fish-eye” free solutions of powders (thickeners, stabilizers, flour, starches, caseinates, powdered milk, clays, etc.). Suppliers have various names for the systems used for such applications, but in principle, they include the same basic components: an in-line rotor–stator mixer, a high-pressure centrifugal pump, a venturi assembly, and a powder feed hopper system. The centrifugal pump conveys the fluid through a venturi with a TEE connection to the powder feed line. The resulting low pressure created in the throat of the venturi by the flowing fluid causes the powder to be sucked into the flowing liquid, which immediately enters the rotor–stator mixer. Here, the intense mechanical and hydraulic shear in the mixer head quickly disperses the powder into the liquid.

The authors are not aware of any published fundamental studies of solid dispersion in rotor–stator mixers but have seen videos supplied by vendors that show preparation of “fish-eye” free solutions of hard-to-wet powders such as Carbopol, with apparently remarkable ease. Scale-up and design of these systems would be based on operational information obtained with vendor equipment. This is highly recommended.

Although rotor–stator devices are not considered highly energy intensive relative to some types of solids size-reduction devices, they are sometimes employed when more controlled size reduction is desired, especially when it is desirable to minimize the production of fines. One example of such use is the conditioning of seed particles for crystallization operations when the seed is a portion of the previous batch (“heel” crystallization).

### 8-4.3 Chemical Reactions

The authors are not aware of any nonproprietary industrial application of rotor–stator mixers for carrying out fast or mixing controlled chemical reactions, even though they appear suitable for the purpose. Indeed for this purpose, Bourne and co-workers (Bourne and Garcia-Rosas, 1986; Bourne and Studer, 1992) evaluated the suitability of a commercially available rotor–stator mixer using the fast competitive azo-coupling reactions of 1-naphthol and diazotized sulfanilic acid. The rationale was to accelerate micromixing by exploiting the general characteristics of rotor–stator mixers, that is, they generate a locally intense turbulence in a small volume with a short residence time.

Their results indicate that it is necessary to feed at or near the shear gap to take advantage of the high-energy dissipation rate in the gap. This was difficult to do. They estimated the turbulent energy dissipation in the shear gap to be on the order of 1000 W/kg. By comparing the product distribution at two scales, they concluded that scale-up on the basis of tip speed was better than on the basis of power per volume.

#### 8-4.4 Additional Considerations for Scale-up and Comparative Sizing of Rotor–Stator Mixers

As noted earlier, a fundamental understanding of the underlying principles and of the important variables that affect the performance of rotor–stator mixers is only now being developed and disseminated. However, most suppliers possess proprietary know-how and experience acquired over many years with their specific equipment and can make recommendations for the selection and scale-up for specific process applications. To do this effectively, the suppliers will require information on the process and fluid properties. Often, they will insist on testing their lab or pilot equipment on actual process fluids before making recommendations. Suppliers will treat process information shared with them, or acquired with their test equipment, as confidential. For additional protection, however, consider secrecy agreements before sharing proprietary process information. Most suppliers may not have facilities to handle hazardous chemicals and high viscosity fluids. For these situations the client can instead suggest model fluids or rent the equipment for installation and testing in-house.

Vendors often design and scale-up rotor–stator mixers based on equal rotor tip speed,  $V_{tip} = \pi ND$ , where  $N$  is the rotational speed of the rotor and  $D$  is the rotor diameter. This criterion is equivalent to equal nominal shear rate in the rotor–stator gap,  $\dot{\gamma}$ . In most industrial rotor–stator mixers, the shear gap width  $\delta$ , remains the same on scale-up, making the two criteria equivalent.

The nominal shear rate in the rotor–stator gaps is calculated as follows:

$$\dot{\gamma} = \frac{\pi ND}{\delta} \quad (8-1)$$

It is important to recall from the discussion above that for turbulent flow the actual shear rate in the rotor–stator gap varies substantially from  $\dot{\gamma}$ , and that gap shear rate does not directly control power draw and dispersion. However, tip speed may control turbulence characteristics, especially if the spacing between stator elements (teeth or stator openings) as well as the shear gap width do not change on scale-up.

Usually, different rotor–stator devices are compared on the basis of the throughput, the amount of fluid that one can move through a specific unit. The throughput listed in catalogs is based on the volumetric flow rate of water as the fluid pumped and mixed under specified conditions of rotor speed and gap width. When viscous fluids are processed, the actual flow rate will be less and the unit's capacity will be lower. Modifications to the rotor may be needed to handle greater throughput and/or higher viscosity fluids. These modifications can enhance throughput but may reduce the mixing performance. They often add significantly to the cost of the unit.

Units may also be compared on the basis of the residence time in the mixing zone. It is important to determine the true volume of the mixing zone for this type of comparison to be meaningful.

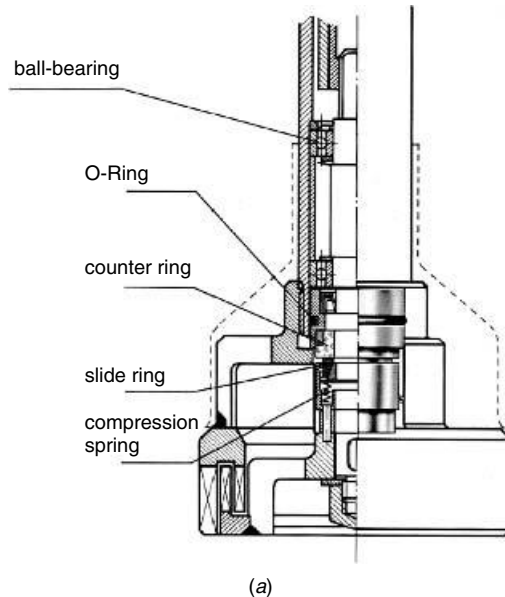
## 8-5 MECHANICAL DESIGN CONSIDERATIONS

As illustrated in Figure 8-14, rotor–stator mixers are complex mechanical devices with many potentially wearable components. Their reliability requires careful attention to important mechanical details, including:

- Proper selection of the motor and design for variable speeds
- Gearbox selection and design
- Seal design
- Pressure rating of the unit
- Bearing design and tolerances, especially for a larger mixer with a cantilevered shaft
- Use of appropriate materials of construction

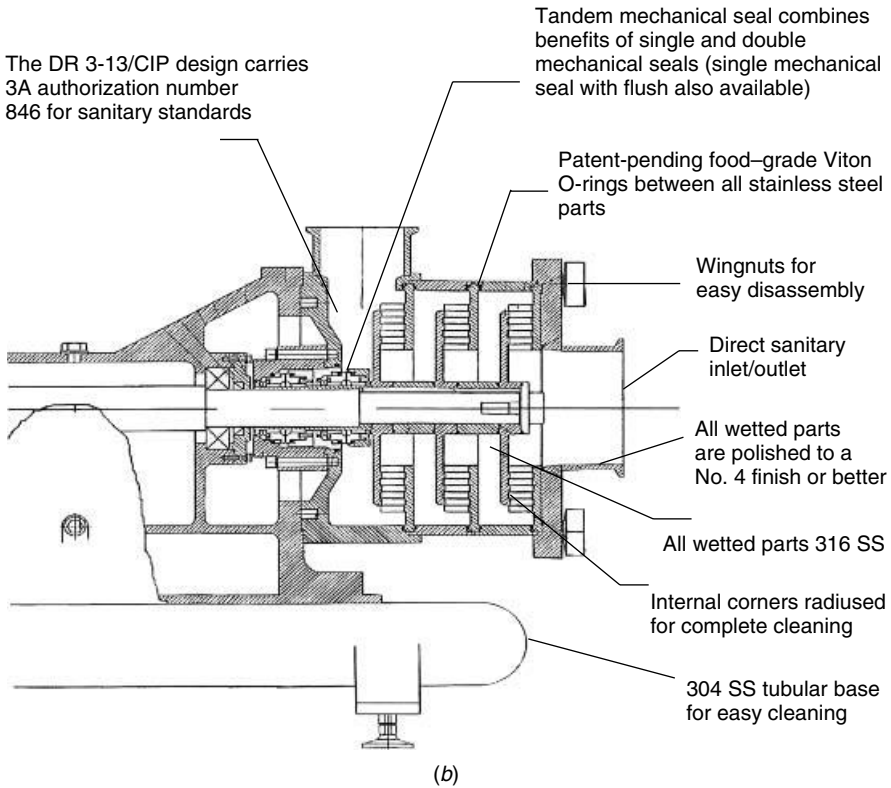
Most rotor–stator devices are equipped with variable speed drives to allow operations at different rotor speeds. There is sufficient know-how about variable speed drives. However, one must pay careful attention to:

- Operations at the low end of the speed range because of the required higher torque and associated mechanical instability
- Mechanical limits specified by the supplier, including shaft and seal design, to be sure the design torque and speed are not exceeded



**Figure 8-14** Mechanical complexity of rotor–stator mixers: (a) seal of IKA Ultra Turrax batch rotor–stator. (Continued)





**Figure 8-14** (b) Multistage mixing head.

It is important to verify, a priori, that the specific equipment has been used at the specified conditions without any known problems. Because of the close tolerances or small gaps between the rotor and stator parts, one must confirm that the gap tolerances will be maintained as the components heat up during processing.

Details of mechanical design, fabrication, and assembly must be entrusted to the suppliers of the mixer. Users must resist the temptation to mechanically redesign or modify standard mixers without a supplier's recommendation and approval.

## 8-6 ROTOR-STATOR MIXING EQUIPMENT SUPPLIERS

There are many vendors of rotor-stator mixers. The reader may refer to current editions of *Thomas's Register*, *Chemical Engineering Buyers' Guide*, and so on, for an up-to-date list of major suppliers. There are a number of other suppliers whose equipment is used to produce dispersions or emulsions but which does not fit the definition of having a rotor and stator. These include high-pressure devices such as valve homogenizers, whistles, and cavitation-based devices.

## NOMENCLATURE

$d_{\max}$	maximum stable drop size ( $\mu\text{m}$ )
$D$	rotor diameter (m)
$D_{\max}$	maximum diameter of a Greerco rotor (m)
$N$	rotor rotational speed (rps)
$P$	power (W)
$Po$	power number, $\frac{P}{\rho N^3 D^5}$
$Re$	Reynolds number, $\frac{\rho N D^2}{\mu}$
$V_{\text{tip}}$	rotor tip speed (m/s)

### Greek Symbols

$\dot{\gamma}$	nominal shear rate ( $\text{s}^{-1}$ )
$\delta$	width of shear gap between rotor and stator (m)
$\mu$	viscosity ( $\text{Pa} \cdot \text{s}$ )
$\rho$	density ( $\text{kg}/\text{m}^3$ )

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