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| SUPERINTENDENCIA DE INDUSTRIA Y COMERCIO          |                                            |
| RAD: 14-59449- -5                                 | FECHA: 2015-09-29 15:46:27                 |
| DEP: 2020 DIRECCION DE NUEVAS CREACIONES          | EVE: 378 FASE NACIONAL INCLUIDO CAPITULO I |
| TRA: 11 PCT-PATENTE DE INVENCION CAPITULOS I Y II | FOLIOS: 8                                  |
| ACT: 519 TRASLADOINFORME                          |                                            |

DIRECCIÓN DE NUEVAS CREACIONES

2020  
Bogotá, D.C.,

Abogado(a)/Señor(a)  
CARLOS R. OLARTE GARCIA

Asunto:      Radicación:      14-59449- -5  
                 Trámite:            11  
                 Evento:            378  
                 Actuación:            519  
                 Oficio:                    11735  
                 Folios:                    8

|                                     |                   |            |   |                               |  |
|-------------------------------------|-------------------|------------|---|-------------------------------|--|
| Patente de Invención                | X                 |            |   | Patente de Modelo de Utilidad |  |
| Vía Nacional                        |                   |            |   |                               |  |
| Vía PCT (Fase Nacional)             | X                 | Capítulo I | X | Capítulo II                   |  |
| No. Solicitud Internacional         | PCT/EP2012/068387 |            |   |                               |  |
| Fecha de Presentación Internacional | 19/09/2012        |            |   |                               |  |

Como resultado del examen de fondo realizado, con fundamento en el Artículo 45 de la Decisión 486 de la Comisión de la Comunidad Andina, se comunica:

1. Documentos estudiados

|                   |                         |               |
|-------------------|-------------------------|---------------|
| Reivindicaciones: | Rad. N° 14 059449 00000 | 19/Marzo/2014 |
| Descripción:      | Rad. N° 14 059449 00000 | 19/Marzo/2014 |
| Dibujos:          | Rad. N° 14 059449 00000 | 19/Marzo/2014 |
| Prioridad:        | EP 11182096.5           | 21/09/2011    |

2. Análisis de la Prioridad

Se acepta la reivindicación de prioridad porque ésta solicitud fue presentada dentro del plazo establecido (Art. 9 D 486) y su contenido técnico corresponde con el de la prioridad reivindicada.

### 3. Objeto de la invención

De acuerdo con lo establecido en el numeral 1.2.2.2 del Capítulo Primero del Título X de la Circular Única para efectos del examen de patentabilidad, se tendrá en cuenta la tasa pagada por el solicitante al momento de comenzar el examen para las reivindicaciones adicionales; esto es que cuando la solicitud contenga más de diez (10) reivindicaciones y no se haya pagado la tasa complementaria, sólo se estudiarán las cubiertas por la tasa pagada. Por lo tanto, se estudiarán las reivindicaciones 1 a 11.

Título de la solicitud: “Dispositivo para mezclar y enfriar dos líquidos reactivos y método para hacer ácido peroximonosulfúrico con el mismo” (Pág. 1, Radicado N° 14-059449-0000).

El problema planteado en esta solicitud consiste en la necesidad de contar con un dispositivo que mezcle eficientemente dos corrientes de reactivos líquidos y enfríe rápidamente la mezcla resultante.

Para resolver este problema técnico, la solicitud en estudio proporciona un dispositivo para mezclar y enfriar dos reactivos líquidos, el cual incluye un dispositivo mezclador acoplado con un intercambiador de calor del tipo de tubos y coraza, así como también un método para hacer ácido peroximonosulfúrico, partiendo de ácido sulfúrico y peróxido de hidrógeno, empleando dicho dispositivo.

El objeto de la presente solicitud se relaciona con la Clasificación Internacional de Patentes (CIP): B01F 3/08, B01F 15/06, B01F 5/06, B01F 5/04, B01J 19/24.

### 4. De la solicitud de patente

#### Reivindicaciones (Art 30 D 486, Art 22 PCT)

Las reivindicaciones 1, 3, 7 y 8 no son claras pues hacen referencia a términos generales, como “multiplicidad”, “esencialmente”, “mezcladores estáticos”. Se sugiere sustituirlos por términos precisos o rangos concretos.

La reivindicación 11 no es clara, porque el término “menor de 80 °C” es impreciso, lo cual no permite distinguir la invención del estado de la técnica. Se sugiere sustituirlo por un término preciso o un rango concreto.

La reivindicación 8 no es clara, ya que hace referencia a un método pero no enseña una secuencia lógica de pasos por medio de los cuales se transformen los reactivos ni tampoco enseña las condiciones técnicas a las que deban ejecutarse cada una de esas etapas. Se sugiere hacer la aclaración respectiva.

La reivindicación 8 presenta falsa dependencia ya que menciona el dispositivo de acuerdo a las reivindicaciones 1 a 6, lo cual no es correcto. Al respecto no debe olvidarse que las reivindicaciones de producto y de procedimiento son diferentes y, por ello, deben redactarse de manera independiente en el pliego reivindicatorio. Sobre las mismas, el Tribunal Andino de Justicia ha dicho lo siguiente:

“El objeto de la invención de producto está constituido por un cuerpo cierto destinado a llenar una necesidad industrial que el estado de la técnica aún no ha satisfecho. En cambio, el objeto de la invención de procedimiento se encuentra conformado por “una serie o sucesión de operaciones que inciden sobre una materia (sólida, líquida o

gaseosa) que se encamina a la obtención de una cosa o un resultado”. (Otero Lastres, José Manuel)<sup>1</sup>

La reivindicación 8 no es clara, pues por su redacción sugiere que se introducen unas cantidades de ácido sulfúrico o de peróxido de hidrógeno que son un cierto porcentaje de una corriente cuando, al parecer, dichos porcentajes corresponden a la concentración de dichos reactivos. Se sugiere hacer la aclaración respectiva.

### Descripción (Art 28 D 486)

La descripción del método para hacer ácido peroximonosulfúrico no es clara, pues por su redacción sugiere que se introducen unas cantidades de ácido sulfúrico o de peróxido de hidrógeno que son un cierto porcentaje de una corriente cuando, al parecer, dichos porcentajes se refieren es a la concentración de dichos reactivos (Pág. 3, Párr. 2 y Pág. 6, Párr. 3). Se sugiere hacer la aclaración respectiva.

### 5. Determinación del Estado de la Técnica

El estado de la técnica se determinó con base en la fecha de declaración de prioridad: 21 de septiembre de 2011, y se encontraron los siguientes documentos que anticipan el objeto de esta solicitud:

| Ítem | Documento    | Título                 | Fecha de Publicación* |
|------|--------------|------------------------|-----------------------|
| D1   | WO 92/007791 | Peroxoacid manufacture | 14 de mayo de 1992    |
| D2   | US 3,818,938 | Fluid mixing apparatus | 25 de junio de 1974   |
| D3   | Libro 1*     | Heat and Mass Transfer | 2006                  |

\* Heat and Mass Transfer. H. D. Baehr, K. Stephan. 2º Ed. Springer-Verlag Berlin Heidelberg 2006.

El documento D1<sup>2</sup> divulga la manufactura de peroxiácidos y más particularmente, un proceso y un aparato para la manufactura de ácido peroximonosulfúrico. (Pág. 1, líneas 2 a 4).

El documento D2<sup>3</sup> divulga un aparato para mezclar dos fluidos e inyectar la mezcla resultante de los dos fluidos en un tercer fluido circulante (Pág. 1, Resumen).

El documento D3<sup>4</sup> describe diferentes tipos de intercambiadores de tubos y coraza. (Apartado 1.3, Pág. 40–64).

### 6. Examen de Patentabilidad (Art 14 D486)

#### Novedad (Art 16 D486)

La reivindicación 8 consiste en “*Un método para hacer ácido peroximonosulfúrico, CARACTERIZADO porque...*” (Ver Pág. 10 del Rad. N° 14-059449-00000-0000)

Comparación de las características técnicas esenciales, con las del Estado de la Técnica:

1 Proceso 129-IP-2007

2 [http://worldwide.espacenet.com/publicationDetails/originalDocument;jsessionid=CXqPVqgJWR0dqbOT2qBywdnq.espacenet\\_levelx\\_prod\\_3?CC=WO&NR=9207791A1&KC=A1&FT=D&ND=&date=19920514&DB=&&locale=en\\_EP](http://worldwide.espacenet.com/publicationDetails/originalDocument;jsessionid=CXqPVqgJWR0dqbOT2qBywdnq.espacenet_levelx_prod_3?CC=WO&NR=9207791A1&KC=A1&FT=D&ND=&date=19920514&DB=&&locale=en_EP)

3 <http://patft.uspto.gov/netacgi/nph-Parser?Sect1=PTO1&Sect2=HTO&f=PAL&p=1&u=%2Fnetacgi%2FPTO%2Fsrchnum.htm&r=1&f=G&l=50&s1=3,818,938.PN.&OS=PN/3,818,938&RS=PN/3,818,938>

4 <http://link.springer.com/book/10.1007/3-540-29527-5>

| Características<br>esenciales                                                       | D1                                                                                                                                                                                                                                                                                                                                                                |
|-------------------------------------------------------------------------------------|-------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------|
| Un método para hacer ácido peroximonosulfúrico, caracterizado porque                | La presente invención se relaciona con la manufactura de peroxiácidos y más particularmente, a un proceso y un aparato para la manufactura de ácido peroximonosulfúrico. (Pág. 1, líneas 2 a 4).                                                                                                                                                                  |
| se sigue el procedimiento de: introducir entre 85 y 98 % en peso de ácido sulfúrico | La ruta de reacción comprende la reacción entre el peróxido de hidrógeno acuoso y el ácido sulfúrico o el óleum o el SO <sub>3</sub> gaseoso. (Pág. 1, líneas 26 a 27).<br><br>La concentración del ácido sulfúrico es, normalmente, de al menos 90% p/p y frecuentemente de 92 a 99% p/p. (Pág. 11, líneas 28 a 30).                                             |
| introducir entre 50 y 80 % en peso de peróxido de hidrógeno acuoso                  | La ruta de reacción comprende la reacción entre el peróxido de hidrógeno acuoso y el ácido sulfúrico o el oleum o el SO <sub>3</sub> gaseoso. (Pág. 1, líneas 26 a 27).<br><br>La concentración del peróxido de hidrógeno empleado es, normalmente, de al menos 50% y especialmente de 60% hasta 75% p/p. (Pág. 11, líneas 30 a 31).                              |
| y enfriar la mezcla que se obtiene como resultado.                                  | Alternativamente, y en un conjunto adicional de ventajosas implementaciones, al producto se le permite fluir a través de una tubería localizada dentro del tanque de tratamiento la cual se cubre por el líquido en su interior de tal forma que el producto se enfría por intercambio de calor a través de las paredes de la tubería. (Pág. 12, líneas 33 a 37). |

De acuerdo con la tabla anterior, las características esenciales del método de la reivindicación 8 habían sido divulgadas previamente en el documento D1.

En consecuencia la reivindicación 8 no es nueva, según lo divulgado en el documento D1.

Asimismo, el documento D1 divulga las características de las reivindicaciones dependientes así:

Reivindicación 9: El proceso de la invención es adecuado para la reacción entre el ácido sulfúrico concentrado y el peróxido de hidrógeno acuoso en donde la relación molar del ácido sulfúrico al peróxido de hidrógeno se selecciona en el rango de 0.5:1 a 5:1 y es particularmente adecuada cuando la relación molar está dentro del rango de 1:1 a 3:1. (Pág. 11, líneas 23 a 28).

Reivindicación 10: De acuerdo a un aspecto de la presente invención, se provee un proceso continuo para la manufactura de ácido peroximonosulfúrico por la reacción entre el ácido sulfúrico concentrado y el peróxido de hidrógeno concentrado el cual se caracteriza por introducir los dos reactivos bajo presión dentro de una cámara de reacción tubular cerrada que tiene una entrada para la solución del ácido sulfúrico en un extremo o adyacente al mismo, y una salida para la mezcla de reacción en el extremo opuesto a la entrada del ácido sulfúrico y una entrada para la solución del peróxido de hidrógeno que está posicionada entre la entrada del ácido sulfúrico y la salida de la mezcla, por lo cual, la solución de peróxido de hidrógeno se introduce en un flujo de la solución de ácido sulfúrico, en donde dicha cámara de reacción se dimensiona con relación a las velocidades de flujo de los reactivos de tal forma que la salida por minuto sea de al menos cerca de 20 veces el volumen interno entre la entrada de la solución del peróxido de hidrógeno y la salida. (Pág. 4, líneas 26 a 37 y Pág. 5, líneas 1 a 5).

El tiempo de residencia se calcula como el volumen de la cámara de reacción dividido entre el caudal de los reactivos. De lo anterior, se calcula que el tiempo de retención descrito en el documento D1 es de 3 segundos o menos.

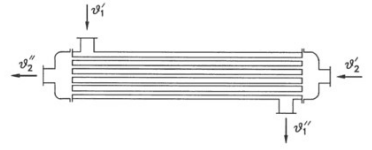
Reivindicación 11: Si se desea, el producto ácido de Caro producido en el generador de acuerdo con la presente invención puede pasarse a través de una unidad de enfriamiento dedicada, si, por ejemplo, se desea almacenar el producto antes que usarlo

inmediatamente, y en tales circunstancias, es preferible efectuar suficiente enfriamiento para bajar la temperatura del producto por debajo de 60 °C para mejorar su estabilidad en almacenamiento. (Pág. 13, líneas 1 a 7).

### Nivel inventivo (Art18 D486)

La reivindicación 1 consiste en “*Un dispositivo para mezclar y enfriar dos líquidos reactivos, caracterizado porque...*” (Ver Pág. 9 del Rad. N° 14-059449-00000-0000)

Comparación de las características técnicas **esenciales**, con las del Estado de la Técnica:

| Características <b>esenciales</b>                                                                                                                                                                                                                                          | <b>D2</b>                                                                                                                                                                                                                                                                                                                                                                                             | <b>D3</b>                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                |
|----------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------|-------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------|--------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------|
| Un dispositivo para mezclar y enfriar dos líquidos reactivos, caracterizado porque comprende                                                                                                                                                                               | Un aparato para mezclar dos fluidos (Pág. 1, Resumen).                                                                                                                                                                                                                                                                                                                                                | No se menciona.                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                          |
| un intercambiador de calor del tipo haz de tubos, donde dicho intercambiador de calor comprende una multiplicidad de tubos paralelos (1) dispuestos dentro de un mismo recinto (2), un espacio de cabeza (3) que se abre a un extremo de todos los tubos del haz de tubos, | No se menciona.                                                                                                                                                                                                                                                                                                                                                                                       | <p>Un aparato conocido como intercambiador de calor se usa para transferir energía de u fluido a otro. (Pág. 40, Párr. 1).</p> <p>En las aplicaciones prácticas, el diseño de intercambiador de calor de tubos y coraza es el diseño más común. Uno de los fluidos fluye por los muchos tubos paralelos que configuran un haz de tubos. El haz de tubos está rodeado por una coraza. El segundo fluido fluye alrededor del exterior de los tubos dentro de esta coraza. (Pág. 41, Párr. 2 y Pág. 42, Párr. 1).</p>  <p>Fig. 1.20: Shell-and-tube heat exchanger (schematic)</p>  <p>Fig. 1.21: Shell-and-tube heat exchanger with baffles.</p> |
| y una primera admisión (4) que da hacia el espacio de cabeza para introducir un primer líquido (5),                                                                                                                                                                        | Se muestra unido al tubo exterior <b>4</b> una entrada de un segundo fluido <b>3</b> , el cual está unido a un medio de suministro de fluido para permitir el paso de un segundo fluido dentro del volumen anular <b>6</b> entre los dos tubos concéntricos. (Col. 2, líneas 48 a 52).                                                                                                                | No se menciona.                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                          |
| donde el dispositivo comprende una segunda admisión (6) que da hacia el espacio de cabeza con una multiplicidad de boquillas (7) para introducir un segundo líquido (8)                                                                                                    | Se muestra unido al tubo interior <b>5</b> un plato <b>11</b> el cual conecta el tubo interior con una entrada de fluido <b>2</b> . Esta entrada de fluido es una tubería u otro recipiente portador de un fluido que permite que el fluido se mueva desde una fuente externa de suministro hasta dentro del volumen interno <b>13</b> , en el interior del tubo <b>5</b> . (Col. 2, líneas 37 a 41). | No se menciona.                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                          |

|                                                                                                                                                                                             |                                                                                                                                                                                                                                                                                                                                                                                   |                 |
|---------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------|-----------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------|-----------------|
| donde dichas boquillas están situadas dentro del espacio de cabeza y se las orienta para dirigir al líquido que se introduce en dirección transversal al eje de los tubos del haz de tubos. | El corazón de mi aparato se muestra en los puertos de salida de fluido <b>7</b> y <b>8</b> . Los puertos de salida <b>7</b> son agujeros que se taladran en las paredes laterales del tubo interno <b>5</b> para permitir el paso del fluido desde el interior de este tubo, a través de estos puertos, hasta el interior del volumen anular <b>6</b> . (Col. 2, líneas 53 a 57). | No se menciona. |
|---------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------|-----------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------|-----------------|

Los documentos D2 y D3 se consideran el estado de la técnica cercano a la invención definida en la reivindicación 1 porque divulgan un dispositivo para mezclar dos fluidos (D2, Pág. 1, Resumen) y un aparato para la transferencia de calor de un fluido a otro (D3, Pág. 41, Párr. 2 y Pág. 42, Párr. 1).

La diferencia entre la invención, definida en la reivindicación 1, y el dispositivo divulgado en el documento D2 es que en éste no se incluyen medios para enfriar la mezcla que sale del dispositivo.

La diferencia entre la invención, definida en la reivindicación 2, y el intercambiador de calor de tubos y coraza descrito en el documento D3 radica en que no se describe el equipamiento que suministra el fluido al que se quiere modificar su temperatura, por ejemplo, disminuir su temperatura.

El efecto que logra la introducción de la mezcla dentro del espacio de cabeza de un intercambiador de tubos y coraza es el de minimizar el tiempo entre la mezcla de los líquidos y su enfriamiento en los tubos (Pág. 4, Párr. 3).

Por lo tanto, el problema técnico objetivo que pretende resolver esta invención se puede formular así: “¿Cómo modificar el dispositivo conocido en D2 para minimizar el tiempo entre la mezcla de los líquidos y su enfriamiento en los tubos?”.

Sin embargo, la utilización de un intercambiador de tubos y coraza para modificar la temperatura de un fluido ya está descrita en el documento D3 que se refiere a un aparato conocido como intercambiador de calor que se usa para transferir energía de un fluido a otro. (Pág. 40, Párr. 1).

En consecuencia, la persona del oficio normalmente versada en la materia estaría motivada a incluir el intercambiador de tubos y coraza descrito en el documento D3, como parte del dispositivo divulgado en el documento D2 para llegar así al objeto de la reivindicación 1 de la solicitud, por lo cual se considera obvia y carente del requisito de nivel inventivo.

Así mismo, el documento D2 divulga las características de las reivindicaciones dependientes, así:

**Reivindicación 2:** Se muestran sobre el tubo interior los puertos de salida del fluido (**7**) los cuales están axialmente alineados desde el eje central del tubo interior. (Col. 3, líneas 7 a 9).

**Reivindicación 3:** Específicamente, el tubo interno es preferiblemente un tubo elongado al igual que el tubo exterior. La ubicación de los tubos internos y externos es tal que ellos están axialmente alineados. (Col. 3, líneas 35 a 38).

**Reivindicación 4:** En una implementación alternativa, el tubo interno puede tener su extremo sellado posicionado de tal manera que no está adyacente al extremo del tubo exterior. (Col. 4, líneas 40 a 42).

Los puertos de salida localizados en las conducciones internas y externas pueden ser cualquier tipo de aberturas que sean fácilmente mecanizadas o ubicadas sobre las conducciones externas o internas. (Col. 4, líneas 54 a 56).

Se prefiere ubicar uno o más anillos de hoyos sobre el tubo interior. Pueden utilizarse ranuras u otros tipos de espacios vacíos como puertos de salida. (Col. 4, líneas 59 a 64).

Reivindicación 5: Los componentes esenciales de mi aparato comprende tubos concéntricamente posicionados que tienen puertos en ellos para permitir el paso de un fluido por el tubo interno, a través de sus respectivos puertos, y dentro del interior de un tubo exterior a través del cual pasa un segundo fluido, el cual se mezcla entonces con el material que sale del tubo interior. (Pág. 3, Col. 1, líneas 36 a 42).

Por otra parte la reivindicación 7 se afecta con el conocimiento del técnico, el cual conoce que:

Reivindicación 7: Los mezcladores estáticos, también conocidos como mezcladores sin movimiento, se han convertido en equipos estándar en las industrias de procesos. Las operaciones de mezclado en tales industrias incluyen el mezclado clásico de fluidos miscibles en un flujo en una sola fase así como también el mejoramiento de la transferencia de calor, la dispersión de un gas en una fase líquida continua y la dispersión de una fase orgánica inmiscible como gotas en una fase acuosa continua<sup>5</sup>.

7. Conclusión

Los requisitos de Patentabilidad de la presente solicitud están afectados así:

| Item | Documento    | Reivindicaciones afectadas | Requisito que afecta    |
|------|--------------|----------------------------|-------------------------|
| D1   | WO 92/007791 | 8 a 11.                    | Novedad/Nivel Inventivo |
| D2   | US 3,818,938 | 1 a 5 y 7.                 | Nivel Inventivo         |
| D3   | Libro 1*     | 1                          | Nivel Inventivo         |

5 Static Mixers in the Process Industries—A Review. R.K. Thakur, Ch. Vial, K.D.P. Nigam, E.B. Nauman, G. Djelveh. Chemical Engineering Research and Design, 2003, 81, 7, 787-826.

En cumplimiento del artículo 45 de la Decisión 486 de la Comisión de la Comunidad Andina, se le indica que debe dar respuesta dentro del término de sesenta (60) días contados a partir del día siguiente de la fecha de notificación de la presente comunicación. En caso de no dar respuesta al presente requerimiento, o si a pesar de la respuesta subsistieran los impedimentos para la concesión, se denegará la patente.

Se invita al solicitante se sirva indicar, si así lo considera, su renuncia al término faltante de sesenta (60) días para dar respuesta a este requerimiento y presentarla por escrito.

Finalmente, se le recuerda al solicitante que si como respuesta a este requerimiento llegara a modificar el capítulo reivindicatorio y éste contenga más de diez (10) reivindicaciones, no pagará la tasa de modificación voluntaria pero sí la correspondiente al pago por reivindicación adicional.

Notifíquese el presente oficio conforme a lo dispuesto en el numeral 5.2, literal e), del capítulo quinto del título primero de la Circular Única No. 10 de 2001.



**JOSÉ LUIS SALAZAR LÓPEZ**  
Director de Nuevas Creaciones (c)

El anterior requerimiento fue notificado por fijación en lista, de fecha 2015-10-06

Elaboró: Diego Fernando Combita Merchan  
Revisó: Olga Ximena Benavides Salazar

REVIEW PAPER

# STATIC MIXERS IN THE PROCESS INDUSTRIES—A REVIEW

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This paper summarizes the field of static mixers including recent improvements and applications to industrial processes. The most commonly used static mixers are described and compared. Their respective advantages and limitations are emphasized. Efficiencies of static mixers are compared based both on theory and experimental results from the literature. The operations, which can benefit from the use of static mixers, are explored, namely, mixing of miscible fluids, liquid–liquid and gas–liquid interface generation, liquid–solid dispersion and heat transfer. Design parameters governing the performance of the various mixers in these applications are reported. The key parameters needed for the selection of a suitable mixer are highlighted.

*Keywords:* mixing; static mixers; inline mixing; process intensification; blending; interface generation; heat transfer; radial mixing; axial mixing.

## INTRODUCTION

Static mixers, also known as motionless mixers, have become standard equipment in the process industries. However, new designs are being developed and new applications are being explored. Static mixers are employed inline in a once-through process or in a recycle loop where they supplement or even replace a conventional agitator. Their use in continuous processes is an attractive alternative to conventional agitation since similar and sometimes better performance can be achieved at lower cost. Motionless mixers typically have lower energy consumptions and reduced maintenance requirements because they have no moving parts. They offer a more controlled and scaleable rate of dilution in fed batch systems and can provide homogenization of feed streams with a minimum residence time. They are available in most materials of construction. The some potential advantages of static mixers over conventionally agitated vessels are given in Table 1.

Although static mixers did not become generally established in the process industries until the 1970s, the patent is much older. An 1874 patent describes a single element, multilayer motionless mixer used to mix air with a gaseous fuel (Sutherland, 1874). An early French patent used staged, helical elements to promote mixing in a tube (Les Consommateurs de Petrole, 1931), and another French patent shows a multielement design for solids blending (Bakker, 1949). Staged elements to promote heat transfer were patented in the early 1950s (Lynn, 1958). Major petrochemical companies had development efforts and presumably utilized their designs internally in the decades proceeding the commercial availability (Stearns, 1953; Veasey, 1968; Tollar, 1966).

There are now approximately 2000 US patents and more than 8000 literature articles that describe motionless mixers and their applications. More than 30 commercial models are currently available.

The prototypical design of a static mixer is a series of identical, motionless inserts that are called *elements* and that can be installed in pipes, columns or reactors. The purpose of the elements is to redistribute fluid in the directions transverse to the main flow, i.e. in the radial and tangential directions. The effectiveness of this redistribution is a function of the specific design and number of elements. Commercial static mixers have a wide variety of basic

Table 1. Potential advantages of static mixer compared to mechanically agitated vessels.

| Static mixer                                               | CSTR                                                    |
|------------------------------------------------------------|---------------------------------------------------------|
| Small space requirement                                    | Large space requirement                                 |
| Low equipment cost                                         | High equipment cost                                     |
| No power required except pumping                           | High power consumption                                  |
| No moving parts except pump                                | Agitator drive and seals                                |
| Small flanges to seal                                      | Small flanges plus one large flange to seal             |
| Short residence times                                      | Long residence times                                    |
| Approaches plug flow                                       | Exponential distribution of residence times             |
| Good mixing at low shear rates                             | Locally high shear rates can damage sensitive materials |
| Fast product grade changes                                 | Product grade changes may generate waste                |
| Self-cleaning, interchangeable mixers or disposable mixers | Large vessels to be cleaned                             |

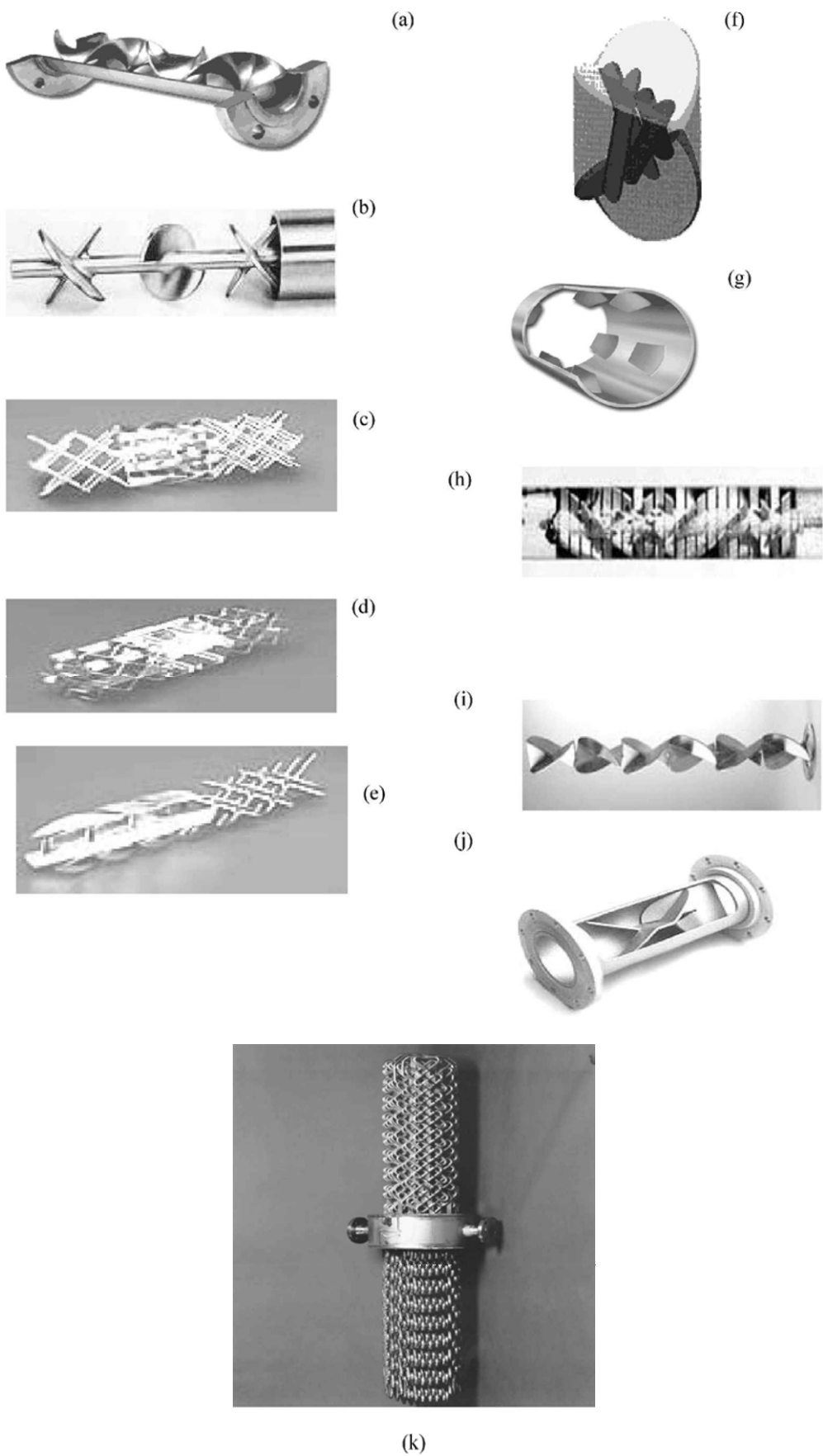


Figure 1. Elements of different commercial static mixers: (a) Kenics (Chemineer Inc.); (b) low pressure drop (Ross Engineering Inc.); (c) SMV (Koch-Glitsch Inc.); (d) SMX (Koch-Glitsch Inc.); (e) SMXL (Koch-Glitsch Inc.); (f) Interfacial Surface Generator-ISG (Ross Engineering Inc.); (g) HEV (Chemineer Inc.); (h) Inliner series 50 (Lightnin Inc.); (i) Inliner series 45 (Lightnin Inc.); (j) Custody transfer mixer (Komax systems Inc.); (k) SMR (Koch-Glitsch, Inc.).

Table 2. Commercially available static mixers.

| Company            | Mixers                                                             |
|--------------------|--------------------------------------------------------------------|
| Chemineer-Kenics   | Kenics mixer (KM), HEV (high efficiency vortex mixer)              |
| Koch-Sulzer        | Sulzer mixer SMF, SMN, SMR, SMRX, SMV, SMX, SMXL                   |
| Charles Ross & Son | ISG (interfacial surface generator), LPD (low pressure drop), LLPD |
| Wymbs Engineering  | HV (high viscosity), LV (low viscosity)                            |
| Lightnin           | Inliner Series 45, Inliner Series 50                               |
| EMI                | Cleveland                                                          |
| Komax              | Komax                                                              |
| Brann and Lubbe    | N-form                                                             |
| Toray              | Hi-Toray Mixer                                                     |
| Prematechnik       | PMR (pulsating mixer reactor)                                      |
| UET                | Heliflo (Series, I, II and III)                                    |

geometries and many adjustable parameters that can be optimized for specific applications. The number of elements in series is routinely adjusted. Another important parameter is the aspect ratio, defined as the ratio of length to diameter of a single element. Commercial designs typically use standard values for the various parameters that provide generally good performance throughout the range of applications and for which experimental data are available. Ordinarily, the use of standard designs is recommended. As a group, motionless mixers exhibit far higher efficiencies than rudimentary mixing devices such as elbows or tees and their performance is better characterized. Figure 1 illustrates commercial designs, and Table 2 lists manufacturers. Tables 3 and 4 summarize applications of static mixers in the process industries.

Table 3. Industrial applications of commercial static mixers.

| Mixer                     | Flow regime                                  | Area of application                                                                                                                                                                    | References                                                                                                                   |
|---------------------------|----------------------------------------------|----------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------|------------------------------------------------------------------------------------------------------------------------------|
| Kenics                    | Laminar/turbulent                            | Thermal homogenization of polymer melt<br>Gas-liquid dispersion<br>Dilution of feed to reactor                                                                                         | Chen (1975)<br>Smith (1978)                                                                                                  |
|                           | Turbulent<br>$980 < Re < 8500$               | Dispersion of viscous liquids<br>Enhancement of forced flow boiling heat exchanger                                                                                                     | Berkman and Calabrese (1988)<br>Azer and Lin (1980)                                                                          |
| SMX                       | Laminar                                      | Mixing of high viscosity liquids and liquids with extremely diverse viscosity, homogenization of melts in polymer processing                                                           | Koch-Glitsch Inc. (2001)<br><a href="http://www.kochglitsch.com/frmain_mixers.htm">www.kochglitsch.com/frmain_mixers.htm</a> |
| SMV                       | Turbulent                                    | Low viscosity mixing and mass transfer in gas-liquid systems                                                                                                                           | Koch-Glitsch Inc. (2001)<br><a href="http://www.kochglitsch.com/frmain_mixers.htm">www.kochglitsch.com/frmain_mixers.htm</a> |
| SMXL                      | Laminar                                      | Liquid-liquid extraction<br>Homogeneous dispersion and emulsions                                                                                                                       | Koch-Glitsch Inc. (2001)<br><a href="http://www.kochglitsch.com/frmain_mixers.htm">www.kochglitsch.com/frmain_mixers.htm</a> |
|                           |                                              | Heat transfer enhancement for viscous fluids                                                                                                                                           |                                                                                                                              |
| SMF                       |                                              | Sludge conditioning, pulp stock blending, bleaching and dilution, bleaching of suspension and slurries                                                                                 | Koch-Glitsch Inc. (2001)<br><a href="http://www.kochglitsch.com/frmain_mixers.htm">www.kochglitsch.com/frmain_mixers.htm</a> |
| SMR                       | Laminar                                      | Polystyrene polymerization and devolatilization                                                                                                                                        | Chemineer Inc. (1988)                                                                                                        |
| HEV                       | Turbulent                                    | Low viscosity liquid-liquid blending, gas/gas mixing                                                                                                                                   |                                                                                                                              |
| LPD                       | Laminar                                      | Blend two resins to form a homogeneous mixture                                                                                                                                         |                                                                                                                              |
|                           | Turbulent                                    | Blending grades of oil or gasoline                                                                                                                                                     |                                                                                                                              |
| LLPD                      | Turbulent                                    | Liquid-liquid dispersions                                                                                                                                                              | Ross Engineering Inc. (2001)<br><a href="http://www.mixers.com">www.mixers.com</a>                                           |
|                           | Laminar                                      | Blend out thermal gradient in viscous streams                                                                                                                                          | Ross Engineering Inc. (2001)<br><a href="http://www.mixers.com">www.mixers.com</a>                                           |
| ISG                       | Laminar                                      | Blending catalyst, dye or additive into viscous fluid                                                                                                                                  | Ross Engineering Inc. (2001)<br><a href="http://www.mixers.com">www.mixers.com</a>                                           |
|                           |                                              | Homogenization of polymer dope                                                                                                                                                         | Ross Engineering Inc. (2001)<br><a href="http://www.mixers.com">www.mixers.com</a>                                           |
|                           |                                              | Pipeline reactor to provide selectivity of product                                                                                                                                     | Ross Engineering Inc. (2001)<br><a href="http://www.mixers.com">www.mixers.com</a>                                           |
| Inliner mixer series 45   | Turbulent                                    | Waste water neutralization                                                                                                                                                             | Lightnin (2001)<br><a href="http://www.lightnin-mixer.com">www.lightnin-mixer.com</a>                                        |
|                           | Turbulent                                    | Fast reaction and blending application including widely differing viscosity, densities and fluid with unusual properties, such as polymer                                              |                                                                                                                              |
| Inliner mixer series 50   |                                              | Chemical and petrochemical systems, hydrocarbon refining, caustics, pulp and fast reactions                                                                                            | Lightnin (2001)<br><a href="http://www.lightnin-mixer.com">www.lightnin-mixer.com</a>                                        |
| SMV-4                     | $1400 < Re < 3700$<br>$16,000 < Re < 58,000$ | Fine liquid-liquid dispersions (water-kerosene)<br>Dispersion of immiscible fluids, e.g. water-kerosene<br>Phase inversion in liquid-liquid system, e.g. water-organic, water- $CCl_4$ | Al Taweel and Walker (1983)                                                                                                  |
| Static-mixer woven screen | Turbulent                                    | Dispersion of kerosene in water                                                                                                                                                        | Sembria <i>et al.</i> (1986)<br>Tidhar <i>et al.</i> (1986)                                                                  |
| Komax SM                  |                                              | Mixing food products such as margarine and tomato pastes, viscous liquids like syrups and light fluids like juices                                                                     | Al Taweel and Walker (1983)<br>Komax Systems Inc. (2001)<br><a href="http://www.komax.com">www.komax.com</a>                 |

Table 4. Applications of static mixers in the process industries.

| Industry                            | Application                                                       |
|-------------------------------------|-------------------------------------------------------------------|
| Chemical and agricultural chemicals | Reaction enhancement                                              |
|                                     | Gas mixing                                                        |
|                                     | Organic–aqueous liquid–liquid dispersions                         |
|                                     | Fertilizer and pesticide preparation                              |
|                                     | Steam injection                                                   |
| Grain processing                    | Acid–base neutralization                                          |
|                                     | Continuous production and conversion of starch                    |
|                                     | Livestock feed mixing                                             |
| Food processing                     | Liquid blending and emulsification                                |
|                                     | Starch slurry cooking                                             |
|                                     | Heating and cooling of sugar solutions                            |
|                                     | Solid ingredient blending                                         |
| Minerals processing                 | Mineral recovery by solvent extraction                            |
|                                     | Slurry dilution                                                   |
|                                     | Oxidation and bleaching                                           |
|                                     | Chemical addition and bleaching                                   |
|                                     | Gaseous reactant blending                                         |
| Petrochemicals and refining         | Gasoline blending                                                 |
|                                     | Caustic scrubbing of H <sub>2</sub> S and CO <sub>2</sub>         |
|                                     | Lube oil blending                                                 |
| Pharmaceuticals and cosmetics       | Emission monitoring and control                                   |
|                                     | Mixing of trace elements                                          |
|                                     | Blending of multicomponent drugs                                  |
|                                     | Dispersion of oils                                                |
|                                     | Sterilization                                                     |
| Polymer, plastics and textiles      | pH control                                                        |
|                                     | Continuous production of polystyrene                              |
|                                     | Mixing of polymer additives                                       |
|                                     | Preheating polymers                                               |
|                                     | Thermal homogenization                                            |
|                                     | Fiber spinning                                                    |
| Paints, resins and adhesives        | Tubular finishing reactors                                        |
|                                     | Dilution of solids (e.g. TiO <sub>2</sub> )                       |
|                                     | Coloring and tinting                                              |
|                                     | Adhesive dispensing and heating                                   |
| Pulp and paper                      | Pulp bleaching                                                    |
|                                     | Stock dilution and consistency control                            |
| Water and waste water treatment     | pH control                                                        |
|                                     | Addition of coagulating agent                                     |
|                                     | Disinfection (Cl <sub>2</sub> , O <sub>2</sub> , O <sub>3</sub> ) |
|                                     | Dechlorinating                                                    |
|                                     | Sludge dewatering process                                         |
|                                     | pH control                                                        |

Mixing operations are essential in the process industries. They include the classical mixing of miscible fluids in single-phase flow as well as heat transfer enhancement, dispersion of gas into a liquid continuous liquid phase, dispersion of an immiscible organic phase as drops in a continuous aqueous phase, three-phase contacting and mixing of solids. Static mixers have been applied to all these applications, including liquid–liquid systems (e.g. liquid–liquid extraction), gas–liquid systems (e.g. absorption), solid–liquid systems (e.g. pulp slurries) and solid–solid systems (e.g. solids blending). Some processes that use commercial static mixers are summarized in Table 3. Static mixers are now commonly used in the chemical and petrochemical industries to perform continuous operations. They have also found applications in the pharmaceutical, food engineering and pulp and paper industries (Table 4). The effectiveness of static mixers for the mixing of miscible fluids or to enhance heat transfer is due to their ability to perform radial mixing and to bring fluid elements into close proximity so that diffusion or conduction becomes rapid. In laminar flows,

static mixers divide and redistribute streamlines in a sequential fashion using only the energy of the flowing fluid. In turbulent flows, they enhance turbulence and give intense radial mixing, even near the wall. In both cases, they can significantly improve heat and mass transfer operations.

The various types of static mixers behave quite differently, and classification schemes have been proposed to explain these differences based on the geometry of the mixing elements (Baker, 1991; Myers *et al.*, 1997). The commercial static mixers can be divided into five main families: namely, open designs with helices (Figure 1a), open designs with blades (Figure 1b, g–j), corrugated-plates (Figure 1c), multi-layers designs (Figure 1d and e), and closed designs with channels or holes (Figure 1f). The applications of these designs can be classified in four groups:

- group 1—mixing of miscible fluids;
- group 2—interface generation between non-miscible phases;
- group 3—heat transfer operation and thermal homogenization;
- group 4—axial mixing.

Group 1 can be divided into two subgroups, depending on whether the prevailing flow regime is laminar or turbulent. The mixers in group C1 are intended to achieve composition homogeneity in the directions transverse to the predominant flow, e.g. in the radial direction. Applications of group 2 depend on the nature of the phases: gas–liquid, immiscible liquid–liquid, liquid–solid and solid–solid operations can be distinguished. Group 1 includes applications to homogeneous reactions. Group 2 includes multiphase reactions coupled with separation processes, such as reactive absorption. Structured packings, used to replace trays and random packings in distillation and other mass transfer operations, are a form of static mixer and are briefly discussed. Group 3 includes traditional thermal homogenization and heat transfer in heat exchangers involving viscous fluids in the laminar regime, such as polymer solutions. Static mixing elements can also be used in turbulent flow to reduce the exchanger size. Group 3 mixers can be used for highly exothermic chemical reactions. Group 4 mixers are an entirely new type with the design intention of promoting mixing and specifically to approximate the residence time behavior of a continuous flow stirred tank with moving parts.

Coupled with the examples in Table 3, Figure 2 provides a first basis for selecting a mixer appropriate to a specific operation. However, such a simplified procedure provides no quantitative information on mixing effectiveness, pressure drop, the optimum number of elements and scale-up. The situation is far more complex when two operations are performed simultaneously, e.g. for two-phase reacting flows or for highly exothermic reactions with unmixed feed streams. The main purpose of this paper is to give guidelines for the selection of static mixers in industrial processes. The paper is divided into four parts. The first part discusses mixing mechanisms and local phenomena in static mixers to better understand how mixing, heat transfer or interface generation proceed. The second part reviews well-established applications using inline mixers for process intensification and highlights potential applications that are still under development. The third part deals with the key parameters

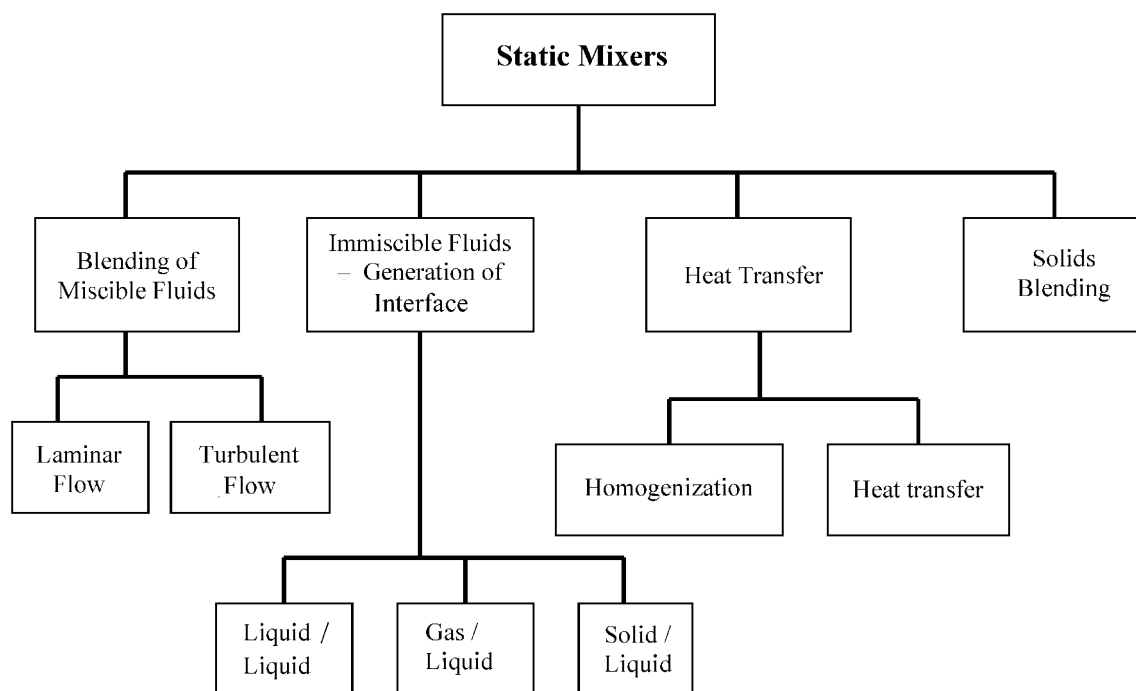


Figure 2. Classification of unit operations using static mixers.

and the methodology for proper selection, evaluation and scale-up for various unit operations. Methods for estimating the performance of static mixers from experimental data, literature correlations and computational fluid dynamics (CFD) are discussed. The last part of the paper discusses ongoing and future developments in the field.

## FUNDAMENTALS

This section deals with fundamental aspects and local phenomena in static mixers. Groups 1 and 3 are discussed together because similar phenomena are responsible for mixing and heat transfer enhancement. However, laminar and turbulent flows must be distinguished because the microscopic mechanisms of mixing differ widely between them. Later, fundamental aspects of interface generation operations, group 2, are discussed. There are three main families of applications for group 3 mixers, gas-liquid, liquid-liquid and solid-liquid, and two main objectives, to obtain stable dispersions and to increase interfacial area for mass transfer.

### Mixing of Miscible Fluids—Distributive Mixing

#### *Static mixers in laminar flow*

Pressure limitations prevent turbulent flow in high viscosity fluids. Suppose an empty pipe has two feed streams, a black liquid and a white liquid, each liquid occupying a semicircular section of the tube as shown in Figure 3(a). In undisturbed laminar flow, the streamlines are straight and the two liquids will emerge from the tube exactly as they entered except for a graying at the interface where they mixed by molecular diffusion. There is no convective mixing in either the tangential or radial directions. Thus a similar situation arises when the two fluids are separated

in the radial direction as illustrated in Figure 3(b). In either case, heat transfer or reaction between the liquids and heat transfer to the tube wall will be poor. Molecular diffusion can provide mixing in capillary tubes, but the effects of diffusion lessen upon scale-up and are rarely sufficient to provide adequate mixing in industrial-scale equipment.

The low levels of mixing in undisturbed laminar flow have the obvious consequence of giving *spatial* inhomogeneities in composition. Conventional static mixers are designed to homogenize the fluid by redistributing it in the radial and tangential directions. Undisturbed laminar flow also gives *temporal* inhomogeneities in the sense that molecules leaving the tube at some instant will have entered at different times. The same redistribution of fluid that gives spatial mixing also gives temporal mixing. In the ideal case of plug flow, which is also known as piston flow, the black and white liquids will be uniformly gray when they leave the tube, and all the molecules leaving together

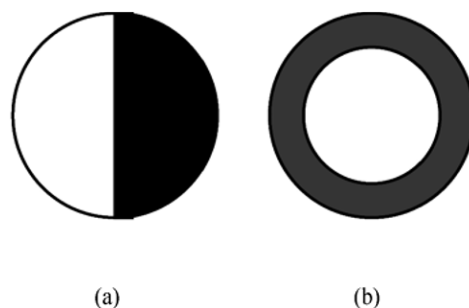


Figure 3. Spatial inhomogeneities: (a) tangential variations; (b) radial variations.

will have entered together. The extent to which actual static mixers approach is ideal can be characterized using the residence time distribution for temporal variations and using the striation thickness distribution for spatial variations.

Figure 4 shows response curves for several flow geometries to a sudden, step input of an inert tracer. The resulting, monotonically increasing curve is the cumulative distribution function of residence times,  $F(t)$ . The reference case is an empty pipe under laminar flow (Poiseuille flow) in the absence of diffusion. The broad distribution of residence times for this case is due to the parabolic velocity profile, and the redistribution caused by the static mixing elements gives a residence time distribution that more closely approaches the sharp distribution of plug flow. The nearness of the approach to plug flow can be characterized by the first appearance time,  $t_{\text{first}}$ , where the tracer first appears at the outlet of the mixer. The dimensionless first appearance time,  $t_{\text{first}}/\bar{t}$  where  $\bar{t}$  is the mean residence time, is 0.5 for the empty pipe since the highest velocity in parabolic flow is twice the average velocity. It is 1.0 for plug and will approach this value to the extent that the static mixing elements successfully redistribute flow in the radial direction. When the first appearance time is fuzzy, a 5% response time, defined as the time when  $F(t) = 0.05$ , provides a more accurate measure that also will approach 1.0 (from above) as the mixing environment approaches plug flow. Using this standard, the first appearance time for undisturbed laminar flow is 0.513. Table 5 shows values of  $t_{\text{first}}/\bar{t}$  for several flow geometries. The results for the Kenics static mixer are based on a model that, for each four mixing elements, inserts one plane of complete radial mixing into a tube that is otherwise in undisturbed laminar flow (Nauman, 1982). The Kenics does not excel by this measure of performance since 40 Kenics elements are needed to increase  $t_{\text{first}}/\bar{t}$  (measured by the 5% response method) to 0.676. The highest reported values for laminar, isoviscometric flow in a circular tube are those of Saxena and Nigam (1984), who achieved values in excess of 0.85 using a helically coiled tube with periodic,  $90^\circ$  changes in the direction of the coil axis.

Another common metric is the dimensionless variance of the residence time distribution. It can be calculated using

$$\sigma^2 = \frac{2 \int_0^\infty [1 - F(t)]t dt}{(\bar{t})^2} - 1 \quad (1)$$

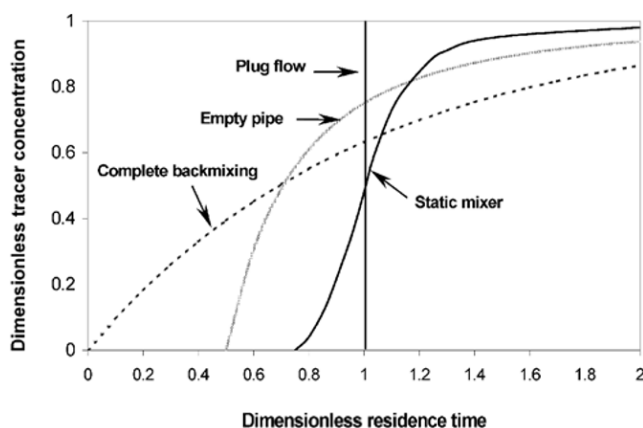


Figure 4. Cumulative residence time distributions for various flow systems.

Table 5. First appearance times for various laminar flow systems.

| Geometry                                                         | $t_{\text{first}}/\bar{t}$ |
|------------------------------------------------------------------|----------------------------|
| Equilateral-triangular ducts                                     | 0.450                      |
| Square ducts                                                     | 0.477                      |
| Straight, circular tubes                                         | 0.500                      |
| Straight, circular tubes (5% response)                           | 0.513                      |
| 16-element Kenics mixer                                          | 0.598                      |
| Helically coiled tubes                                           | 0.613                      |
| Annular flow                                                     | 0.500–0.667                |
| Parabolic flow between flat plates                               | 0.667                      |
| 40-element Kenics mixer (5% response)                            | 0.676                      |
| Complete flow inversion                                          | 0.707                      |
| Single screw extruder                                            | 0.750                      |
| Helical coils with changes in the direction of centrifugal force | > 0.85                     |

The dimensionless variance is zero for plug flow. It is theoretically infinite for laminar flow without diffusion but becomes finite in all real systems due to molecular diffusion (Nauman, 1981; Nauman and Buffham, 1983). An exponential extrapolation of experimental data is recommended for numerical integration of equation (1).

Static mixers were initially developed for blending fluids in laminar flow. Applications to heat transfer, turbulence and multiphase systems appeared later. The first static mixers were designed to achieve good mixing in the cross-section of a circular tube for a fluid in laminar flow. This naturally requires a greater pressure drop than for the empty pipe, but the additional power is lower than needed for similar mixing efficiencies with mechanical agitation or to achieve turbulence. Motionless inserts such as blades or corrugated plates induce changes in the fluid streamlines. Inserts with holes, channels, helical elements and oblique blades cause local acceleration and stretching of the fluid. They split the incoming fluid into layers and then recombine the layers in a new sequence. Multilayer designs (Figure 1d and e) with blades and baffles split the fluid in multiple layers. These various mixing actions cause *distributive mixing*. It is mixing caused by convection rather than diffusion, although to the extent that distributive mixing is high, diffusion is better able to achieve homogeneity on a molecular scale.

The *striation thickness* (Mohr *et al.*, 1957) is used to quantify distributive mixing. Figure 5 illustrates the concept and shows how the striation thickness,  $S$ , decreases when the fluid is sheared in a direction perpendicular to the initial striations. Correspondingly, the interfacial area between the white and black fluids increases. With a large enough displacement of the upper surface, the striation thickness will drop below the resolving power of the eye and become small enough that molecular diffusivity will eliminate concentration differences. However, the efficiency of simple shear becomes quite low as the striations become oriented in the direction of the shear. Much greater efficiency is possible if the direction of shear is periodically changed so that it again becomes perpendicular to the striations. Splitting and recombining the striations into new patterns also improves the efficiency. These various mechanisms are illustrated in Figure 6 in which layer generation, stretching, splitting and layer recombination are shown. Figure 6 represents a  $2^N$  mixer where the number of fluid layers is increased by a factor 2 by each

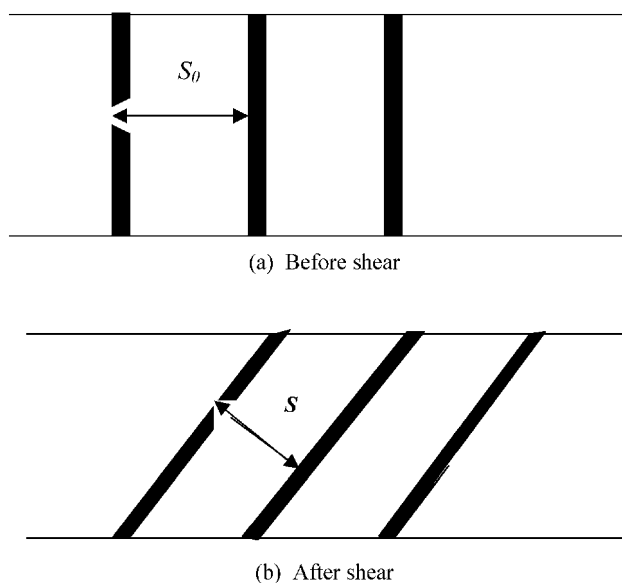


Figure 5. Response of striation thickness to simple shear. (a) Initial configuration; (b) after shear.

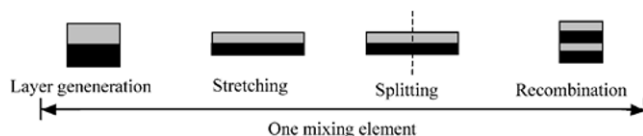


Figure 6. Two-step mechanism for mixing in laminar flow using static mixers.

element and therefore increases by a factor of  $2^N$  for  $N$  elements. The Kenics and the LPD (Low-pressure drop) static mixers (Figure 1a and b) are classified as  $2^N$  mixers because they split the incoming stream into two flow paths as illustrated in Figure 6. The ISG mixer has four channels (Figure 1f) and is therefore classified as a  $4^N$  mixer, but the higher theoretical efficiency comes at the price of greater pressure drop. Multilayer designs, such as the SMX (Figure 1d) and the SMV (Figure 1c) have still more flow paths. Godfrey (1992) tabulated the relationship between the number of striations and the number of elements for several commercial static mixers (Table 6). Figure 7 explains the mechanism of laminar mixing proposed by Edwards (1992). He considered three idealized flow situations: simple shear, uniaxial extension and planar extension. Edwards observed that extensional flows are slightly more effective for distributive mixing at low strain than simple shear flows and become much more effective when the strain is large, due to the unfavorable orientation of striations that occurs during

Table 6. Theoretical number of striations generated by commercial static mixers.

| Kenics | ISG   | Inliner      | SMV               |
|--------|-------|--------------|-------------------|
| $2^N$  | $4^N$ | $3(2)^{N-1}$ | $n_c(2n_c)^{N-1}$ |

$n_c$ , number of channels.

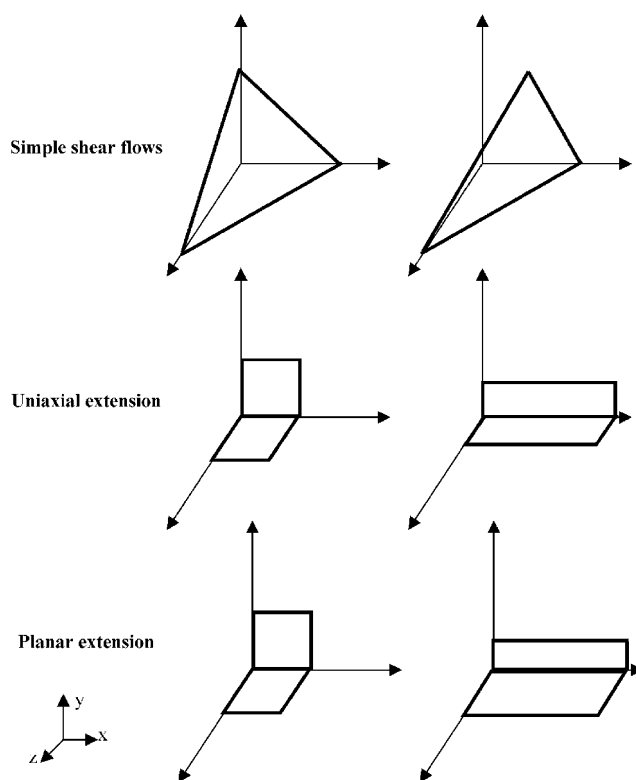


Figure 7. Increase in surface area of fluid elements under ideal laminar flow conditions.

simple shear. However, Edwards concluded that simple shear flows are more effective at intermediate strain levels, especially with non-Newtonian fluids. In deep laminar flow, i.e. in creeping flow, the flow patterns around a static mixing element are completely deterministic. The mixers work very effectively in this regime, using the mechanisms illustrated in Figures 5–7. Indeed, their greatest benefit compared to an empty pipe for miscible fluid blending and heat transfer occurs in the limit of creeping flow. At Reynolds numbers greater than a few hundred (based on the empty tube), flow instabilities lead to downstream oscillations and pseudorandom behavior. Even in creeping flow, mixing elements in series asymptotically approach a condition known as *chaos* where the downstream location of a fluid element becomes essentially unpredictable based on its upstream location. There are several ways of characterizing these complex flows. Some practical measures include the degree of completion of a fast chemical reaction, the residence time distribution and heat transfer coefficients. These practical measurements, expressed as correlations of experimental data, are by far the most valuable to the process engineer. However, mathematical characterization techniques will become more useful as CFD (computational fluid dynamics) becomes better developed. Mapping methods have proved useful (Kruijt *et al.*, 2001a, b). Residence time distributions can be determined from CFD results, although the reader is cautioned that the calculations must be weighted by volumetric or mass flow rate rather than area (Nauman, 1991). Also, some CFD codes exhibit numerical diffusion that can become significant in detailed calculations, particularly those involving reactions.

A more sophisticated characterization scheme was provided by Manas-Zlockzower (1994) using the *extensional efficiency parameter*,  $\alpha$ . The velocity gradient tensor  $\bar{\bar{s}}$  is divided into its symmetric and anti-symmetric parts. These are, respectively the rate of deformation tensor  $\bar{\bar{\gamma}}$  and the vorticity tensor  $\bar{\bar{\omega}}$ . If  $\mathbf{v}$  is the local velocity vector,  $\bar{\bar{s}}$ ,  $\bar{\bar{\gamma}}$  and  $\bar{\bar{\omega}}$  are defined as:

$$\begin{aligned}\bar{\bar{s}} &= \frac{1}{2} \cdot [(\nabla \mathbf{v} + \nabla \mathbf{v}^T) + (\nabla \mathbf{v} - \nabla \mathbf{v}^T)] \\ \bar{\bar{s}} &= \bar{\bar{\gamma}} + \bar{\bar{\omega}}\end{aligned}\quad (2)$$

The extensional efficiency  $\alpha$  is then expressed as the ratio:

$$\alpha = \frac{|\bar{\bar{\gamma}}|}{|\bar{\bar{\gamma}}| + |\bar{\bar{\omega}}|} \quad (3)$$

where  $|\bar{\bar{\gamma}}|$  and  $|\bar{\bar{\omega}}|$  are the norms of the tensors. For pure rotational flow,  $|\bar{\bar{\gamma}}| = \alpha = 0$ ; for pure extensional flow,  $|\bar{\bar{\omega}}| = \alpha = 1$ ; for a simple shear flow (for example, in an empty pipe),  $|\bar{\bar{\omega}}| = |\bar{\bar{\gamma}}|$  and  $\alpha = 0.5$ .

The extensional efficiency in static mixers ranges between 0.5 and 1, but can exhibit large variations along the length of an element (Rauline *et al.*, 1998).

For the process design engineer, the most useful statistic might be the reduction in maximum striation thickness as a function of the number of elements. The maximum striation thickness at the inlet to the mixer is the tube diameter (see Figure 3) and decreases by a factor of  $2^N$  according to simple theory. In practice, this prediction is overly optimistic, particularly when the fluids being blended have different viscosities. Correlations based on CFD calculations would be helpful for design, optimization and scale-up of process equipment. Detailed CFD calculations are still too expensive for complex optimization studies, but may be suitable for confirming a final design.

#### Static mixers in turbulent flow

When turbulence can be achieved, eddy diffusion gives sufficient mixing for most industrial processes. Consequently, static mixers are less used and less studied in turbulent flow systems. Mixer vendors often claim that static mixers can significantly reduce contact time or increase heat transfer compared to an empty pipe. This is true for laminar flow, but it is less certain for turbulent flow. It is true, however, that static mixers can increase the level of turbulence without changing pipe diameter and flow rate, albeit with a higher pressure drop.

Relatively few studies have been aimed at explaining how mixing proceeds in static mixers under turbulent conditions. We cite here the work of Goldschmid *et al.* (1986) and of Bourne and coworkers (Bourne and Maire, 1991; Bourne *et al.*, 1992; and Baldyga *et al.*, 1997). We distinguish two mechanisms by which energy is dissipated in turbulent flow: the energy dissipation,  $\varepsilon_1$ , due to boundary layers at the walls and surfaces of the inserts, and the energy dissipation in the bulk fluid,  $\varepsilon_2$ , i.e. in the region of

approximately homogeneous core turbulence. It is this second form of energy dissipation that is most important in distributive mixing. Consider an unmixed feed stream containing black and white liquid. Turbulence will quickly intermingle the initially unmixed feed and disperses it down to the size of the smallest eddies. In concept, the fluid in these smallest eddies remains black or white before becoming gray due to molecular diffusion. The size of the smallest eddies is proportional to the Kolmogoroff length scale:

$$\eta = \left( \frac{\mu^3}{\rho^3 \varepsilon'} \right)^{1/4} \quad (4)$$

where  $\mu$  is the viscosity,  $\rho$  is the density, and  $\varepsilon'$  is the power dissipation per unit mass of fluid.

The weak dependence on power dissipation, i.e. to the 0.25 power, means that  $\eta$  varies over a relatively narrow range, from 5 to 50  $\mu\text{m}$  for the great majority of industrial processes. It happens that the constant of proportionality between the Kolmogoroff scale and the smallest eddy size is on the order of 1. Given typical diffusion coefficients for low viscosity fluids (i.e. those for which turbulence is possible), complete mixing will occur in milliseconds. Except for very fast reactions, the reaction rate will be governed by the intrinsic kinetics without regard for mixing effects.

The total energy dissipation,  $\varepsilon'$ , is divided between that in the boundary and core regions:

$$\varepsilon' = \varepsilon_1 + \varepsilon_2 \quad (5)$$

Only the second contribution,  $\varepsilon_2$ , enhances the rate of fast chemical reactions. The objective is therefore to achieve high  $\varepsilon_2$  values at a low pressure drop level, which means that the ratio  $\varepsilon_2/\varepsilon'$  must be high. Bourne and his coworkers have devised a set of fast, competitive-consecutive reactions, the selectivity of which is very sensitive to mixing on the molecular scale. They used these reactions (Bourne *et al.*, 1992) to show that  $\varepsilon'$  in an empty pipe is about  $5 \text{ W kg}^{-1}$ , while SMV and SMXL static mixers can generate respectively 793 and  $509 \text{ W kg}^{-1}$  when the fluid velocity is  $2 \text{ m s}^{-1}$ . For these mixers,  $\varepsilon_2$  values calculated by Bourne *et al.* were 33 and 66% of the  $\varepsilon'$  values:  $\varepsilon_2 = 262 \text{ W kg}^{-1}$  for the SMV and  $\varepsilon_2 = 363 \text{ W kg}^{-1}$  for the SMXL. This shows that the SMV mixer, which uses a corrugated-plate design and causes the highest pressure drop, is less efficient for reaction enhancement than the SMXL mixer, which uses a multilayer design. The SMXL reduces the Kolmogoroff scale,  $\eta$ , by a factor of about 3. The characteristic time for diffusive mixing varies as  $\eta^{-2}$  and thus decreases by a factor of about 9. Baldyga *et al.* (1997) have characterized the influence of static mixing elements on the characteristic times for micromixing and mesomixing. A model to estimate these times and to predict yields of fast complex reactions was proposed, but has not yet been validated in other systems.

A difficulty with using static mixer to enhance fast chemical reactions is that the resulting turbulence is less homogeneous than in an empty tube. The Kolmogoroff scale depends on the local rate of turbulent energy dissipation and thus varies from point to point within

the bulk fluid, leading to localized regions where the reaction occurs preferentially.

### Solids blending

Static mixers fed by gravity are used for the blending of particulate solids such as cereal grains, bread and cake mixes, and concrete components. A theoretical analysis of homogeneity or the avoidance of segregation was conducted by Wang and Fan (1977). As a practical matter, relatively satisfactory mixing can usually be accomplished in a few division and recombination steps (Bakker, 1949).

### Interface Generation—Dispersive Mixing

Dispersing a secondary phase in a continuous phase is a common operation in the process industries. Static mixers can produce locally high shear rates, and thus can be used as interface generators in multiphase flows. The objective is generally to increase mass transfer in unit operations such as gas–liquid absorption, liquid–liquid (liquid–liquid) extraction or formation of polymer dispersion. However, the two-phase mixture may be the actual end product, as many emulsions are to be found in food, paint, dairy, cosmetic, adhesive and detergent industries. The performance of a static mixer for either objective is highly dependant on the physico-chemical properties of the phases. The volumetric flow rate ratio,  $\phi$ , the viscosity ratio  $R_\mu$  and the density ratio  $R_\rho$  for liquid–liquid and gas–liquid systems arise from dimensional analysis. These parameters are defined as:

$$\phi = \frac{Q_d}{Q_c}; \quad R_\mu = \frac{\mu_d}{\mu_c}; \quad R_\rho = \frac{\rho_d}{\rho_c} \quad (6)$$

The subscripts d and c represent the dispersed and the continuous phases and where  $Q$  is the volumetric flow rate of a phase. For viscoelastic systems, an elasticity ratio,  $R_e$ , is important as well.

The mechanisms of breakage and coalescence of gas bubbles and liquid drops are similar. They exhibit however three main differences:

- (i) liquid drops can be far smaller than bubbles—minimum diameter is around 10–100  $\mu\text{m}$  for bubbles and 0.1–1  $\mu\text{m}$  for drops;
- (ii)  $R_\mu$  and  $R_\rho$  are quite different for liquid–liquid and gas–liquid systems;  $R_\rho$  is in the range  $10^{-3}$ – $10^{-2}$  when gas is the dispersed phase and about 1 when a liquid is dispersed in another liquid (static mixers have been used to achieve uniformity for liquid-in-gas systems, e.g. in carburetion, but not to generate new interfacial area);
- (iii) In gas–liquid systems, it is generally easy to know whether gas is the dispersed based on gas and liquid flow rates. In liquid–liquid systems, phase inversion is possible, especially under shear conditions, and surface chemistry is more important than in gas–liquid systems. Flow stability is an additional parameter that must be considered.

Surface chemistry, and particularly the presence of strong surfactants, often plays a dominant role in multiphase systems. Static mixers are commonly used to create emulsions and dispersions and emulsions that are partially stabilized by surfactants. However, system performance is very

case-specific and difficult to generalize. For this reason, most published results have been confined to clean, binary systems. Surface forces are also important in liquid–solid and gas–solid systems where agglomeration or flocculation can occur. Colloidal forces predominate when liquid drops or solid particles or drops are lower than 1  $\mu\text{m}$ , for example in stable emulsions stabilized by surface-active agents (Peters, 1992). This is also the case in semi-dilute and concentrated suspensions of small particles that exhibit complex behavior such as sol–gel transitions. Moreover, solid particles have more complex shapes than small bubbles or drops that are generally spherical. Steric and electrosteric stabilization can play a role with solid particles (So *et al.*, 2001). The behavior of solid suspensions is less well understood than that of gas–liquid and liquid–liquid dispersions and is therefore more difficult to predict. Applications of static mixers to liquid–solid and gas–solid systems include solids blending in a fluid phase, the dispersion of additives in a suspension, e.g. for clarification or sludge treatment, and solid dispersion by breaking agglomerates in a fluid phase. The last application generates interfacial area. The others are blending operations similar to those described in the first section, although they technically involve multiple phases.

### Gas–liquid and liquid–liquid systems

Surface tension opposes drop break-up and favors coalescence. In a clean, quiescent system, the equilibrium configuration is complete phase separation with minimal interfacial area. In a flow system, fluid motion causes drop break-up, and a dynamic equilibrium is possible. When strong surface agents are present, particularly when they are ionic surfactants that give rise to electrostatic forces, liquid drops can become thermodynamically stable even in quiescent systems. This is true for oil-in-water micro-emulsions that have particle sizes less than 0.1  $\mu\text{m}$ . Even ordinary emulsions with sizes from about 0.2 to 0.5  $\mu\text{m}$  have long shelf lives. Motionless mixers are suitable for blending such systems and may be used to create them in chemically favorable cases. However, most uses are to augment the breakup of drops greater than about 1  $\mu\text{m}$ . Even here, surface tension,  $\sigma$ , is a key parameter and may be beneficially lowered using surface agents. Operation is dramatically affected by the viscosity ratio,  $R_\mu$ , and by whether the continuous phase is in turbulent or laminar flow. Break-up in either flow regime becomes increasingly difficult as  $R_\mu$  exceeds 1. Non-Newtonian viscous behavior (e.g. shear thinning) in either phase is relatively unimportant, but elasticity can seriously impede break-up. Its effects are difficult to predict quantitatively. Fluids suitable for turbulent flow are typically Newtonian.

Static mixers are used in multiphase systems to decrease dispersed phase drop sizes and to increase interfacial mass transfer. Ideally, a more uniform distribution of drop size might also be possible, compared say to what is possible in an agitated vessel. In practice, mixer performance is typically characterized by a single parameter such as the mass transfer coefficient,  $K_{La}$ , or the Sauter mean drop diameter,  $d_{32}$ . Peters (1992) summarized the phenomena that may cause the disruption of a plane surface separating two phases in order to create liquid drops. They include turbulent eddies, surface ripples and the associated Rayleigh–Taylor instabilities, gravitational instabilities arising from density

differences, and Kelvin–Helmholtz instabilities arising from relative motion of the phases. Similar phenomena are reported for gas bubbles (Lin *et al.*, 1998).

#### Continuous phase in laminar flow

The Rayleigh–Taylor instability, which governs the break-up of elongated fluid structures due to surface tension, is the most important in laminar flow. Elongated structures are metastable and will break into drops of a characteristic size, although smaller satellite drops are often formed as well. Drops that closely approach each other due to Brownian or convective motion will coalesce. A dynamic equilibrium can be achieved in which the break-up and coalescence rates are equal and the mean particle size becomes constant. The key parameter regulating this equilibrium is the Weber number:

$$\text{We} = \frac{\tau_c d_{32}}{\sigma} \quad (7)$$

where  $\tau_c$  is the shear stress in the continuous phase. An alternative to the Weber number is based on the tube diameter,  $D$ , rather than the drop diameter,  $d_{32}$ , and is called the capillary number

$$\text{Ca} = \frac{\tau_c D}{\sigma} \quad (8a)$$

For a Newtonian fluid,

$$\text{Ca} = \frac{\mu_c u}{\sigma} \quad (8b)$$

where  $u$  is the velocity of the continuous phase. The Sauter mean diameter is found from a correlation of the form

$$\frac{d_{32}}{D} = f(\text{Re}_c, \text{We}, \phi, R_\mu, R_\rho) \quad (9a)$$

or

$$\frac{d_{32}}{D} = f(\text{Re}_c, \text{Ca}, \phi, R_\mu, R_\rho) \quad (9b)$$

where  $\text{Re}$  is the Reynolds number based on the mean drop diameter but on physical properties of the continuous phase:

$$\text{Re}_c = \frac{\rho_c u d_{32}}{\mu_c} \quad (10)$$

The directions in which physical properties and operating conditions affect these correlations is reasonably well known. An increase in flow rate or in the viscosity of the continuous phase  $\mu_c$  tends to decrease  $d_{32}$  while an increase in the viscosity of the dispersed phase  $\mu_d$  tends to increase it. It is far easier to divide drops when  $R_\mu < 1$  than when  $R_\mu > 1$ . In some static mixers, drop breakup becomes essentially impossible when  $R_\mu$  exceeds about 10 (Grace, 1982). Increasing  $R_\rho$  tends to decrease bubble size, as stable gas bubbles are known to be larger than stable liquid drops. The influence of the dispersed phase volume fraction is not always included in correlations for  $d_{32}$ , but coalescence is clearly favored by higher values of  $\phi$ . At very high values, changes in the flow pattern or phase inversion are possible. If the dispersed phase is a viscoelastic liquid, deformation and breakup are more difficult. In contrast, a viscoelastic continuous phase induces higher deformation (Mighri *et al.*, 1997, 1998). When both phases

are elastic, deformation decreases as the elasticity ratio increases and the drops resulting from break-up are more uniform in size.

Figure 8 shows some drop deformation and breakup patterns suggested by Peters (1992) for idealized laminar flow. Either shear or extensional flows will elongate the drops as a preliminary step to break-up by Rayleigh–Taylor instability. Grace (1982) and Rallison (1984) found that simple shear flows (for example Couette flow) are less effective than extensional flows to break drops in liquid–liquid systems. The same conclusion is assumed to hold for gas–liquid systems. Based on Figure 8, the deformation  $DF$  is expressed as:

$$DF = \frac{L_p - B}{L_p + B} \quad (11)$$

Rupture occurs when  $DF$  is higher than a critical deformation value. Figure 9 shows the stability diagram proposed by Grace (1982) for drops in shear and extensional flows. The lines in that figure define the critical capillary numbers that separate breakage and coalescence, breakage occurring in the region above the lines. Grace (1982) defined a characteristic time  $t^+$  of drop breakage as:

$$t^+ = \frac{t\sigma}{\mu_c d_{32}} \quad (12)$$

He concluded that  $t^+$  is approximately proportional to  $R_\mu$ .

Taylor (1932, 1934) used a four roller apparatus to elongate drops. He found that very large deformations could be achieved without break-up under steady-state conditions, but that break-up occurred when the rolls were stopped. The experiments of Grace (1982) and Rallison (1984) studied only steady-state laminar flows. Stone and Leal (1989a,b) and Stone (1994) have shown that drops can be broken at much lower values of the capillary number if the flow is unsteady. This has been confirmed for a variety of unsteady flow systems by Khakar and Ottino (1986, 1987), Muzzio and Tjahjadi (1991), Tjahjadi and Ottino (1991), Tjahjadi *et al.* (1992), and Jansen and Meijer (1993).

The flow fields generated by a series of static mixing elements are periodic and thus favor drop break-up compared to flow in an empty pipe. High values of the extensional efficiency,  $\alpha$ , are preferred, but they must be

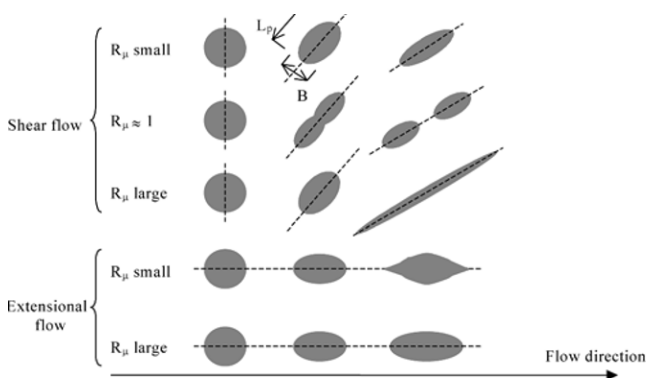


Figure 8. Drop deformation under idealized flow conditions (adapted from Peters, 1992).

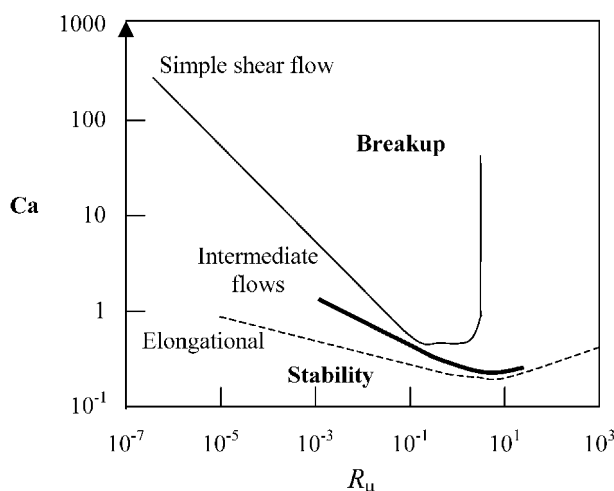


Figure 9. Stability diagram for drops (adapted from Grace, 1982).

accompanied by zone of relaxation. The parameter  $\alpha$  cannot be directly used to compare static and dynamic mixers, as it is not an invariant under a change of reference frame (Rauline *et al.*, 1998). The performance of static mixing element for interface generation in the laminar regime is very sensitive to the local flows of the continuous phase, and it has been suggested that calculations of  $\alpha$  in single-phase flows can be used to estimate the performance of these mixers for interface generation. It is becoming possible to perform these calculations using CFD codes. Examples of such calculations are the studies by Avalosse and Crochet (1997) with helical Kenics elements, Fradette *et al.* (1998) with the SMX mixer, Rauline *et al.* (2000) and especially Rauline *et al.* (1998) with six commercially available static mixers as shown in Figure 10. According to Rauline *et al.* (1998), a multilayer design, such as the SMX geometry,

leads to higher variations of  $\alpha$  along an element than found with a helical element of the Kenics type. As a result, SMX elements are probably more effective for bubble or drop generation than Kenics elements.

#### Continuous phase in turbulent flow

In turbulent flows, as the influence of the viscosity of the continuous phase  $\mu_c$  is not significant provided that drops are larger than the Kolmogoroff length scale, a different version of the Weber number is used to characterizes  $d_{32}$  at dynamic equilibrium:

$$We = \frac{\rho_c u^2 d_{32}}{\sigma} \quad (13a)$$

However, when the viscosity of the dispersed phase is high, equation (13a) has to be modified by introducing a viscosity group  $Vi$  and a correction function  $f(Vi)$  as follows:

$$We = \frac{\rho_c u^2 d_{32}}{\sigma} \cdot [1 + f(Vi)] \quad (13b)$$

Several definitions can be found for  $Vi$ . Berkman and Calabrese (1988) suggested:

$$Vi = \frac{\mu_d u}{\sigma} \cdot \sqrt{R_\rho} \quad (14)$$

The function  $f(Vi)$  tends to 0 when  $Vi$  approaches 0 (Peters, 1992). The dimensionless diameter  $d_{32}/D$  decreases as  $We$  increases. Another way to take turbulence into account is to introduce the turbulent energy dissipation  $\varepsilon'$  and to use  $Vi$  and  $\varepsilon'$  as correlating variables instead of the Weber number. An equivalent relation is obtained, which means that  $We$  depends essentially on turbulence intensity. As a conclusion, static mixers act mainly as turbulence promoters in turbulent flows. Bubble or drop break-up is mainly due to turbulent eddies and the criterion for good bubble and drop size reduction is that the mixer generate a high intensity of turbulence.

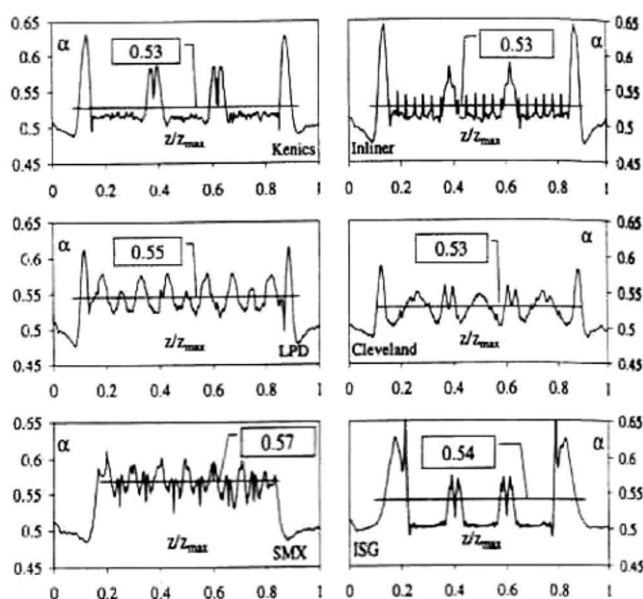


Figure 10. Evolution of surface-averaged extensional efficiency for six commercially available static mixers (from Rauline *et al.*, 1998).

#### Liquid-solid and gas-solid systems

Dispersing a solid in a fluid is a complex operation. We consider only laminar flow systems due to the high viscosities of moderately concentrated solid suspensions. For a dry solid, the first step is solid wetting which depends both on the physico-chemical properties of the mixture and on the initial dispersion of the particles which depends on the method used for injecting the solid phase. The following steps are common for solids already present or formed directly in the fluid phase, for example by a precipitation process. Many powders produced in industry have a characteristic size below 1  $\mu\text{m}$ . Colloidal forces may cause essentially irreversible agglomeration due to linkages by strong forces. The agglomerates can form larger but less stable structures by flocculation due to Van der Waals attractions. Shear forces are able to break flocculates. Rwei *et al.* (1990) investigated the dispersion of carbon black agglomerates in a simple shear flow, while Mifflin and Schowalter (1988) studied the influence of viscoelastic forces on flocculation. See also the recent paper by Furling *et al.* (2000).

A dimensionless number similar to the capillary number [equation (8a) or (8b)] may be used to quantify the relative

influence of viscous forces and of internal attractive forces. The surface tension in equations (8) is replaced by a parameter  $\sigma'$  which is analogous to surface tension but is more difficult to measure. It can be estimated using theoretical models, such as the Rumpf's model which was validated by Lee *et al.* (1993) and Li *et al.* (1997b). However, the applicability of such models depends on the nature of the solid (Kendall, 1988).

There are two mechanisms for size reduction: breakage and erosion. Breakage requires a high shear stress and leads to daughter particles far smaller than the initial floculates. Erosion gradually decreases the particle size, but the daughter particles tend to flocculate. The net effect is a modest reduction in average floculate size. Hydrodynamic forces can also cause flocculation so there is a continuous competition between flocculation, erosion and breakage (Figure 11). The situation is also complicated by the complex rheological behavior of concentrated suspensions, which is only partially understood. Einstein's model ignores particle-to-particle interactions and is limited to dilute suspensions. Many empirical or theoretical models have been proposed (Barnes and Holbrook, 1993), but there is no general way to predict the apparent viscosity of a suspension. It is generally safe to assume that suspensions are Newtonian at low shear rate, exhibit shear-thinning behavior and present a Newtonian plateau at high shear rates. Some of them become shear-thickening at very high shear rates (Boersma *et al.*, 1990). A better understanding of the rheological behavior of suspensions is the objective of recent research, but is complicated by sensitivity to physico-chemical parameters, such as pH, ionic strength, and the presence of surface active agents (So *et al.*, 2001).

The applicability of static mixers for dispersing a solid into a fluid phase has been demonstrated (Isom, 1994), but such applications remain largely empirical. An increase in solid concentration simultaneously increases the viscosity and the probability of collision between particles. There is a simultaneous increase in flocculation rate and in the intensity of the hydrodynamic forces that are responsible for breakage. For static mixers, it seems that the effect of viscosity generally prevails. Dilute suspensions, which have a low apparent viscosity, are difficult to mix using static mixers.

### Heat Transfer

Flow inserts have long been used to enhance heat transfer to fluids flowing inside tubes. These devices function in two

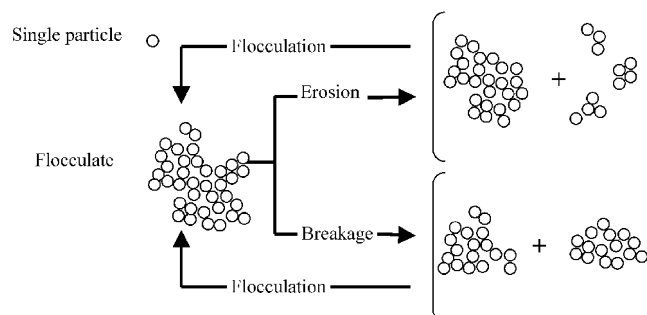


Figure 11. Schematic representation of erosion, breakage and flocculation mechanisms for liquid-solid systems.

ways. They increase the metal to fluid surface area in a manner analogous to the use of externally finned tubes. They also change the hydrodynamics, and the combination of effects can significantly increase heat transfer coefficients, albeit at the cost of higher pressure drops compared to empty tubes. Inserts can be used to increase the intensity of turbulent systems, but the major use of inserts is in systems that, at least in the absence of inserts, would be in laminar flow. Here, we consider primarily the benefit that comes from redistribution of the flow within the tube cross-section and regard the increase in metal-to-fluid contact area as a secondary benefit, although it may be the dominant factor in designs where supplemental mixing is not needed.

The important factors for evaluating a motionless mixer as a heater exchanger are the Nusselt number, **Nu**, and the Fanning friction factor, *f*. Physical properties of the system are incorporated into the Reynolds number, **Re**, the Prandtl number, **Pr**, the viscosity ratio,  $\mu_{\text{wall}}/\mu_{\text{bulk}}$ , and the *L/D* ratio of the tube. The Reynolds number is based on the tube diameter and the superficial velocity. The correlation has the form

$$\text{Nu} = f(\text{Re}, \text{Pr}, L/D, \mu_{\text{wall}}/\mu_{\text{bulk}}) \quad (15)$$

If viscous heating is important, the Brinkman number, **Br**, should be included. The indicated dependence on *L/D* is appropriate for laminar flow in an empty tube, but this dependence may disappear if mixing inserts are distributed continuously down the tube. The friction factor will be a function of the Reynolds number and the correlation of data is usually done for isothermal flow. When cooling, the actual pressure drop will be lower than predicted for the isothermal case because cool, viscous material near the wall is replaced with warmer, less viscous material. The opposite is true for heating.

As a class, the commercial static mixers have marginal benefits for heat transfer enhance. Most of them are designed for general mixing applications, and if the entering fluid is homogeneous, tangential mixing has no benefit for enhancement of heat transfer. Optimal radial mixing is not simple homogenization but is a selective interchange of material between the wall and centerline in a process known as *flow inversion*. Nauman (1979) defined the concept of complete flow inversion and gave a diagram of a device that could achieve it in principle. He also described a practical device called a two-zone, partial flow inverter that can increase the Nusselt number by 30% with minimal increase in pressure drop, the equivalent of about five tube diameters.

### Axial Mixers

Axial mixers are intended to mix fluid that entered the system at different times. The goal is not a sharp distribution of residence times since this would give to plug flow and is approached in conventional static mixers. Instead, the goal is to approximate the broad distribution of residence times in a CSTR, with the purpose of damping input fluctuations. A CSTR acts as an exponential filter with time constant  $\bar{t}$ . The axial mixer will function in the same way if it has the same residence time distribution as a CSTR. Specifically, the cumu-

lative distribution function should have a low value for  $t_{\text{first}}$  and an overall form that is approximately exponential:

$$F(t) = 1 - \exp(-t/\bar{t}) \quad (16)$$

One way of approximating this function is with a transpired wall reactor (Nauman, 2002, p. 111). A closer approximation can be achieved using an array of tubes in parallel (Nauman *et al.*, 2002).

## APPLICATIONS OF STATIC MIXERS

Applications of static mixers in the process industries classified according to the various groups defined in the Introduction: mixing of miscible fluids, multiphase mixing, heat transfer and axial mixing.

### Group 1: Miscible Fluids

This is the most common use of static mixers in industry. Two or more miscible fluids are blended or a reacting mixture is blended to eliminate concentration gradients that would arise if the reaction occurred in an empty tube. Static mixers are useful whenever radial and tangential mixing and a plug flow reaction environment are desired. The static mixers replace or complement conventionally agitated vessels and mechanically driven inline mixers (Reeder, 1998). They have been optimized for laminar flow and can replace single- and twin-screw extruders for some polymer applications. They remain somewhat limited in blending fluids that have substantially different viscosities. In turbulent flows, motionless mixers are generally used for process intensification. That is, they allow the same operations to be performed with a somewhat smaller, in-process inventory. Three important applications are gas mixing in the turbulent or laminar regimes, blending of aqueous solutions in turbulent flow, especially for water treatment, and blending of polymer melts or solutions in the laminar regime. They are also used as reactors, particularly for polymerizations.

#### *Homogenization in laminar flows*

Additives such as plasticizers and internal lubricants, stabilizers, colorants, fillers and flame retardants are commonly blended into polymer melts. The typical combination of a gear pump and motionless mixer replaces an extruder at the end of a polymerization line. Jurkowski and Olkhov (1997) studied how blending of nearly immiscible polymers (polyamide-6 and low-density polyethylene) could be improved using static mixers. Such operations are closely linked to applications of thermal homogenization since the same device will simultaneously homogenize both concentration and temperature. Motionless mixers are also used to process glues (Schneider *et al.*, 1988), and a familiar household application is the use of disposable static mixers to blend two-part epoxy resins.

Applications of static mixers in the food industry are numerous, but they appear less frequently in the literature. Food products are typically highly viscous and non-Newtonian (Holdsworth, 1993) and are usually processed in the laminar regime. Cybulski and Werner (1986) reported that static mixers are used to mix acids, juices, oils,

beverages, milk drinks or sauces in food formulations. Baker (1991) reported an application to melted chocolate.

#### *Gas mixing*

Static mixers are good tools for mixing gases and pre-vaporized liquid fuels prior to a reaction. Indeed, that is the first recorded use of a static mixer (Sutherland, 1874). Despite the high diffusivities of gases, a mixture will not immediately achieve homogeneity and additional mixing may be needed for good combustion. Increasing the residence time after the gases have been metered together will accomplish the necessary mixing, but this will also increase the in-process inventory and could lead to safety problems in the event of a backfire. Additional, active mixing is therefore required. Static mixers are often used for pre-reactor feed blending to improve reaction yields. Baker (1991) discussed their use in nitric acid production. Static mixers placed upstream of the reactor to mix air with ammonia increased nitric acid yield by almost 1%, and they eliminated hot spots that could damage the costly platinum catalyst. Baker reported that the variation in ammonia concentration at the reactor inlet was reduced by a factor 30 and that the catalyst life was extended up to 20%, which allowed reductions in catalyst turnaround frequency and in production costs. Many chemical reactions involving gases can be improved using static mixers, such as those for making vinyl chloride, ethylene dichloride, styrene, xylene and maleic anhydride (Baker, 1991). Static mixers have been reported to have a great potential in reducing NO emission in combustors (Braun *et al.*, 1998). A less conventional applications for a static mixer is found in the nuclear industry to improve sampling and analysis of contaminants in an air flow (McFarland *et al.*, 1999).

#### *Water clarification and sludge treatment*

Turbidity in potable water is caused by suspended, solid particles in low concentrations. Static mixers are used to disperse a flocculating agent, such as alginate, as a first step in clarification. Flows are in the turbulent regime, but excessive shear, as might be caused by mechanical agitation, can damage the flocculates, leading to higher consumption of the flocculating agent. Baker (1991) reported plants in the USA and Canada that use static mixers in this application. Recent applications to ultra-filtration (Derradji *et al.*, 2000), ultraflocculation and turbulent microflotation (Rulyov, 1999) are also reported. An important operation in water treatment is disinfection. This operation requires both mixing and interface generation as the disinfectants, usually chlorine or ozone, are introduced as gases. Although chlorine dissolution is an easy task, efficient use of ozone requires a preliminary dissolution in water before mixing the ozone solution into the main stream (Baker, 1991; Clancy *et al.*, 1996). Static mixers can also play a positive role in this case. Another application of static mixers in water treatment is dechlorination. Plant effluents may require a dechlorination step prior to discharge to avoid forming carcinogenic trihalomethane compounds (Baker, 1991).

Another aspect of water treatment is sludge conditioning in wastewater treatment. The objective is to dewater the

sludge, and the dewatering process begins with the addition of polymers, alum or ferric chloride as an aid to coagulation. Like the water clarification process, shear must be minimized to avoid breaking the coagulated solids. Static mixers are able to reduce additive requirements (Baker, 1991). Procelli *et al.* (1993) report a decrease of 10% in ferric chloride consumption and a slight increase in solid content using static mixers. They estimated the savings at approximately \$35,000 from December 1989 to November 1990 in the wastewater treatment plant serving the city of Ann Arbor, Michigan, USA. The process requires good pH control because the active agent acts only in a limited range of pH. Static mixers are able to produce homogenized flows with a short residence time. They are therefore reported to be useful tools for an adaptive control of chemical addition in water treatment plants (Lee and Choi, 2000; Galey *et al.*, 2000), but their applicability to improve control is not limited to water treatment.

#### Mixing with reaction

Applications of static mixers to polymerization reactions have long been suggested (Grace, 1971), but relatively few appear to have been implemented on an industrial scale. Sultzer designed a polystyrene process that makes extensive use of motionless mixers, particularly of the SMR type, and a commercial-scale plant has been built in Japan. Tein *et al.* (1985) report some details of this process. Static mixers are also used in post-reactors and in devolatilization preheaters (in which reaction occurs) in other polystyrene processes. An academic study on styrene polymerization in a static mixer reactor was reported by Yoon and Choi (1996). Fleury *et al.* (1992) studied the polymerization of methyl methacrylate while Schott *et al.* (1975), Khac Tien *et al.* (1990), Baker (1991) and Myers *et al.* (1997) describe the use of motionless mixers for making polystyrene, nylon, urethane and sulfonated compounds. A discussion of static and dynamic mixing in polystyrene processing is given by Myers *et al.* (1997).

Most of the suggested applications are to highly exothermic polymerizations. However, the greatest number of industrial installations is for the reaction injection molding (RIM) of polyurethanes where the reaction exotherm is moderate. Commercial RIM machines use an impingement mixer followed by a static mixer to quickly blending the reactive components (Kolodziej *et al.*, 1982). Mixing performance was characterized by the striation thickness distribution. An academic study by Hoefsloot *et al.* (2001) treated polypropylene degradation in a static mixer reactor.

Other types of chemical reactions can benefit from the use of static mixers. A reactive extrusion process for glycol glucoside synthesis can be improved using static mixers (Subramanian and Hann, 1996). An application to the lactase treatment of whole whey has been reported (Fauquex *et al.*, 1984; Metzdorf *et al.*, 1985). Lammers and Beenackers (1994) investigated the use of a static mixer reactor for the production of starch ethers such as hydroxypropyl starch for the food and pulp and paper industry. Grafelman and Meagher (1995) reported liquefaction of starch using a single-screw extruder and a post-extrusion static mixer reactor. Cultivation of attenuated hepatitis A virus antigen in a

titanium static mixer is described by Junker *et al.* (1994). Reaction applications suggested for static mixers include cracking of heavy and crude oils (Jurkias, 1998) or for the controlled hydrolysis of whey proteins by trypsin (Margot *et al.*, 1998). Koch-Sulzer has developed packed reactors using inline mixers containing a catalyst and even using static mixers made of a catalytic material.

Motionless mixers are not restricted to continuous flow systems. Figure 12 illustrates how they can be used for continuous flow, fed-batch and batch reactions. One semi-batch process for phenol alkylation is similar to that shown in Figure 12(c) but has eliminated the agitator in the reaction vessel and uses a proprietary nozzle—a type of motionless mixer—to mix fresh alkene into the recycle stream (Nauman, 2002, p. 389). This is an example of a system that is miscible at equilibrium but is two-phase at the point of introduction.

#### Group 2: Immiscible Systems

This group includes processes for dispersing one phase in another or for increasing the mass transfer coefficient between phases. Applications include liquid–liquid, gas–liquid and solid–fluid systems.

##### Liquid–liquid systems

Static mixers are well suited for co-current extraction processes. They are competitive in this application with mechanically agitated systems such as rotating disc columns or stirred tanks in series. A major advantage is their resistance to flooding, even when the phases have similar densities. The aim is to form drops that are small enough to provide high interfacial area but large enough to avoid formation of an emulsion, and static mixers are well suited to this purpose. Baker (1991) reported industrial applications for amine washing, caustic washing, water washing of organics and extraction of hydrogen sulfide from petroleum fractions using diethanolamine. Recently, static mixers have been used for co-current extractions with supercritical carbon dioxide, for example to carry out fractionation of lipids in order to separate squalene from triglycerides and diacylglycerylethers (Catchpole *et al.*, 2000). Co-current extraction of caffeine from supercritical CO<sub>2</sub> with water using static mixers has been proposed to replace counter-current packed columns (Pietsch and Egers, 2000). Motionless mixers are also used to enhance liquid–liquid reactions. Examples are largely proprietary, but see the contribution of Chamayou *et al.* (1996) on the production of Amiodarone, a widely used anti-arrhythmic drug.

Static mixers can also be used in the classical, counter-current mode for liquid–liquid extractions. Jancic *et al.* (1983) and Streiff and Jancic (1984) studied their application to several test systems: kerosene–water, butanol–succinic acid–water, toluene–acetone–water and carbon tetrachloride–propionic acid–water. They concluded that inserting static mixers reduces coalescence and requires a lower residence time than conventional devices, even with high liquid flow rates. Other applications of static mixers with liquid–liquid systems are reported by Merchuk *et al.* (1980) for copper extraction, and by Le Coze *et al.* (1995) for indium extraction.

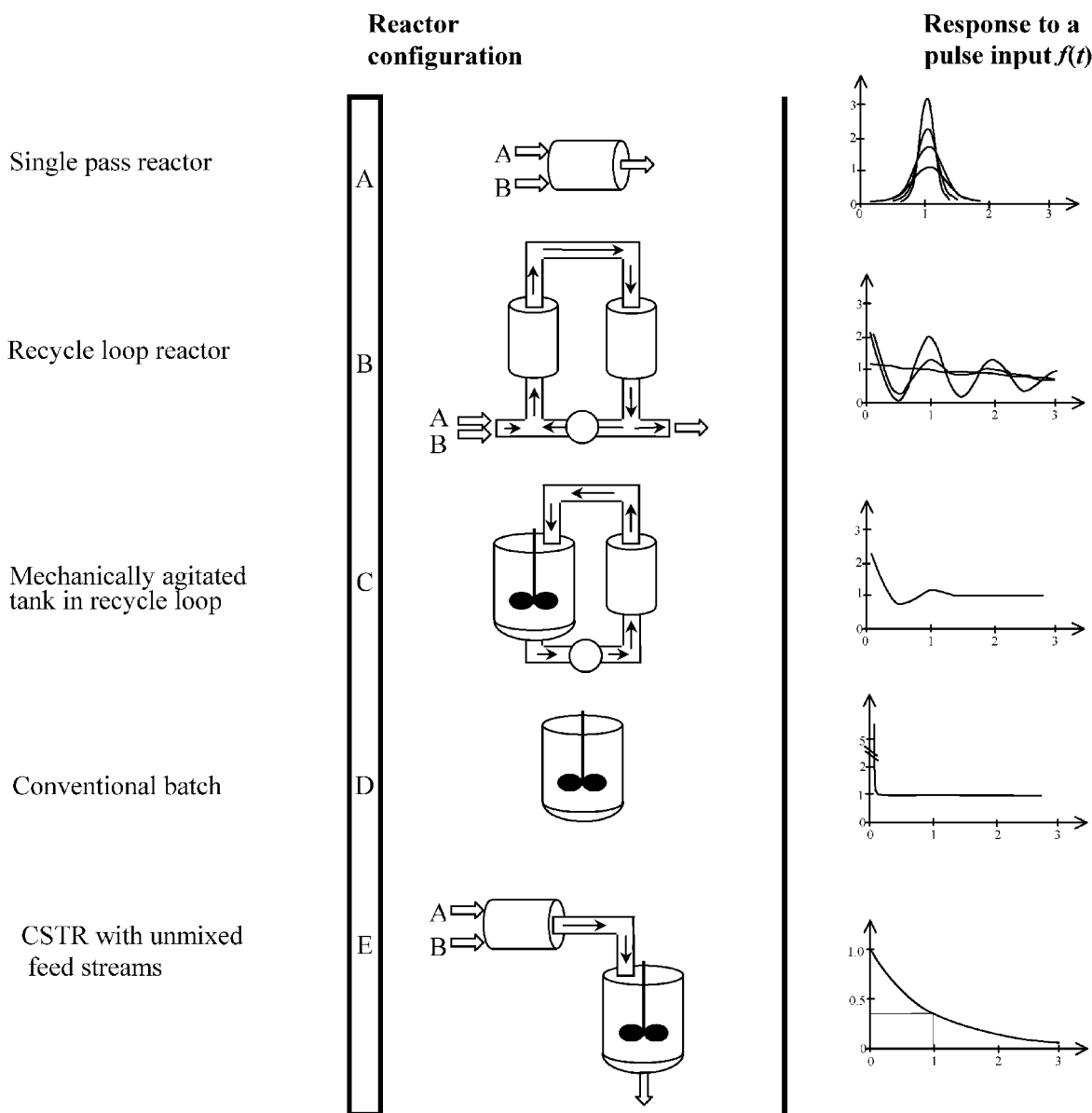


Figure 12. Reactor configurations using static mixers.

Static mixers have potential applications for continuous emulsification process. Emulsions are of utmost importance in many food products (Cybulski and Werner, 1986), cosmetics and pharmaceuticals. Static mixers can produce a primary emulsion and homogenize the emulsifier concentration simultaneously. A typical application is microencapsulation (Powell and Mahalingam, 1992; Maa and Hsu, 1997), but published information is limited in these areas. Motionless mixers are known to have a complex behavior with liquid-liquid systems (Merchuk *et al.*, 1980). It is sometimes difficult to predict whether a water-in-oil or an oil-in-water emulsion will be produced. Static mixers can exhibit multiple steady states in this application, and both types of dispersion may exist alternatively. Motionless mixers are also known to promote phase inversion. Flow-induced phase inversion during emulsification has been reported by Tidhar *et al.* (1986) for water-kerosene and water-carbon tetrachloride

systems and by Akay (1998) for concentrated emulsions of epoxy resins.

A recent application of static mixers to liquid-liquid systems is the coating of very fine particles. In one process, supercritical carbon dioxide is used as a carrier for powder coatings. A high degree of mixing is required to produce coated particles using a high-pressure spray process (Wagner and Eggers, 1996; Weidner, 1999; Weidner *et al.*, 2001). In the Unicarb spray process, the supercritical fluid is mixed with polymer or paint solutions using static mixers (Lee *et al.*, 1990).

At least two commercial processes use motionless mixers to disperse liquid water (under pressure) into molten polystyrene as an aid to flash devolatilization.

#### Gas-liquid systems

Static mixers can be readily adapted to absorption and scrubbing. They are particularly useful for co-current

absorption of highly soluble gases like carbon dioxide, ammonia and chlorine where only a few transfer stages are needed. Inline static mixers with either bubble flows or spray flows are used. They are also used in systems involving absorption followed by chemical reaction, especially when the absorbed gas promptly reacts. Capital cost is lower than counter-current towers especially in high-pressure applications. Static mixers can also enhance multistage counter-current towers, increasing performance and throughput. Rader *et al.* (1989) reported many applications of static mixers to gas–liquid systems, both as co-current and in counter-current devices. Distillation and other column operations now use structured packings instead of trays or random packings. These packings are a form of static mixer with the objective of creating a large interface for mass transfer.

Natural gas processing plants make extensive use of static mixers. They are used to:

- (i) scrub hydrogen sulfide from natural gas using sodium hydroxide solutions, amines or proprietary solvents;
- (ii) scrub carbon dioxide using amines or proprietary solvents;
- (iii) selectively remove hydrogen sulfide in the presence of carbon dioxide;
- (iv) dehydrate gases with glycols.

Some applications of static mixers in chemical and petrochemical industry are:

- (i) purifying the effluent from an oxychlorination reactor with a sodium hydroxide solution in ethylene dichloride production;
- (ii) scrubbing ammonia, hydrogen chloride, hydrogen fluoride or cyanides with water;
- (iii) scrubbing chlorine gas and acid gases with sodium chloride solutions or solvents;
- (iv) scrubbing noxious organic compounds with various solvents;
- (v) pre-quenching for acrylonitrile absorbers.

Static mixers are reported to be more efficient than other contacting devices, such as spray nozzles, venturi scrubbers and randomly packed columns.

Static mixers as co-current devices are used in water sterilization (Baker, 1991; Zhu, 1991; Martin and Galey, 1994; Clancy *et al.*, 1996; De Traversay *et al.*, 2001). Water treatment with ozone has been reviewed by La Pauloue and Langlais (1999). The oxidative absorption of hydrogen sulfide by a solution of the ferric chelate of nitrilotriacetic acid in a co-current column packed with static mixers is described by Demmink *et al.* (1994). Another gas–liquid example is the continuous hydrogenation of vegetable oils (Rusnac *et al.*, 1992).

Static mixers are known to increase mass transfer in bubble columns. Inserts can be added in the riser section of airlift reactors (Chisti *et al.*, 1990; Goto and Gaspillo, 1992; Gavrilescu and Roman, 1995, 1996; Gavrilescu *et al.*, 1997), in the draft tube of bubble columns (Goto and Gaspillo, 1992), in the draft tube of a bubble slurry column (Gaspillo and Goto, 1991), in a mechanically stirred

airlift loop reactor (Lu *et al.*, 2000), directly in packed bubble columns (Fan *et al.*, 1975; Wang and Fan, 1977) and in three-phase fluidized beds (Potthoff and Bohnet, 1993). Industrial applications include the cultivation of a filamentous mold to produce cephalosporin C (Gavrilescu and Roman, 1995) and for ethanol production in an airlift reactor (Vicente *et al.*, 1999).

Structured packings for distillation and similar column operations are manufactured by many companies including some manufacturers of conventional static mixers. Applications range from laboratory distillation columns to the immense columns used to separate styrene and ethylbenzene at production rates in excess of 700,000 tons/year. Structured packings have largely replaced trays and random packings for new column designs. Most design information is proprietary, but a study of Fitz *et al.* (1999) has shown that such packings can improve capacity and performance efficiency. Good column effectiveness can be maintained even at extremely low liquid flow rates; that is, they reduce dewetting and thus provide good turn-down. Test systems included *p*-xylene/*o*-xylene, cyclohexane/*n*-heptane and *i*-butane/*n*-butane from 0.02 to 27.6 bars. However, distillation efficiency was found to deteriorate at pressures above 10 bars.

#### *Solid–fluid systems*

One use of static mixers for solid–liquid systems is the dispersion of a particulate solid into a liquid, sometimes including breaking of aggregates that are bound van der Waals interactions. Applications are found in the chemical industry, for example to disperse a catalyst in a fluid phase, and in pulp and paper processes (Isom, 1984). Jean *et al.* (1987) investigated the continuous production of narrow-sized titanium dioxide particles using static mixers. Barresi *et al.* (1997) used a motionless mixer for the wet mixing of ceramic powders. Similar applications are believed to exist in the food industry, but publications are scarce in this field.

Non-uniform contacting between a reactive gas and solid catalyst particles is a notorious problem in fluidized beds. A variety of inserts have been used to mitigate this problem. Krambeck *et al.* (1987) and Pustelnik and Nauman (1991) described and analysed the use of horizontal baffles to improve contacting in a large, cold-flow model of a methanol-to-gasoline reactor. Metzendorf *et al.* (1991) suggested the use of motionless mixers to reduce axial dispersion of the liquid phase in a liquid–solid fluidized bed reactor.

#### *Solid–solid systems*

Another operation that benefits from the use of static mixers is blending of solids including mixing of dry pigments and ink powders, blending detergent additives, mixing lubricants into powdered metals or polymer pellets, blending dry clays and cements or dry clays with a catalyst (Baker, 1991). When the mixers are fed by gravity, they would function in a vacuum and thus deserve a separate category from the solid–gas mixers in the previous section. One common application is the blending of fine powders prior to a subsequent fabrication stem, but there are few formal publications in this area.

### Group 3: Heat Transfer

Applications of static mixers to improve heat transfer can be divided into three types. The first is thermal homogenization, often coupled with compositional homogenization. The second is pure heat transfer in heat exchangers. The third type combines heat transfer with chemical reaction.

#### *Thermal homogenization*

In undisturbed laminar flow in an empty pipe, thermal diffusion is the only mechanism for heat transfer in the radial direction. A great variety of motionless inserts have been used to promote radial flow and thus to reduce radial temperature gradients in process fluids. One common application is the installations of a static mixer immediately downstream from a screw extruder in order to obtain a thermally homogeneous polymer melt. Myers *et al.* (1997) suggested using elements of the open type for this application and reported that they reduced radial temperature variations by a factor of 10. A typical application of thermal homogenization is for film blowing or sheet extrusion because thermoplastics require a uniform temperatures to eliminate position-dependent variations in the extrudate (Chen, 1975). Schott *et al.* (1975) suggested using motionless mixers for polyethylene, polypropylene, polystyrene and ABS resins processing. Such applications of motionless mixers to extrusion have become standard practice. Their main purpose is thermal homogenization but they will also alleviate composition differences resulting from polymer blending and coloring. The need is primarily for radial mixing, but the complex flow patterns in extruders gives extrudates that lack tangential symmetry. Thus some tangential mixing is needed as well and is automatically provided by most motionless mixers.

Myers *et al.* (1997) reported applications in the turbulent regime to enhance thermal homogenization.

#### *Heat exchangers without reaction*

For reasons discussed in the next section, a single-tube heat exchanger is often used for polymerizations. Conventional, multitube heat exchangers are used in the absence of a reaction or when there is a reaction that causes little change in viscosity. Static mixing elements can be used in either laminar or turbulent flow to improve heat transfer coefficients, but enthusiastic equipment vendors sometimes oversell this application. Mixing inserts have several significant disadvantages compared to empty tubes: higher pressure drop, greater potential for fouling, relative difficulty of cleaning and greater cost. Note that thermal diffusivities are several orders of magnitude higher than molecular diffusivities, so that most heat transfer operations are feasible using tubes of reasonable diameter and length and with reasonable residence times. The use of mixing inserts is justified when there is a strong need to minimize in-process inventory. Examples include the need to suppress detrimental reactions, when the process material is particularly dangerous, or when it is particularly expensive. The first of these reasons is common and is discussed in the next section.

Mixing elements are most beneficial in deep laminar flow (Ishikawa and Kamiya, 1994; Joshi *et al.*, 1995), and most applications have been to this area. Literature descriptions of their use are largely for the reactive flows covered in the next

section. Static mixers have a distinct advantage over empty tubes for such applications because they provide a more uniform distribution of residence times. If interest is limited to pure heat transfer, inserts specifically designed for this purpose seem better than the general purpose devices that are the main topic of this review. The greatest interest has been in twisted tape and offset strip fins (Bergles, 1995). These devices can be used for turbulent flow and boiling heat transfer as well as for laminar flow.

#### *Heat transfer with reaction*

Single- and multitube heat exchanges are widely used as reactors. Motionless inserts are doubly beneficial in this application because they simultaneously improve heat transfer and narrow the residence time distribution. It is primarily the second advantage that justifies their use. However, published examples of industrial reactors using motionless mixers are quite rare. The major exceptions are for styrene polymerization and reaction injection molding of polyurethanes. Urethane polymerizations are two component reactions with low heats of reaction. The role of the static mixer is to intimately blend the reactive components rather than to remove heat.

Vinyl polymerizations such as that for polystyrene have large heats of reaction and large changes in viscosity. These lead to two forms of instability in tubular reactors, the classic thermal runaway and a hydrodynamic instability analogous to the viscous fingering problem in secondary oil recovery (Nauman, 2002, p. 496). Both problems can be (largely) avoided by reacting to low conversion (ca. 15%) in a long, single tube. This is the approach used in the high pressure process for polyethylene. These reactors are literally kilometers long, and the employment of mixing inserts would appear useful, although it is not clear that this is actually done. Multitube reactors fed with a low viscosity monomer are hydrodynamically unstable when operated in the once-through mode because the feed cannot readily displace viscous polymer. Multitube reactors are used in recycle loops where the per-pass change in viscosity is small. Recycle reactors using motionless mixers for nonstyrenic polymerizations are known to be in commercial operation, but details have not been published. The Sultzer process for polystyrene uses several reactors in series, the first of which is a recycle loop containing SMR elements (Tein *et al.*, 1985). On a once-through basis with monomer feed, the SMR would be susceptible to the viscous fingering, but the SMR can be used after the recycle loop because the partially converted (ca. 60%) reactant mixture is sufficiently viscous to avoid hydrodynamic instabilities. Other polystyrene processes use one or more autorefrigerated (boiling) stirred tank reactors or stirred tube reactors in series to obtain conversions of about 70%. At this point, an ordinary shell-and-tube reactor can be used for further conversion and to preheat the reaction mass for a flash devolatilization step. A single tube, operating approximately adiabatically with or without static mixers (Craig, 1987) can be used as a post reactor in some processes, but the final preheating prior to devolatilization is done in a multitubular reactor. For some product grades, it is necessary to minimize conversion in this heat exchanger. Motionless mixers are used in the Sulzer process and have been evaluated on the pilot scale for other processes. The Sulzer process also uses motionless

mixers to reheat the polymer prior to a second flash and to mix in water as a foaming aid.

Outside the area of polymerization, Lammers and Beenackers (1994) suggested using a continuous tubular reactor containing static mixers to produce starch ethers for food and pulp. Static mixers can be used in turbulent flow reactors, for example in the catalyst tubes of a reformer furnace. Static inserts are said to increase heat transfer coefficients, eliminate channeling within the catalyst bed, avoid cocking, prevent catalyst deterioration due to hotspots and improve yield. Applications to condensation (Fan *et al.*, 1978) and boiling heat transfer (Azer and Lin, 1980) have been reported. Another example is provided by Gough and Rogers (1987), who discussed the treatment of coal tar oil residues using static mixer heat exchangers. The residues contain heat sensitive phenolic compounds that can readily polymerize. Similar residues result from distillate bottoms from naphta cracking and can eventually include carbon solids. These fluids are preheated before being burned in a reducing atmosphere to produce carbon black. Gough and Rogers (1987) showed that static mixers could be used to improve pre-heater performance.

Motionless mixers are a potential solution to many problems encountered in heat exchanger operation. In cooling processes, skinning due to boundary-layer solidification may be alleviated due to the better radial mixing (Baker, 1991). Fouling in reactive systems is caused by long residence times at the wall and high temperature differences between the wall and the bulk fluid. Crystallization, polymerization or biological growth may occur, and the results films have low thermal conductivity and cause significant resistance to heat transfer. Gough and Rogers (1987) have shown that static mixers can reduce fouling, coking and enhance heat transfer in oil tar residue treatment.

### Axial Mixers

Industrial applications have not yet been reported for this new class of motionless mixer.

### KEY PARAMETERS FOR STATIC MIXER SELECTION

This section defines and quantifies the key parameters for proper selection of a static mixer. The pertinent parameters depend on the application, but one parameter is common to all applications of static mixers. Process intensification using mixing inserts always gives a higher pressure drop than an open pipe of the same diameter.

#### A General Parameter: Pressure Drop

Pressure drop estimation is the first step for proper selection of a static mixer. Correlations are available for isoviscometric flows in commercial units, but pilot plant measurements may be necessary when dealing with systems of unknown or complex rheology, reacting systems and multiphase systems. Measuring pressure drop is a relatively easy task using the sensitive and low-cost piezoresistive or piezocapacitive sensors that are now available. Signal processing techniques can even be used to gain insight into flow characteristics through the analysis of pressure fluctuations (Vial *et al.*, 2000).

The basic equation appropriate to flow of a homogeneous, isothermal, incompressible, Newtonian fluid in a circular tube is

$$\Delta P = \frac{2f\rho u^2}{D}L = \frac{2f\rho u^2}{D}NL_e \quad (17)$$

where  $N$  is the number of mixing elements,  $L_e$  is the length of one element, and  $f$  is a function of Reynolds number that is determined experimentally for a particular mixer or by CFD. Note that  $u$  is the superficial velocity and that  $Re$  is calculated for the empty pipe. Equation (17) applies to an empty pipe with  $f = Re/16$  for laminar flow and to turbulent flow with  $f = 0.079Re^{-0.25}$ . For a given  $Re$ , the friction factor will be higher for a tube with inserts than for an empty tube, and the transition from laminar to turbulent flow will occur at a much lower Reynolds number than the classic value of 2100. For static mixers installed in non-circular ducts, the functional form of equation (17) is still used, but the tube diameter,  $D$ , is replaced with another characteristic dimension. The hydraulic mean diameter is appropriate for turbulent flows.

Vendor correlations for pressure drops typically report the ratio, denoted here as  $Z$ , of friction factors for the mixer and the empty tube:

$$Z = \frac{f_{\text{mixer}}}{f_{\text{empty}}} = \frac{\Delta P_{\text{mixer}}}{\Delta P_{\text{empty}}} \quad (18)$$

This formulation is convenient for retrofitting static elements into existing tubes since it gives the ratio of pressure drops directly. However, for an initial design, the optimal diameter for a tube with inserts will generally be different than the optimal design for an empty tube. When density or viscosity varies appreciably as a function of axial position, equation (17) can be applied to one element at a time, using values for  $f$ ,  $\rho$  and  $u$  appropriate to each element. The equivalent calculation for the empty tube replaces  $\Delta P/L$  with  $dP/dz$  and then integrates down the tube.

Pandit and Joshi (1998) reviewed pressure drop correlations for Newtonian fluids in fluidized beds, packed beds, and static mixers. For non-Newtonian and especially viscoelastic fluids or for multiphase flows, experimentally based correlations are seldom available for static mixers, but reasonable approximations are sometimes possible using the same type of correction techniques that are used for complex flows in empty tubes.

#### Pressure drop in laminar flow

As a rule of thumb, fluids having viscosities higher than 0.1 Pa s will be in laminar flow under conditions typical of the process industries and fluids exhibiting pronounced non-Newtonian or viscoelastic behavior will almost always be in laminar flow. For flows in empty tubes, behavior for  $Re < 100$  is generally laminar with negligible contribution from the momentum terms in the equations of motion. For  $Re < 2100$ , the flow is usually laminar but small disturbances can lead to wake shedding and other oscillatory behavior. For  $Re > 2100$ , the flow is unstable. An assumption of turbulence is conservative for pressure drop calculations since pressuredrops in turbulent flow are higher than those for laminar flow. An assumption of turbulence is non-conservative for  $2100 < Re < 5000$  when mixing or heat transfer are involved.

The same general concepts apply to flows in static mixers of the open type except that the transition values for **Re** are lower by a factor of about 2. Flows are generally laminar for **Re** < 50 and turbulent for **Re** > 1000. The inserts cause systematic disturbances to the flow field so that complex but fairly reproducible flow behavior can be expected in the intermediate range 50 < **Re** < 1000. More exact numbers depend on the design of the elements including their aspect ratio,  $L_e/D_e$ . For helical, Kenics KM elements, the intermediate region begins when **Re** is about 43 when  $L_e/D_e$  is 0.8, but is delayed to **Re** ≈ 55 when  $L_e/D_e$  is 1 (Jaffer and Wood, 1998). The influence of aspect ratio has been confirmed by Joshi *et al.* (1995). They also concluded that low aspect ratios were better for heat transfer. However, most experimental data in the literature are for a typical commercial design that has an aspect ratio of 1.5 (Rauline *et al.*, 1998). For Sulzer SMX elements, Li *et al.* (1997a) reported that the laminar regime prevails up to **Re** = 15, while the turbulent regime begins when **Re** = 1000.

The simplest method for predicting pressure drop is to use the factor *Z* in equation (18). Some vendors provide data in this form, and *Z* values are available in the reviewed literature for several commercial designs (Pahl and Muschelknautz, 1982; Cybulski and Werner, 1986; Rauline *et al.*, 1998). Values are widely available for the open, helical inserts sold by Kenics (Bor, 1971; Grace, 1971; Alloca, 1982; Pahl and Muschelknautz, 1982; Cybulski and Werner, 1986; Joshi *et al.*, 1995), and *Z* ranges from 5 to 8 in the laminar regime. To a first approximation, *Z* is constant in the laminar regime. Sir and Lecjaks (1982) and Cybulski and Werner (1986) found it to be a weak function of **Re** in the laminar regime:

$$Z = a' + b' \cdot \text{Re}^{m'} \quad (19)$$

The exponent  $m'$  is about 1 when **Re** < 0.1 and tends to 0.5 as **Re** > 100. Joshi *et al.* (1995) used  $m' = 0.5$  for  $L_e/D_e$  ratios between 1.5 and 2.5.

The multilayered SMX design from Koch-Sulzer has been also the subject of literature studies, but reported *Z* values are not accurate enough for proper estimation: Pahl and Muschelknautz (1982) reported *Z* values between 10 and 60, while Cybulski and Werner (1986) reported values from 10 to 100. The corrugated-plate SMV elements give *Z* values between 60 and 300 (Cybulski and Werner, 1986).

Other static mixers have been less studied but similar conclusions are reached. Open designs give low *Z* values, e.g.  $Z \approx 6$  for the LPD mixers (Heywood *et al.*, 1984) and  $Z \approx 9$  for Lightning mixers (Pahl and Muschelknautz 1982). The corrugated-plate (SMV) and multilayer (SMX, SMXL) designs cause much greater pressure losses that are difficult to quantify accurately. Closed designs with holes or channels cause the greatest pressure drops. *Z* is 284 for ISG elements according to Heywood *et al.* (1984), while Pahl and Muschelknautz (1982) reported *Z* values between 250 and 300 for this mixer design. As a conclusion, the parameter *Z* is simple and useful, as it indicates directly the additional pressure drop compared to an empty pipe. In theory, *Z* should be a function only of the Reynolds number and the mixer design, and should be directly scaleable with respect to tube diameter and number of elements provided the small and large mixers maintain geometric similarity. If there are departures from geometric similarity, additional

correlating variables should be introduced. These include the aspect ratio and the void fraction. Small Kenic-type mixers, and particularly those molded from plastic, have smaller void fractions than large mixers fabricated from metal. A possible way to account for this is to use a modified Reynolds number based on the average velocity,  $u/\varepsilon$ , where  $\varepsilon$  is the void fraction rather than the superficial velocity, *u*.

Another way of representing pressure drop data is to directly correlate the friction factor, *f*, rather than the ratio of friction factors, *Z*. The results are equivalent and subject to the same limitations, but a direct correlation for *f* can avoid confusion when comparing mixers to empty tubes of different diameter. In laminar flows, the classic relationship between *f* and **Re** is usually obtained:

$$f = \frac{C_1}{\text{Re}} \quad (20)$$

where  $C_1$  is a constant greater than 16. Specifically,  $C_1 = 16Z$ . This relation is sometimes written in a different way:

$$\text{Ne} \cdot \text{Re} = K_p \quad (21)$$

where  $K_p$  is the power constant and **Ne** is the Newton number, defined as twice the friction factor.

$$\text{Ne} = 2f = \frac{\Delta P}{\rho u^2} \cdot \frac{D}{L} \quad (22)$$

Table 7 gives  $K_p$  values for the SMX, Inliner, LPD, KM and ISG static mixers. One vendor prefers this method for correlating pressure drop data on the grounds that it allows comparison between static mixers and agitated tanks. However, the direct way of comparing powers is by the ratio of average power inputs per mass of flowing fluid. Assuming the comparison is made at the same flow rate and density, this is just the ratio of the pumping power for the mixer,  $Q\Delta P$ , to the power consumption of the impeller,  $P_{\text{impeller}}$ .

$$\frac{\text{Power for static mixing}}{\text{Power for mechanical agitation}} = \frac{Q\Delta P}{P_{\text{impeller}}} \quad (23)$$

where *Q* is the volumetric flow rate. Presumably, this ratio will be less than one at equivalent performance.

Published correlations for *f* are summarized in Table 8. Some of them take the aspect ratio into account (e.g. Lecjaks *et al.*, 1987). Shah and Kale (1991, 1992) use a more accurate form for correlating pressure drop data for single-phase laminar flows:

$$f = \frac{C_1}{\text{Re}} + \frac{C_2}{\text{Re}^m} \quad (24)$$

Table 7. Comparison of  $K_p$  from vendor data, published experimental data and CFD studies.

| $K_p$     | Vendor | Literature | CFD  |
|-----------|--------|------------|------|
| Kenics    | 170    | 195        | 255  |
| Cleveland | 190    | —          | 190  |
| LPD       | 195    | 220        | 225  |
| Inliner   | 240    | 270        | 300  |
| SMX       | 1200   | 1140       | 1120 |
| ISG       | 7210   | 8140       | 8460 |

The second term is intended to reflect the effects of radial flows caused by the mixing elements. The authors provide empirical results for  $C_1$ ,  $C_2$  and  $m$  for Kenics and SMX static mixers, both for Newtonian and power-law fluids. Sir and Lecjaks (1982) found a similar relation for the Kenics static mixer, but exponent  $m$  was equal to 0. However, most predictive correlations are based on the simpler form of equation (20). Correlations of  $f$  vs  $Re$  for Kenics, SMX and SMV mixers are shown in Figure 13, and results for several other commercial static mixers are presented in Figure 14.

Static mixers are widely used with complex fluids in the polymer and food industries, but prediction of head loss with non-Newtonian fluids has been the subject of only a few studies. Shah and Kale (1991, 1992) correlated their data using equation (24) and the Metzner–Reed generalization of Reynolds number:

$$Re_g = \frac{\rho u^{2-n} D^n}{8^{n-1} k' \cdot \varepsilon^{2-n}}$$

(25)

The parameter  $k'$  is the consistency index and  $n$  is the flow index. More recently, Li *et al.* (1997a) suggested another definition of  $Re_g$  which is more general, as it not limited to power-law fluids:

$$Re_g = \frac{\rho u D}{\mu_c^* \cdot \varepsilon}$$

(26)

Here,  $\mu_c^*$  is the apparent viscosity at the wall corresponding to the shear rate at the wall. Example calculations are given by Li *et al.* (1997a), who based their correlation on equation (17) and thus used only one adjustable constant.

Limited data are available in the literature for viscoelastic fluids (Shah and Kale, 1991, 1992; Chandra and Kale, 1992, 1995). Shah and Kale (1991, 1992) compared viscoelastic solutions of polyacrylamide to inelastic solutions of carboxymethylcellulose and concluded that elasticity always increases the friction factor. This is expected since elasticity is important in entrance flows, and the sequential elements in a static mixer create a sequence of entrance flows. The Fanning friction factor for a viscoelastic fluid can be deduced from that for an inelastic fluid with the same viscous behavior by introducing the Weissenberg number. This approach is limited to shear-thinning, power-law fluid and requires measurements of the primary normal stress difference as a function of shear rate. See Shah and Kale (1991, 1992) for details.

Pressure drop in turbulent flow

Owing to the limited use of static mixers in turbulent flow, fewer correlations of pressure drop are available. Pahl and Muschelknautz (1982) and Cybulski and Werner (1986) give correlations for the friction factor for two ranges of Reynolds number,  $1200 < Re < 7000$  and  $7000 < Re < 30,000$ . A correlation used for turbulent flow is:

$$f = \frac{C_t}{Re^q}$$

(27)

where  $C_t$  is a constant. The exponent  $q$  itself is a function of Reynolds number, typically decreasing as higher values of  $Re$ . Cybulski and Werner (1986) give results for the Kenics, LPD and Komax mixers. At high Reynolds numbers,  $q$  approaches 0 and  $f$  becomes constant. A similar behavior is observed in empty pipes with  $f_{empty} \rightarrow 0.02$  as  $Re \rightarrow \infty$ .

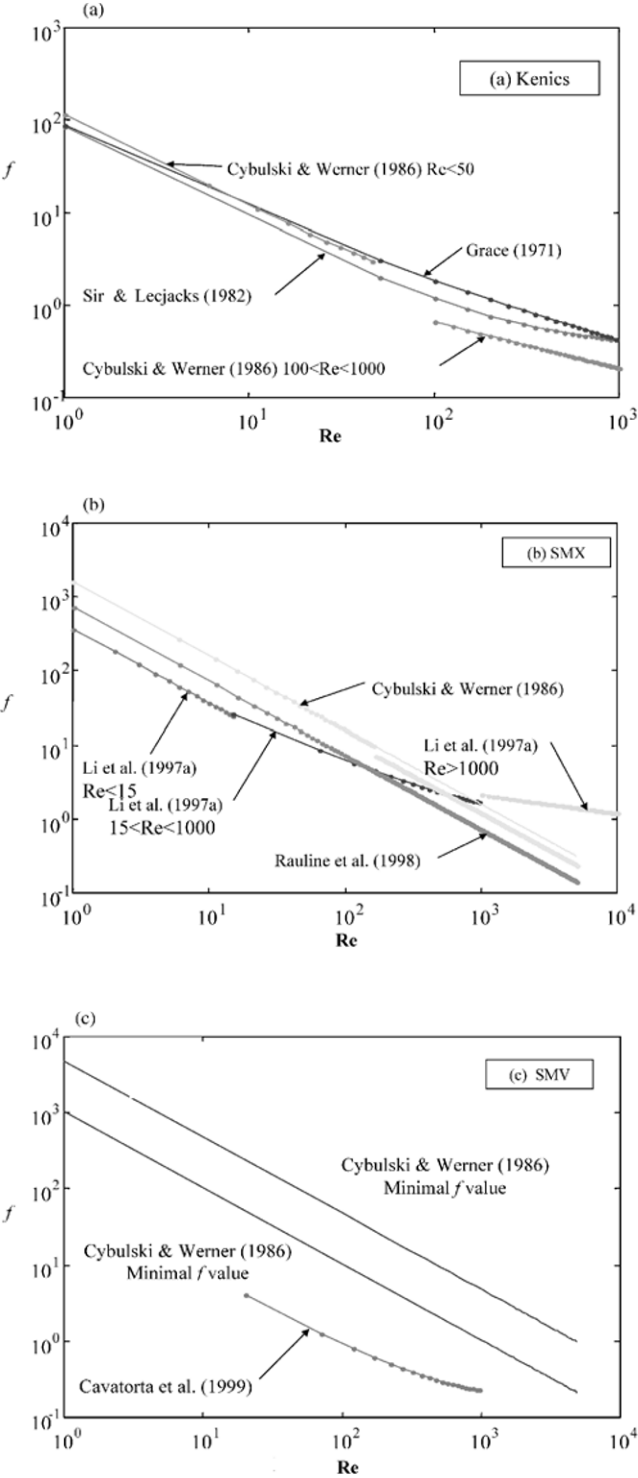


Figure 13. Friction factor vs Reynolds number,  $Re$ . (a) Kenics; (b) SMX; (c) SMV mixers.

Limiting  $f$  values for Kenics, Hi-Toray, SMX and SMV mixers are respectively 3, 11, 12 and 6–12 (Pahl and Muschelknautz, 1982). The ratio  $Z = f/f_{empty}$  is also the ratio of pressure drops, so that static mixers in high turbulence consume several hundred times as much pumping energy as flows in an empty tube.

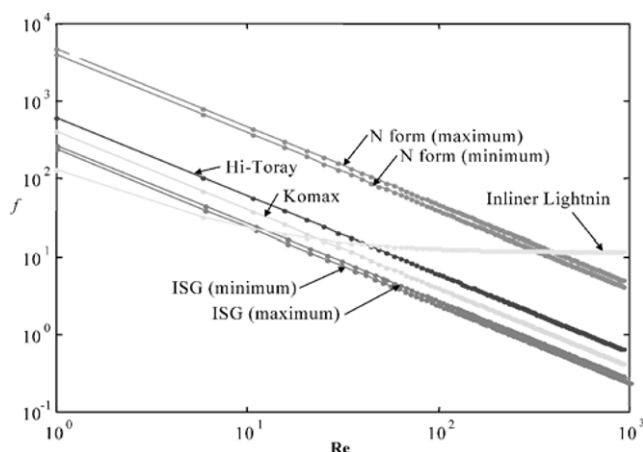


Figure 14. Comparison of friction factors for commercial mixers.

### Pressure drop in gas–liquid flows

The method proposed by Lockhart and Martinelli (1949) to estimate pressure drop in gas–liquid flows,  $\Delta P_{G/L}$ , provides the starting point for most work on multiphase flows in static mixers.  $\Delta P_{G/L}$  is estimated using single-phase pressure drop calculations for the gas phase alone,  $\Delta P_G$ , and for the liquid phase alone,  $\Delta P_L$ :

$$\frac{\Delta P_{G/L}}{L} = \phi_L^2 \cdot \frac{\Delta P_L}{L} = \phi_G^2 \cdot \frac{\Delta P_G}{L} \quad (28)$$

In equation (28),  $\phi_L$  and  $\phi_G$  are correction factors which can be found in the literature for various combinations of laminar and turbulent flow in the two phases (e.g. in Cybulski and Werner, 1986). These correction factors can be correlated using the ratio of pressure drops of the two phases which is defined as:

$$X^2 = \frac{\Delta P_L}{\Delta P_G} = \frac{\Delta P_L/L}{\Delta P_G/L} \quad (29)$$

Chisholm (1967) suggested the following relation for  $\phi_L$  and  $\phi_G$ :

$$\phi_L^2 = 1 + \frac{C}{X} + \frac{1}{X^2} \quad \text{and} \quad \phi_G^2 = 1 + C'X + X^2 \quad (30)$$

In empty pipes,  $C = 20$  in turbulent–turbulent flows,  $C = 5$  in laminar/laminar flows and  $C = 12$  for laminar/turbulent flows. Bao *et al.* (1994) confirmed that this is an effective method for  $\Delta P$  prediction in gas–liquid flows in static mixers, especially in the laminar regime. However, the value of  $C$  must be adjusted for the various kinds of static mixer. Lockhart and Martinelli's method has been applied to SMX mixers (Streiff, 1977; Shah and Kale, 1991, 1992; Chandra and Kale, 1995) and to Kenics and Komax mixers (Chandra and Kale, 1995). Values of  $C$  in the laminar–laminar regime are 3.4, 2.85 and 2.6 for Kenics, Komax and SMX mixers, respectively. A special treatment has been suggested by Chandra and Kale (1995) for non-Newtonian and viscoelastic liquid phases. For power-law liquids, the constant  $C$  is a function of flow index  $n$ :

$$C = \frac{n}{n+1} \cdot C'' \quad (31)$$

Since  $n = 1$  for a Newtonian fluid,  $C'' = 2C'$  where  $C'$  now represents the value obtained in the same mixer with a Newtonian fluid. Thus Chandra and Kale (1995) report  $C'' = 6.8$ , 5.7 and 5.2 for Kenics, Komax and SMX mixers respectively. Elasticity causes additional pressure losses. For a viscoelastic solution, Chandra and Kale (1995) found that  $C'' = 7.1$ , 6.0 and 6.2 for Kenics-, Komax- and SMX-based mixers respectively.

### Pressure drop in liquid–liquid flows

Lockhart and Martinelli's approach is known to be inaccurate for liquid–liquid systems. Bao *et al.* (1994) recommended a methodology proposed by Beattie and Whalley (1982). The liquid–liquid dispersion is considered to be a pseudo-homogeneous phase with an apparent density  $\rho_{L/L}$  and an apparent viscosity  $\mu_{L/L}$ , defined as:

$$\frac{1}{\rho_{L/L}} = \frac{x_1}{\rho_1} + \frac{x_2}{\rho_2} \quad (32)$$

and

$$\mu_{L/L} = \mu_1 \cdot (1 - \zeta) \cdot (1 + 2.5\zeta) + \mu_2 \cdot \zeta \quad (33)$$

where  $x$ ,  $\rho$  and  $\mu$  are respectively the mass fraction, the density and the viscosity of a phases and subscript 1 refers to the heavy phase, while subscript 2 refers to the light phase.  $\zeta$  is defined as

$$\zeta = x_2 \frac{\rho_{L/L}}{\rho_2} \quad (34)$$

The friction factor  $f$  is computed as in single-phase flows using the apparent density and viscosity given by equations (32) and (33) to determine the Reynolds number. The ordinary, single-phase equations for  $f$  vs  $Re$  are then used. This approach can be applied using existing correlations for the various kinds of motionless mixer, but results have not been reported in the literature.

### Pressure drop calculations using CFD

Pressure drops can be calculated from numerical solutions to the equations of motion. However, due to the complex geometry of the mixing elements and the resulting three-dimensional flows, such calculations have only recently become feasible. Attempts were made in the early 1980s to model the flows in the helical Kenics mixer because it was the first of the commercial motionless mixers and its geometry is relatively simple. Approximate analytical solutions for the local velocity fields were also proposed by Dackson and Nauman (1987). Helical elements were sometimes approximated by two-dimensional models such as the partitioned-pipe mixer shown in Figure 15. This geometry is a pipe divided into a sequence of semi-circular ducts by means of rectangular plates that intersect at right angles. A more rigorous example of three- to two-dimensional reduction was done (Dackson and Nauman, 1987). The fully developed flow field was obtained by transforming a nonorthogonal coordinate system and using the stream function to eliminate the axial pressure field. This approach is rigorous except that it is unable to account for entrance effect as the fluid moves from one element to another. The method has been used extensively by Ottino's group (e.g. Khakar *et al.*, 1987).

Table 8. Friction factor correlations for commercial static mixers.

| Mixer            | Correlation                                                                       | References                     |
|------------------|-----------------------------------------------------------------------------------|--------------------------------|
| Helical elements | $f = \frac{85.5}{\text{Re}} + 0.34$                                               | Sir and Lecjaks (1982)         |
|                  | $f = \frac{77.76}{\text{Re}} + \frac{10.88}{\text{Re}^{0.5}}$                     | Grace (1971)                   |
|                  | $f = 0.5 + \frac{115.2}{\text{Re}} \text{ (Re} \leq 50)$                          | Cybulski and Werner (1986)     |
|                  | $f = \frac{6.592}{\text{Re}^{0.5}} \text{ (100} \leq \text{Re} \leq 1000)$        | Cybulski and Werner (1986)     |
| Sulzer SMX       | $f = \frac{368}{\text{Re}_g} \text{ (Re}_g \leq 15)$                              | Li <i>et al.</i> (1997a)       |
|                  | $f = \frac{220}{\text{Re}_g^{0.8}} + 0.8 \text{ (15} \leq \text{Re}_g \leq 1000)$ | Li <i>et al.</i> (1997a)       |
|                  | $f = \frac{12}{\text{Re}_g^{0.25}} \text{ {Re}_g > 1000}$                         | Li <i>et al.</i> (1997a)       |
|                  | $f = \frac{160}{\text{Re}} \text{ to } \frac{1600}{\text{Re}}$                    | Cybulski and Werner (1986)     |
|                  | $f = \frac{160}{\text{Re}} \text{ to } \frac{960}{\text{Re}}$                     | Rauline <i>et al.</i> (1998)   |
| Sulzer SMV       | $f = 0.148 + \frac{78.4}{\text{Re}} \text{ (20} \leq \text{Re} \leq 1000)$        | Cavatorta <i>et al.</i> (1999) |
|                  | $f = \frac{1040}{\text{Re}} \text{ to } \frac{4800}{\text{Re}}$                   | Cybulski and Werner (1986)     |
| Inliner Lightnin | $f = \frac{144}{\text{Re}}$                                                       | Cybulski and Werner (1986)     |
|                  | $f = \frac{118.4}{\text{Re}} + 11.2$                                              | Cybulski and Werner (1986)     |
| N form           | $f = \frac{240}{\text{Re}} \text{ to } \frac{272}{\text{Re}}$                     | Cybulski and Werner (1986)     |
| Komax            | $f = \frac{400}{\text{Re}}$                                                       | Cybulski and Werner (1986)     |
| Hi-Toray         | $f = \frac{608}{\text{Re}}$                                                       | Cybulski and Werner (1986)     |
| ISG              | $f = \frac{4000}{\text{Re}} \text{ to } \frac{4800}{\text{Re}}$                   | Cybulski and Werner (1986)     |
|                  | $f = 5407.5 \left(\frac{D}{L}\right) \cdot \frac{1}{\text{Re}}$                   | Rauline <i>et al.</i> (1998)   |
| LPD              | $f = 146.08 \left(\frac{D}{L}\right) \cdot \frac{1}{\text{Re}}$                   | Rauline <i>et al.</i> (1998)   |
|                  | $f = \frac{139.5}{\text{Re}}$                                                     |                                |

With improvements in CFD codes and particularly with much faster computers, it is now possible to obtain reasonably accurate estimates of laminar flow pressure drops in motionless mixtures. Results have been published on the LPD mixer (Tanguy *et al.*, 1990), the SMRX (Mickaili-Huber *et al.*, 1996), the Kenics mixer (Hobbs *et al.*, 1998) and the SMX mixer (Rauline *et al.*, 2000) on three-dimensional grids. Rauline *et al.* (1998) demonstrated the applicability of CFD to predict pressure drop for six commercially available static mixers (Kenics, Inliner, LPD, Cleveland, SMX and ISG). A comparison of  $K_p$  values obtained from the vendor, from published experimental results and from CFD calculations is given in Table 7. Quite good agreement is achieved for each type of mixer. CFD methods have reached a reasonable state of maturity with respect to pressure calculations in single-phase, isoviscometric, laminar flows. However, CFD is still restricted to a small

subset of the applications for static mixers in industry. Even where the calculations are feasible, there remains an intellectual issue in that the internal workings of most CFD codes are unpublished so that independent verification is impossible.

Key Parameters for Assessing Mixing Homogeneity  
*Mixing homogeneity in nonreacting flows*

There are a great variety of parameters that can be used to assess mixing homogeneity. Grosz-Röll (1980) tabulates more than 50. Unfortunately, these parameters are not always clearly defined and nor easy to compare to each other. There is no single criterion suitable for all applications, and all the criteria have advantages and disadvantages. The first analysis of mixing effectiveness in static mixers used the striation model illustrated in Figure 5. This

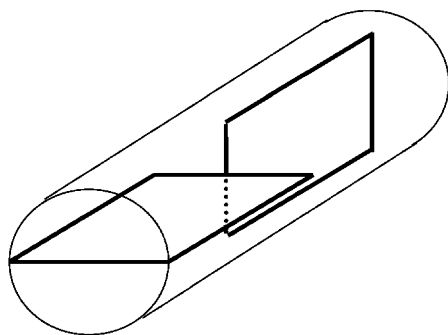


Figure 15. Geometry of the partitioned-pipe mixer.

model remains highly satisfactory from a theoretical viewpoint because it is independent of molecular diffusivity and problems associated with sample size. Given two fluids that are identical except for some measurable characteristic such as color, the performance of a static mixer should depend only on the initial distribution of fluids at the reactor inlet, the geometry of a mixing element, and the number of elements in series. In transitional flows, there may be a dependence on Reynolds number. If the fluids are miscible but have different physical properties, their relative viscosities and volume fraction will also influence the spatial distribution of components at the mixer outlet, but the maximum striation thickness and the striation thickness distribution remain well-defined concepts. Given an accurate tracking of striations and residence times, reaction engineering calculations can be superimposed on a numerical solution. Unfortunately, striations thicknesses are difficult to measure, and even CFD calculations pose certain problems due to numerical diffusion and sampling problems.

Point-to-point concentrations are relatively easy to measure and constitute the basis for most experimental studies on the uniformity of mixing. These measurements are used to calculate the coefficient of variation, COV, and the relative standard deviation, RSD, of concentrations in the outlet stream of a motionless mixer. There are several subtleties associated with the measurements and their interpretation that are often ignored in the literature. The first point is that the sampling scheme must be weighted by volumetric flow rather than cross-sectional area since there will be a difference between the mixing-cup average concentration at the outlet and the spatial average concentration. The second point involves the size of the sample. Too large a sample will mask point-to-point variations in concentration. Too small a sample will give sampling errors. The sampling error problem can be serious in CFD studies where the sampled entities are a relatively small number of tracer particles. It is not a problem in physical measurements of concentration due to the large number of molecules that will be present in any sample. We suppose that the analysis itself is highly accurate. A tutorial on these topics is given in Nauman and Buffham (1983). See also Nauman (1991) for a discussion of flow vs area sampling.

Consider a two-component mixture and denote the scaled concentration of component 1 as  $c_1$  where  $0 < c_1 < 1$ . Then the concentration of component 2 is  $c_2 = 1 - c_1$ . Consider

the cross-sectional plane at the end of the last mixing element and divide the cross-section into a number,  $J$ , of sampling areas. The sampling areas should be sized such that the area multiplied by the local velocity normal to that area is the same for each sampling point. This is sampling according to volumetric flow rate and will give relatively large sampling areas near the tube walls. Using these sampling areas, the mixing-cup or flow-rate averaged concentration is given by:

$$\bar{c}_1 = \frac{1}{J} \sum_{j=1}^J c_{1,j} = \frac{Q_1}{Q_1 + Q_2} \quad (35)$$

where  $c_{1,j}$  denotes the concentration at sampling area  $j$ , and where the volumetric flow rates of species 1 and 2 are denoted  $Q_1$  and  $Q_2$ , respectively. The standard deviation of concentration is:

$$S_1 = \sqrt{\frac{1}{J-1} \cdot \sum_{j=1}^J (c_{1,j} - \bar{c}_1)^2} \quad (36)$$

This scheme can be modified to use equally sized sampling areas, but then the measured concentration must be weighted by the normal velocities associated with each area.

The relative standard deviation is:

$$\text{RSD} = \frac{S_1}{\sqrt{\bar{c}_1(1 - \bar{c}_1)}} \quad (37)$$

The definition supposes that species 1 and 2 were completely segregated at the mixer inlet (i.e. all black or white, no gray). If they remain unmixed,  $\text{RSD} = 1$ . If gray emerges due to molecular diffusion or if the striation thickness becomes so small that the sampling area contains many striations, then RSD can approach zero and the mixture will be homogeneous as the level of scrutiny corresponding to the sampling area.

An alternative measure of homogeneity is provided by the coefficient of variation:

$$\text{COV} = \frac{S_1}{\bar{c}_1} \quad (38)$$

$\text{COV} = 0$  for a complete distributive mixing, while  $\text{COV} = 1$  represents total segregation. Grosz-Röll (1980) considered a system to be homogeneous if  $\text{COV} < 0.05$ , but this definition is arbitrary and insufficient for applications such as blending colors to visual uniformity that may require  $\text{COV} < 0.01$  (Myers *et al.*, 1997). Precision extrusion may require an even smaller COV for homogeneity of temperature.

For a two-component system, RSD and COV are related to each other the ratio of volumetric flow rates:

$$\text{COV} = \left( \frac{Q_2}{Q_1} \right)^{0.5} \text{RSD} \quad (39)$$

A very general correlation for COV has the following form (Grosz-Röll, 1980):

$$\text{COV} = f \left( \text{Re}; \text{Sc}; \frac{\mu_1}{\mu_2}; \frac{|\rho_1 - \rho_2|}{\rho_2}; \frac{Q_1}{Q_2}; \frac{L}{D} \right) \quad (40)$$

COV is a function of Reynolds number, **Re**, Schmidt number, **Sc**, viscosity ratio  $\mu_1/\mu_2$  ( $\mu_1 > \mu_2$ ), density ratio, volumetric flow rate ratio  $Q_1/Q_2$  and the  $L/D$  ratio for the

mixer. The geometric parameters of the mixing elements have been excluded from the correlation variables and the number of mixing elements is reflected by the  $L/D$  ratio. The Schmidt number contains the molecule diffusivity:

$$Sc = \frac{\mu}{\rho \cdot D}$$

(41)

where  $D$  is the binary diffusion coefficient for the mixture. An alternative to the Schmidt number is the dimensionless diffusion number, defined as

$$Di = \frac{D\bar{t}}{D^2}$$

(42)

where  $\bar{t}$  is the mean residence time in the mixer.  $Di$  combines operating and physical variables in a manner similar to that for the Reynolds number.  $Di$  is used in reaction engineering, and Merrill and Hamrin (1970) concluded that molecular diffusion could be ignored in undisturbed laminar flow.

Grosz-Röll (1980) noted that the dependence on Reynolds number can be neglected for laminar flow, that the density ratio has little influence in most applications, and that molecular diffusion is typically negligible. Furthermore, the ratio of flow rates can be replaced by the average concentration of component 1 without loss of generality. Thus, the coefficient of variation becomes a function of three dimensionless variables:

$$COV = f\left(\frac{\mu_1}{\mu_2}; \bar{c}_1; \frac{L}{D}\right)$$

(43)

We ignore for the moment the effect of viscosity ratio. Then a plot of COV vs  $L/D$  for fixed values of  $\bar{c}_1$  shows the approach to homogeneity in a motionless mixer (Alloca,

1982; Pahl and Muschelkautz, 1982). A functional form used to fit COV data is:

$$COV = b \cdot \exp\left(-\frac{Z' \cdot L}{D}\right)$$

(44)

where  $b$  and  $Z'$  are adjustable constants. This relation is useful in for both laminar and turbulent flow. The parameter  $Z'$  represents the rate of decrease in COV per unit of mixer length. Table 9 summarizes experimental result of Pahl and Muschelkautz (1982), Alloca (1982), Grosz-Röll (1980) and Cybulski and Werner (1986). The data in Table 9 are for  $\mu_1/\mu_2 = 1$  and laminar flow. Results for turbulent flow are given by Grosz-Röll (1980) for the corrugated-plate Sulzer SMV that is primarily used in the turbulent regime.

As a first approximation,  $b$  depends only on  $\bar{c}_1$  or, equivalently the volumetric flow ratio,  $Q_1/Q_2$ , in laminar flows, while  $Z'$  depends only on mixer geometry. Grosz-Röll (1980) observed that the COV decreases when  $\bar{c}_1$  is increased. Indeed, the COV should have a minimum at  $\bar{c}_1 = 0.5$  and be symmetric about this point, at least for the case of  $\mu_1/\mu_2 = 1$ , because the components are conceptually interchangeable when they have the same physical properties. Grosz-Röll (1980) specified component 1 to be the high viscosity component so  $\mu_1/\mu_2 \geq 1$ . Symmetry is lost when  $\mu_1/\mu_2 > 1$  because it is easier to mix a low viscosity fluid into a high viscosity than conversely.

The following methodology for choosing a distributive mixer is suggested:

- (i) select a mixer type and diameter;

(ii) specify the desired value of  $\bar{c}_1$ ;

(iii) calculate  $Re$  to determine the flow regime;

(iv) estimate  $b$  and  $Z'$  using Table 9 or data from manufacturers;

Table 9. Parameters  $b$  and  $Z'$  for commercial static mixers.

| Mixer      | $\bar{c}_1$ | $b$   | $Z'$  | $L/D$ to obtain COV = 0.05 | References                   |
|------------|-------------|-------|-------|----------------------------|------------------------------|
| Kenics     | 0.1         | 3.04  | 0.171 | 24.0                       | Pahl and Muschelkautz (1982) |
|            | —           | 5.04  | 0.155 | 29.8                       | Alloca (1982)                |
| Komax      | 0.1         | 4.14  | 0.151 | 29.2                       | Pahl and Muschelkautz (1982) |
|            | —           | 6.24  | 0.125 | 38.6                       | Alloca (1982)                |
| Hi-mixer   | 0.1         | 3.53  | 0.412 | 10.3                       | Pahl and Muschelkautz (1982) |
|            | —           | 3.73  | 0.326 | 13.2                       | Alloca (1982)                |
| Lightnin   | 0.1         | 3.30  | 0.059 | 71.0                       | Pahl & Muschelkautz (1982)   |
|            | —           | 4.28  | 0.040 | 111.2                      | Alloca (1982)                |
| SMX        | 0.1         | 2.88  | 0.472 | 8.6                        | Pahl and Muschelkautz (1982) |
|            | —           | 3.73  | 0.469 | 9.2                        | Alloca (1982)                |
|            | 0.001       | 48.73 | 0.500 | 13.8                       | Grosz-Röll (1980)            |
|            | 0.01        | 15.11 | 0.505 | 11.3                       | Grosz-Röll (1980)            |
|            | 0.1         | 4.52  | 0.505 | 8.9                        | Grosz-Röll (1980)            |
|            | 0.5         | 1.51  | 0.505 | 6.7                        | Grosz-Röll (1980)            |
|            | 0.001       | 53.30 | 0.518 | 13.5                       | Cybulski and Werner (1986)   |
|            | 0.01        | 14.99 | 0.501 | 11.4                       | Cybulski and Werner (1986)   |
|            | 0.1         | 5.54  | 0.526 | 9.0                        | Cybulski and Werner (1986)   |
|            | 0.5         | 1.64  | 0.510 | 6.8                        | Cybulski and Werner (1986)   |
| SMV        | 0.001       | 5.17  | 0.752 | 6.2                        | Grosz-Röll (1980)            |
|            | 0.01        | 1.70  | 0.755 | 4.7                        | Grosz-Röll (1980)            |
|            | 0.1         | 0.51  | 0.742 | 3.1                        | Grosz-Röll (1980)            |
|            | 0.5         | 0.17  | 0.741 | 1.7                        | Grosz-Röll (1980)            |
| SMXL       | 0.001       | 39.97 | 0.163 | 41.0                       | Cybulski and Werner (1986)   |
|            | 0.5         | 1.25  | 0.163 | 19.7                       | Cybulski and Werner (1986)   |
| Etoflow HV | —           | 4.40  | 0.176 | 25.4                       | Alloca (1982)                |
|            | —           | 5.58  | 0.138 | 34.2                       | Alloca (1982)                |

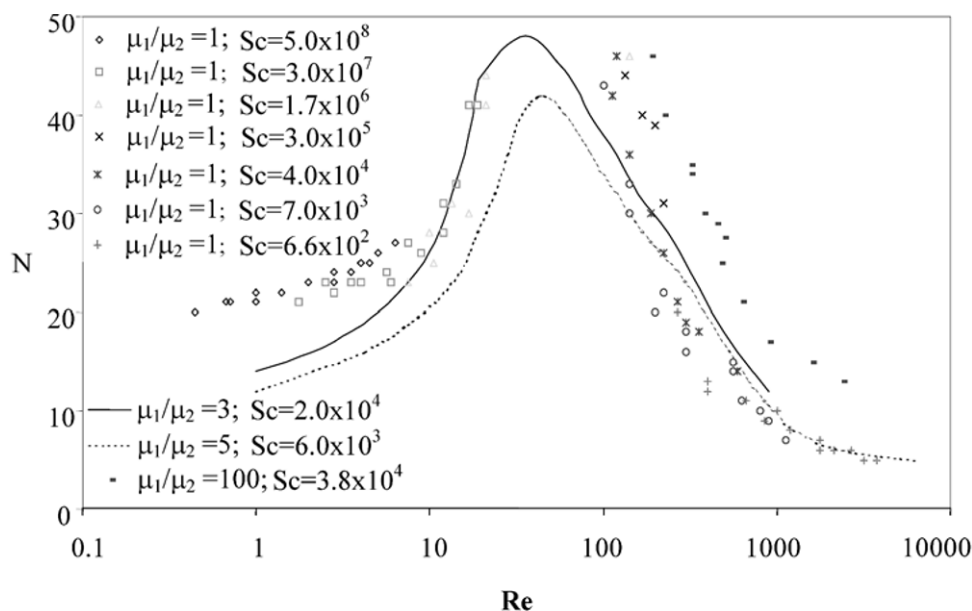


Figure 16. Number of helical elements necessary to achieve homogeneity (adapted from Sir and Lecjaks, 1982).

- (v) use equation (44) to determine  $L$  and of the number of elements necessary to achieve the desired COV value.

Some special cases may require more attention:

- when  $Re$  is in the transition range;
- when  $\mu_1/\mu_2 > 1$ ;
- when  $\bar{c}_1 < 0.01$  or  $\bar{c}_1 > 0.99$ ;
- when the components are only partially miscible even though the homogenized mixture would be single phase.

Situations (i) and (ii) are addressed in Figure 16. This figure shows the number of mixing elements needed to achieve homogeneity ( $COV = 0.05$ ) using Kenics-type helical inserts, for several values of  $\mu_1/\mu_2$  and  $Sc$  (Cybulski and Werner, 1986). It is seen that  $N$  increases with increasing  $Re$  up to about  $Re = 50$ .  $N$  then passes through a maximum and decreases for higher  $Re$ , becoming approximately constant in the turbulent regime. This agrees with the observations of Grosz-Röll (1980) and indicates that homogeneity is more difficult to achieve in the transition region than in the laminar or turbulent regions. Similar behavior is expected with other types of static mixer and has been explained using CFD studies as discussed in the next section. Figure 16 also shows that molecular diffusion is relatively unimportant in the transition and turbulent regimes.  $Sc$  does

affect flows at very low  $Re$  values that are not included in Figure 16. In contrast, there is an effect of viscosity ratio. The continuous curves for  $\mu_1/\mu_2 = 3$  and  $\mu_1/\mu_2 = 5$  in Figure 16 suggests that achieving homogeneity is actually easier when  $\mu_1/\mu_2$  is increased, but this conclusion cannot be generalized to very high values of  $\mu_1/\mu_2$ . Cybulski and Werner (1986) studied the influence of viscosity ratio on  $N$  for several commercial static mixers, and their results are presented in Table 10. See also Rauline *et al.* (2000). When  $\mu_1/\mu_2$  is around  $10^3$ ,  $N$  may be twice the value required when viscosity ratio is around 1. Finally, when  $\mu_1/\mu_2$  is higher than 100,000,  $N$  becomes so high that it is more cost effective to use dynamic mixers than motionless inserts (Myers *et al.*, 1997). Similar results are obtained for volumetric flow rate ratio. This parameter becomes quite important when  $Q_1/Q_2$  is outside the range of 0.01 to 100 ( $0.01 < \bar{c}_1 < 0.99$ ). Special injection devices may be required for the minor component or else dynamic mixers may be necessary (Baker, 1991).

Sir and Lecjaks (1982) recommended the following equation to estimate the optimal number of Kenics helical elements:

$$N = a' Re^{a_1} \cdot Sc^{a_2} \cdot \left(\frac{Q_1}{Q_2}\right)^{a_3} \cdot \left(\frac{\mu_1}{\mu_2}\right)^{a_4} \quad (45)$$

Table 10. Number of mixing elements needed to achieve homogeneity.

|                            | Number of elements, $N$ |                         |                             |                      |
|----------------------------|-------------------------|-------------------------|-----------------------------|----------------------|
|                            | $\mu_1/\mu_2 \approx 1$ | $\mu_1/\mu_2 \approx 1$ | $\mu_1/\mu_2 < 10^3$        | $\mu_1/\mu_2 > 10^3$ |
| Mixer type                 | $Re < 10$               | $10 < Re < 10^3$        | $10^3 < Re < 2 \times 10^3$ | $Re > 2 \times 10^3$ |
| Kenics $L_c/D_c = 1.7$     | 10                      | 20                      | 31                          | 41                   |
| Sulzer SMV $L_c/D_c = 1$   | 2                       | —                       | —                           | —                    |
| Sulzer SMX $L_c/D_c = 1.5$ | —                       | 8                       | 11                          | 14                   |
| N form $L_c/D_c = 1.2$     | 7                       | 13–19                   | 19–25                       | 25                   |

The values of the empirical parameters  $a'$ ,  $a_1$ ,  $a_2$ ,  $a_3$  and  $a_4$  depend on the flow regime. For example, Sir and Lecjaks (1982) showed that  $a_2$  reaches a minimum value for  $Re > 400$ . The functional form of equation (45) should be useful for mixers other than the Kenics.

#### Mixing in reactive flows

When static mixers are used as reactors, the residence time distribution becomes an important parameter. The approach to piston flow with increasing  $N$  has been the subject of many studies (Bor, 1971; Nigam and Vasudeva, 1980; Nauman, 1982). A variety of models have been proposed, generally for the Kenics helical mixer, using Newtonian and non-Newtonian fluids (Wen and Fan, 1975; Nigam and Naumann, 1985; Pustelnik, 1986; Kembrowski and Pustelnik, 1988). More recently, RTD measurements and modeling with SMX static mixers have been published (Li *et al.*, 1996; Fradette *et al.*, 1998; Yoon and Choi, 1996). Manufacturer's data are also available, sometimes expressed as the Peclet number that is the key parameter in the well-known axial dispersion model:

$$Pe = \frac{uL}{D_{ax}} \quad (46)$$

where  $D_{ax}$  is the axial dispersion coefficient.  $D_{ax}$  has the same units as molecular diffusion but is intended to reflect the combined effects of convection, molecular diffusion and, for turbulent flow, eddy diffusion. The yield of an isothermal, first-order reaction depends only on the residence time distribution, and any model is adequate for predicting the yield of such reactions provided it adequately the residence time distribution function. The axial dispersion model will do this when  $N$  is large and the residence time distribution approaches piston flow, but the axial dispersion model is unsuited for use with complex reactions in the laminar transitional regimes. It is also unsuited for reactions with unmixed feeds.

The necessary approach for laminar flows is a rigorous solution of the convective diffusion equations for mass and heat, combined with the equations of motions that are coupled due to the dependence of viscosity on temperature and composition. Such calculations are feasible for undisturbed laminar flow in tubes with premixed feed. See for example Nauman (2002, Chapter 8). They remain infeasible for motionless mixers. Modern CFD codes are accurate for pressure and velocity calculations but are notoriously inaccurate for diffusion. The difficulty is partly due to numerical diffusion induced by the convergence acceleration techniques used in most codes and the difficulty in accurate material balance closure using the finite element or finite volume techniques that are now preferred for CFD. Work is in progress, but a new generation of CFD codes will be necessary to enable accurate reaction calculations.

The use of static mixers for fast chemical reactions with unmixed feed streams has received substantial literature attention. Readers will find details in the contributions of Bourne and Maire (1991), Bourne *et al.* (1992), Penney *et al.* (1995) and Baldyga *et al.* (1997), but none of the methods proposed have yet received general acceptance. Fast reactions with unmixed feed streams pose a currently insurmountable problem for CFD codes.

#### Predicting mixing using CFD

Owing to the complex geometry of static mixers, analytic solutions for the velocity field are infeasible. However, numerical solutions can provide a starting point for understanding mixing performance. In particular, simulations can provide qualitative insights that can be used to improve mixer designs. For example, alternate configurations of the Kenics mixer can be generated by varying the geometric factors such as the aspect ratio and twist angle. The velocity fields for these alternate configurations obtained from CFD computations can then be analyzed in a number of ways to characterize performance. Lagrangian tracking of particles is a standard tool for this analysis. The results can be used to determine the residence time distribution, and various inlet-to-outlet mappings including Poincaré sections and stretching histories. Conceptually at least, these measurements can be used to understand the action of the mixer for applications such as heat or mass transfer, coalescence and break-up of drops, and chemical reactions. The coefficient of variation can be calculated subject to the caveats above. Recall also that static mixers are flow devices and that particle tracking experiments should be weighted by flow rate rather than area, a fact sometimes forgotten by otherwise sophisticated researchers.

CFD studies on helical inserts of the Kenics type are fairly extensive. Early approaches were mentioned above. A commercial finite element, FLUENT, was used by Bakker and LaRoche (1993) and Bakker *et al.* (1994) to study the Kenics KM and HEV mixers. Gyenis and Blickle (1992) performed stochastic simulations of unsteady-state particle flows. Hobbs and Muzzio (1997a,b) did significant work in numerically characterizing the Kenics static mixer using a commercially available CFD, FLUENT/UNS<sup>TM</sup>. The numerical approach accounts for transitions between mixer elements and the finite thickness of the mixer elements, factors that were neglected in earlier studies. Flow transitions at the entrance and exit of each element affect the velocity field for about 25% of the element length under creeping flow conditions. The magnitude of the rate of strain tensor was roughly uniform over the central 75% of a mixing element, but shifted toward higher values in the end regions where the element-to-element transition occurs. Particle tracking simulations were used to compute residence time distributions, striation evolution and the coefficient of variation as a function of the number of mixer elements for low Reynolds number flows (Hobbs and Muzzio, 1998). The average stretching of material elements increased exponentially with the number of elements, which is a signature of chaotic flows. The logarithm of stretching intensity had a Gaussian distribution over the central spectrum of stretching intensities, with no deviation from the Gaussian profile at low stretching intensities, suggesting a globally chaotic flow. For creeping flow conditions ( $Re < 1$ ), the flow in Kenics elements is globally chaotic, and mixing performance is independent of  $Re$ . For  $Re = 100$ , significant islands of regular motion develop. These islands do not exchange material with the remainder of the flow and act as barriers to uniform mixing. For  $Re = 1000$ , the flow is again predominantly chaotic but small islands still lead to less effective mixing than under creeping flow conditions. This finding is broadly consistent with the experimental results shown in Figure 16.

Hobbs *et al.* (1998) used CFD to study variations on the standard Kenics mixer geometry. They investigated the twist direction of the adjacent elements, the length to diameter ratio of an element and the amount of twist per element. While flow in a mixer with the standard configuration—where elements have alternating twist direction—is generally chaotic; the flow in a mixer where all elements have the same twist direction displays large segregated areas that prevent homogenization. The extent of mixing per element was found to be independent of aspect ratio, suggesting the use of shorter elements. This finding is also consistent with experimental results (Joshi *et al.*, 1995). In creeping flow, a substantial increase in mixer efficiency can be achieved using elements with less twist than the standard 180° design. Thus agrees with the suggestions by Dackson and Nauman (1987) who also studied the creeping flow regime. Variations on element twist angle  $\theta$  were investigated over a range  $\theta = 30^\circ$  to  $\theta = 210^\circ$ . These variations affected the nature of the flow, and a distinct maximum in mixer efficiency was found near 120° of twist per element. The mixer was 44% more energy efficient than the standard mixer with a 180° twist.

Hobbs and Muzzio (1997b) investigated the performance of the Kenics static mixer for adding small streams of an inert tracer into the bulk flow. Flow ratios of 1/99 and 10/90 were simulated at nine different injection points. Two alternative geometries are considered in addition to the standard Kenics geometry. The spread of the tracer was visually evaluated by examining the cross-sectional slices at various axial positions. The coefficient of variation was also computed as a function of axial position. For the standard Kenics geometry, the extent of mixing depends on the location of the injection for the first few elements, but subsequently becomes independent of injection location. In a sufficiently long mixer, material injected at any location spreads to the entire flow, but the least effective injection locations require up to four elements more than the most effective locations to achieve the same COV value. The COV decreases more rapidly for a flow ratio of 1/99 than 10/90. An alternative geometry in which the elements have a 120° of twist instead of the standard 180° of twist shows a similar dependence on injection location and flow ratio, but is more energy-efficient than the standard Kenics geometry. When all elements have the same direction of twist, segregated islands exist in the flow. If the injection into these segregated zones, there is virtually no mixing. For injection outside of the segregated zones, the tracer spreads to the remaining flow, but does not penetrate the zones.

Byrde and Sawley (1999) studied the optimization of a Kenics static mixer for **Re** above the noncreeping flow region. Contrary to the previous investigations for creeping flow, it was found that the standard twist angle of 180° is indeed optimal. It is clear that the Kenics geometry behaves differently (and generally worse) in transitional flows than in deep laminar flows.

Turning to other types of mixers, Bertrand *et al.* (1994) used a commercial code, POLY3D™ from Rheotek Inc., to investigate the residence time distribution in LPD and ISG mixers. They set  $N = 2$  so the distribution was little changed from undisturbed laminar flow. The ISG mixer was found to be more efficient than the LPD mixer in shifting the residence time distribution toward that of piston flow.

Turbulent mixing in a Sulzer SMV static mixer was studied by Lang *et al.* (1995). A finite volume scheme was used to solve the continuity, momentum and energy equations. The application was an industrial denitrification process. The simulation showed that maldistributions of concentration and temperature were reduced by the SMV mixers, but that a considerable part of the mixing occurs in the wake of the SMV mixer. The mixer induces vortices that continue the mixing downstream of the SMV.

Tanguy *et al.* (1990, 1993) produced a preliminary analysis of flow in SMRX static mixer. Mickaily-Huber *et al.* (1996) also studied the flow in SMRX with the aim of optimizing the design. A finite element method was used, but commercial grid generators were unable to create a grid for the complex geometry of the SMRX, and a special grid generator had to be developed. The effect of twist angle between elements, which is generally neglected in experimental investigations, on pressure drop, mixing and intensity of segregation was studied numerically. A 90° inner element twist angle was found to provide the most efficient mixing.

Fradette *et al.* (1998) performed a three-dimensional finite element simulation of fluid flow through a SMX static mixer. Calculated pressure drops for both Newtonian and non-Newtonian fluids were compared to the experimental measurements of Li *et al.* (1996). Good agreement between simulations and experiments showed that finite element simulation can properly represent the very complex velocity field generated by the SMX mixer. The energy levels at various points in the mixer and axial elongation rates were also calculated. Visser *et al.* (1999) used CFD to calculate flow velocities and residence time distributions in a three dimensional model of the SMX. The residence time calculations determined by particle tracking gave Peclet numbers of 4.2 per SMX element in good agreement with experimental values.

The mixing performance of the helical Kenics mixer and the SMX mixer were compared in the creeping flow regime by Rauline *et al.* (2000) using three-dimensional numerical simulations. Several criteria were used: mixer length, Lyapunov exponent, mean shear rate and intensity of segregation. The SMX mixer was found to be more efficient than the Kenics when the mixing task is difficult or when installation space is restricted.

### Key Parameters for Interface Generation

Interface generation and mass transfer enhancement using static mixers is much more difficult to predict than mixing effectiveness for miscible fluids. Key parameters for gas-liquid systems such as gas hold-up and average bubble size are difficult to predict even in empty pipes or in conventional gas-liquid contactors like bubble columns due to the large number variables that affect performance: two densities, two viscosities, surface tension, two flow rates, co- or counter-current operation and device geometry. As a consequence, parameters are easy to put into evidence, but only a few correlations are available for their prediction. In this section, we review the parameters important for characterizing interface generation in static mixers for multiphase systems and the methods available for design purposes.

### Gas-liquid systems

In gas-liquid systems, static mixers can be used in bubble, spray and stratified annular flows. A key parameter to be estimated is the liquid-side mass transfer coefficient,  $K_L a$ . General methods for measuring  $K_L a$  in gas-liquid systems can be found in Márquez *et al.* (1994). Some correlations of experimental data have been published for commercial motionless mixers, but their range of applicability is limited. As a result, using motionless inserts for multiphase mass transfer requires either experimental investigations or proprietary data from vendors or other sources. Wang and Fan (1978) used the following correlating equation for a fixed length bubble column equipped with motionless mixers:

$$K_L a = a' \cdot u_L^{b'} \cdot \varepsilon_G^{c'} \quad (47)$$

Here,  $u_L$  is the liquid superficial velocity,  $\varepsilon_G$  is the gas hold-up, and  $a'$ ,  $b'$ ,  $c'$  are adjustable constants. The authors showed that the static mixers increase both the interfacial area,  $a$ , and the liquid-side mass transfer coefficient,  $K_L$ . The increase in interfacial area is due to an increase in gas hold-up due to the lower relative velocity and the higher residence time of small bubbles. A strong influence of  $u_L$  on  $K_L$  was also reported, contrary to conventional bubble columns. This is mainly due to the positive influence on mixing of motionless elements in the continuous liquid phase

A more general correlating equation for a given gas and liquid replaces the gas holdup with the gas superficial velocities,  $u_G$ , and adds dependence on the  $L/D$  ratio for the system:

$$K_L a = a' \cdot u_L^{b'} \cdot u_G^{c'} \cdot \left(\frac{L}{D}\right)^{d'} \quad (48)$$

where  $d'$  is another adjustable constant. Existing data show that  $K_L a$  in a bubble column can be increased 30–400% through the use of static mixers. An additional advantage of static mixers for gas-liquid mass transfer is that the volume occupied by the static mixers can be much lower than the volume occupied by conventional packings. Static mixers with a high void fraction are generally desirable in gas-liquid flows. Radar *et al.* (1989) showed that the hydraulic diameter  $d_h$  of the mixing elements is an important parameter for gas-liquid applications, as  $d_h$  has a strong influence on interfacial area.

In absorption processes, correlations can be made directly for the height of a transfer unit, HTU. Cybulski and Werner (1986) reported the following equations for an air/naphthalene column equipped with Sulzer SMV mixers:

$$\begin{aligned} \text{HTU}_G &= 0.66 \cdot d_h \cdot \text{Re}_G^{0.4} \cdot \text{Sc}_G^{0.66} \\ \text{HTU}_L &= 0.00245 \cdot \text{Re}_L^{0.22} \cdot \text{Sc}_L^{0.5} \end{aligned} \quad (49)$$

In these expressions,  $\text{Sc}_L$ ,  $\text{Re}_L$ ,  $\text{Sc}_G$  and  $\text{Re}_G$  are the Schmidt and Reynolds numbers for the liquid and the gas phases. As a rule of thumb, mass transfer effectiveness using SMV corrugated plates is about four times higher than that of Pall rings, and  $K_L a$  is 3–10-fold higher than in an absorption column using Raschig rings.

Internal and external loop reactors have received considerable attention. It is clear that static mixers are able to increase interfacial area, but they also cause a greater pressure loss that reduces liquid circulation in loop reactors

(Goto and Gaspillo, 1992). This could in turn decrease  $K_L$ , but an increase in  $K_L a$  is the usual result from using static mixing elements. For loop reactors,  $u_L$  is not an independent variable, and with a fixed reactor size, equation (48) becomes a function of  $u_G$  alone:

$$K_L a = a' \cdot u_G^{b'} \quad (b' > 0) \quad (50)$$

For an external loop airlift reactor, the performance of a helical corrugated sheet with a 90° twist angle (Goto and Gaspillo, 1992) and SMV elements (Chisti *et al.*, 1990) has been compared to that of conventional loop reactors (see Figure 17). Chisti *et al.* (1990) also studied also the influence of the rheological properties of the continuous phase for shear-thinning fluids. They found that  $K_L a$  increased more for fluids with high consistency indices. This effect decreases however at high gas flow rate, as highly viscous liquids are known to favor coalescence. Results are shown in Figure 17.

Interfacial area is generally easy to estimate for annular flows. In this case, only the mass transfer coefficient has to be predicted. For Kenics helical elements, Morris *et al.* (cited by Joshi *et al.*, 1995) proposed expressions for the Sherwood number  $\text{Sh}$ .

$$\begin{aligned} \text{Sh} &= \frac{K_L \cdot d_h}{D} = 1.86 \cdot (1 + 0.32 \cdot N) \cdot \left[ \text{Re}_h \cdot \text{Sc} \cdot \frac{d_h}{L} \right]^{0.88} \\ &\quad 500 < \text{Re} < 1600 \\ \text{Sh} &= \text{Sh}_0 \cdot [1 + 0.06 \cdot N^{0.283} \cdot \text{Re}_h^{0.32} \cdot \text{Sc}^{-0.4}] \\ &\quad 9000 < \text{Re} < 30,000 \end{aligned} \quad (51)$$

In these expressions,  $\text{Re}_h$  is a modified Reynolds number based on the hydraulic diameter  $d_h$ ,  $D$  is the liquid diffusivity and  $\text{Sh}_0$  is the Sherwood number in an empty pipe.  $\text{Sh}_0$  can be estimated from:

$$\text{Sh}_0 = 0.023 \left[ 1 + 6 \frac{D}{L} \right] \text{Re}^{0.8} \text{Sc}^{0.4} \quad \frac{L}{D} > 20 \quad (52)$$

### Liquid-liquid systems

In welcome contrast to gas-liquid systems, many studies have been aimed at estimating the Sauter-mean diameter in co-current, liquid-liquid systems using static mixers. The

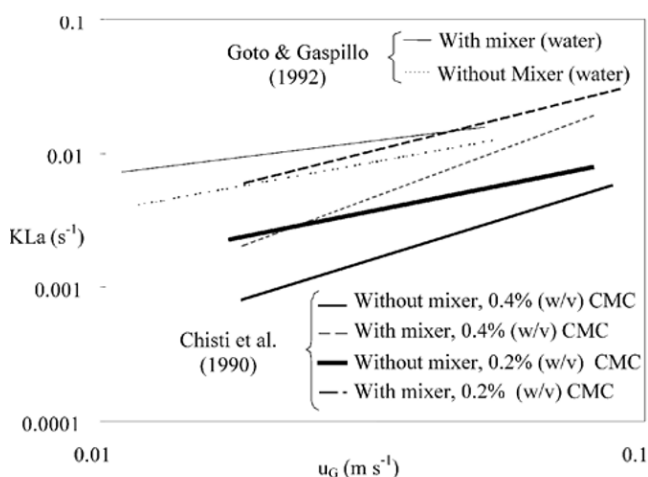


Figure 17. Influence of static mixers on  $K_L a$  in gas-liquid bubble columns.

first contribution to the prediction of  $d_{32}$  was that of Middleman (1974) who studied the Kenics static mixer. He proposed the following relation:

$$\frac{d_{32}}{D} = A \cdot \text{We}^{a_1} \cdot \text{Re}^{a_2} \quad (53)$$

Since his pioneering work, many other correlations have been published that include a variety of commercial static mixers. Results are summarized in Table 11. Most show an explicit dependence on the Weber number with an exponent that ranges from  $-0.74$  to  $-0.4$ . The negative exponent means that increasing the flow rate of the continuous phase or decreasing the surface tension leads to smaller drops. The exponent on Reynolds number is slightly positive so that an increase in viscosity of the continuous phase favors drop size reduction. Only a few of the studies varied the mixer geometry, so most results are restricted to the specific device that was used. For the three studies in which  $N$  was varied, the exponent ranged from  $-0.825$  to  $-0.436$  so that increasing the number of mixing elements can significantly lower the drop size. Figure 18(a) shows the influence of  $\text{We}$  on  $d_{32}$  for these studies, and Figure 18(b) shows how the ratio of dispersed to continuous phase viscosity affects  $d_{32}$ . The published correlations in Table 11 and cover a variety of liquid-liquid systems. The dispersed phases were typical organics including kerosene, benzene, toluene and oil. The various correlations include the effects of dispersed phase

properties such as density, viscosity and volume fraction. Water was used as the continuous phase in all or nearly all of the studies, so considerable care is recommended if another continuous phase is used.

When the operational goal is mass transfer, e.g. for co-current liquid-liquid extraction, the key parameter is the liquid-liquid transfer coefficient,  $K_D a$ . Literature studies include extractions in systems of benzene-water-acetic acid, toluene-water-acetic acid systems, butanol-water-succinic acid, toluene-water-acetone and carbon tetrachloride-water-propionic acid systems (Jancic *et al.*, 1983; Streiff and Jancic, 1984). A relation equivalent to equation (48) for  $K_L a$  can be used for  $K_D a$  (Cybulski and Werner, 1986):

$$K_D a = a' \cdot u_c^{b'} \cdot u_d^{c'} \cdot \left(\frac{D}{L}\right)^{d'} \quad (54)$$

where  $u_c$  and  $u_d$  are the superficial velocities of the continuous and the dispersed phases, respectively. Experimental results on toluene-water and benzene-water systems showed that  $d'$  has only a slight dependence on the physical properties of the fluids. As a result,  $K_D$  and Sherwood number can be assumed to be independent of aspect ratio  $L/D$  when the number of elements is higher than 4 (Joshi *et al.*, 1995). In contrast,  $b'$  and  $c'$  are quite different for toluene-water and benzene-water systems and cannot be predicted from measurements made on different fluids. An

Table 11. Correlations for Sauter-mean diameter for drops in liquid-liquid systems.

| Type of mixer | Correlation                                                                                                                                                                                                               | References                     |
|---------------|---------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------|--------------------------------|
| Kenics        | $\frac{d_{32}}{D} = C \text{We}^{-0.6} \text{Re}^{0.1}$                                                                                                                                                                   | Middleman (1974)               |
| Sulzer SMV    | $\frac{d_{32}}{d_h} = 0.21 \text{We}^{-0.5} \text{Re}^{0.15}$                                                                                                                                                             | Streiff (1977)                 |
| Lightnin      | $\frac{d_{32}}{D} = C \text{We}^{-0.6} \text{Ne}^{-0.4}$                                                                                                                                                                  | Al-Taweel and Walker (1983)    |
| Sulzer SMV    | $\frac{d_{32}}{d_h} = 0.12 \text{We}^{-0.5} \text{Re}^{0.15}$                                                                                                                                                             | Streiff and Jancic (1984)      |
| Sulzer SMV    | $\frac{d_{32}}{d_h} = 0.047 + 0.197 \text{We}^{-0.5} \text{Re}^{0.15}$ (SS316)                                                                                                                                            | Sembira <i>et al.</i> (1986)   |
|               | $\frac{d_{32}}{d_h} = -0.3 + 0.91 \text{We}^{-0.5} \text{Re}^{0.15}$ (Teflon)                                                                                                                                             |                                |
| Kenics        | $\frac{d_{32}}{D} = 1.12 \text{We}^{-0.65} \text{Re}^{0.2} \cdot R_\mu^{0.5}$                                                                                                                                             | Haas (1987)                    |
| Kenics        | $\frac{d_{32}}{D} = 0.49 \text{We}^{-0.6} \left\{ 1 + 1.38 \text{Vi} \left( \frac{d_{32}}{D} \right)^{1/3} \right\}^{0.6}$                                                                                                | Berkman and Calabrese (1988)   |
| SMR           | $d_{32} = C \cdot (1 + K\phi) \cdot \left\{ \frac{\text{We}_c(1 + B\text{vi})}{2(1 + ((7d_{32}/d_h)))} \right\} \left( \frac{\sigma}{\rho_c} \right)^{0.6} \left( \frac{\rho_c}{\rho_d} \right)^{0.2} \varepsilon^{-0.4}$ | Streiff and Kaser (1991)       |
| Lightnin      | $d_{32} \propto \text{We}^{-a}$ , $0.71 < a < 0.83$ $d_{32} \propto \varepsilon^{-b}$ , $0.49 < a < 0.6$                                                                                                                  | El-Hamouz <i>et al.</i> (1994) |
| 'Screen Mesh' | $d_{32} = 0.682 \cdot [\text{We}_{\text{jet}}^{-0.859} \phi^{0.875}] \cdot \left( \frac{l}{L} \right)^{0.33}$                                                                                                             | Al Taweel and Chen (1996)      |
| Koflo         | $d_{32} = 0.483 \cdot D^{1.202} \cdot u^{-0.556} \cdot \sigma^{0.556} \cdot \mu_c^{-0.56} \cdot \mu_d^{0.004} \cdot N^{-0.436} \cdot \phi^{0.663}$                                                                        | Maa and Hsu (1996)             |
| Komax         | $d_{32} = 0.794 \cdot D^{2.112} \cdot u^{-0.556} \cdot \sigma^{0.556} \cdot \mu_c^{-0.56} \cdot \mu_d^{0.004} \cdot N^{-0.660} \cdot \phi^{0.663}$                                                                        | Maa and Hsu (1996)             |
| Sulzer        | $d_{32} = 0.483 \cdot D^{2.112} \cdot u^{-0.556} \cdot \sigma^{0.556} \cdot \mu_c^{-0.56} \cdot \mu_d^{0.004} \cdot N^{-0.825} \cdot \phi^{0.663}$                                                                        | Maa and Hsu (1996)             |
| Sulzer SMV    | $d_{32} = C \cdot (1 + K\phi) \left( \frac{\text{We}_c(1 + a \cdot \text{Vi})}{2} \right)^{0.6} \left( \frac{\sigma}{\rho_c} \right)^{0.6} \left( \frac{\rho_c}{\rho_d} \right)^{0.1} \varepsilon^{-0.4}$                 | Streiff <i>et al.</i> (1997)   |

$\text{We}_c$  is the Weber number of based on maximum stable diameter of droplets

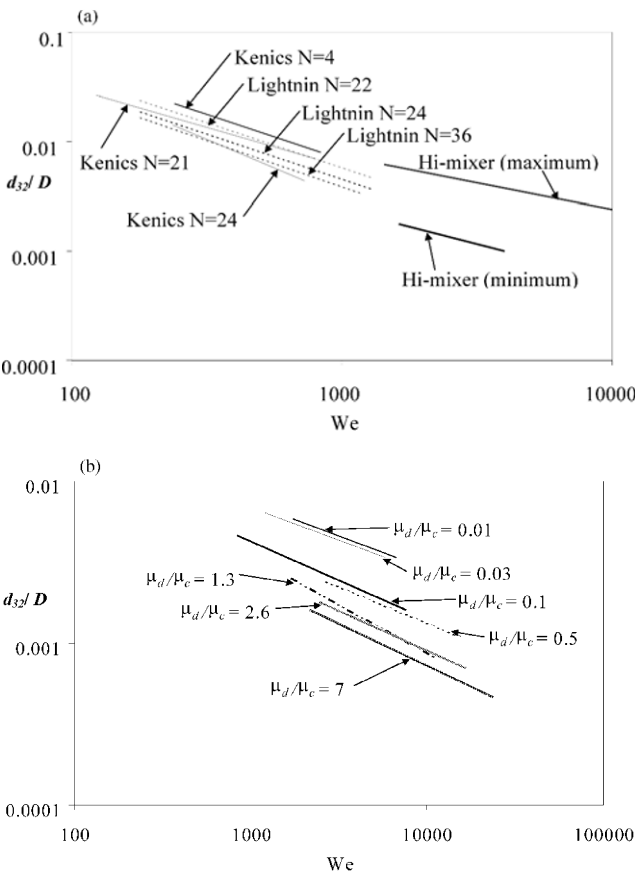


Figure 18. Sauter-mean drop size vs Weber as a function of (a) mixer type (b) viscosity ratio (adapted from Matsumura *et al.*, 1981).

alternative to equation (54) has been used by Streiff *et al.* (1997):

$$K_D a = a' \cdot u_c^{b'} \cdot (1 - \phi) \tag{55}$$

This equation uses the volume fraction of the dispersed phase,  $\phi$ , and has the advantage of using fewer adjustable parameters.

Experimental data and empirical correlations are scarce on mass-transfer in liquid-liquid systems using static mixers. Little is known about shear-induced phase inversion

except than it often favored by motionless inserts. Experimental investigations are still necessary to understand the utility of static mixers for liquid-liquid mass transfer and to improve tools for process design and scale-up. Computational studies using CFD remain in the future.

*Multiphase flows with a dispersed solid phase*

When the design objective is the dispersal and deagglomeration of solid particles, there are no generalized design methods. When static mixers are used to break flocculates, for example in pulp and paper or in water-treatment industries, the final size of the particles is the key parameter. However, as the phenomena involved depend essentially on properties of the solid phase, particle size and shape in the exit stream are rather difficult to predict and no reliable data seem to be available on this topic in the literature. When motionless mixers are used to enhance chemical reactions, e.g. in liquid-solid and gas-liquid-solid fluidized beds or slurry reactors, the key parameters are the liquid-solid mass transfer coefficient  $K_S$  and the bed expansion  $\varepsilon_P$ . In the case of liquid-solid fluidized beds,  $\varepsilon_P$  is often related to liquid velocity using the following expression:

$$\varepsilon_P = a' \cdot u_L^{1/m} \tag{56}$$

Even at low liquid velocities, static mixers can decrease bed expansion by a factor 2–5 using SMV static mixer as compared to a conventional fluidized bed. The parameter  $m$  varies from 2.5 in turbulent flow to 4.5 in laminar flow for beds without inserts (McCabe *et al.*, 1985). Metzdorf *et al.* (1991) found  $m = 4.65$  using SMV elements as inserts. The advantage of static mixing elements is that they reduce bed expansion when liquid velocity is increased. This leads to better control of bed expansion and allows a larger range of liquid flow rates (see Figure 19; Metzdorf *et al.*, 1991). Figure 19 also shows the influence of column diameter  $D$  and hydraulic diameter of the mixing elements. Cavatorta *et al.* (1999) found that static mixers only slightly increase the liquid-solid mass transfer coefficient in fluidized beds. However, Goto and Gaspillo (1992) report that static mixers do improve liquid-solid mass transfer in slurry bubble columns, but no correlation is available for design and scale-up calculations.

Key Parameters for Heat Transfer

Heat transfer encompasses thermal homogenization and heat exchange. The mechanisms and key parameters for homogenization are similar to those described above for mixing miscible fluids. Here, we only consider heat transfer to a wall. As already mentioned, static inserts can be added to the tubes of conventional shell and tubes heat exchangers. The classical procedures for heat exchanger design are not affected by static mixers except the estimation of the pressure drop as discussed earlier and the inside heat transfer coefficient,  $h_i$ . The key parameter is the Nusselt number:

$$Nu = \frac{h_i D}{k} \tag{57}$$

where  $k$  is the thermal conductivity of the process fluid. For comparison purposes, Table 12 gives the classical correlations for heat transfer inside an empty tube. A viscosity correction factor,  $(\mu_{wall}/\mu_{bulk})^{0.14}$ , is usually added and a

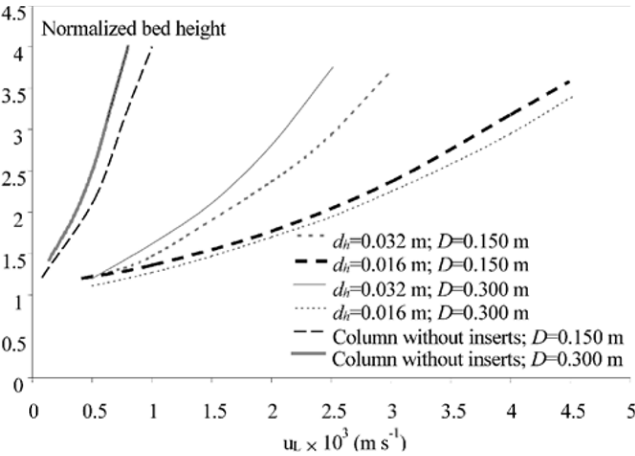


Figure 19. Evolution of bed height in fluidized bed reactors equipped with inserts.

Table 12. Correlations for Nusselt number in an empty pipe.

|                  |                           |                                                         |                                          |
|------------------|---------------------------|---------------------------------------------------------|------------------------------------------|
| Laminar regime   | Constant heat flux        | $Nu = 1.95 \cdot \left( RePr \frac{D}{L} \right)^{1/3}$ | $Nu = \frac{48}{11} (L \gg D)$           |
|                  | Constant wall temperature | $Nu = 1.62 \cdot \left( RePr \frac{D}{L} \right)^{1/3}$ | $Nu = 3.66 (L \gg D)$                    |
| Turbulent regime | General                   | —                                                       | $Nu = 0.023 Re^{0.8} Pr^{0.4}$ (heating) |
|                  |                           | —                                                       | $Nu = 0.023 Re^{0.8} Pr^{0.3}$ (cooling) |

buoyancy correction is occasionally needed as well. A general relation to predict **Nu** when static mixers are inserted in pipes has been proposed by Cybulski and Werner (1986):

$$Nu = Nu_0 + a' \cdot Re^{b'} \cdot Pr^{c'} \cdot \left( \frac{D}{L} \right)^{d'} \quad (58)$$

In this expression, **Nu**<sub>0</sub> represents pure conductive, and **Pr** is the Prandtl number, defined as:

$$Pr = \frac{\mu c_p}{k} \quad (59)$$

where *c<sub>p</sub>* is the specific heat of the process fluid.

Table 13 gives published correlations for the Nusselt number for flow inside tubes that contain static mixers. The correlations are assumed to be independent the ratio of wall viscosity to bulk viscosity, presumably on the grounds that there is good renewal of liquid at the walls. However, Li *et al.* (1996) found this assumption to be incorrect of SMX mixers in laminar flow and it is probably incorrect for other mixers given a large enough temperature difference. For laminar flow, all but one of the correlations in Table 13 show a dependence on *L/D*, indicating that renewal of liquid at the wall is limited and that a thermal boundary layer does develop in static mixers. As a rule of thumb, heat transfer coefficient can be increased by a factor of 2–3 using helical elements and by a factor of 5 using Sulzer SMX elements. A comparison of heat transfer enhancement using commercially available motionless mixers is presented in Figure 20. Joshi *et al.* (1995) give a detailed design comparison of

Kenics mixers compared to open tubes and conclude that, for the same duty, pressure drop and in-process inventory, the Kenics should be installed in a tube about 40% larger than would be used for an empty tube.

Fan *et al.* (1978) proposed a surface renewal model for condensation or vaporization inside a tube equipped with static mixers:

$$Nu = a' [Pr] [Re_L] \left[ \frac{\mu_L}{\mu_m} \right] \left[ \frac{D}{L} \right] \left[ \frac{\rho_m}{\rho_L} \right]^{1/2} \times \left[ \frac{(1 - x') H_{fg} + H_{sp}}{C_{pm}(T_b - T_w)} \right]^{1/2} \quad (60)$$

In equation (60) the subscript m represents the gas–liquid mixture, *H<sub>fg</sub>* is the latent heat, *H<sub>sp</sub>* is the sensible heat, the subscript w represents properties at the wall and *x'* is the mass fraction of vapor at the condenser exit, namely vapor quality. Azer and Lin (1980) showed that helical elements could increase heat transfer up to a factor of 10. These authors defined a new key parameter *H* which takes both heat transfer enhancement and pressure drop increase into account:

$$H = \frac{(Q_{heat}/A_q \cdot \Delta P)_{with \text{ mixer}}}{(Q_{heat}/A_q \cdot \Delta P)_{without \text{ mixer}}} \quad (61)$$

In this expression, *Q<sub>heat</sub>* is the heat flux while *A<sub>q</sub>* is the heat transfer area. The experimental results of Azer and Lin (1980) showed that *H* decreases when Reynolds number is increased, as pressure drop increases dramatically with **Re** in

Table 13. Correlations for Nusselt number in tubes containing static mixers.

|                |                                                                                                       |                                                                                 |                                    |
|----------------|-------------------------------------------------------------------------------------------------------|---------------------------------------------------------------------------------|------------------------------------|
| Helical Kenics | $Nu = 1.87(RePr)^{1/3}$                                                                               | Laminar regime, constant temperature                                            | Lammers <i>et al.</i> (1994)       |
|                | $Nu = 3.65 + 3.89 \left( Re \times Pr \frac{D}{L} \right)^{1/3}$                                      | $Re < 2000$                                                                     | Grace (1971)                       |
|                | $Nu = 1.44(Re \times Pr)^{1/3}$                                                                       | $Re < 200$                                                                      | Cybulski and Werner (1986)         |
|                | $Nu = 4.65 \left( Re \times Pr \frac{D}{L} \right)^{1/3}$                                             | $Re < 2000$                                                                     |                                    |
|                | $Nu = 0.075 \cdot Re^{0.8} \cdot Pr^{0.4}$                                                            | $Re > 2000$                                                                     |                                    |
|                | $Nu = 6.1 \left( RePr \frac{D}{L} \right)^{1/3} \cdot \left( \frac{D}{L} \right)^{0.71}$              | $Re < 700$                                                                      | Joshi <i>et al.</i> (1995)         |
|                | $Nu = 0.1 Re^{0.8} Pr^{0.4} \left( \frac{D}{L} \right)^{0.71}$                                        | $700 < Re < 1000$                                                               |                                    |
|                | $Nu = 0.46 Re^{0.58} Pr^{0.4} \left( \frac{D}{L} \right)^{0.71}$                                      | $Re > 1000$                                                                     |                                    |
| Sulzer SMX     | $Nu = 0.078 Re^{0.8} Pr^{0.4}$                                                                        | Turbulent regime                                                                | Myers <i>et al.</i> (1997)         |
|                | $Nu = 3.55 Re^{0.4} Pr^{0.4} \left( \frac{D}{L} \right)^{0.25} \left( \frac{Pr}{Pr_w} \right)^{0.14}$ | Laminar regime $\{Pr\}_w$ is $\{Pr\}$ number calculated at the wall temperature | Li <i>et al.</i> (1996)            |
| Sulzer SMXL    | $Nu = 2.6 \cdot Re^{0.36} \cdot Pr^{0.36}$                                                            | $1000 < Re < 100,000$                                                           | Cybulski and Werner (1986)         |
| Sulzer SMV     | $Nu = 0.98 \cdot Re^{0.38} \cdot Pr^{0.38}$                                                           | $1000 < Re < 100,000$                                                           | Cybulski and Werner (1986)         |
|                | $Nu = a' \cdot (RePr)^{1/3} \quad 1.7 < a' < 3.6$                                                     | Laminar regime, constant temperature,<br>$K_h$ depends on $D/L$ and $d_h$       | Van der Meer and Hoogendorn (1978) |
|                | $Nu = 2 Re^{0.4} Pr^{0.4}$                                                                            | Turbulent regime                                                                | Cybulski and Werner (1986)         |

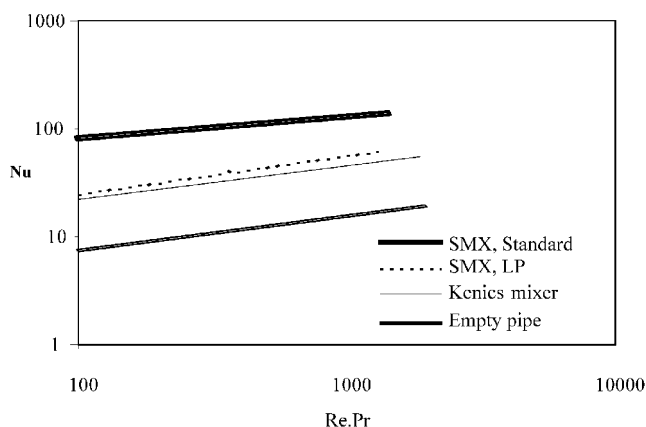


Figure 20. Nusselt numbers for commercial static mixers.

the turbulent regime (see Figure 21). Static mixers are highly effective during the transition between forced convection, single-phase flow and subcooled, nucleate boiling because heat transfer enhancement is high without an excessive increase in pressure drop. The heat flux per unit pressure drop can be improved by increasing the number of element at low *Re*.

The use of static mixers to improve heat transfer in chemical reactors was discussed above. Quantitative generalizations are difficult.

Visser *et al.* (1999) used CFD to predict pressures, velocities and temperatures in SMX elements. The calculated pressure drop agrees with experimental results. Calculated heat transfer coefficients for cooling through the wall were lower than experimental values. This is attributed to the fact that heat conduction in the plates was neglected. Simulations with infinite conductivity of the plates resulted in higher heat transfer coefficients than observed experimentally. CFD is a promising tool for optimizing heat transfer enhancement using static mixers.

### Key Parameters for Axial Mixing

One key parameter for an axial mixer is the first appearance time of an inert tracer. The device tested by Nauman

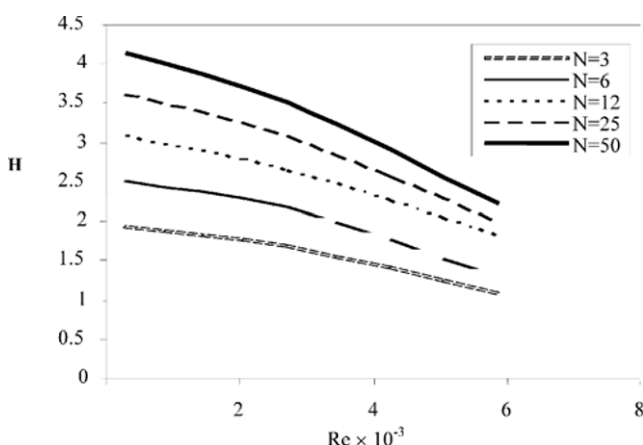


Figure 21. Key parameter *H* for heat transfer in condensation and evaporation processes.

(2002) gave  $t_{\text{first}}/\bar{t} < 0.1$ . A more important parameter is the damping ratio for inlet concentration fluctuations. The extent of damping depends on the period of the input signal, and can be estimated from the residence time distribution which in turn can either be measured experimentally or calculated from the design parameters. The four-zone mixer described by Nauman (2002) decreases the amplitude of sinusoidal disturbances by factors of 0.48, 0.74 and 0.89 when the input signals are sin waves with periods of  $0.5\bar{t}$ ,  $\bar{t}$  and  $2\bar{t}$ , respectively. These attenuations are very close to the values of 0.45, 0.71 and 0.89 that would be achieved in a CSTR.

### GUIDELINES FOR SELECTING A STATIC MIXER

Selecting a static mixer for a specific operation is not an easy task. Perhaps more difficult is deciding whether a static mixer should be used at all. Much of the literature on static mixers has been written by enthusiasts who advocate their use, sometimes to excess. There is no doubt that static mixers have been sold into applications where they do not belong. Some engineers have specified their use on the flimsy grounds that they will do no harm and might do some good. Having said that, we also say that sound design practice is to at least consider the use of static mixers whenever mechanical agitation appears necessary or when heat or mass transfer imposes limits on process performance.

There are relatively few applications that demand the use of static mixers. There is usually a traditional method for solving the same problem. Proper specification of a static mixer requires a comparison between the solution that uses conventional technology and the solution that uses a static mixer. Operating conditions that favor the use of motionless mixers include continuous flow with constant flow rates, the desire for a low residence time, and the desire to avoid back-mixing, but there are many exceptions to these generalizations. Comparisons must also be made within the class of motionless mixers because there are many commercial types, and once in a while, a special, new design is justified.

General guidelines for choosing a particular type of the static mixers can be summarized as follows:

- (i) Mixing inserts with blades and large sections available for flow (e.g. HEV, LPD, LLPD, Komax) are suitable for simple turbulent flow applications such as mixing, heat transfer or thermal homogenization. They can also be used for solids blending and dispersion when the fluid phase viscosity is low. These designs tend to remain relatively clean in operation and, by enhancing wall turbulence, may prevent plugging and fouling.
- (ii) An open design with helical elements is optimal for laminar flow, inline mixing and thermal homogenization (e.g. helical Kenics elements), as the helices redistribute and mix the flow without creating vortices.
- (iii) Static mixers with inserts of the corrugated plate type (e.g. the SMV mixer) are ideal for creating very homogeneous blends and for achieving co-current gas-liquid and liquid-liquid mass transfer because they induce an intense radial mixing. Corrugated-plate static mixers may also be used for heat transfer enhance.

- (iv) The multilayer design (e.g. SMX, SMXL) is used for highly viscous fluids in the laminar regime, but also seem suitable in multiphase liquid–liquid, liquid–solid and gas–liquid flows when the continuous phase is viscous or viscoelastic.
- (v) The Koch–Sulzer SMR is a specialized design for heat transfer during a polymerization.
- (vi) Specialized types of static mixers are optimal for packing and flow redistribution in distillation columns. They are generally superior to the general purpose mixers that are the primary topic of this paper. A similar remark can be made regarding inserts for heat exchanger tubes.

Figure 22 (adapted from Myers *et al.*, 1997) presents some of these guidelines in the form of a logic diagram. Note however, that contrary to the diagram, static mixers are used in batch and fed batch systems.

Once a particular type of static mixer has been selected, its more detailed evaluation is based on the following factors:

- (i) pressure drop and power requirements;
- (ii) specific key parameters for operation;
- (iii) miscellaneous practical considerations.

Predictions of pressure drop and the other key parameters are discussed above. The most important practical consideration is probably capital cost. Operating costs including utilities and maintenance may differ from those of the traditional solution. Process latitude, especially for turn-down, can be an important factor in evaluating a design. The ease of retrofitting can be important in process upgrades. The ease of cleaning or of quickly interchanging

units can be the dominant factor in facilities that make multiple products.

## SCALE-UP CONSIDERATIONS

There are three conceptually different ways of increasing the capacity of a production facility that uses motionless mixers:

- (i) add identical mixers in parallel—the shell-and-tube designs used for heat exchangers are common and inexpensive ways of increasing capacity;
- (ii) make the mixer longer—adding tube length and additional mixing inserts to a single-tube will increase capacity in approximate proportion to the increase in length;
- (iii) increase the tube diameter, either to maintain a constant pressure drop or to scale with (approximate) geometric similarity—geometric similarity for a tube means keeping the same length-to-diameter ratio,  $L/D$ , upon scale-up. Scaling with a constant pressure drop will lower the length-to-diameter ratio if the flow is turbulent.

The throughput scale-up factor,  $S$ , is defined as

$$S = \frac{\text{throughput for full-scale unit}}{\text{throughput of pilot unit}} \quad (62)$$

A scale-up with  $S=10$  is considered moderate while a scale-up with  $S=100$  is aggressive but reflects a modern trend made possible by our increasing understanding of process fundamentals.

## Scaling in Parallel

Scaling tube flows in parallel gives an exact duplication of process conditions. The number of tubes,  $N_{\text{tubes}}$ , increases in direct proportion to the desired increase in throughput:

$$\frac{(N_{\text{tubes}})_{\text{full-scale}}}{(N_{\text{tubes}})_{\text{pilot}}} = S \quad (63)$$

All other process variables including  $\bar{t}$ ,  $\Delta P$ ,  $Re$  and  $Nu$  are independent of  $S$ . The only worry about scaling static mixers in parallel is the possibility of tube-to-tube instabilities and maldistributions that can arise when the tubes are fed from a common pressure source. The viscous fingering problem already discussed can occur between tubes just as it can within a tube when there is a large change in viscosity due to reaction. Sound practice is to use three or more tubes in parallel in the pilot unit, but this may force undesirably high flow rates at the pilot scale. A publication by Hoftyzer and Zwietering (1961) discusses stability of a tubular polymerization reactor and provides some guidance, but there is still no satisfactory means for predicting tube-to-tube instabilities. Of course, using a separate pump for each tube will eliminate any concerns about tube-to-tube interactions.

Multiphase columns are often scaled by increasing the column diameter while keeping the same size of packing, the same superficial velocity, and the same column length or at least the same pressure drop. This form of scaling in parallel is generally satisfactory for gas–liquid contacting operations. It is assumed that heat transfer through the column walls is unimportant. Depending on the type of packing, the voidage near the wall may be higher than in the center of the column, but such wall effects can usually be

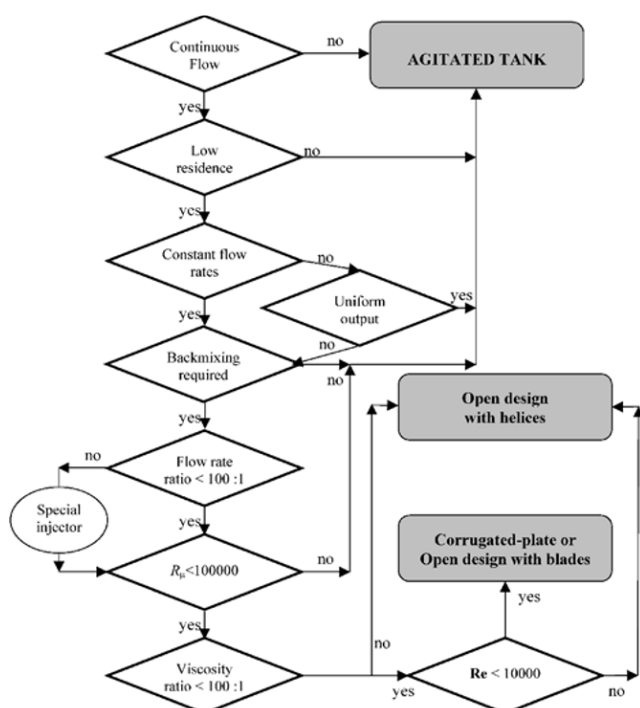


Figure 22. Logic diagram for static mixer selection (adapted from Myers *et al.*, 1997).

ignored. Flow distribution across the column can be a problem, and redistribution plates are usually needed at periodic axial positions in a large diameter column.

### Scaling in Series

Scaling in series is usually a conservative way of scaling when the process fluids are incompressible. The key limitation is, of course, pressure drop. The Reynolds number increases proportionately to the flow rate:

$$\frac{(\text{Re})_{\text{full-scale}}}{(\text{Re})_{\text{pilot}}} = S \quad (64)$$

as does the ratio of tube lengths and the number of mixing elements. Existing correlations show that the increase in **Re** will improve heat and mass transfer coefficients. The pressure drop will increase as  $S^2$  in the laminar regime and as  $S^{2.75}$  in fully turbulent flows. Depending on the process objectives, it might be possible to increase tube length somewhat less than proportionately to flow rate. Reactions, however, will require the same residence time and thus a proportionate increase in flow rate.

Other than the marked increase in pressured drop, the major concern with scaling static mixers in series is represented by the rather anomalous mixing behavior at intermediate Reynolds. As shown by the experimental results in Figure 16 and confirmed by CFD studies, mixing performance, as measured by the number of elements needed to achieve a given COV, can decrease with an increase in **Re**. However, the total number of elements will increase proportionately to flow rate, and this increase should be adequate to retain desirable mixing characteristics.

It is concluded that scaling in series will be satisfactory from the viewpoint of product quality in liquid and liquid-liquid systems. It is generally inappropriate for gas and gas-liquid systems due to the large increase in in-process inventory that will accompany the higher pressures needed for a series scale-up.

### Scaling with a Change in Diameter

The most common way of changing the tube diameter upon scale-up is scaling with geometric symmetry. The aspect ratios of the tube,  $L/D$ , and of the mixing elements,  $L_e/D_e$  are held constant upon scale-up. All linear dimensions such as  $L$  and  $D$  scale as  $S^{1/3}$  so that a factor of 2 increase in diameter gives a factor of 8 increase in throughput. Specifically,

$$\frac{L_{\text{full-scale}}}{L_{\text{pilot}}} = \frac{D_{\text{full-scale}}}{D_{\text{pilot}}} = S^{1/3} \quad (65)$$

The volume scales as  $S$  so that the in-process inventory also scales as  $S$  and the mean residence time is constant if the fluid is incompressible. The Reynolds number scales as  $S^{2/3}$ . For laminar flows with  $f$  given by equation (20), the pressure drop is constant upon scale-up. This means that a geometrically similar scale-up works as well for gas flows as for liquid flows.

There are several problems associated with geometrically similar scale-ups. The most important limitation is that the wall surface area scales as  $S^{2/3}$  so that this design has the common problem of surface area rising more slowly than

heat generation or heat exchanger duty. Thus, if  $S$  is large, geometrically similar scale-ups are limited to systems that are adiabatic or nearly adiabatic. A second problem is that the increase in **Re** may drive the system into the transition region with a consequent less in mixing efficiency. With a geometric similar scale-up, the number of mixing elements is constant, and thus the COV could rise unacceptably. A similar but smaller problem arises when the striation thickness model is used to characterize mixing. If the tube diameter is increased, more elements are needed to maintain the same quality product. The extra number of elements is given by

$$N_{\text{extra}} = \frac{\log S}{3 \log 2} \quad (66)$$

assuming the  $2^N$  model for striations. This is a slowly increasing function of  $S$ , e.g. a factor of 512 increase in throughput requires only three extra elements. Note that the  $\log 2$  term in equation (66) is replaced by  $\log 4$  for mixers governed by the  $4^N$  model.

A final problem associated with geometric scale-ups is due to the diminishing effects of molecular diffusion upon scale-up. The diffusion number, **Di**, will decrease with any form of scale-up that increases tube diameter while keeping a constant value for  $\dot{t}$ . Molecular diffusion has a generally beneficial effect on performance that will decrease on scale-up. The conservative approach is to use a large enough diameter tube that molecular diffusion is negligible in the pilot plant, e.g. **Di**  $< 8 \times 10^{-4}$ . Using a large-diameter tube has the added advantage of giving a low **Re** that will remain relatively low upon scale-up. The disadvantage of using a large diameter mixer in the pilot plants is that the number of mixing elements may be too small or the in-process inventory too large.

For geometrically similar scale-ups in highly turbulent flows inside tubes, with or without mixing inserts, the pressure drop will increase approximately as  $S^{1/2}$ :

$$\frac{\Delta P_{\text{full-scale}}}{\Delta P_{\text{pilot}}} = S^{1/2} \quad (67)$$

The extra pumping energy results in greater eddy diffusion, contributing to mixing and heat transfer, but not enough to overcome the relative loss in surface area. The reaction will become adiabatic if  $S$  is large.

Scale-ups with geometric similarity give equation (67) for both turbulent gas and liquid flows provided equation (65) is satisfied. However, the compressibility of the gas means that the in-process inventory of the gas will increase faster than  $S$ , and thus the mean residence time will increase upon scale-up. A scale-up using geometric similarity is possible for gas flows, but the exponent in equation (65) must be replaced by a value lower than  $1/3$  to reflect the change in average density between the pilot plant and full-scale plant. Constant-pressure scale-ups are generally preferred for gas flows, and these depart from geometric similarity.

In laminar flow, scaling with constant pressure drop is identical to scaling with geometric similarity. In turbulent flow, they are different. The following analysis assumes that the friction factors varies as **Re**<sup>-0.25</sup>. This is a good approximation for turbulent flows in empty tubes and in SMX mixers at high Reynolds numbers (Li *et al.*, 1997a). It is likely to be a reasonable approximation for other mixers

in the fully turbulent region. Under this assumption, a scale-up with constant  $\Delta P$  will result when

$$\frac{D_{\text{full-scale}}}{D_{\text{pilot}}} = S^{11/22} \quad \text{and} \quad \frac{L_{\text{full-scale}}}{L_{\text{pilot}}} = S^{5/22} \quad (68)$$

This form of scale-up gives volume and in-process inventory proportional to  $S$  so that  $\bar{t}$  is constant. The external surface area scales as  $S^{16/27}$  compared to  $S^{2/3}$  for geometric similarity. The Reynolds number also scales as  $S^{16/27}$ . The aspect ratio scales as  $S^{-6/22}$  and thus decreases upon scale-up. The aspect ratio of the mixing inserts will presumably not be changed upon scale-up so the number of inserts will decrease. As a numerical example, suppose a pilot-scale mixer contains 24 elements with  $L_e/D_e = 1.5$ . For a scale-up of  $S = 100$ , equation (68) gives a relative diameter of 10 and a length of 102. Assuming  $L_e/D_e$  is unchanged upon scaleup, only seven mixing elements will fit into the large tube. This result typifies the difficulty in scale-up for gas phase static mixers.

### WHAT THE FUTURE HOLDS

Static mixers manipulate flow fields by cleverly designed geometries of bounding surfaces. Until the 1980s, the clever designs were based on physical insight and intuition. These designs became the first generation of static mixers. They were immensely successful, stimulating a host of industry applications and academic investigations, not all of which have survived the test of time. Most of the current review paper has dealt with these first generation motionless mixers.

The second generation retains the conceptual designs of the first generation but refines the geometries and their applications. The emergence of computational fluid dynamics has allowed this to occur at a much faster rate than would have been possible using physical (as opposed to numerical) experiments. For the specific example of helices of the Kenics type: what is the optimal angle of twist for an element; what  $L_e/D_e$  ratio should be used to achieve this twist, what should be the offset angle for the next element in series, and should a sequence of elements all have the same geometric parameters or should they be varied as a function of position in the sequence? Some of these questions have already been answered. All are feasible to answer, at least for a specific application in single-phase, laminar flow, given modern CFD codes and computers. We anticipate that the answers will be forthcoming both in the open literature and as vendor-supplied responses to customer queries.

The third generation of motionless mixers will use CFD to explore new conceptual designs, and particularly designs that are optimal, in some sense, for a given application. Consider for example heat transfer to a fluid in laminar flow in a tube of some fixed length. Suppose that one motionless insert can be installed within the tube. What inlet-to-outlet transformation should the insert perform in order to maximize heat transfer? Complete mixing within the cross-section of the tube is not the right answer. It has been conjectured but not proven that the optimal transformation is complete flow inversion (Nauman, 1979). The proof can presumably be answered using functional optimization, but the answer will beg the question as to what specific mixer geometry will best approach the optimal transformation

subject to a constraint on pressure drop. A more general question begins with a tube of diameter  $D$  and length  $L$  and asks what internal alterations should be made to optimize heat transfer given a maximum allowable pressure drop. Similar questions can be asked for thermal or compositional homogenization. For example, suppose the in-process inventory and the pressure drop are limited to some maximum values. Subject to these constraints, what static geometry will minimize the (maximum) striation thickness? It is clear from the results in Figure 16 that motionless mixers do not perform as well as predicted by the simple  $2^{-N}$  or  $4^{-N}$  models for striation thickness even when blending physically identical fluids since  $N = 30$  should reduce the striation thickness to molecular dimensions. There is still substantial room for improvements both in mixing characterization techniques and physical designs. Given a specific application, the third generation of static mixers will represent global design optimization rather than local optimization of first generation designs.

Computer speeds continue to increase and memory sizes grow. The hardware has or soon will become powerful enough to simulate third generation designs for simple applications such as striation thick minimization and heat transfer. In the near future, software improvements will give CFD codes that can treat heat transfer and molecular or eddy diffusion with sufficient accuracy for reaction engineering calculations. Longer term, the codes will become useful for multiphase flows, and these improvements will have tremendous impact on the design and utilization of motionless mixers both as self-contained devices and as packings and flow distributors in columns and beds.

The basic concept of motionless mixers—that is, of manipulating flow fields through cleverly designed geometries of bounding surfaces—will find applications beyond those traditionally associated with motionless mixers. For example, a recent design promotes axial rather than radial mixing by approximating the residence time distribution of a CSTR (Nauman *et al.*, 2002). For two-phase flows, filters with random configurations of fibers or particles have long been used to promote coalescence. Operational improvements should be possible with more structured designs. These and other two-phase contacting devices will also benefit from using non-metallic or coated elements that have favorable fluid-to-surface interactions.

As a general conclusion, computational fluid dynamics has already become an essential tool for understanding static mixer performance. Experiments cannot provide the precise, local measurements that are needed to understand the effects of complex geometries. Clever people will continue to be needed to pose the right questions, and experiments will be needed to confirm truly novel designs and applications. However, the future truly lies with CFD. Current codes are complicated both in terms of their mathematical formulation and the user interface. The trend for some years has been to finite elements and finite volume codes; but as mentioned above, these techniques are not particularly well suited to solving convective diffusion problems. As hardware becomes faster, there may be movement back to finite difference codes due to their inherent simplicity for large sets of coupled, partial differential equations for which material and energy balance closure is important. There will certainly be movement toward massively parallel computing systems.

## NOMENCLATURE

|                          |                                                                                     |
|--------------------------|-------------------------------------------------------------------------------------|
| $a$                      | interfacial area per unit length of mixer, $\text{m}^{-1}$                          |
| $a', a_1, a_2, a_3, a_4$ | constants                                                                           |
| $A_q$                    | heat-transfer area, $\text{m}^2$                                                    |
| $b$                      | constant                                                                            |
| $b'$                     | mixing constant                                                                     |
| $B$                      | width of an ellipsoidal drop, $\text{m}$                                            |
| $Br$                     | Brinkman number                                                                     |
| $\bar{c}_1$              | average volumetric concentration of species $i$ , $\text{mol m}^{-3}$               |
| $c_{ij}$                 | volumetric concentration of species $i$ at sampling point $j$ , $\text{mol m}^{-3}$ |
| $c_p$                    | specific heat at constant pressure, $\text{J kg}^{-1} \text{K}^{-1}$                |
| $c'$                     | constant                                                                            |
| $C_1, C_2, C_t$          | constants                                                                           |
| $C, C', C''$             | constants                                                                           |
| $C_{\text{pm}}$          | mean specific heat of vapor–condensate mixture, $\text{J kg}^{-1} \text{K}^{-1}$    |
| <b>Ca</b>                | capillary number                                                                    |
| COV                      | coefficient of variation                                                            |
| $d$                      | diameter of dispersed phase, $\text{m}$                                             |
| $d'$                     | constant                                                                            |
| $d_{\text{in}}$          | inside diameter of condensate tube, $\text{m}$                                      |
| $d_h$                    | Hydraulic diameter, $\text{m}$                                                      |
| $d_p$                    | diameter of solid particle, $\text{m}$                                              |
| $d_{32}$                 | Sauter mean diameter of bubble or drop, $\text{m}$                                  |
| <b>D</b>                 | molecular diffusivity, $\text{m}^2 \text{s}^{-1}$                                   |
| $D$                      | diameter of column, mixer or pipe, $\text{m}$                                       |
| $D_{\text{ax}}$          | axial dispersion coefficient, $\text{m s}^{-2}$                                     |
| $D_e$                    | diameter of element, $\text{m}$                                                     |
| $D_i$                    | diffusion number                                                                    |
| $D_v$                    | molecular diffusivity, $\text{m}^2 \text{s}^{-1}$                                   |
| $DF$                     | bubble or drop deformation                                                          |
| $f$                      | fanning friction factor with inserts                                                |
| $f_{\text{empty}}$       | fanning friction factor in an empty pipe                                            |
| $f(t)$                   | residence time distribution                                                         |
| $f_v$                    | friction factor of a viscoelastic fluid                                             |
| $F$                      | cumulative residence time distribution function                                     |
| $g$                      | acceleration due to gravity, $\text{m s}^{-2}$                                      |
| $h$                      | heat transfer coefficient, $\text{W m}^{-2} \text{K}^{-1}$                          |
| $h_i$                    | inside heat transfer coefficient, $\text{W m}^{-2} \text{K}^{-1}$                   |
| $H$                      | key parameter for heat transfer estimation                                          |
| $H_{\text{fg}}$          | latent heat, $\text{J kg}^{-1}$                                                     |
| $H_{\text{sp}}$          | sensible heat, $\text{J kg}^{-1}$                                                   |
| $HTU$                    | height of a transfer unit, $\text{m}$                                               |
| $k$                      | thermal conductivity, $\text{W m}^{-1} \text{K}^{-1}$                               |
| $k'$                     | consistency index of a power-law fluid, $\text{Pa s}^n$                             |
| $K_D$                    | liquid–liquid mass transfer coefficient, $\text{m s}^{-1}$                          |
| $K_L$                    | gas–liquid mass transfer coefficient, $\text{m s}^{-1}$                             |
| $K_p$                    | power constant                                                                      |
| $K_S$                    | liquid–solid or gas–solid mass transfer coefficient, $\text{m s}^{-1}$              |
| $L$                      | length of column, mixer or pipe, $\text{m}$                                         |
| $L_e$                    | length of one element, $\text{m}$                                                   |
| $L_p$                    | length of an ellipsoidal drop, $\text{m}$                                           |
| $m, m'$                  | constants                                                                           |
| $n$                      | flow index for power-law fluids                                                     |
| $n'$                     | number of sampling points                                                           |
| $n_c$                    | number of channels of a mixing element                                              |
| $N$                      | number of mixing elements                                                           |
| $N_{\text{extra}}$       | number of extra tubes                                                               |
| $N_s$                    | rotation speed, $\text{s}^{-1}$                                                     |
| $N_{\text{tubes}}$       | number of tubes in parallel                                                         |
| <b>Ne</b>                | Newton number                                                                       |
| <b>Nu</b>                | Nusselt number                                                                      |
| <b>Nu<sub>0</sub></b>    | Nusselt number due to conductive effects                                            |
| $p$                      | viscoelastic flow index                                                             |
| $P$                      | pressure, $\text{Pa}$                                                               |
| $\Delta P$               | pressure drop, $\text{Pa}$                                                          |
| $P_{\text{impeller}}$    | impeller power, $\text{W}$                                                          |
| <b>Pe</b>                | Peclet number                                                                       |
| <b>Pr</b>                | Prandtl number                                                                      |
| $q$                      | constant                                                                            |
| $Q$                      | volumetric flow rate, $\text{m}^3 \text{s}^{-1}$                                    |
| $Q_{\text{heat}}$        | heat transfer rate, $\text{W}$                                                      |
| <b>Re</b>                | Reynolds number                                                                     |

|                       |                                                                   |
|-----------------------|-------------------------------------------------------------------|
| <b>Re<sub>h</sub></b> | modified Reynolds number                                          |
| <b>Re<sub>g</sub></b> | generalized Re number                                             |
| $R_\mu$               | ratio of dispersed to continuous phase viscosity                  |
| $R_\rho$              | ratio of dispersed to continuous phase density                    |
| <b>RSD</b>            | relative standard deviation                                       |
| <b>s</b>              | velocity gradient tensor, $\text{s}^{-1}$                         |
| $S$                   | scale-up factor for throughput                                    |
| $S_1$                 | standard deviation                                                |
| $S, S_0$              | stirration thickness and initial stirration thickness, $\text{m}$ |
| <b>Sc</b>             | Schmidt number                                                    |
| <b>Sh</b>             | Sherwood number                                                   |
| <b>Sh<sub>0</sub></b> | Sherwood number in an empty pipe                                  |
| $t$                   | residence time, $\text{s}$                                        |
| $\bar{t}$             | mean residence time, $\text{s}$                                   |
| $t^+$                 | dimensionless time                                                |
| $t_{\text{first}}$    | first appearance time in residence time distribution, $\text{s}$  |
| $T$                   | temperature, $\text{K}$                                           |
| $T_b$                 | bulk temperature, $\text{K}$                                      |
| $T_w$                 | temperature of heat transfer surface, $\text{K}$                  |
| $u$                   | fluid velocity in an empty pipe, $\text{m s}^{-1}$                |
| $u_G$                 | superficial gas velocity, $\text{m s}^{-1}$                       |
| $u_L$                 | superficial liquid velocity, $\text{m s}^{-1}$                    |
| $V$                   | velocity vector, $\text{m s}^{-1}$                                |
| $Vi$                  | viscosity group                                                   |
| <b>We</b>             | Weber number                                                      |
| $x'$                  | mass vapor quality                                                |
| $x_i$                 | mass fraction                                                     |
| $X$                   | pressure ratio for gas–liquid pressure drop estimation            |
| $z$                   | constant                                                          |
| $Z$                   | pressure drop factor                                              |
| $Z'$                  | mixing coefficient, $\text{m}^{-1}$                               |

## Greek symbols

|                                |                                                                             |
|--------------------------------|-----------------------------------------------------------------------------|
| $\alpha$                       | extensional efficiency                                                      |
| $\underline{\hat{\gamma}}$     | rate of deformation tensor, $\text{s}^{-1}$                                 |
| $\dot{\gamma}$                 | mean shear rate, $\text{s}^{-1}$                                            |
| $\dot{\gamma}_w$               | wall shear rate, $\text{s}^{-1}$                                            |
| $\varepsilon$                  | mixer void fraction                                                         |
| $\varepsilon'$                 | energy dissipation per unit mass, $\text{W kg}^{-1}$                        |
| $\varepsilon_1, \varepsilon_2$ | energy dissipation in boundary and core regions, $\text{W kg}^{-1}$         |
| $\varepsilon_G$                | gas hold-up                                                                 |
| $\varepsilon_P$                | bed expansion in fluidized bed reactors                                     |
| $\zeta$                        | composition parameter for liquid–liquid dispersion                          |
| $\eta$                         | Kolmogoroff length scale, $\text{m}$                                        |
| $\theta$                       | twist angle, $\text{deg}$                                                   |
| $\phi$                         | ratio of volumetric flow rates in multiphase flows                          |
| $\phi_G, \phi_L$               | correction factors for Lockhart and Martinelli's method                     |
| $\mu$                          | viscosity, $\text{Pa s}$                                                    |
| $\mu^*$                        | viscosity based on $\dot{\gamma}_w$ for non-Newtonian fluids, $\text{Pa s}$ |
| $\mu_{\text{bulk}}$            | viscosity of the bulk fluid, $\text{Pa s}$                                  |
| $\mu_{L/L}$                    | apparent viscosity of a liquid–liquid dispersion, $\text{Pa s}$             |
| $\mu_m$                        | mean dynamic viscosity of the vapor–condensate mixture, $\text{Pa s}$       |
| $\mu_{\text{wall}}$            | viscosity of fluid at the wall (effect of temperature), $\text{Pa s}$       |
| $\nu$                          | power dissipation per unit mass, $\text{W kg}^{-3}$                         |
| $\rho$                         | density, $\text{kg m}^{-3}$                                                 |
| $\rho_L$                       | mean density of condensate, $\text{kg m}^{-3}$                              |
| $\rho_{L/L}$                   | apparent density of a liquid/liquid dispersion, $\text{kg m}^{-3}$          |
| $\rho_m$                       | mean density of the vapor–condensate mixture, $\text{kg m}^{-3}$            |
| $\sigma^2$                     | variance of residence time distribution, $\text{s}^2$                       |
| $\sigma$                       | surface tension or interfacial tension, $\text{N m}^{-1}$                   |
| $\sigma'$                      | cohesion forces in agglomerates, $\text{N m}^{-1}$                          |
| $\sigma_0$                     | initial standard deviation                                                  |
| $\tau$                         | shear stress, $\text{Pa}$                                                   |
| $\tau_{11}, \tau_{22}$         | normal stress for viscoelastic fluids, $\text{Pa}$                          |
| $\varphi$                      | volume fraction of the dispersed phase                                      |
| $\omega$                       | vorticity tensor, $\text{s}^{-1}$                                           |

Subscripts

|         |                  |
|---------|------------------|
| c       | continuous phase |
| d       | dispersed phase  |
| e       | element          |
| G       | gas              |
| L       | liquid           |
| m       | mixer            |
| o       | without mixer    |
| s       | solid            |
| 1, 2, i | components       |

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