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Effect of Shear Produced by Pipe Fittings on the Drop Size Distributions in Turbulent Flow of Kerosene/Water Mixtures

تأثير الجهد المنبعث من قطع تركيب الأنابيب على توزيع قطرات مزيج من الماء والبترول في حالة تدفق مضطرب

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Abstract

Drop size distribution data for kerosene-water dispersion were obtained in I."I.D. pipe at a range of velocities in turbulent flow for a straight horizontal pipe. U shaped pipe and an offset pipe fitting oriented horizontally and vertically (upward and downward) to the main flow. A Lightnin in line static mixer was used as a premixer and the drop size distribution was measured by a Malvern 2600 analyzer. By changing the number of internal elements from 4 to 18 the mixer produced a primary dispersion with the mean drop sizes in the range of 50-700 μm for the flow rates of 20 to 84 l/minute. The Sauter mean diameter, d_{32}, was found to decrease as the number of elements was increased until an equilibrium drop size was reached. This equilibrium drop size varied with the fluid velocity through the mixer. For a dispersion of ~0.5% kerosene in water, the correlation of drop size with energy dissipation rate, e, was found to give a reasonable agreement with Kolmogoroff's theory with an exponent in the range of -0.47 to -0.56 for a horizontal pipe and -0.60 to -0.72 for U-shaped and offset pipe fittings. The Sauter mean diameter was also correlated against Weber number with an exponent in the range of -0.71 to -0.83 for all the fittings used.

Keywords: Sauter mean diameter, dispersion, mixing, on-line measurement, oil-water, fittings.

تم في هذا البحث جمع نتائج من عملية مزج قطرات من مادة الكاز (مادة بترولية) مع الماء داخل أنابيب حديدية ذات قطر داخلي ١" في حالة تدفق مضطرب. كانت هذه الأنابيب ذات أطوال وأشكال مختلفة من مستقيم أو حرف U ، أو على شكل أنابيب ذات وصلة جانبية موصولة عمودياً أو افقياً بالنسبة لمستوى التدفق قبل دخولهم الى قطع الأنابيب. وقد تم استخدام خلاط ثابت من نوع Lightnin لعملية مزج الماء والبترول وتمت عملية قياس حجم القطرات وتوزيعها الطبيعي عن طريق جهاز Malvern 2600. وكانت من أغراض البحث الحصول على أحجام مختلفة من القطرات وتم ذلك بتغير عدد القطع الموجودة في الخلاط الثابت من ٤ الى ١٨ قطعة حيث تم الحصول على حجم قطرات تتراوح ما بين ٥٠-٧٠٠ مايكروميتر عندما كان تغير التدفق من ٢٠-٨٤ لتر/ الدقيقة.

وقد وجد أن حجم هذه القطرات المتشتته تقل مع زيادة عدد القطع في الخلاط الثابت الى أن تصل الى حالتي اتزان ما بين عدد القطرات المتكسرة وعدد القطرات الملتحمة وكانت حالة الاتزان هذه تختلف باختلاف سرعة المزيج وباختلاف تركيز المادة المتشتتة.

وباستخدام تركيز منخفض من الكاز (٠,٥) كمادة متشتته في الماء فقد تم ربط علاقة ما بين حجم القطرات وكمية الطاقة المتشتته في تكوين هذه القطرات، ε ، مرفوعة الى قوة تتراوح ما بين ٠,٤٧- و ٠,٧٢- وقد تم كذلك ربط علاقة أسية ما بين حجم القطرات ورقم Weber الذي يمثل النسبة ما بين قوة تكسير القطرات/ قوة لزوجة هذه القطرات لجمع الأنابيب المستعملة وهذه النتائج متوافقة لحد ما مع نظرية كمولوغروف.

Introduction

The detailed design of the removal of water droplets from oil or from any kind of liquid-liquid separation requires knowledge of dispersion properties such as the droplet size distribution, dispersed phase concentration and the physical properties of the system. However, the value of these parameters in a section of a pipe does not necessarily represent the dispersion properties in the whole system. Fluids passing through pipes and fittings will experience various forms of turbulence which will have a great influence on these properties. Hence it is essential to determine the effect of these fittings in order to gain a better understanding of the dispersion conditions at a specific position in a pipe or plant.

The major aims of this work were:

1. To produce data on the effect of the shear through fittings on the dispersion properties in a heterogenous mixture with preconditioning mixer.
2. to produce data on drop size distributions for use in the design of oil-water separators.
3. to produce a correlation of the dependence of d_{32} on the energy dissipation rate and Weber number for a kerosene/water system.

Drop Breakup in Turbulent Pipe flow

When two liquid phases are mixed, droplets are formed by 'break-up' of the dispersed phase in the shear field, while simultaneously, in other parts of the flow, droplets are coalescing.

Kolmogoroff's theory (1949) on isotropic turbulence was pioneering in explaining this drop break-up. The theory was found to yield the following expression, Hinze[1], for maximum stable particle size of the dispersed phase.

$$d_{max} = k \frac{\sigma^{0.6}}{\rho^{0.6} \varepsilon^{0.4}} \qquad (1)$$

where the constant k must be determined from experimental data. The determination of the maximum drop diameter from the drop size distribution is somewhat ambiguous, if not difficult, as the largest drop size in a distribution is hardly reproducible. It is common practise to use a cut off size in the cumulative drop size distribution as a measure of the maximum size. Thus, Hinze[1] defined d_{95}, the diameter below which 95% of the cumulative drop volume was confined as the maximum drop size and reported a value of k=0.725 in a Couette flow field.

The assumption made is that the smaller eddies produced by the dissipative process are statistically independent, in size, from the primary eddies and that they are isotropic. A lot of experimental work, (Mlynek and Resnick[2], Shinnar[3] and Sprow[4]), has been done and a satisfactory agreement has been achieved between the experimental data and equation (1) with discrepancy only in the value of the constant k for agitated vessels.

Collins and Knudsen[5] have published data on drop size distribution of an oil-water dispersion in well defined turbulent pipe flow. They developed a stochastic mathematical model in which d_{max} is a basic input parameter which could predict both the shape of the observed distributions and the kinetics of the droplet breakup process for the distribution produced by their experimental turbulent flow field.

Sleicher[6] *and Paul and Sleicher*[7] have reported experimental data for the maximum stable drop size for two immiscible liquids of different physical properties such as viscosity in the range of 0.5 to 32cp, density from 700 to 1585 kg/m^3 and interfacial tension from 8 to 45 dyne/cm. flowing in a pipe. The dispersed phase volume fraction was very small (less than 1%) so that the coalescence process was negligible. Hughmark[8] used

the data from Sleicher[6] and Paul and Sleicher[7] and produce a correlation to calculate the maximum drop size for two pipe sizes (0.5 and 1.5" I.D).

Karabelas[9] carried out experimental work with water dispersed in hydrocarbon at various flow rates to measure the drop size distribution in well-mixed dilute liquid-liquid dispersion (of a maximum of 0.26% water volume concentration) across the vertical profile of horizontal pipeline using photographic and droplet encapsulation techniques to measure the maximum stable drop size. where N_{We} is the Weber number, which is the ratio of the viscous force in surface tension force defined by

$$N_{W_e} = \frac{\rho V^2 D}{\sigma} \qquad (2)$$

In systems where mass transfer is important it is useful to have a knowledge of the interfacial area, a, given by

$$a = \frac{6\phi}{d_{32}} \qquad (3)$$

Therefore the Sauter mean diameter, d_{32}, has been used to replace d_{max} in equation (1), which reconciled the discrepancies among the findings of various investigators. Equation 1 can then be written as

$$d_{32} = k_1 \frac{\sigma^{0.6}}{\rho^{0.6} \varepsilon^{0.4}} \qquad (4)$$

This equation is valid for low dispersed phase hold up. A number of workers, Mersmann & Grossmann[10], have proposed equations of a similar form based on experimental results,.

Hanzevack and Demetriou[11] studied the effect of flow velocity and pipeline configuration of 1% water-kerosene dispersion in turbulent flow using laser image processing. Studying the concentration profile in a horizontal 8.2cm I.D. pipe, they claimed that the transition from stratified to adequately dispersed flow occurred at about 2.3 m/s velocity.

Middleman[12] correlated the Sauter mean diameter as a function of Weber number and Reynolds number in the following equation

$$\frac{d_{32}}{D_h} = k_2 N_{We}^{-0.6} N_{\mathrm{Re}}^{0.1} \qquad (5)$$

Matsumura et al[13] reported a variation of the exponent of Weber number in equation 6 to be between -0.57 to -0.67 and that the value of k_2 is dependent on the viscosity of the emulsion.

Experimental Procedure

A schematic diagram of the flow loop used in the experimental work is shown in (Figure 1). The two liquid phases were stored in polypropylene tanks of $2m^3$ volume (a small storage tank of 20 litre capacity was also linked to the rig for use at low concentration of oil phase) from which the fluids were pumped, in the required ratios, at rates of up to $8m^3/h$. to the rig's tessection. The two streams were joined at the inlet of a "Lightnin" in-line static mixer. The purpose of this device was to precondition the feed dispersion. The flow rate of the continuous phase was 20-84 l/min. (corresponding to 0.65-2.76 m/s superficial velocity through 1" I.D pipe fitting giving Reynolds numbers for water continuous dispersions up to $7*10^4$) and regulated by a control valve. The kerosene/water volumetric ratio was maintained at 1:200 through the entire set of experiments. The mixture was then fed into a fibre bed coalescer where it was separated into two constituent phases which were then recycled to the feed tanks.

The continuous flow rate, temperature and pressure across the fittings were logged by a computer. Quantitative information on the flow rate and the pressure drop across the fittings was required to calculate the energy dissipated through the fitting.

The drop size distributions were measured using a laser diffraction technique, Stewart et al [14]. The analysis of the data collected is based on Faunhofer diffraction theory (Malvern[15]). Experiments have been carried out to obtain drop size distribution profiles for different pipes geometries.

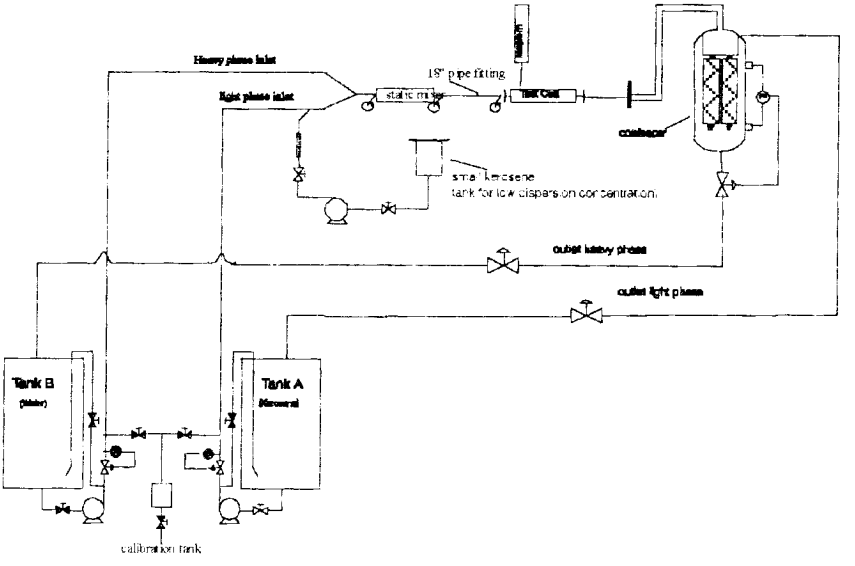

Figure (1): Experimental rig in the Pilot Plant.

Fittings Used

The geometries studies can be divided into four cases:

Case one: static mixer placed directly before the test cell. In this case the number of elements in the static mixer was varied from 0 to 18.

Case two: horizontal pipe of length 18d (d=1") fitted after the static mixer.

Case three: two separate "U" shape pipe fittings of 42d and 18d total length oriented horizontally and vertically (upward and downward) to the main flow, positioned after the static mixer.

Case four: an offset pipe fitting of 18d total length and including two normal radius elbows.

This was positioned after the static mixer.

In cases two to four the static mixer was fitted with 18 elements.

The geometry of each of these cases is illustrated in Figure 2.

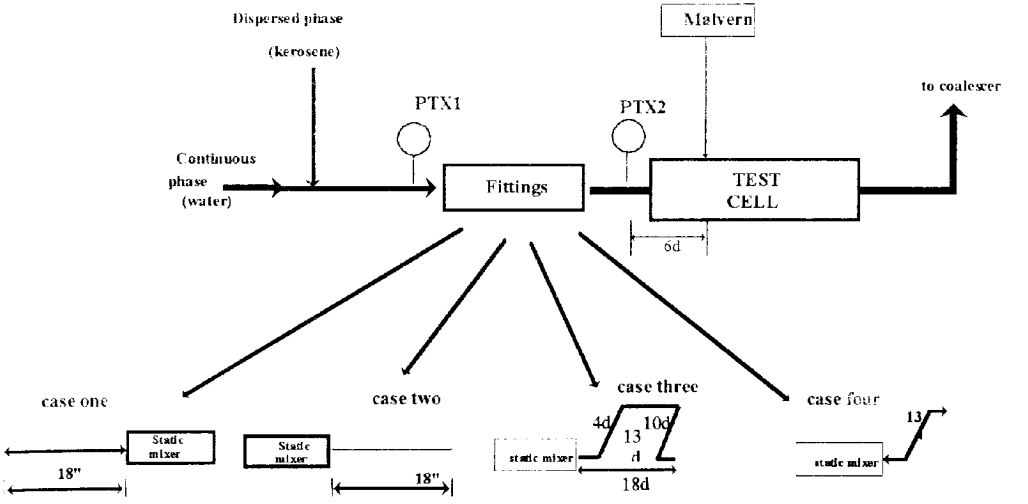

Figure (2): Dispersion rig with all the fittings used.

Results and Discussion

The drop size distribution in a pipe or fitting may be related to the shear of the system. Previous workers, Collins and Knudsen[5], have based correlations of drop size on Kolmogoroff's theory which relates the maximum stable drop size, d_{max}, to the energy dissipation rate, e. In this work the Sauter mean diameter, d_{32}, was correlated with energy dissipation rate, e, for all of the cases described above.

For different oil-water systems the density and interfacial tension will change. According to Kolmogoroff's theory, drop size distribution will also be a function of these properties. Therefore, results from this work have also been correlated with Weber number, N_{We}, defined by equation 2 which is dependent on the physical properties of the system. The correlation is particularly useful for systems where chemical additives were present resulting in a significant decrease in interfacial tension. The effect of such additives on the characteristics of kerosene-water dispersions is reported elsewhere, (Stewart et al [14]).

The shear is also a function of Reynolds number

$$N_{R_e} = \frac{\rho v D}{\mu} \qquad (6)$$

In this work, only one oil-water system and a constant pipe diameter (1") was used, therefore velocity is the only variable in the above equation.

If a fluid of volumetric flow rate Q passes through a fitting of length L and diameter D then the residence time t is given by

$$\tau = \frac{\pi D^2 L}{4Q} = \frac{L}{V} \qquad (7)$$

where V is the velocity of the fluid passing through the mixer of diameter D. The rate of energy dissipation per unit mass contained in the mixer is

$$\varepsilon = \frac{4Q\Delta P}{\pi D^2 L \rho_c} = \frac{V \Delta P}{L \rho_c} \qquad (8)$$

and the energy dissipated per unit mass of the passing fluid through the mixer is

$$E = \tau\varepsilon = \frac{\Delta P}{\rho_c} \qquad (9)$$

Effect of Number of Elements in the Static Mixer on the Equilibrium value of d_{32}.

The static mixer used was flexible in the fact it was possible to change the number of elements and alter their orientation depending on the dispersion required and the state of the fluid stream (laminar or turbulent). The Reynolds number was calculated for all experimental conditions and it was varied from 16500 to 70,000 and it was found that the flow was in the turbulent regime. Therefore the mixer elements were arranged as recommended by the manufacture for this flow regime as shown in Figure 3.

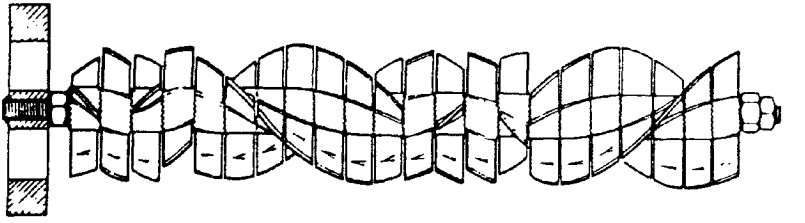

Figure (3): In-line mixing elements

A set of experiments was carried out to obtain a profile of the change of droplet size distribution with the number of static mixer elements (static mixer length) for case one described above. The number of elements was varied from 4 to 18. For the same number of elements, the higher the velocity the more the shear produced and the lower the d_{32}. As the fluid mixture passes through a field of constant turbulence intensity, the dispersed phase elements disintegrate into finer droplets. This process of break-up and coalescence of the dispersed phase continues as long as the turbulence is maintained. The time scale of the equilibrium rate of the dispersion and coalescence processes is still uncertain. The equilibrium rate has a special relevance in operations in which the residence time is short and is of crucial importance in static mixers. The position of this equilibrium is controlled by the turbulence intensity and hence the value of e for a liquid-liquid system. Figure (4) shows the equilibrium Sauter mean diameter, d_{32}, as a function of the flow velocity through the mixer. For a constant velocity, the residence time increases as the number of elements increases. This results in decreasing d_{32} up to a point where a dynamic equilibrium is achieved. At this point the rate of coalescence is equal to the rate of dispersion resulting in a constant average drop size at each velocity, 400μm at 0.75 m/s and 70mm at 2.5 m/s. The presence of mixer elements greater than the number required to achieve equilibrium, serves merely to maintain that equilibrium.

The rate at which drops are dispersed, the residence time required to achieve equilibrium, as well as the equilibrium value of d_{32} were found to be strongly dependant on the rate of energy dissipation per unit of mixer volume. Higher fluid velocities induce large values of E (as a result of higher pressure drop across the static mixer) which in turn result in faster

disintegration, finer equilibrium dispersions and shorter residence time requirements to obtain equilibrium dispersions.

The effect of energy dissipation rate on Sauter mean diameter for different static mixer elements is shown in Figure (5). The logarithmic dependence of d_{32} as predicted by Kolmogoroff's theory (equation 5) is demonstrated in this figure.

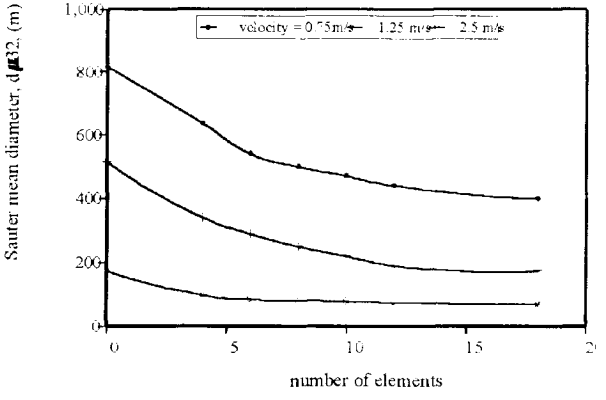

Figure (4): Kinetics of drop dispersion

Comparison of Present Work with Published Data

Kolmogoroff's theory predicts an exponent of -0.4 to the energy dissipation rate and -0.6 for N_{we} whereas the present experimental data has an exponent in the range of -0.47 to -0.56 for energy dissipation rate and -0.71 to -0.83 for Weber number. This discrepancy is most likely due to the departure from the homogenous, isotropic turbulence structure assumed in the theoretical development and is similar to that found from the experimental work of AL-Taweel and Walker[16]. Their data produced exponents of -0.6 and -0.75 for energy dissipation rate and Weber number respectively. The data obtained in this work was for 0.5% dispersed phase concentration whereas the work presented by AL-Taweel and Walker[16] was for 1% phase concentration. Figure (6) shows that the equilibrium value of d_{32} is effected by the phase ratio. The higher the phase ratio the larger the value of d_{32}.

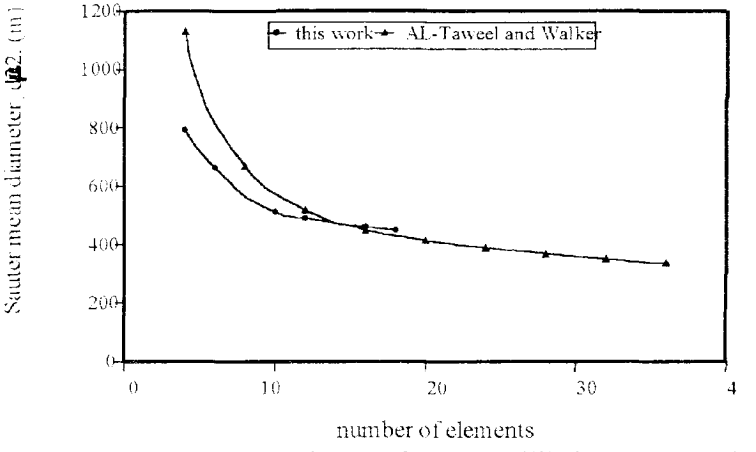

Figure (5): Comparison of the equilibrium value of d_{32} with published data.

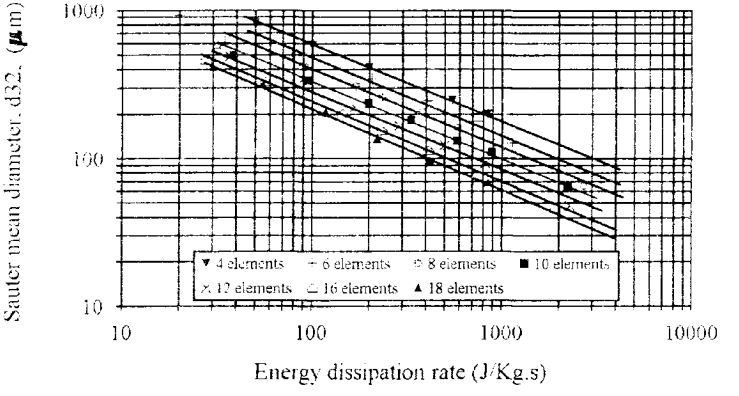

Figure 6: Effect of Energy dissipation rate on sauter mean diameter

Chen and Libby[17] reported that using the Sulzer static mixer for an oil-water system, the following correlation could be used to predict the Sauter mean diameter, d_{32}

$$\frac{d_{32}}{D} = 1.14 N_{We}^{-0.75} \left(\frac{\mu_d}{\mu_c} \right)^{0.18} \qquad (10)$$

When equation 10 is used to predict the d_{32} as a function of Weber number for the kerosene-water system, good agreement between our results and this prediction as shown in Figure (7). The discrepancy may be due to the value of the constant (k=1.14) in the correlation. From the present data the exponent on the Weber number was in the range -0.71 to -0.83 which is lower than that predicted by the widely used theory of drop break up, Kolmogoroff's theory, in the inertial sub-range. Recently Baldyga and Bourne[18-19] and Bourne[20] explained that this exponent could be smaller than -0.6 down to a value of -0.93 due to the fact that the theory ignores intermittency and employs a time averaged energy dissipation rate. However this and other turbulence characteristics are more accurately represented by a distribution which Baldyga and Bourne[18] predicted using multifractals.

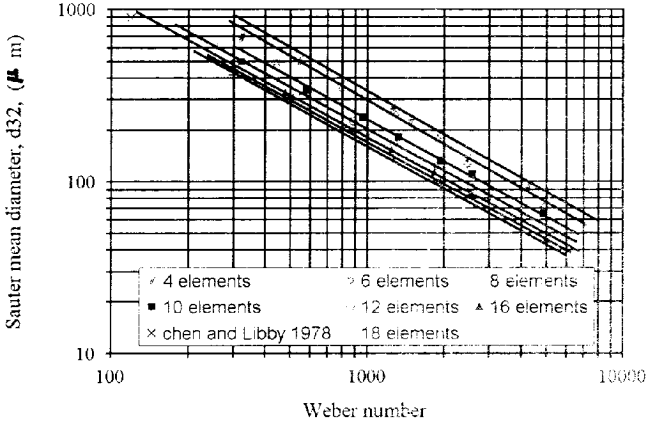

Figure (7): Sauter mean diameter as a function of Weber number.

Although mixing elements such as bends and pipeline configurations are often counted on to provide sufficient turbulent mixing to insure representative sampling, they have not been studied thoroughly. The only work reported in the open literature in which the effect of fittings on drop size distributions for liquid-liquid dispersion has been studied is that of Hanzevack and Demetriou[11]. They measured the effect of velocity and pipe

line configuration on the maximum drop size (d_{99}) and concentration profile for a 1% water-kerosene dispersion. No comparison could be made between the results and those of Hanzevack and Demetriou[11] for several reasons; firstly their continuous phase was kerosene, whereas water was the continuous phase in this work. Secondly, the pipe diameter in their work is 8.2 cm whereas in ours it was 2.5cm. Thirdly, they did not use a premixer such as a static mixer and finally, a different drop size distribution measurement technique was used (laser image processing).

The energy expended in the process of phase dispersion, E, is correlated with interfacial area generated in Figure (8). It can be seen that the same interfacial area can be produced using different combinations of velocity and mixer elements, or expressed in more fundamental terms, by using different combinations of turbulence intensity and residence time.

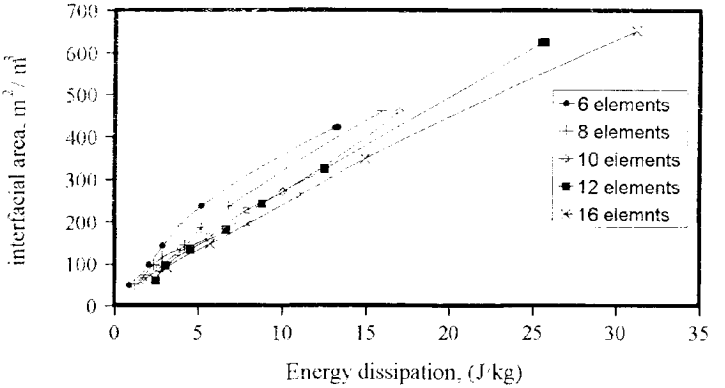

Figure (8): Interfacial area generated by the static mixer.

Therefore, it is of interest to know the effect of increasing the number of elements at a particular velocity on the efficiency of conversion of mechanical to surface energy. The efficiency of utilisation can be obtained by comparing the free energy of the newly generated surface to the mechanical energy that is expended to generate it. Thus where d_{320} is the Sauter mean diameter at the entrance of the static mixer which has been measured experimentally. The higher efficiency of dispersion, under conditions of high velocity and small number of elements, can be attributed

$$\eta = \frac{6\phi\sigma}{\rho_c E}\left(\frac{1}{d_{32}} - \frac{1}{d_{320}}\right)100\% \qquad (11)$$

to the rapid rate of droplet break-up when the average drop size is much larger than the equilibrium size. Conversely after the equilibrium conditions have been established, the rate of interfacial area production and the efficiency of energy utilisation both tend to zero. At any point in the mixer, the driving force for droplet break-up is the difference between the local average drop size and the local equilibrium drop size. As the dispersion flows through the mixer, this driving force decreases to a point where further reduction in drop size requires disproportionately larger increments in residence time. The efficiency calculated using Equation (11) is shown in Figure (9) for different numbers of mixer elements and velocities.

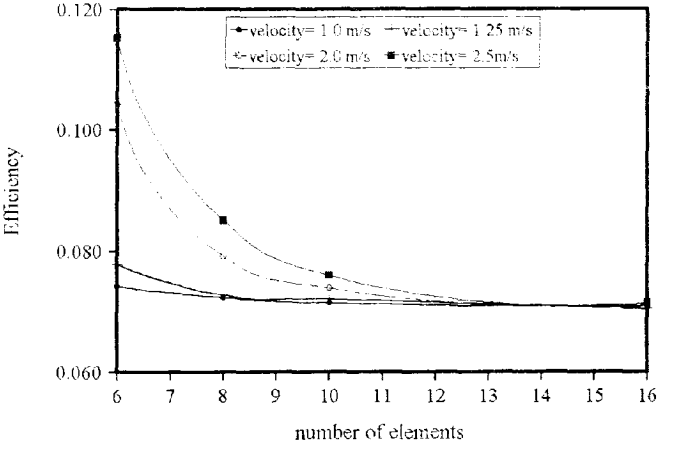

Figure (9): Efficiency utilization of static mixer at different flow velocities.

Effect of Horizontal Pipe on Sauter Mean Diameter, d_{32}.

The drop size distribution profile for a static mixer fitted after 18d (d=1") horizontal pipe and before the test cell have been compared with drop size distribution profile of 18d horizontal pipe fitted after the static mixer. A plot of Sauter mean diameter against the fluid velocity for this case

shows that at the same value of velocity, d_{32} is higher when the 18d pipe is fitted between the static mixer and the test cell. The results plotted in Figure(10), show a significant degree of coalescence takes place even at high fluid velocities when the residence time is extremely short. For example when the fluid velocity is changed from 0.8 to 2.25m/s, the residence time decreases from 0.6 to 0.25s. In both cases the Sauter mean diameter was found to increase between the exit of the static mixer and the end of the 18d horizontal pipe. At fluid velocity of 0.8m/s there was a 42% increase observed compared to a 25% increase at 2.25m/s. These changes are proportional to the change in the residence time.

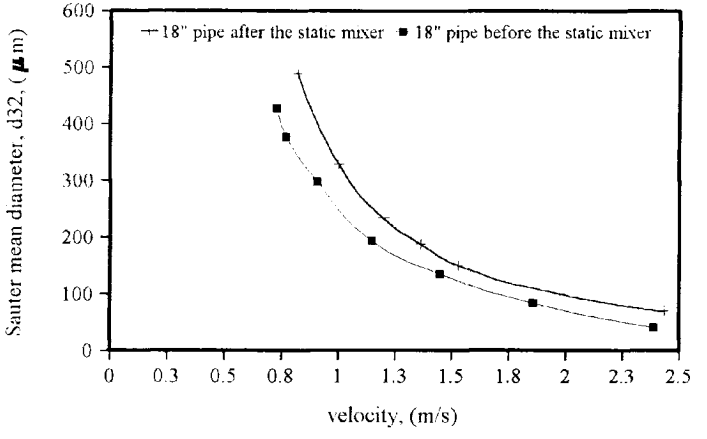

Figure (10): Effect of horizontal pipe on Sauter mean diameter.

Effect of 'U' Shaped Pipe on Sauter Mean Diameter, d_{32}.

Two 'U' shaped pipe fittings with four standard radius bends of total length of 18d and 43d were fitted after the static mixer. The 18d 'U' pipe fitting was chosen to get a 'U' pipe fitting with a total length equal to the 18d horizontal pipe without any elbows. The 43d 'U' pipe fitting was chosen to get a 'U' pipe fitting with a distance of 18d between the inlet and outlet. The geometry of the 'U' pipe fittings used, together with all dimensions, is shown in Figure (2). In all cases, the 'U' pipe fittings were fitted after the static mixer and were oriented horizontally and vertically (upward and downward) to the main flow.

Figure (11) shows the change in d_{32} as a function of velocity for an 18d horizontal pipe fitted after the static mixer and 18d 'U' pipe fitting oriented horizontally to the main flow after the static mixer so that gravity will play no role in the breakup or coalescence of droplets. It can be clearly seen that at the same value of flow rate, the Sauter mean diameter, d_{32}, is smaller at the exit of the 'U' pipe fitting than at the exit of the horizontal pipe as a result of the increased shear produced by the four standard elbows in the 'U' pipe fitting.

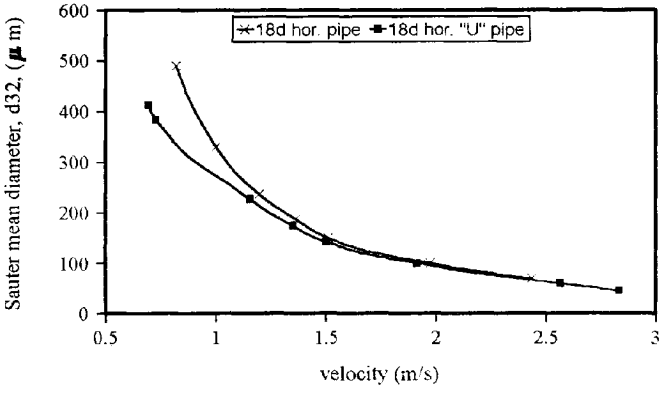

Figure (11): Effect of geometry on Sauter mean diameter.

Drop size distributions were measured for the 43d 'U' pipe fitting and results were compared with that of the18d 'U' pipe fitting. The only difference in the two fittings is the total length of the pipe, both have four standard elbows and were oriented horizontally to the main flow. At the same flow rate, the Sauter mean diameter, d_{32}, of the drop size distribution produced by the large 'U' pipe fitting is larger than that produced by the small 'U' pipe fitting. This is due to the higher rate of coalescence of drops in the large 'U' which is more than twice the length of the small 'U' pipe fitting. This is well demonstrated in Figure (12). The exponent of the correlation relating the Sauter mean diameter and the energy dissipation was in the range of -0.6 to -0.72. The error bars in this figure show the 95% confidence limits, which indicate although the effect of the increased length is small it is significant.

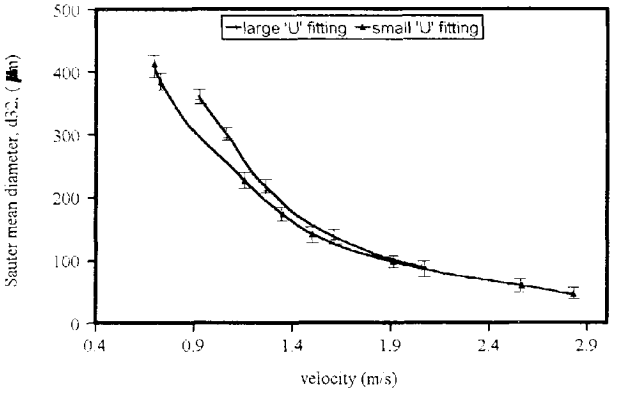

Figure (12): Comparison of Sauter mean diameter for small and large 'U' fittings.

Effect of an Offset Pipe on Sauter Mean Diameter, d_{32}.

The offset pipe fitting had two standard elbows and was 13d in length which is equal to the length of one of the arms of the 43d 'U' pipe fitting. The offset pipe was oriented horizontally and vertically (upward and downward) to the main flow after the static mixer as shown in Figure (2). When drop size distribution profiles of the offset pipe and the 'U' pipe fitting, oriented vertically upward to the main flow, against fluid velocity were compared, (Figure 13), a marked difference in the value of d_{32} was noticed, the error bars again indicate the 95% confidence limits. At the same velocity, the Sauter mean diameter for the 'U' pipe fitting is larger than that of the offset pipe. This indicates that the rate of drop coalescence due to the extra 30d of pipe length was greater than the rate of drop breakup from the two extra elbows in the 'U' pipe fitting. When the fittings were vertically oriented, the pressure drop across the offset pipe fitting was greater than that for the 'U' pipe fitting at the same flow rate. This is due to the fact that the net pressure drop across the first and second arm of the 'U' pipe fitting is negligible especially when it is vertically oriented (due to the flow in the direction of gravity in one arm and opposite to the gravity in the other arm of the 'U' fitting).

The equivalent length of a fitting is the length of straight pipe which

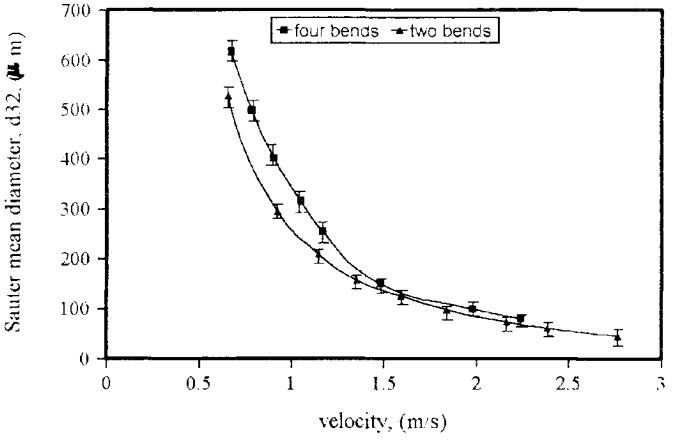

Figure (13): Effect of two and four bends oriented upward to the main flow on the Sauter mean diameter.

would produce the same pressure drop as obtained across the fitting. Different fittings may produce different levels of shear due to the variation of frictional loss. For a turbulent flow, the additional frictional loss for fittings can allowed by expressing the loss as an equivalent length of a straight pipe in pipe diameter, L_e / D. For any complex series pipes and fittings in which the friction factor is constant, the total equivalent pipe length will equal to the actual length of horizontal section of pipe plus the equivalent length of any pipe fittings such as elbows, valves, ... etc.). Table (1) shows the equivalent length of all the fittings used in this experimental work. The value of L_e used was taken from the literature (Chopey and Hicks[21]), in which it is presented as a fixed, flow-independent value. In reality it is a function of liquid velocity.

Table 1: Comparison between the length and equivalent length, L_e, of all the fittings used.

Fitting used (d=1")	Length (feet)	Equivalent length (feet)
18d horizontal pipe	1.5	1.5
18d 'U' normal elbows	1.5	11.9
43d 'U' normal elbows	3.6	14.27
13d offset pipe	1.1	6.3

Plotting d_{32} as a function of pressure drop over the equivalent length, $(\Delta P/L_e)$, for the offset pipe (Figure (14)), we can see that d_{32} for the vertically upward offset pipe is higher than that of the horizontal offset pipe, which is greater than that of the vertically downward offset pipe. When d_{32} was correlated with $(\Delta P/L_e)$, the exponent shows a value between -1.03 and -1.1. The reason for this may be the assumption that the equivalent length of the offset pipe fitting is same in all orientations, however, the pressure drop is different in each case. At the same value of d_{32}, then the pressure drop across the vertically upward offset pipe fitting is higher than that of the horizontally offset which is higher than that of the vertically downward offset. In theory the three curves (on Figure 14) should be on one line, such that if the pressure drop is given then d_{32} could be predicted regardless of its orientation. However, since the flow is not homogenous but two phase heterogeneous flow, there are complications in describing and quantifying its nature. Thus in calculating the equivalent length for multiphase pipe flow, there should be a third term added to the two terms previously discussed (the actual length and the equivalent pipe length of any pipe fitting) the size of which will vary depending on the direction of flow due to the hold up of the second phase. Further work on the stability of flow, the relative volumes of the two phases present in the pipeline and the pressure drop caused by the presence of an additional phase should be carried out to understand and evaluate the two phase flow.

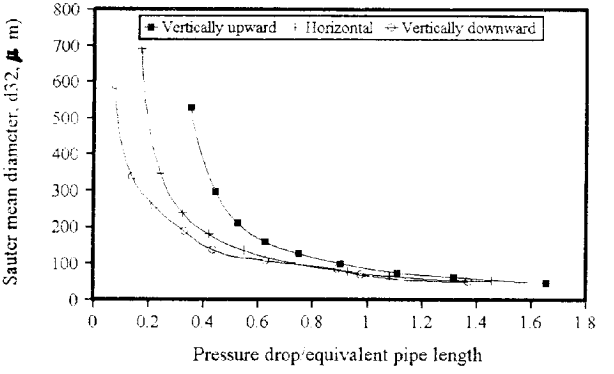

Figure (14): Effect of an offset pipe fittings on Sauter mean diameter.

Conclusions

Drop size distribution measurements were carried out on-line using a modified Malvern 2600 laser diffraction instrument aligned through a specially constructed glass-walled test cell. This instrument was found to provide reproducible and accurate data for dispersed phase concentrations ~ 0.5%. From the analysis of the drop size distribution data obtained, a number of conclusions can be drawn:

1. Correlation of the mean drop size with the energy dissipation rate, E, were produced and gave an exponent of -0.47 to -0.56 for a horizontal pipe configuration and -0.6 to -0.72 for the 'U' pipe fittings. The energy dissipation rate, E, strongly affects the equilibrium value of d_{32} and the rate at which it is achieved.
2. The Weber number (the ratio of inertia to surface tension forces) was also correlated with Sauter mean diameter and shows an exponent in the range of -0.71 to -0.83 for both horizontal pipe configurations and 'U' pipe fittings of the lengths tested.
3. Correlations of the mean drop size with the pressure drop over equivalent length, $(\Delta P/L_e)$, for all the fittings oriented horizontally and vertically (upward and downward) to the main flow were produced and found to give an exponent in the range of -1.11 to -1.03.
4. d_{32} is approximately 25% larger at 18d downstream of the static mixer than that at the exit of the mixer. This indicates that a significant degree of coalescence occurs even in a relatively short length of horizontal pipe. This result has important implications for the design procedures used for mixing plant and for the process requiring the separation of liquid-liquid dispersions as it suggests that the order of pipes and fittings in a flow network can have a profound effect on the nature of the dispersion at the exit stream.

References

1. Hinze J.O., Fundamentals of the Hydrodynamic Mechanism of Splitting in Dispersion Processes, *AIChE J.* **1(3),** (1955) 289.
2. Mlynek Y.; Resnick W., Drop Sizes in Agitated Liquid-Liquid Systems, *AIChE J.*, **18(1),** (1972), 122.
3. Shinnar R., On the Behaviour of Liquid Dispersions, *J. Fluid Mech.* **10,**

(1961) 259.
4. Sprow F. B., Distribution of Drop Sizes Produced in Turbulent Liquid-Liquid Dispersion. *Chem. Eng. Sci.*, **22**, (1967), 435.
5. Collins S. B.; Knudsen J. G., Drop-Size Distributions Produced by Turbulent Pipe Flow of Immiscible Liquids, *AIChE J.*, **16(6)**, (1970) 1072.
6. Sleicher C. A., Maximum Stable Drop Size in Turbulent Flow, *AIChEJ*, **8 (4)**, (1962) 471.
7. Paul H. I.; Sleicher C. A., The Maximum Stable Drop Size in Turbulent Flow: Effect of Pipe Diameter, *Chem. Eng. Sci.*, **20** (1965) 57.
8. Hughmark G. A., Drop Breakup in Turbulent Pipe Flow, *AIChE J.*, **17(4)**, (1971), 1000.
9. Karabelas A. J., Droplet Size Spectra Generated in Turbulent Pipe Flow of Dilute Liquid-Liquid Dispersions, *AIChE J.*, **24(2)**, (1978), 171.
10. Mersmann A; Grossmann H., Dispergieren in Flussigen Zweiphasen system, *Chem. Ing. Tech.*, **52**, (1980), 621.
11. Hanzevack E.L; Demetriou G.D., Effect of Velocity and Pipeline Configuration on Dispersion in Turbulent Hydrocarbon-Water Flow Using Laser Image Processing., *Int J. Multiphase Flow*, **15(6)**, (1989), 985.
12. Middleman S., Drop Size Distributions Produced by Turbulent Pipe Flow of Immiscible Fluids Through a Static Mixer, *Ind. Eng. Chem. Process. Des. Dev*, **13**, (1974), 78.
13. Matsumura K; Morishima Y.; Masuda, K; Ilcenago, H., Some Performance Data of the Hi-Mixer, an In-Line Mixer., *Chem. Ing. Tech*, **53**, (1981), 52.
14. Stewart A.C; EL-Hamouz A.M.; Davies G.A., Effect of Chemical Additives on the Interfacial Properties of Oil-Water Mixture, J.Disp. *Sci. Tech.*, **17(6)**, (1996), 675.
15. Malvern Instruments. 2600 Series user Manual, (1992).
16. AL-Taweel A.M.; Walker L.D., Liquid Dispersion in Static In-Line Mixers, *The Canadian Journal of Chemical Engineering*, **61**, (1983), 527.
17. Chen S.J.; Libby D.R.., Gas, Liquid and Liquid-Liquid Dispersions in a Kenics Mixer, *71st Annual AIChE Meeting*, (1978).
18. Baldyga J.; Bourne J. R., Drop Breakup and Intermittent Turbulence, *Journal of Chemical Engineering of Japan*, **26(6)**, (1993), 738.
19. Baldyga J.; Bourne J. R., Drop Breakup in the Viscous Subrange: A Source of Possible Confusion, *Chem Eng. Sci.*, **49(7)**, (1994), 1077.
20. Bourne J.R., Personal Communication (1994).
21. Chopey N.P.; Hicks T. G., "Handbook of Chemical Engineering Calculations", First Edition, McGraw Hill Book Company, New York (1984).

Appendix A: Nomenclature

a Interfacial area, $m^2 m^{-3}$
D pipe diameter, m
D_h hydraulic pipe diameter, m
d_{max} maximum drop diameter, μm
d_{32} Sauter mean drop diameter, μm
d_{320} Sauter mean drop diameter upstream from the mixer, μm
E energy dissipation per unit mass of fluid, J kg^{-1}
f friction factor in a pipe, dimensionless
K_i proportionality constant, i=1,2,3
L mixer length, m
L_e pipe equivalent length, m
N_{Re} Reynolds number, dimensionless
N_{We} Weber number, dimensionless
DP pressure drop, kp_a
Q volumetric flow rate, $m^3 s^{-1}$
V velocity, m s^{-1}

Greek letters

e energy dissipation rate per unit mass of fluid, J $kg^{-1} s^{-1}$
s interfacial tension, mN m^{-1}
f dispersed phase volumetric fraction
r_c density of continuous phase, kg m^{-3}
t residence time, s
m_c viscosity of continuous phase, N s m^{-2}
m_d viscosity of dispersed phase, N s m^{-2}
r density of the fluid, kg/m^{-3}
h efficiency of conversion of mechanical to surface energy, dimensionles
Subscripts
s normal radius elbow

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