www.czasopisma.pan.pl PAN www.journals.pan.pl
Chemical and Process Engineering 2021, 42 (2), 77–90
DOI: 10.24425/cpe.2021.137342

High flow, low shear impellers versus high shear impellers; dispersion of oil drops in water and other examples

Andrzej W. Pacek, Alvin W. Nienow

School of Chemical Engineering, University of Birmingham, Edgbaston, Birmingham, B15 2TT, UK

Flow patterns generated by two ChemShear impellers, CS 2 and CS 4 have been measured and flow numbers calculated; $Fl = 0.04$ for both impellers. Transient and equilibrium drop sizes, d_{32} µm of 3 different viscosity silicone oils agitated by a high-shear Rushton turbine, RT, a low-shear, high-flow HE3 impeller and the two ChemShears were determined. The equilibrium d_{32} are correlated for all impellers, by $d_{32} = 1300(\varepsilon_T)^{-0.58}_{\text{max.sv}}\nu^{0.14}$ with an $R^2 = 0.94$. However, the time to reach steady state and the equilibrium size at the same specific power do not match the above descriptors of each impeller's characteristics. In other literature, these descriptors are also misleading. In the case of mixing time, a high shear RT of the same size as a high flow HE3 requires the same time at the same specific power in vessels of $H/T = 1$. In bioprocessing, where concern for damage to cells is always present, free suspension animal cell culture with high shear RTs and low-shear impellers is equally effective; and with mycelial fermentations, damage to mycelia is greater with low shear than high. The problems with these descriptors have been known for some time but mixer manufacturers and ill-informed users and researchers continue to employ them.

Keywords: high shear/high flow descriptors, drop sizes, mixing time, bioprocessing

1. PREAMBLE

In 1993 as part of a project on catastrophic phase inversion (Pacek et al, 1993), we developed a combined video-strobe-computer technique that gave sharp images at concentrations up to 70% by volume dispersed phase down to about 25 µm in size, able to follow both the size change and structure of droplets with time during that process (Pacek et al., 1994a). It consisted of a stereo microscope with a very shallow depth of field attached to a video camera and gave sharp images of droplets even in intensely agitated liquid-liquid dispersions by using a strobe light pulsing at the camera framing rate. One of the most interesting features that emerged from the observations of structure was the presence of complex dispersions for water-in-oil dispersions (W/O) above about 0.25 volume fraction up to phase inversion where droplets of oil were found in the drops of water (O/W/O) (Pacek et al., 1994b). Such structures were not seen for O/W dispersions. This drops-in-drops structure in systems without surfactants present had not been reported in depth before and was used to explain qualitatively the lower concentration in W/O dispersions at which phase inversion occurs compared to O/W. Thus, it showed, also for the first time, that the behaviour of O/W and W/O dispersions is asymmetric (Pacek and Nienow, 1995).

* Corresponding author, e-mail: a.w.nienow@bham.ac.uk

Using the same video technique, we also began a program measuring drop sizes in a variety of situations, two of which are very relevant to this Special Issue to commemorate the life of Professor Jerzy Baldyga. In one study, we used it to measure the steady state drop sizes at two different scales with geometrically-similar systems stirred with a so-called high shear Rushton turbine. This work indicated that the same size drops were obtained at each scale when an equal impeller tip speed was used. Such a result is contrary to the standard analysis of drop break-up based on Hinze–Kolmogorov turbulence theory, which implies equal drop sizes should be found at the same mean specific energy dissipation rate for geometrically-similar systems. On the other hand, a number of other experimental studies had also reported equal drop sizes at equal tip speeds. We were very fortunate to establish a collaboration with Professor Baldyga, who was able to explain the results using an extension of the basic turbulence theory using the concept of intermittency (Baldyga et al., 2001).

At that time, almost all studies of liquid-liquid systems had used Rushton turbines, perhaps because the manufacturers called them high shear impellers and recommended them for liquid-liquid dispersions implying that they would give smaller dispersed drops for the same specific power compared with other impellers. Both these points were still valid at the time of the most recent reviews (Ghotli et al., 2013; Leng and Calabrese, 2016). So in collaboration with Dr Andre Bakker, then with Chemineer, we commenced a study comparing drop break-up with different impellers, which they marketed, under various descriptors implying their potential usage. Similar descriptors are still used by Chemineer and many other mixer manufacturers. Some of that work has been published (Pacek et al., 1999) but some of it has not. Therefore, given the connection to our collaboration with Professor Baldyga, it seems to be very appropriate to publish the extended data in this Special Edition. We also take the opportunity to set the findings in a wider context where such descriptors cause problems and can lead to incorrect impeller choices.

2. INTRODUCTION AND EXPERIMENTAL DETAILS

2.1. The vessel and impellers

The vessel was flat bottomed, baffled, diameter $T = 0.15$ m as was the liquid height H, with four, $0.1T$ standard baffles, fitted with a lid so that all air was excluded. Three impellers were supplied by Chemineer, all of which were placed $0.4T$ off the vessel base and had the same diameter, $D =\sim 0.4T$. One of these was the low shear, high flow efficiency, HE3 impeller, $D = 0.063$ m, recommended for efficient blending. The other two were the ultra high shear ChemShear impellers, Style 2, CS 2, and Style 4, CS 4, each of which are recommended for emulsification processes where very small drops are required (Fondy and Bates, 1963). These impellers are shown diagrammatically in Figure 1 and futher details of the Chemshear impellers are given in Table 1. The Chemineer impellers were designed to have a power number, $Po = 0.33$ and measured to be 0.30, (HE3), 0.32 (CS 2) and 0.36 (CS 4) (Pacek et al., 1999). The other was a Rushton turbine, RT, $D = 0.063$ m, traditionally known as a high shear impeller, $Po = 5.0$ and also recommended for producing small drops in liquid-liquid dispersion by Chemineer. It is also recommended in the *Handbook of Industrial Mixing* (Leng and Calabrese, 2004) and its updated version, *Advances in Industial Mixing* (Leng and Calabrese, 2016).

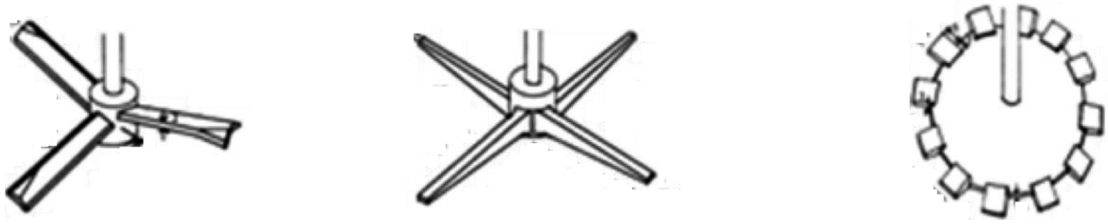

Fig. 1. The Chemineer impellers used in this study: High flow, low shear HE3, ChemShear CS 2, ChemShear CS 4

Table 1. Dimensions and number of the blades of the CS 2 and CS 4 ChemShear impellers and the swept volume of each of the impellers

CS 2	CS 4	Impeller swept volume V_{sv} ($\text{m}^3 \times 10^{-5}$)			
		RT	HE3	CS 2	CS 4
D = 61 mm 4 blades length 22 mm blade depth 10 mm tapering to 1 mm blade width 2 mm	D = 60 mm 12 blades length 7 mm blade depth 3 mm tapering to 2 mm blade width 7 mm	3.75	1.25	0.73	0.84

2.2. Drop size measurement and fluids used

The technique used for measuring the drop sizes was essentially the same as that described in the Preamble and published in detail elsewhere (Pacek et al., 1994a; 1994b). Deionised, distilled water was used as the continuous phase and the dispersed phase consisted of 5% by volume of three silicone oils (ρ_d = 970 kg/m^3, σ = 35 mN/m) with kinematic viscosities, ν of 10^{-6} m^2/s (1 cSt), 20×10^{-6} m^2/s (20 cSt) and 100×10^{-6} m^2/s (100 cSt). The vessel was filled with the required amount of water and oil and great care was taken to remove all air. In a similar manner to the earlier work, transient drop size distributions were measured until a steady state was reached. The speeds chosen, N rev/s, gave mean specific energy dissipation rates, ($\bar{\varepsilon}_T$) W/kg from about 0.4 to 1 W/kg. The previous work had shown that in this range of specific powers, the size distributions were the same near the top and the bottom of the vessel (Pacek et al., 1999). Therefore, measurements were only made at the mid-height of the vessel in this work. From the size distributions, Sauter mean sizes, d_{32} μm, were determined as a function of time and after steady state had been reached.

2.3. Flow numbers

In the earlier work (Pacek et al., 1999), the flow numbers of the Chemshear impellers were not measured; nor were they available from Chemineer. Here, the flow pattern in the vertical baffle plane of the discharge of the impellers was measured using PIV as done previously for Rushton turbines (Dyster et al., 1993) and HE3 impellers (Jaworski et al., 1996) both using LDA. From these flow patterns, the dimensionless flow number, Fl was determined, where

$$Fl = \frac{Q}{ND^3} \tag{1}$$

and Q (m^3/s) is a measure of the amount of fluid pumped by the impeller defined for radial flow impellers by

$$Q = \pi(D + 2s) \int_{-w/2}^{w/2} V_r(z)\,\mathrm{d}z \tag{2}$$

where s is the distance off the tip of the impeller blade and tends to zero for the evaluation of Fl, w is the blade width and V_r is the ensemble average radial velocity at each Z level where Z is the vertical direction. For the HE3, Fl = 0.6 (Jaworski et al., 1996); for the RT, Fl = 0.8 (Dyster et al., 1993).

3. RESULTS

3.1. Flow patterns and flow numbers for the two ChemShear impellers

The vectorial representation of the resultant mean velocity in the $r-z$ plane is shown in Figure 2 for both impellers. The rear of the tail represents the point at which the measurement was taken and its direction and the distance to the small circular head indicates the magnitude normalised by V_{tip} where $V_{tip} = \pi ND$. It is clear that in each case, the discharge flow is radial and extremely constrained in the vertical direction, especially compared to that seen with the relatively-wide blade Rushton turbine (Dyster et al., 1993). On the other hand, the maximum velocity of about $0.8V_{tip}$ for the CS 2 is a little higher than with the Rushton turbine ($\sim 0.75V_{tip}$) (Dyster et al., 1993), whilst that for the CS 4 ($\sim 0.5V_{tip}$) is a little lower.

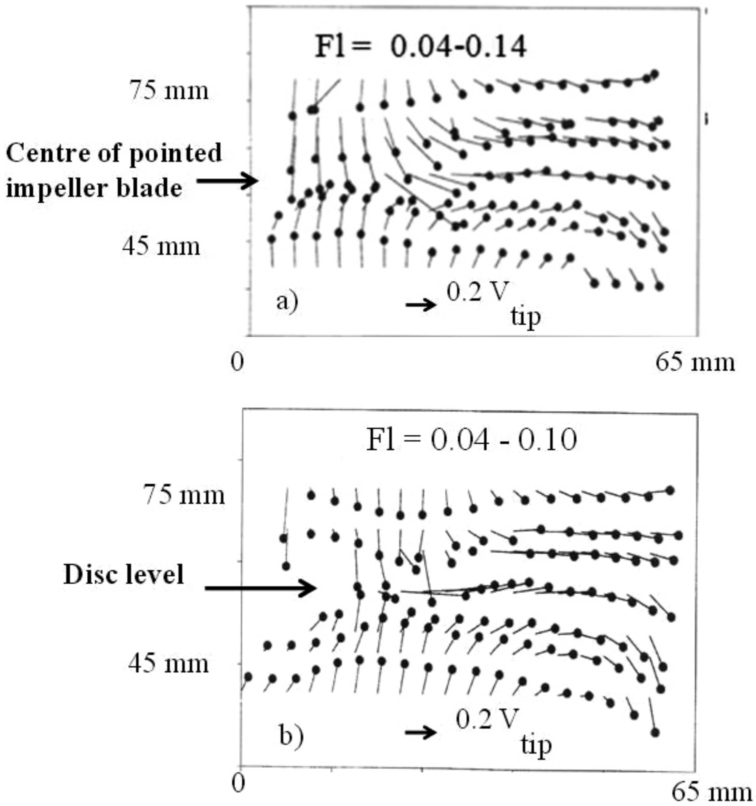

Fig. 2. Plot of mean velocity vectors in the $r-z$ plane for the ChemShear impellers: a) CS 2; b) CS 4

Based on the width of the tapered blades at their outward extremity and the outward velocities normal to this depth, the Flow Number of each of the impellers was the same, $Fl = 0.04$. It is also interesting to determine a "flow number" based on the velocities normal to a depth at the radius of the tip of the blade at which the flow is horizontal. This depth approximates that of the blades at their root. Using this depth the "flow numbers" are 0.14 and 0.10 for the CS 2 and the CS 4 respectively, still much lower than the Rushton turbine and HE3 impeller.

3.2. Steady state drop sizes

Figure 3 shows the equilibrium Sauter mean drop sizes, d_{32} μm, for the three different viscosity silicone oils. For the highest viscosity oil (100 cSt) (Fig. 3a), all of the impellers have been used. Because the two ChemShear impellers gave similar drop sizes (as they had in the earlier work (Pacek et al., 1999)), only one of the ChemShear impellers was used for the 20 cSt (Fig. 3b) and 1 cSt oils (Fig. 3c). Consistently, the high shear RT gives drops of the order of 2 to 3 times the size of the other three impellers. On the other hand, the low shear HE3 impeller gives sizes only up to some 20% greater than the ChemShear impellers and in some cases, the drop size is less with the HE3 than with the CS 2. The other consistent feature is that the higher viscosity oils give somewhat larger drops compared to the low viscosity at the same mean specific energy dissipation rates (specific power input). The raw data for the impellers used with all 3 different viscosity oils is shown in Table 2.

Table 2. Steady state Sauter mean drop sizes, d_{32} and specific power input, P/M, W/kg, based on: a) the vessel volume; b) the mass in the impeller swept volume, $(\varepsilon_T)_{\max.sv}$, W/kg

P/M, W/kg $(\varepsilon_T)_{\max.sv}$, W/kg	100 cSt	20 cSt	1 cSt
RT			
0.39 28	350 µm	250 µm	230 µm
0.60 42	290 µm	220 µm	150 µm
0.78 55	240 µm	200 µm	115 µm
HE 3			
0.44 93	175 µm	120 µm	90 µm
0.68 144	125 µm	100 µm	70 µm
0.87 184	95 µm	80 µm	50 µm
CS 2			
0.39 141	185 µm	120 µm	70 µm
0.60 193	120 µm	100 µm	65 µm
0.76 275	95 µm	90 µm	55 µm
CS 4			
0.35 108	175 µm		
0.54 170	140 µm		
0.70 220	90 µm		

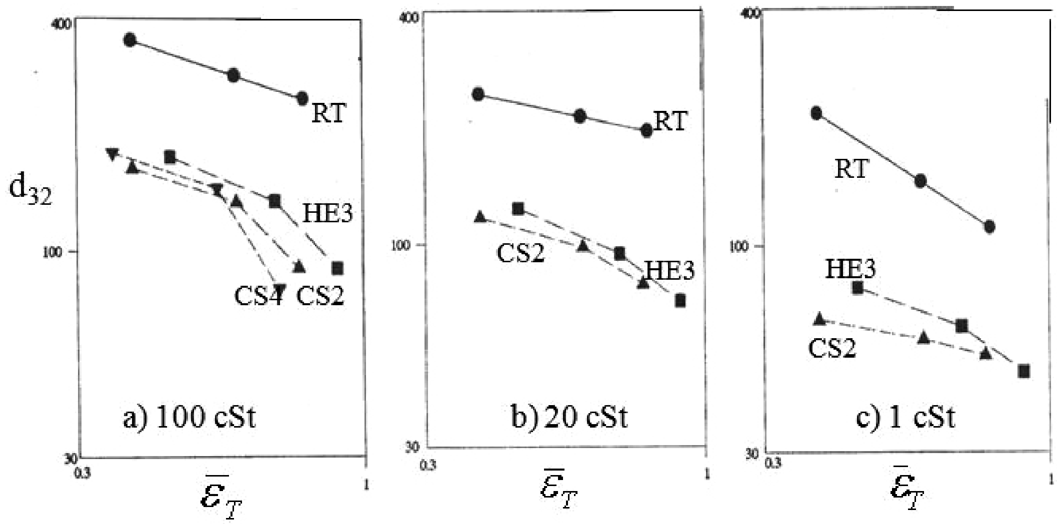

Fig. 3. Steady state Sauter mean drop diameter, d_{32}, versus specific power input, P/M, W/kg, (≡ mean specific energy dissipation rate, $(\bar{\varepsilon}_T)$) for the different impellers studied: a) 100 cSt; b) 20 cSt; c) 1 cSt

3.3. Dynamics of drop break-up

Figure 4 shows the Sauter mean transient drop size, d_{32}, at the same three specific powers as were used for the equilibrium drop sizes given in Figure 3a for up to 1 hr at each, for the most viscous 100 cSt oil for the RT and the CS 4. The drop sizes were always bigger for the RT as already shown in Fig. 3a but what is also revealed is that the transients are rather short and very similar for the two impellers in spite of their very different flow numbers.

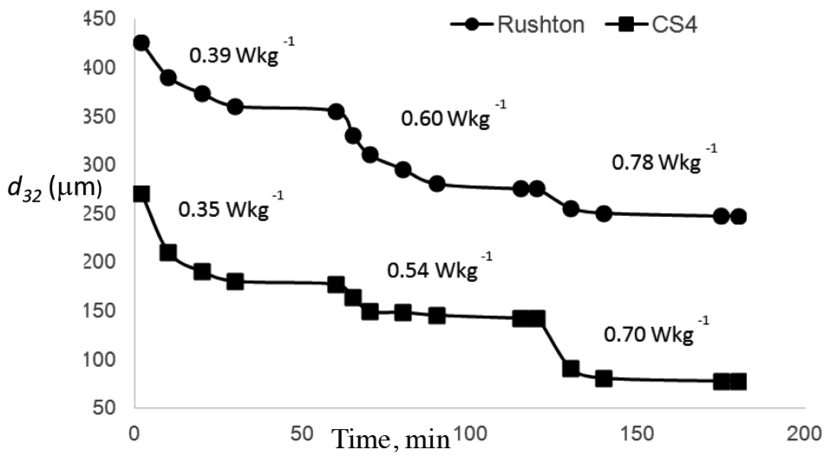

Fig. 4. Sauter mean drop size transients for the RT and CS 4 for the 100 cSt silicone oil as a function of the specific power input, P/M W/kg (mean specific energy dissipation rate, $(\bar{\varepsilon}_T)$, W/kg), for the Rushton turbine and CS 4 impeller

4. DISCUSSION

4.1. Steady state drop sizes

It is immediately obvious in Figure 3 that not only does the high shear RT not give the smallest drops, the low shear, high flow HE3 produces drops which are nearly as small as those produced by the ChemShear impellers at each viscosity. Such results are similar to those obtained with sunflower oil dispersions in our previous paper (Pacek et al., 1999). Since then there have been a few other papers which have shown that at the same specific power, so-called low shear impellers give significantly smaller drops than so-called high shear Rushton turbines. For example, a propeller ($d_{32} \sim 0.75$ times smaller) (Wille et al., 2001), and low shear Lightnin A310 ($d_{32} \sim 0.5$ smaller) (Musgrove and Ruzkowsski, 2000).

A particularly interesting example compared the drop sizes produced when using a RT, pitched blade turbine (PBT), HE3 and A310 (Padron and Okonkwo, 2018. They also found that the two low shear impellers, HE3 and A310 gave significantly smaller drops than the Rushton turbine and the PBT (by a factor of about 2 to 3). As usual, they considered that the equilibrium drop size should be related to the maximum specific energy dissipation rate found close to the impeller (Leng and Calabrese, 2004) rather than the specific power (mean specific energy dissipation rate). They determined $(\varepsilon_T)_{\max}$ from the relationship

$$(\varepsilon_T)_{\max} = 0.54k^{1.5}/l \tag{3}$$

where k is the turbulent kinetic energy, m^2/s^2, and l is the integral scale of turbulence, m, generally related to the depth of the blades. Both were obtained on the basis of laser Doppler anemometry (LDA) measurements made earlier by BHR Group and not generally available. The second approach was the simple method first proposed by McManamey (1979) for different diameter Rushton turbines and later used by Davies (1985) to correlate drop sizes from a wide variety of different configurations designed to produce liquid-liquid dispersions. This method simply assumes that all of the energy dissipation occurs in the zone swept out by the impeller, which was the technique used in our previous paper (Pacek et al., 1999). There is some similarity in the two approaches as deep blades will have a larger integral scale and also larger swept volumes. Both approaches in Padron et al. (2018) gave very similar functionalities but the fit to the data was a little better with the swept volume approach (R^2 = 0.98 compared to 0.90).

Given this finding and the information available, the swept volume approach was again used here to give $(\varepsilon_T)_{\max.sv}$. Table 2 also shows $(\varepsilon_T)_{\max.sv}$ for all runs based on the impeller swept volumes given in Table 1. Figure 5 shows that this approach works very well. The success of the simple swept volume concept is particularly valuable as it has been shown that though in the literature, both PIV and LDA measurements (angle resolved or time-averaged) have been made to determine $(\varepsilon_T)_{\max}$, the values obtained were very varied, dependent on how the raw data were manipulated. The latter authors had a similar issue with their own measurements.

In theory for the turbulent flow regime,, the exponent a in the relationship

$$d_{32} \propto (\varepsilon_T)^{a}_{\max.sv} \tag{4}$$

should be –0.4 (Leng and Calabrese, 2004) but the value found here was about –0.6 for each viscosity. This exponent is the same as the one found in the earlier work for the low Po impellers (Pacek et al., 1999) though in that case, the exponent for the Rushton turbine was close to the theoretical value of –0.4. In their recent paper, Padron and Okonkwo (2018) found that the data for all the impellers fitted an exponent of –0.4. The reason for this discrepancy is not clear, though there is an extensive discussion elsewhere (Pacek et al., 1999) and some interesting related concepts involving the impact of coalescence on the exponent (Baldyga et al., 2001) to which the reader is referred.

From Fig. 5 and Table 2, the impact of viscosity and $(\varepsilon_T)_{\max.sv}$ on drop size can be combined (Fig. 6) to give the following relationship

$$d_{32} = 1300(\varepsilon_T)^{-0.58}_{\max.sv}\nu^{0.14} \tag{5}$$

where d_{32} is in μm, $(\varepsilon_T)_{\max.sv}$ is in W/kg and ν is in cSt with R^2 = 0.94. Theoretically, increasing viscosity from 1 cSt has an almost negligible impact on drop size as interfacial tension provides the major stabilising force. However, at high viscosity, d_{32} increases with an exponent of 0.75 as viscosity becomes the dominant stabilising force (Arai et al., 1977; Leng and Calabrese, 2004). Over the range used here, Arai et al. (1977)

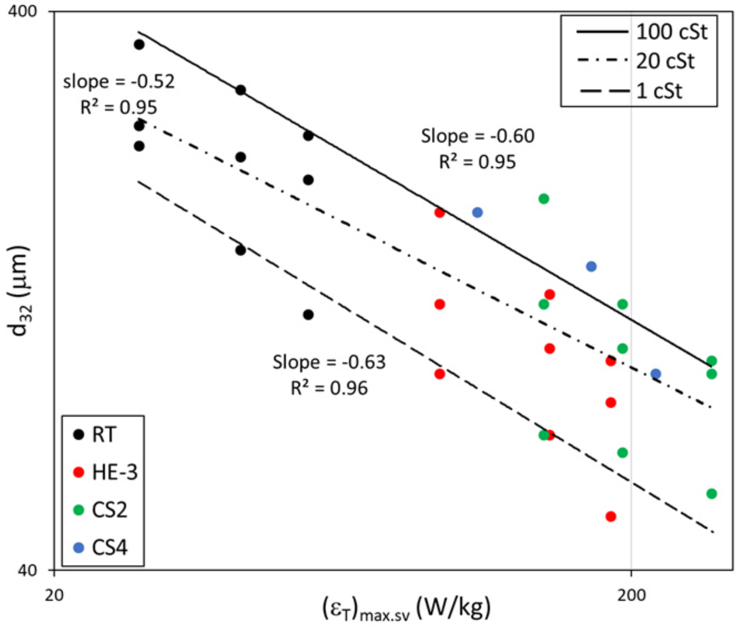

Fig. 5. Steady state Sauter mean drop diameter, d_{32}, versus $(\varepsilon_T)_{\text{max.sv}}$ for each impeller with 1 cSt, 20 cSt and 100 cSt silicone oil

gave an exponent very similar to the present value of 0.14. Interestingly, Equation (5) predicts the drop sizes for 5% sunflower oil (Fig. 12b, Pacek et al., 1999) allowing for the kinematic viscosity of ∼ 60 cSt; and the difference in interfacial tension of 27 mN/m compared to 35 mN/m for silicone oil using the exponent 0.6 as suggested theoretically.

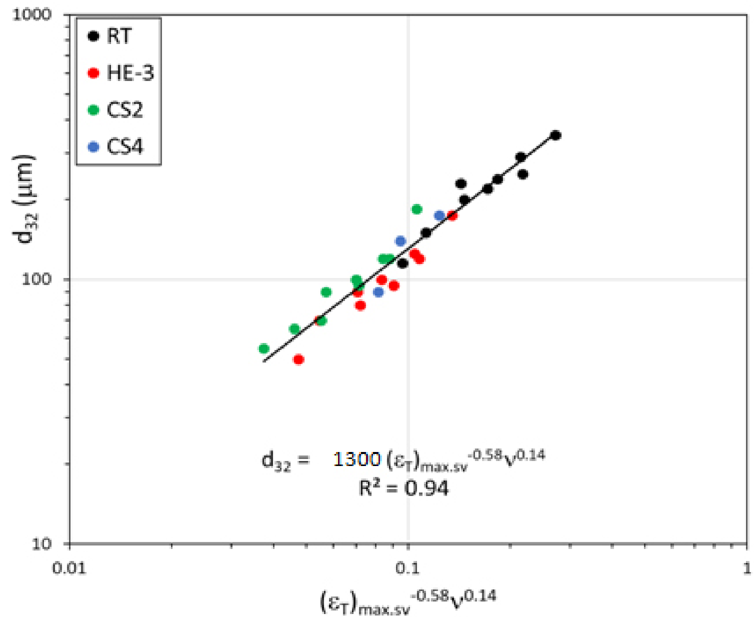

Fig. 6. All the d_{32} μm drop sizes correlated with $(\varepsilon_T)_{\text{max.sv}}$ W/kg and kinematic viscosity, ν cSt

4.2. Dynamics of drop break-up

Figure 4 shows the dynamics of the drop-break-up for the most viscous silicone oil for the Rushton turbine and the CS 4 ChemShear impeller at 3 different specific power inputs. It is similar to Fig. 6 of our previous paper (Pacek et al., 1999). The results there were for the dynamics of sunflower oil break-up for the Rushton turbine at ~ 1 W/kg and the HE3 and CS 4 at ~ 0.5 W/kg where both the latter impellers gave similar transient times, significantly shorter than the Rushton turbine. Here, the time to reach equilibrium falls from about 30 min to about 20 min as the specific power increases for the CS 4 impeller (Fl = 0.04) whilst for the Rushton turbine (Fl = ~ 0.8), the time is certainly no shorter and may be a little longer. Overall, from this and the previous work, it can be seen that both the ChemShear impellers give transients very similar to the high flow HE3 impellers (Pacek et al., 1999) and in general shorter than Rushton turbines.

It is again interesting to compare these findings with literature models. These models (Leng and Calabrese, 2004; Padron and Okonkwo, 2018) suggest that the time to reach the equilibrium stable drop size depends on the number of times that the drops pass through the highest specific energy dissipation rate in the impeller region. Clearly the time to achieve the required number of circulations is inversely proportional to the circulation time, t_c which for equal diameter impellers is given by (Pacek and Nienow, 1995),

$$t_c \propto Po^{\frac{1}{3}}/Fl \tag{6}$$

With the RT and HE3 having similar flow numbers, this model predicts that the low power number of the HE3 should lead to a shorter time to equilibrium as found experimentally (Pacek and Nienow, 1995). However, the very low flow numbers of the ChemShear impeller should lead to much longer equilibrium times which is not the case either in the earlier work with sunflower oil or here with silicone oil. Again the behaviour of the dispersing process with the different impellers does not accord with the descriptor terminology.

In general, the high shear and low shear terms as descriptors of impeller function or effectiveness seem inappropriate when applied to the impeller types available for producing liquid-liquid dispersions.

4.3. Other examples of problems with the high flow/low shear and high shear descriptors

4.3.1. Mixing time

Mixing time is another aspect of mixing where HE3 impellers and other thin blade hydrofoils such as the Lightnin A310 are considered as high flow impellers recommended for bulk blending, implying shorter mixing times at the same specific power. Indeed, if a model based on flow numbers is used to predict mixing times, that is the outcome (Nienow, 1997). However, based on turbulence theory, the mixing time, θ_m s, in baffled vessels with $H = T$ gives the relationship (Nienow, 1997),

$$\theta_m = 5.9T^{2/3}\,(\bar{\varepsilon}_T)^{-1/3}\,(D/T)^{-1/3} \tag{7}$$

This equation implies that the mixing time is independent of the impeller type whether a high shear Rushton turbine, or a low shear, high efficiency hydrofoil such as the HE3 impeller or the Lightnin A310 hydrofoil (Langheinrich et al., 1998), the Prochem Maxflow-T hydrofoil (Hass and Nienow, 1989) or, more recently, a low shear "elephant ear" impeller (Simmons et al., 2007).

A form of Equation (7) is now recommended in Advances in Industrial Mixing for a range of mixing applications including for example the impact of mixing on chemical reactions (Paterson et al., 2015). Clearly, the concept of high flow impellers as being superior for bulk blending has been superceded in the literature but it persists in general practice.

4.3.2. Impeller choice for bioreactors

The bioprocessing community has always been concerned about the shear sensitivity of cells to agitation and so is the literature. That concern was greatly increased with the advent of the production of highly valuable biologicals such as viral vaccines and monoclonal antibodies using animal and insect cells in the 1980s. Because such cells do not have a cell wall, they therefore are perceived as being more easily damaged by turbulent stresses leading to loss of production or lower product quality. Hence, there has been a desire to use "low shear" impellers such as the colloquially known "elepant ear" impeller developed by ABEC (Fig. 7) and now similar geometries are used by other bioreactor manufacturers such as Applikon. Such an impeller is simply a deep bladed pitched blade turbine with turbulence kinetic energy similar to that found with other pitched blade turbines and deep hydrofoils (Simmons et al., 2007). On the other hand, since the "elephant ear" impeller has deep blades, it does have a relatively larger swept volume and therefore potentially, a lower maximum specific energy dissipation rate.

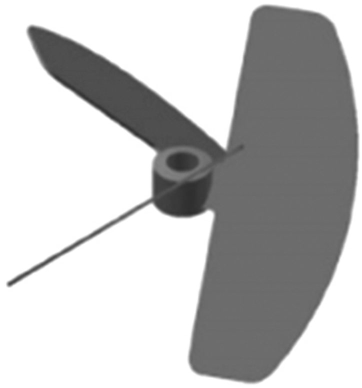

Fig. 7. An ABEC "low shear", "elephant ear" impeller

There is a particular perception that the high shear Rushton turbine is damaging to animal cells in free suspension. However, as early as 1992, such cells were reported to grow successfully at the bench scale with such impellers (Oh et al., 1992). Later, in 2011, work at the industrial scale with Rushton turbines and low shear propellers showed the same growth and product quality with each (Sandadi et al., 2011). More recently, the use of Rushton turbines to cultivate CHO cells has been studied at Genentech and showed not to impact on productivity or product quality at specific powers of ∼ 1 W/kg compared to their industrial scale culture at ∼ 0.02 W/kg (Nienow et al., 2013).

More recent concerns have related to the culture of human and bovine stem cells, which must attach to a surface to grow. The required high surface area is obtained in a stirred tank bioreactor by using a large number of particles with special surface properties of ∼ 200 µm diameter called microcarriers. Here, Rushton turbines have not been used because the specific power required to suspend particles (here low density microcarriers (specific gravity <∼ 1.1)) is greater with radial flow Rushton turbines than with axial flow impellers, whether hydrofoils or simple pitched blade turbines (Ibrahim and Nienow, 2004). Thus, it is the axial flow pattern which aids suspension that is the important impeller characteristic, not its low or high shear designation (Nienow et al., 2016; Hanga et al., 2020).

The same considerations apply when growing both T-cells and CAR-T cells in stirred bioreactors, which have also been perceived as shear sensitive. In these cases, particles of ∼ 5 µm diameter but relatively dense (specific gravity ∼ 1.4) called Dynabeads have to be employed to activate the cells in order for cell growth to be achieved. For satisfactory growth with the right product qualities, these Dynabeads have to be suspended to ensure effective contact between them and cells. Thus again, it is the most effective way of achieving particle suspension through generating an axial flow pattern that determines the agitator type and overall configuration of the bioreactor rather than the choice of a high or low shear impeller (Costariol et al., 2020; Rotondi et al., 2021).

In all the above biological systems, a critical feature is that the magnitude of the Kolmogorove scale of turbulence at the maximum specific energy dissipation rate should be greater than the size of the biological entity or a cell/microcarrier combination if satisfactory culture is to be achieved (Nienow, 2021). With mycelial fermentations, the cells tend to grow as clumps, which may be of the order of a few mm, so it is much greater than the Kolmogorov scale and breakage occurs. However, this process has been successfuly modelled by an energy dissipation, circulation function, which shows, as found in practice, that the so-called low shear impellers damage the mycelia more than the so-called high shear Rushton turbines (Justen et al., 1996; Nienow, 2021). Once again, the high shear, low shear terminology is misleading.

5. CONCLUSIONS

PIV studies have been made for the first time of the flow patterns produced in the impeller discharge region of two ChemShear impellers, namely, the CS 2 and CS 4 models. Both produce very strong radial flow, rather constrained compared to those produced by the deeper blades of the Rushton turbine. From these, the flow numbers have been determined and found to be, as expected, very low, $Fl = 0.04$. In spite of these very low flow numbers, as in the previous related work (Pacek et al., 1999), the high flow HE3 impeller and the high shear CS 4 impeller reached the equilibrium drop size in approximately the same time, both shorter than the Rushton turbine but less so than in the previous paper. With respect to the equilibrium drop sizes, at the same specific power, the high shear RT gave much larger drops than the low shear, high flow HE3 impellers which were only slightly higher than those from the ChemShear impellers for each viscosity; with drop sizes increasing in line with dispersed phase viscosity. For all impellers drop size is well correlated with the power dissipated in the impeller swept volume as was recently also shown by Padron and Okonkwo (2018). The present work alongside that of Pacek et al. (1999) and Padron and Okonkwo (2018) strongly suggests that when choosing impellers for an industrial liquid-liquid dispersion application, the concepts of high shear and low shear/high flow impellers is not a helpful one; and that a Rushton turbine is an inappropriate choice.

These descriptors are also discussed in relation to turbulent mixing time where a high shear RT impeller of the same D/T as a high flow impeller gives the same mixing time at the same specific power in a vessel where $H/T = 1$; as do other radial flow impellers, pitched blade turbines and other hydrofoil impellers (Nienow, 1997; Simmons et al., 2007). Again, the descriptors are misleading.

Finally, these descriptors are discussed in relation to bioprocessing. The issue here is mainly the perception that cells are shear sensitive, especially those without a cell wall. Here the "high shear" RT is particularly considered damaging but it has been shown to grow free suspension cells quite successfully giving similar cell densities and product qualities as with low shear impellers both at the bench and industrial scale. With mycelia, where the organism can grow into clump-like structures much bigger than the Kolmogorov scale, "low shear" impellers cause more clump fragmentation leading to smaller ones at the same specific power. In bioreactors, when particles are also required to be suspended, then it is the axial flow that is important not the "shear" designation. Much of the problem of the use of the descriptors high shear/lowshear in bioprocessing is because of the over concern for the sensitivity of cells to fluid dynamic stress and the use of outdated literature perpetrated in some textbooks (Nienow, 2021).

As can be seen from some of the references, the misleading nature of these descriptors has been recognised for some considerable time, yet the mixer manufacturer websites still use them; for example (Indco, 2021; NOV Chemineer, 2021; Sepro Mixing and Pumping, 2021; SPX Flow, 2021) and there are more. The high shear/high flow descriptor was developed by Oldshue in Lightnin documents in the 70's and published in his book (Oldshue, 1983). Other manufacturers followed suit and have continued to do so. The use of these descriptors is holding back the use in industry of the developments in academic mixing research, which many of the users do not follow either, especially in bioprocessing.

The only advantage of just using these two misleading descriptors is simplicity. Flow patterns are another simple way; but axial flow impellers such as those with wide blades and a high solidity ratio may be very effective for gas dispersion (Nienow and Bujaski, 2002; Nienow, 2014) whilst narrow blades are very poor (Chapman et al., 1983). Unfortunately, there is not any simple way of building on the very substantial developments in fluid mixing technology of recent years. A leading figure in those developments has been Professor Jerzy Baldyga. His work was excellent academically and industrially very relevant; but generally complex and almost never simple, as exemplified in our work with him involving intermittency (Baldyga et al., 2001).

SYMBOLS

a	exponent
d_{32}	Sauter mean drop size, µm
D	impeller diameter, m
Fl	flow number, dimensionless
H	liquid height in vessel, m
k	turbulent kinetic energy, m^2/s^2
l	integral scale of turbulence, m
M	mass of fluid in the vessel, kg
N	agitator speed, rev/s
P	power, W
P/M	specific power input, W/kg
Po	power number, dimensionless
r	radial direction
t_c	circulation time, s
T	vessel diameter, m
V_{SV}	impeller swept volume m^3
V_T	volume of liquid in vessel, m^3
V_{tip}	tip velocity, m/s
z	vertical direction

Greek symbols

$\bar{\varepsilon}_T$	mean specific energy dissipation rate, W/kg
$(\varepsilon_T)_{\text{max.sv}}$	specific enery dissipation rate in the impeller swept volume, W/kg
ν	kinematic viscosity, cSt
ρ_d	dispersed phase density, kg/m^3
σ	interfacial tension, mN/m
θ_m	mixing time, s

REFERENCES

Arai K., Konno M., Matumagata Y., Saito S., 1977. The effect of dispersed phase viscosity on the maximum stable drop size in turbulent flow. *J. Chem. Eng. Jpn.*, 10, 325–330. DOI: 10.1252/jcej.10.325.

Baldyga J., Bourne J.R., Pacek A.W., Amanullah A., Nienow A.W., 2001. Effects of agitation and scale-up on drop size in turbulent dispersions: Allowance for intermittency. *Chem. Eng. Sci.*, 56, 3377–3385. DOI: 10.1016/s0009-2509(01)00027-6.

Chapman C.M., Nienow A.W., Cooke M., Middleton J.C., 1983. Particle-gas-liquid mixing in stirred vessels: Part 2 – gas-liquid mixing. *Chem. Eng. Res. Des.*, 61, 82–95.

Costariol E., Rotondi M.C., Amini A., Hewitt C.J., Nienow A.W., Heathman T.R.J., Rafiq Q.A., 2020. Demonstrating the manufacture of human CAR-T cells in an automated stirred-tank bioreactor. *Biotechnol. J.*, 15, 2000177. DOI: 10.1002/biot.202000177.

Davies J.T., 1985. Droplet sizes of emulsions related to turbulent energy dissipation rates. *Chem.Eng. Sci.*, 40, 839–842. DOI: 10.1016/0009-2509(85)85036-3.

Dyster K.N., Koutsakos E., Jaworski Z., Nienow A.W., 1993. An LDA study of the radial discharge velocities generated by a Rushton turbine: Newtonian fluids, Re $\geq$ 5. *Chem. Eng. Res. Des.*, 71, 11–23.

Fondy P.L., Bates R.L., 1963. Agitation of liquid systems requiring a high shear characteristic. *AIChE J.*, 9, 338–342. DOI: 10.1002/aic.690090312.

Ghotli R.A., Raman A.A.A., Ibrahim S., Baroutian S., 2013. Liquid – liquid mixing in stirred vessels: A review. *Chem. Eng. Comm.*, 200, 595–627. DOI: 10.1080/00986445.2012.717313.

Hanga M.P., Ali J., Moutsatsou P., de la Raga F.A., Hewitt C.J., Nienow A.W, Wall I., 2020. Bioprocess development for scalable production of cultivated meat. *Biotech. Bioeng.*, 117, 3029–3039. DOI: 10.1002/bit.27469.

Hass V.C., Nienow A.W., 1989. A new axial pumping agitator for the dispersion of gas in liquids. *Chem. Ing. Technik*, 61, 152–154. DOI: 10.1002/cite.330610213.

Ibrahim S., Nienow A.W., 2004. Suspension of microcarriers for cell culture with axial flow impellers. *Chem. Eng. Res. Des.*, 82, 1082–1088. DOI: 10.1205/cerd.82.9.1082.44161.

Indco, 2021. The difference between high shear and low shear mixing. Available at: https://www.indco.com/blog/indco/2020/03/05/the-difference-between-high-shear-and-low-shear-mixing.

Jaworski Z., Nienow A.W., Dyster K.N., 1996. An LDA study of the turbulent flow field in a baffled vessel agitated by an axial, down-pumping hydrofoil impeller. *Can. J. Chem. Eng.*, 74, 3–15. DOI: 10.1002/cjce.5450740103.

Justen P., Paul G.C., Nienow A.W., Thomas C.R., 1996. Dependence of mycelial morphology on impeller type and agitation intensity. *Biotech. Bioeng.*, 52, 672–684. DOI: 10.1002/(SICI)1097-0290(19961220)52:6<672::AID-BIT5>3.0.CO;2-L.

Langheinrich C., Nienow A.W., Eddleston T., Stevenson N.C., Emery A.N., Clayton T.M, Slater N.K.H., 1998. Liquid homogenisation studies in animal cell bioreactors of up to 8m^3 in volume. *Food Bioprod. Process.*, 76, 107–116. DOI: 10.1205/096030898531873.

Leng D.E., Calabrese R.V., 2004. Immiscible liquid-liquid systems. In: Paul E.L., Atiemo-Obeng V.A., Kresta S.M. (Eds), *Handbook of industrial mixing: Science and Practice,* John Wiley, New York. 639–753. DOI: 10.1002/0471451452.ch12.

Leng D.E., Calabrese R.V., 2016. Immiscible liquid-liquid systems. In: Kresta S.M., Etchells III A.W., Dickey D.S., Atiemo-Obeng V.A. (Eds), *Advances in industrial mixing: A companion to the handbook of industrial mixing*. John Wiley, New York. 12, 457–464.

McManamey W.J., 1979. Sauter mean and maximum drop diameters of liquid-liquid dispersions in turbulent agitated vessel at low dispersed phase hold-up. *Chem. Eng. Sci.*, 34, 432–434. DOI: 10.1016/0009-2509(79)85081-2.

Musgrove M., Ruszkowski S., 2000. Influence of impeller type,and agitation conditions on the drop size of immiscible liquid dispersions. *Proceedings of the 10th European Conference on Mixing*. Delft, the Netherlands, 2–5 July 2000, 165–172. DOI: 10.1016/b978-044450476-0/50022-4.

Nienow A.W., Coopman K., Heathman T.R.J., Rafiq Q.A., Hewitt C.J., 2016. Chapter 3 – Bioreactor engineering fundamentals for stem cell manufacturing. In: Cabral J.M.S. et al (Eds.) *Stem Cell Manufacturing*. Elsevier Science, Cambridge, USA. 43–75. DOI: 10.1016/B978-0-444-63265-4.00003-0.

Nienow A.W., Scott W.H., Hewitt C.J., Thomas C.R., Lewis G., Amanullah A., Kiss R., Meier S.J., 2013. Scale-down studies for assessing the impact of different stress parameters on growth and product quality during animal cell culture. *Chem. Eng. Res. Des.*, 91, 2265–2274. DOI: 10.1016/j.cherd.2013.04.002.

Nienow A.W., 2021. The impact of fluid dynamic stress in stirred bioreactors-the scale of the biological entity: A personal view. *Chem. Ing. Tech.*, 93, 17–30. DOI: 10.1002/cite.202000176.

Nienow A.W., 1997. On impeller circulation and mixing effectiveness in the turbulent flow regime. *Chem. Eng. Sci.*, 52, 2557–2565. DOI: 10.1016/S0009-2509(97)00072-9.

Nienow A.W., 2014. Stirring and stirred tank reactors. *Chem. Ing Tech.*, 86, 2063–2074. DOI: 10.1002/cite.2014 00087.

Nienow A.W., Bujalski W., 2002. Recent studies on agitated three phase (gas-liquid-solid) systems in the turbulent regime. *Chem. Eng. Res. Des.*, 80, 832–838. DOI: 10.1205/026387602321143363.

NOV Chemineer, 2021. Available at: https://www.chemineer.com/literature.html.

Oh S.K.W., Nienow A.W., Emery A.N., Al-Rubeai M., 1992. Further studies of the culture of mouse hybridomas in an agitated bioreactor with and without continuous sparging. *J. Biotechnol.*, 22, 245–270. DOI: 10.1016/0168-1656(92)90144-X.

Oldshue J.Y., 1983. *Fluid mixing technology and practice*, McGraw Hill, New York.

Pacek A.W, Nienow A.W., 1995. A problem for the description of turbulent dispersed liquid-liquid systems. *Int. J. Multiphase Flow*, 21, 323–325. DOI: 10.1016/0301-9322(94)00068-u.

Pacek A.W, Nienow A.W., Moore I.P.T., 1994b. On the structure of turbulent liquid-liquid dispersed flows in an agitated vessel. *Chem. Eng. Sci.*, 49, 3485–3498. DOI: doi.org/10.1016/s0009-2509(94)85027-5.

Pacek A.W., Chamsart S., Nienow A.W., Bakker A., 1999. The influence of impeller type on mean drop size and drop size distribution in an agitated vessel. *Chem. Eng. Sci.*, 54, 4211–4222. DOI: 10.1016/s0009-2509(99)00156-6.

Pacek A.W., Moore I.P.T., Calabrese R.V., Nienow A.W., 1993. Evolution of drop size distributions and average drop diameters in liquid-liquid dispersions before and after phase inversion. *Chem. Eng. Res. Des.*, 71, 340–341.

Pacek A.W., Moore I.P.T., Nienow A.W., Calabrese R.V., 1994a. A video technique for the measurement of the dynamics of liquid-liquid dispersion during phase inversion. *AIChE J.*, 40, 1940–1949. DOI: 10.1002/aic.690401203.

Padron D.A., Okonkwo D.A., 2018. Effect of impeller type on drop size of turbulent, non-coalescing liquid-lquid dispersions. *16th European Conference on Mixing*, Toulouse, France, 9–12 September 2018.

Paterson G.K., Pauls E.L., Kresta S.M., Etchells A.W., 2015. Mixing and chemical reaction. In: Kresta S.M., Etchells III A.W., Dickey D.S, Atiemo-Obeng V.A. (Eds.), *Advances in industrial mixing: A companion to the handbook of industrial mixing*. John Wiley, New York. 465–478.

Rotondi M., Grace N., Betts J., Bargh N., Costariol E., Zoro B., Hewitt C.J., Nienow A.W., Rafiq Q.A., 2021. Design and development of a new ambr250® bioreactor vessel for improved cell and gene therapy applications. *Biotechnol. Lett.*, 43, 1103–1116. DOI: 10.1007/s10529-021-03076-3.

Sandadi S., Pedersen H., Bowers J.S., Rendeiro D., 2011. A comprehensive comparison of mixing, mass transfer, Chinese hamster ovary cell growth, and antibody production using Rushton turbine and marine impellers. *Bioprocess. Biosyst. Eng.*, 34, 819. DOI: 10.1007/s00449-011-0532-0.

Sepro Mixing and Pumping, 2021. Mixing basics: Shear and flow. Available at: https://mixing.seprosystems.com/mixing-basics-shear-and-flow/.

Simmons M.J.H., Zhu H, Bujalski W., Hewitt C.J., Nienow A.W., 2007. Mixing in bioreactors using agitators with a high solidity ratio and deep blades. *Chem. Eng. Res. Des.*, 85, 551–559. DOI: 10.1205/cherd06157.

SPX Flow, 2021. Lightnin Mixers. General Overview. Available at: https://www.spxflow.com/assets/pdf/LGT_A400_B-937_US.pdf.

Wille M., Langer G., Werner U., 2001. The influence of macroscopic elongational flow on dispersion processes in agitated tanks. *Chem. Eng. Technol.*, 24, 119–127. DOI: 10.1002/1521-4125(200102)24:2<119::aid-ceat119>3.0.co;2-g.

Received 6 June 2021
Received in revised form 7 July 2021
Accepted 8 July 2021