

## **PREDICTION OF DROP SIZES FOR LIQUID-LIQUID SYSTEMS IN STIRRED SLIM REACTORS – PART II: MULTI STAGE IMPELLERS**

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### **ABSTRACT**

Based on the successful predictions of transient drop sizes in the first part of this research (Maaß et al., *Prediction of drop sizes for liquid-liquid systems in stirred slim reactors-Part I: Single stage impellers. Chem. Eng. J., 162 (2010) 792-801*), this part is a straight continuation and extension of the earlier work. The predictive capabilities of the used population balance equation model are increased for single stage impellers and transferred to scale-up procedures of such applications. Therefore different scale-up rules for liquid-liquid systems are tested experimentally and by simulations in two different sized, geometrically similar vessels. The multi stage impellers are tested against comparable single stage impellers in terms of power consumption, mixing time and minimum impeller speed. Especially for high aspect ratios (larger than three), multi stage impellers successfully compete with the single stage ones. The measured drop size distributions in slim reactors with multi stage impellers showed no dependency on the local position, although the dispersion process is tedious due to the compartmentalization. The simulations are not able to reflect this initial phase of the dispersion process, but are in close agreement with the experiments after complete dispersion is fulfilled. Based on these experiences the aspect ratio is increased up to five and the resulting drop size can be predicted with reasonable deviations (lower 10 than percent). The results of the scale-up of this multi stage impeller liquid-liquid system do not lead to a clear conclusion. Although the simulations recommend the use of constant power input, the experiments could not support this. None of the other traditional scale-up rule are supported by the experiments. Overall, the results of power consumption, mixing time and dispersion behavior show the great potential of multi stage impellers for process optimization and intensification in slim reactors.

### **KEYWORDS:**

Drop size prediction, high aspect ratio reactors, multi stage impeller, population balance equation, liquid-liquid dispersion and scale-up

## 1 INTRODUCTION

In the previous part of this work [1] the major investigation objective was the influence of the liquid level on transient drop sizes in highly concentrated liquid-liquid systems. Diverse experiments were carried out to determine the impact of an increasing H/T ratio (liquid level height  $H$  versus tank diameter  $T$ ) using a single stage stirrer system. Different modeling approaches have been tested to predict the Sauter mean diameter for such specific scale-ups. The energy law [2] using the average energy dissipation only roughly predicts the Sauter mean diameter. The population balance equation (PBE) using a one-zone modeling approach is only slightly better in its prediction of results. Very satisfying results were achieved by using the PBE with a two-zone model; the overall deviations between calculated and predicted Sauter mean diameters were less than ten percent.

In this part of the work, the major objectives are to test the usability of multi stage stirrers for dispersing liquid-liquid systems occurring in polymerization processes and to provide scale-up criteria for the selected multi stage impeller system based on experiments and PBE simulation results.

## 2 STATE OF THE ART

### 2.1 Scale-up of liquid-liquid systems

Mersmann et al. [3] provide a broad overview of scale-up and design rules for stirred vessels, and they also differentiate between the main tasks occurring in an agitated system. In addition, they introduce basic scale-up rules for dispersion processes. In this context, successful scale-up means that larger scale operations are anticipated and understood. It is crucial to address several interdependent, flow-sensitive physical processes occurring simultaneously. These are for example, dispersion, dispersion kinetics, coalescence, and drop suspension. Scale-up is process dependent, and in this study, only turbulent systems are discussed. Therefore, the stirrer Reynolds number ( $Re = \rho n d^2 / \eta$ ) must be greater than  $10^4$ . Under these conditions, the rates of drop deformation, collision and breakage, and also the coalescence determining film drainage are governed by the small scale turbulence structures. These factors are somewhat insensitive to tank size [3] if a certain diameter is exceeded. However, the amount of time that drops spend in the high-shear and quiescent regions depends on the mean circulation time, which is governed by impeller pumping rate and macro scale turbulence phenomena. These rates are strongly influenced by tank size [4].

Many processes have been scaled successfully using equation (1). This simple rule contains different approaches for exponent  $Z$  between 0.0 and 2.0. Some major examples are given in Table 1.

$$nd^Z = \text{const.} \quad (1)$$

These rules emphasize different demands in liquid-liquid processes, such as polymerization (dispersion, heat transfer or mixing time), and the assumption of turbulent flow. Additionally, the demand for geometrical similarity is very important for the proper use of these scale-up rules.

**Table 1 – Example scale-up rules for turbulent liquid-liquid systems based on equation (1)**

value of exponent Z	Scale-up criteria
Z = 2.0	constant Reynolds number (Re = const.)
Z = 1.0	constant stirrer tip speed ( $w_{\text{Tip}} = \text{const.}$ )
Z = 2/3	constant power input per volume ( $P/V = \varepsilon = \text{const.}$ )
Z = 0.0	constant circulation time ( $\tau_C = \text{const.}$ )

Numerous authors have published experimental results to prove one or the other scale-up rules [3, 5-10]. Unfortunately, the results of these studies are not coherent with each other, probably due to the different kinds of systems. Leng and Calabrese [4] mention the different stages of difficulty in scaling-up liquid-liquid systems. Dilute and therefore low coalescing systems, like the ones used by Baldyga et al. [5], are the easiest processes to scale up [4]. The more difficult ones, concentrated systems, involve simultaneous coalescence and breakage, like those used by Sprow [10]. While suspension polymerization processes are always stabilized, only coalescence hindered examples from literature will be taken into account for the comparative analysis.

As already mentioned, Baldyga et al. [5] hindered coalescence by using a dilute system. Only one percent of chlorobenzene was used as the disperse phase in the continuous water phase. They carried out a scale-up of geometrically similar systems by a factor of two (vessel diameter  $T_1 = T_2/2 = 0.15$  m) using  $Z = 2/3$  since they assumed constant power input per volume as the scale-up rule. In conclusion, when scaling up, the constant power per unit volume criterion leads to smaller drops, whereas the constant tip speed criterion leads to larger drops in the large tank to an extent which depends upon the duration of agitation [5]. Upon scale-up, the Reynolds number is increased and therefore the inhomogeneity of turbulence as well. Increasing inhomogeneity causes larger velocity fluctuation in the vicinity of the stirrer and lower velocity fluctuations at high distances from the stirrer. Both changes still cumulate to a con-

stant energy dissipation rate which leads to smaller drops due to the increased breakage in the stirrer region and the domination of this phenomena due to the coalescence hindering.

Özkaya et al. [8] used a highly concentrated emulsion (dispersed phase fraction  $\phi_d = 50\%$ ), performing a full suspension polymerization process stabilized with surfactants. They were analyzing the size of the solid PVC particles at the end of the process. A reactor volume scale-up factor of two ( $V_{R;1} = V_{R;2}/2 = 14 \text{ m}^3$ ) was used in their investigations. They came to the conclusion that constant stirrer speed is a suitable criteria for such applications.

Numerical investigations to analyze the scale-up effects on the drop size distribution of liquid-liquid dispersions in agitated vessels using the population balance equation (PBE) have been carried out by Podgorska and Baldyga [11]. A multifractal formalism describing the fine-scale structure of turbulence, including its intermittency, was applied to express the drop breakage and coalescence frequencies for drops whose diameter falls within the inertial sub-range of turbulence. This model was used to predict the transient drop size distributions of dispersions ( $\phi_d = 0.05$ ) in stirred tanks and to test three out of the four methods of scaling-up agitated vessels named in Table 1 ( $Z = 0, 2/3$  and  $1$ ) [11]. They differentiate between slow and fast coalescing systems in their analysis. Scaling-up results in simulations performed for the slow-coalescing system (Hamaker constant  $A = 10^{-22} \text{ J}$ ), thus forming smaller drops when employing a constant energy dissipation rate and circulation time. Larger drops are formed when using constant impeller tip speed. They conclude that there is no valid simple scaling-up method. The presented model helped them only to predict a trend of the scaling-up influence on the drop size. To enlighten this difficult situation, comparative experimental and numerical analysis in two different scales will be carried out in this study with a slow coalescing system.

## 2.2 Multi stage stirrer

The standard single impeller stirred tanks are often criticized for the uneven distribution of shear, energy dissipation or heat transfer, all of which are crucial parameters for liquid-liquid processes such as polymerizations. Multiple-impeller systems should generate a more homogeneous distribution of shear, energy and heat transfer than a single impeller for the same power input. For  $H/T$  ratios at least larger than two, multiple impeller systems are strongly recommended with one impeller used per  $H/T = 1.0$  compartment [4, 12]. Increasing reactor volume by increasing liquid level up to  $H/T$  ratios of four or five with a single impeller system demands an extreme increase of the impeller speed for a fully dispersed reactor. This leads to a high power input with significant energy costs. These costs could be saved with a multiple impeller system.

Gogate et al. [12] provide an excellent overview of the existing work on multiple impeller systems with a special emphasis on bioreactors. They compare and recommend literature sources on several reactor characterizations. Impeller design and choice are critical for main-

taining proper liquid-liquid processes. Any impeller in a vessel capable of pumping fluid and providing shear can produce liquid-liquid dispersions. The impeller-system needs to fulfill the demands caused by the change of viscosity and the crucial need for sufficient heat transfer during the whole polymerization process. An overview of existing stirrers can be found in several literature sources [12, 13].

### 2.2.1 Flow regimes and mixing

The results for the flow field in a multiple impeller system with its overlaying macro structures are mostly analyzed for dual impeller systems [12, 13]. As each impeller produces its own flow pattern, the coupling becomes more intricate and the regularities in single impeller systems are no longer valid. Detailed experimental or numerical results of the fluid dynamics in systems with more than two impeller stages are relatively rare. Due to the extreme rising number of combination possibilities (impeller type and stage distance) these results are very specific [14, 15]. There is a major need for quantitative results of the flow pattern and regimes of multiple impeller systems depending on impeller type and impeller clearance. The same is true for the fluid mixing, especially taking into account the advanced possibilities of computational fluid dynamics (CFD).

Most of the mixing time results have been achieved for dual impeller systems with  $H \leq 2T$ . For these liquid level heights, the results of the single impeller system were always lower than those of the multiple impeller system. A multiple impeller system can be considered as a number of independent circulation loops, promoted by each impeller. The total mixing time is dependent on the rate of transfer between these loops [12]. While the mass-flow rate between the stirrer compartments is significantly lower than the circulation flow rates induced by the impeller, for vessel sizes with a certain H/T ratio the single impeller system is faster than the multi impeller system. Gogate et al. [12] recommend the use of equation (2) originally published by Vasconcelos et al. [16] with a high degree of confidence for the prediction of mixing time in multiple impeller systems.

$$\theta_{95} = c_1 \cdot \left(\frac{P}{V}\right)^{-1/3}, \quad c_1 = 1.5 \cdot 10^2; \text{ recommended value by Vasconcelos et al. [16]} \quad (2)$$

It calculates the mixing time for a 95% degree of mixing in terms of the specific power input P/V. The value of the constant c depends on the type and geometry of the impeller. The recommended value has been given as an optimized value, considering all the impellers used in their study.

### 2.2.2 Power uptake

While the scale-up will take place under fully turbulent flow, only these conditions will be taken into account for the power uptake analysis. In this flow regime, the power number Ne is

found to be constant and specific for every type and geometry of the impeller. For multi stage systems this needs to be expanded by the number of impellers and their clearance.

Generally it is assumed that the  $Ne$  number increases linearly with the number of impellers such that the multiple impeller system's power consumption is the sum of each individual impeller [15]. This has been reported by many authors and is summarized in the works of Armenante et al. [17, 18]. Their work should always be considered as a basis for the analysis of power uptake in multiple impeller systems. On one hand, they have investigated radial impellers (disk turbines DT) with up to a triple impeller system. On the other hand, they studied axial impellers as up to three downward pumping  $45^\circ$  six-blade pitched-blade turbines. The results for the two different types of stirrers are contradictory. The multi stage disk turbines showed a strong influence of the interspacing between the stirrer stages on the power number. Only the power uptake of a single disk turbine is dependent on the bottom clearance. For axial impeller systems they showed that there was no influence of either parameter on their power uptake.

Armenante and Chang [17] found an increase of 20 percent in the power uptake under  $H/T = 1.0$  by increasing the impeller clearance  $s$  from  $0.6d$  up to  $2d$  of a double DT system. For the triple system at  $H/T = 2.0$  with a flat bottom vessel,  $Ne$  has its minimum at  $0.6d$ , runs through its maximum at an impeller clearance of  $2.5d$  (25 % increase) and goes down again (20 % decrease). The maximum increase of  $Ne$  by the bottom clearance was almost 70 percent. Gogate et al. [12] summarize existing work from literature as follows: power consumption behavior of multiple-impeller systems seems to be a function of the impeller clearances, at least for lower liquid level ( $H/T \leq 2.0$ ). The power number increases with increasing distance between the impeller stages. For certain critical impeller clearance, the power number turns out constant and can be estimated by simple addition of individual power numbers. For low distances ( $s < 0.7d$ ), the total power is considerably less (30-50 %) than the additive estimation.

Unfortunately, no review is given about the works on investigations of liquid-liquid systems with multiple impeller systems. Only works directly related to polymerization processes present results of liquid-liquid systems with multi stage (mainly dual) impellers [19]. As known to the authors, no direct comparisons between drop sizes of single and multi impeller systems have been published in detail yet. In contrast to this is the amount of available studies on bubble sizes in stirred bioreactors. CFD simulations coupled with PBE were employed to study flow field conditions and optimum operation conditions in various applications [20].

### 3 MATERIAL AND METHODS

#### 3.1 Experimental set-up

The major objectives of the following investigations are to test the usability of multi stage stirrers for dispersing liquid-liquid systems occurring in polymerization processes and to provide scale-up criteria for the selected multi stage stirrer based on experiments and tests with a PBE model. Figure 1 and Table 2 show the main features of the selected stirred tanks used for the comparative scale-up experiments. To ensure geometrical similarity of the set-ups, main dimensions for both the pilot plant vessel ( $T = 460$  mm) and the lab scale vessel ( $T = 155$  mm) were provided with tolerances lower than three percent.

**Table 2 – Dimensions of the used stirred tanks and stirrers**

vessel type	T	$H_{\max}/T$	$l_B/H$	d/T	h/d	$h_{St}/d$
lab scale	155 mm	5.0	$\approx 0.95$	0.6	0.65 & 1.8	0.06-0.24
pilot plant scale	460 mm	2.8	$\approx 0.95$	0.6	0.65 & 1.8	0.06-0.24

The used maximum liquid level in part one ( $H/T = 2.3$ ) was increased ( $H/T = 2.8$  for the pilot plant vessel and even up to 5.0 for the lab scale vessel). The baffle length  $l_B$  was kept constant at  $l_B/H = 0.95$  (fully baffled) for all experiments. The number of the used baffles was always two with a width of  $0.08T$ . As a consequence of the multi purpose demands on the stirrer system fulfilling all needs for a polymerization process, a flat blade impeller was chosen for the multi impeller system. This impeller combines the strengths of producing a radial flow with high turbulence and has a unique simplicity in its geometry. The diameter ratio of  $d/T = 0.6$  and a blade thickness of  $0.02 \cdot d$  were kept constant for all experiments as the constant geometry parameter. The stirrer height  $h_{St}$  could be varied from  $h_{St}/d = 0.06$  to  $0.24$  by the number of used blades. However,  $h_{St}$  here determines the cumulative height of all used stirrer blades in one application. They could be arranged at the stirrer axis in cross layering or super-imposed, always with adjustable interspaces  $s$ . They were always equal over the different stages in one application. The maximum distance between the top agitator and the bottom agitator was  $5.7 \cdot d$  for a four stage impeller at  $H/T = 5.0$ . Minimum clearance between stirrer and vessel bottom  $h$  was  $h/d = 0.65$  to avoid any contact between the stirrer and the torospherical bottom.

The same model system that was used in the first part of this study (see [1]) was used for all drop size experiments. All tested scale-up rules depend on geometrical similarity. The considered scale-up parameters are equal power input per volume ( $\varepsilon = P/V = \text{const.}$ ), equal circulation time ( $\tau_C = \text{const.}$ ) or equal impeller tip speed ( $w_{Tip} = \pi n d = \text{const.}$ ). All three parameters

are adjusted by the stirrer speed  $n$  in the lab scale reactor while only one stirrer speed per parameter configuration is carried out in the pilot plant scale vessel. The measurement program for the analysis of the drop sizes is given in Table 3.

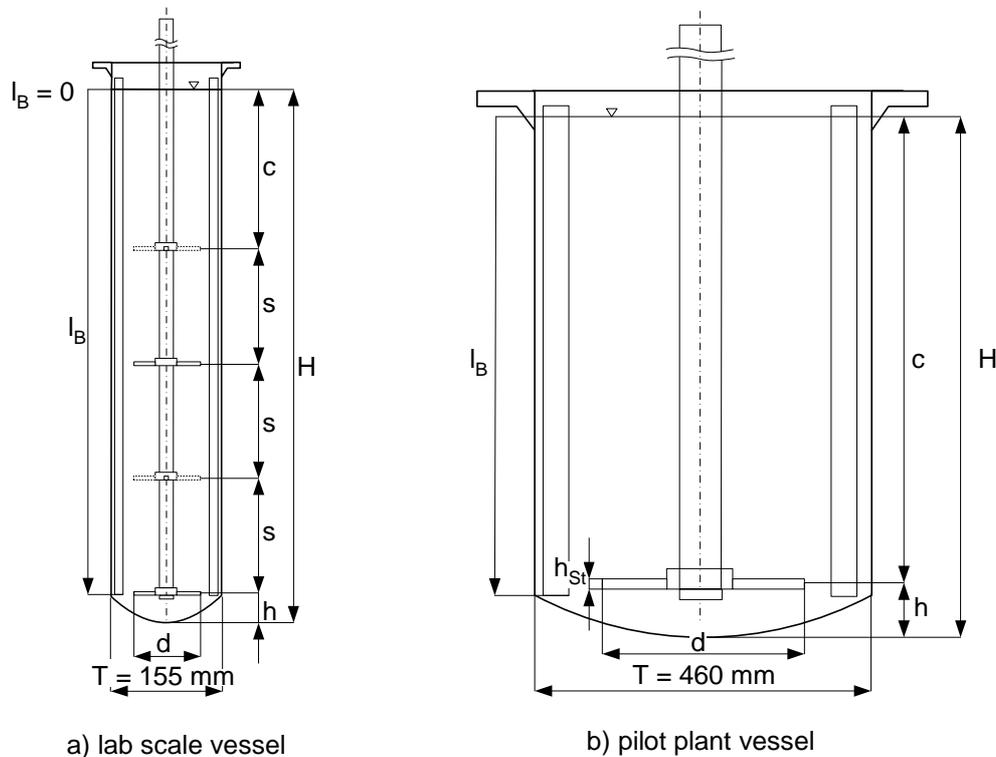


Figure 1 – Experimental set-up and dimensions of the used stirred tanks and stirrers

Table 3 – Overview of the measurement program analyzing the liquid-liquid system

H/T	number of stirrer stages	$h_{St}/d$ [-]	$h/d$ [-]	$\phi$ [-]	pilot plant		Lab scale plant	
					$n$ [rpm]	$n$ [rpm]	$\epsilon = \text{const.}$ [W/m <sup>3</sup> ]	$w_{Tip} = \text{const.}$ [m/s]
1.4	one	0.06	1.80	0.25	200	200	233	5.78
2.8	one	0.24	0.65	0.25	200	-	602	5.78
2.8	four	0.24	0.65	0.25	200	200	368	5.78
4.0	four	0.24	0.65	0.25	-	-	257	-
5.0	four	0.24	0.65	0.25	-	-	203	-

### 3.2 Measurement Procedure

The measurement technique for the power analysis of each set-up is the same as that used in part one of this study [1] and which recommended in the literature for small scale vessels [21]. The transient drop size distributions (DSD) are determined using an in-situ photo optical method [22] with automatic drop detection [23]. The accuracy has been discussed by Maaß et al. [1, 22], where a more detailed presentation of the method used in the present work can be found. This technique was also used to determine the space dependency of the DSD over the height of the vessel.

Mixing time measurements were carried out by a decolorization method [24, 25], in which the reactor contents are first colored by alkali and phenolphthalein and subsequently decolorized again by adding an excess of acid. The time from the addition of acid to the disappearance of the last trace of color inside the vessel corresponds to the mixing time and was determined manually. A five percent excess of acid was used corresponding under the selected conditions to a degree of mixing of 95 percent. This value is called the 95% mixing time  $\theta_{95}$  and is given in equation (3). Upon the conclusion of each mixing time measurement, the actual excess was determined precisely by titration in order to allow a correction [24] to compensate the influence of any possible deviations in concentration. This correction shows the deviation between the desired and the actual stoichiometric excess ( $0.05/\delta$  - see equation (3)). Since the addition site exerts a marked influence on mixing time [26], the acid was always fed at the same point on the liquid surface at the border of the vessel in the middle of the two baffles to gain higher values for the mixing time [27]. All experiments were carried out three times with resulting standard deviations always lower than three percent. To achieve comparability with other vessels, the dimensionless mixing time  $\Theta_{95}$  was calculated out of the stirrer speed  $n$  and the 95% mixing time  $\theta_{95}$  (see eq. (3)).

$$\Theta_{95} = n \cdot \theta_{95} = n \cdot \frac{\theta_{1-\delta}}{1 + 0.56 \cdot \log\left(\frac{0.05}{\delta}\right)} \quad (3)$$

For further evaluations the dimensionless mixing time is set into relation with the specific power input (derived from equation (4)). All experiments were carried out in the lab scale reactor with  $T = 155$  mm. While  $n^3 \cdot d^5$  is kept constant for different stirrer applications, the dimensionless mixing time (independent from  $n$  in the turbulent regime) is only a function of  $Ne$  and  $H$  for the single phase water system.

$$\varepsilon = P/V \propto \frac{Ne \rho n^3 d^5}{HT^2} \quad (4)$$

In addition to the mixing time measurements, the stirrer speed required to reach complete dispersion, also called minimum impeller speed  $n_{min}$ , was also determined visually. Measurements of this parameter are subject to errors, and it is extremely hard to relate published re-

sults [8, 28, 29] with one another. This is due to the large difference between “just dispersed,” where no continuous layer of the dispersed phase remains in the vessel, and “completely dispersed,” where the liquid-liquid dispersion is distributed uniformly throughout the vessel. In a liquid-liquid system this can be quite difficult to detect [21]. Özkaya et al. [8] define for vinyl chloride monomer polymerization the minimum impeller speed which will give way to finish all stages of polymerization. Skelland and Seksaria [28] built an excellent foundation with their experimental work on the minimum impeller speed in liquid-liquid systems. While they investigated a fast coalescing system with 50 percent dispersed phase fraction, they also defined the minimum impeller speed as the rotational speed just sufficient to completely disperse one liquid in the other, so that no clear liquid was observed either at the top or the bottom of a mixing vessel with an H/T ratio of 1.0.

However, the experiments here were carried out to determine the impeller speed where the “just drawdown” condition is achieved. Therefore, only a small amount of n-butyl chloride was used, given on top of the continuous phase. The dispersed phase was blended with a non-water soluble black dye to increase visibility. Thus it was possible to determine the stirrer speed which is able to draw down the lighter phase into the continuous phase. The analyzed process is running under low coalescing conditions, so suspended drops will break and their daughter fragments will stay stable.

### 3.3 Population balance equation

In general, the drop size distribution in liquid-liquid dispersions in stirred batch vessels is mainly affected by coalescence and breakage of drops. The population balance equations define how populations of separate drop sizes develop under specific properties of the system over time. Therefore, they belong to a subcategory of equations known as integro differential equations.

The technical details for using and solving those equations for this study can be found in the first part of the paper [1]. There, a brief introduction into population balance equation and about its interconnections with the inhomogeneous flow is given. A major result from the first part of the paper is the necessity of a two-zone model due to the extreme inhomogeneity of the energy dissipation in slim reactors. Therefore the two-zone modeling approach will be considered for further simulations of drop sizes in the investigated vessels. The used model from literature is the frequently quoted one from Coualoglou and Tavlarides [30] for both, the breakage and the coalescence kernels. It was used for the prediction of transient drop sizes with excellent results in the first part of the paper [1]. The information of the applications and their two-zone parameters are given in Table 4. All reactors used in the first and second part of this research are listed in this table. The stirrer region volume  $V_{\text{stirrer}}$  assumed to be the volume with high energy dissipation rates is correlated to the type and height of the stirrer ( $V_{\text{stir-}}$

$v_{rer} \sim C \cdot h_{St}$ ). The flow rate between the stirrer and the bulk region is a function of the stirrer speed and the size of the stirrer region. The absolute values of the listed determinants at a stirrer speed of 410 rpm have been achieved with CFD simulations using STAR-CCM+ from CD-adapco. The technical details of the simulations are presented by Maaß et al. [31].

**Table 4 – System information for the two-Zone modeling**

H/T [-]	$V_{reactor}$ [L]	number and type of stirrers	$h_{St}/d$ [-]	$V_{stirrer}$ [L]	$V_{bulk}$ [L]	$V$ [L/s]	reference
1.4	4	one BI	0.06	0.157	3.843	$1.4 \cdot n \cdot V_{Stirrer}$	this study
1.4	100	one BI	0.06	4.13	95.87	$1.4 \cdot n \cdot V_{Stirrer}$	this study
2.3	6.5	one RCI	0.12	0.423	6.073	$1.4 \cdot n \cdot V_{Stirrer}$	Maaß et al. [1]
2.3	6.5	one RCI	0.24	0.644	5.853	$1.4 \cdot n \cdot V_{Stirrer}$	Maaß et al. [1]
2.8	7	four CBI	0.24	0.604	6.396	$1.4 \cdot n \cdot V_{Stirrer}$	this study
2.8	207	four CBI	0.24	16.51	190.51	$1.4 \cdot n \cdot V_{Stirrer}$	this study
2.8	7	one BI	0.24	0.604	6.396	$1.8 \cdot n \cdot V_{Stirrer}$	this study
2.8	207	one BI	0.24	16.51	190.51	$1.8 \cdot n \cdot V_{Stirrer}$	this study
4.0	10	four CBI	0.24	0.604	9.396	$1.4 \cdot n \cdot V_{Stirrer}$	this study
5.0	12.68	four CBI	0.24	0.604	12.075	$1.4 \cdot n \cdot V_{Stirrer}$	this study

Due to the physical changes through the change of dispersed phase concentration (in comparison with part I), the structure of the used PBE-kernels will be discussed briefly. The coalescence rate  $F(d'_p, d''_p)$  is assumed to be a two step process involving drop collision and film drainage of the continuous fluid. Hence, it is the product of the collision frequency  $h(d'_p, d''_p)$  and the coalescence efficiency  $\lambda(d'_p, d''_p)$ . The dampening of the turbulence due to the dispersed phase fraction is accounted for in both equations. It leads to lower values for the collision frequency, because less turbulence causes a decrease in the number of collisions. The trend of the coalescence efficiency is reverse - the contact of two particles is stabilized with decreasing turbulence and therefore the efficiency is increased. Rising dispersed phase fraction in a system with other constant parameters induce an increase of the coalescence rate because the order of magnitude of  $\phi$  is three times higher in  $\lambda(d'_p, d''_p)$  than in  $h(d'_p, d''_p)$ . Such results have been shown already by Gäbler et al. [32]. They reported that the change of the dispersed fraction in simulations only qualitatively match the experimental results. The analyzed models

were not able to predict the absolute values of the transient Sauter mean diameter correctly. The results were an under prediction of the increase of the drop sizes with increasing dispersed phase fraction. This led to the assumption that for every different dispersed phase fraction, a parameter fitting is necessary (see Maaß et al. [1] for the equations of the models). PARSIVAL® was used in this work to estimate the free parameters by minimizing the difference between calculated and experimental values of the drop size distributions for every measured time step via a damped Gauss-power method [33]. The transient drop size distribution can be computed with the estimated parameters. Those were estimated based on one set of experimental data and their validity was tested by applying on other sets of data.

## 4 RESULTS AND DISCUSSION

### 4.1 Pre-Investigations

The advantages of multi stage stirrers and their usability for dispersing liquid-liquid systems occurring in polymerization processes seem to be obvious. To prove this assumption, some basic experiments (mixing time, minimum dispersion speed) on fluid mixing in slim reactors have been executed as comparative analysis between single and multi stage systems. Power consumption of different stirrer applications has been determined to judge the results on the basis of energy efficiency.

#### 4.1.1 Power uptake

Power number is not only a clear criterion for the energy consumption of a stirred system but also the basis for accurate prediction of drop sizes. Therefore it was necessary to investigate all stirrer settings used for the dispersion experiments. The results of turbulent power numbers ( $Re$  used here was always  $\geq 2 \cdot 10^4$ ) for different single impeller systems are presented in Figure 2. Two different applications have been studied for an increasing stirrer height – parallel and crossed blades. The development of  $Ne$  is as expected: increasing stirrer height leads to increasing  $Ne_{\text{turb}}$  as already shown by Bates et al. [34]. This is more intense for the parallel blade impeller (PBI) than for the crossed blade impeller (CBI) due to the stronger flow resistance of the PBI. Interestingly, doubling the impeller height ( $h_{\text{st}}/d = 0.06$  to  $0.12$  to  $0.24$ ) always leads to an increase of the power number by a factor of 2.3 for the PBI and 1.7 for the CBI. Therefore, the increase of the power uptake by doubling the impeller height of the crossed blade impeller has the same magnitude of the retreat curve impeller, which was studied in the first part of the paper [1].

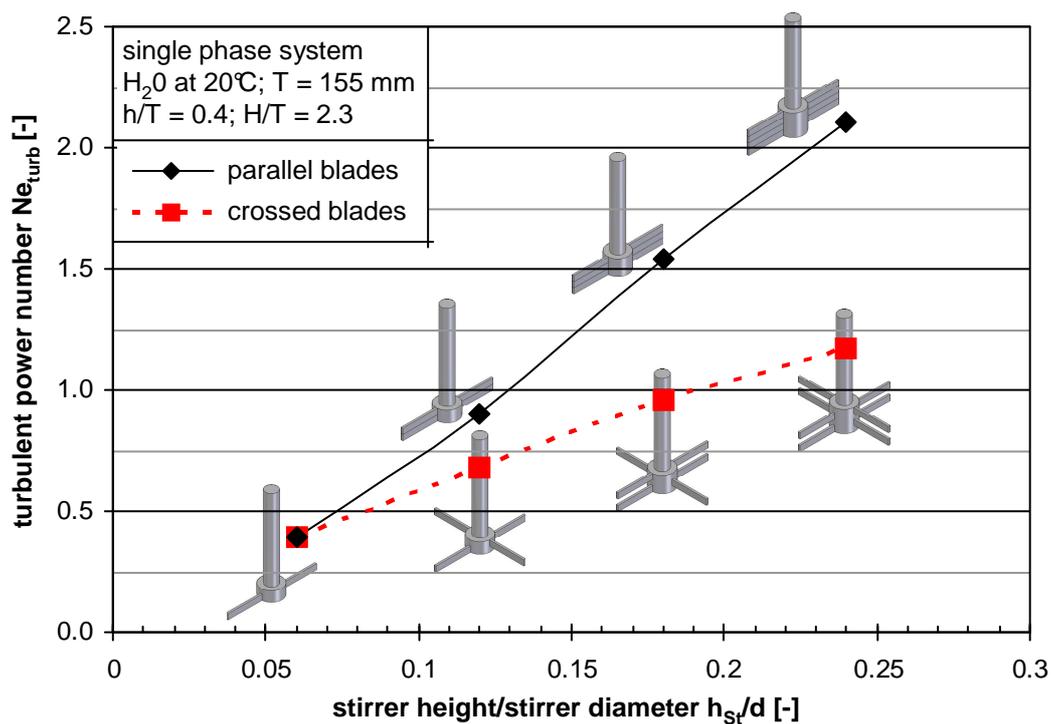


Figure 2 – Development of turbulent power numbers for increasing stirrer height of single stage stirrers.

Figure 3 shows the results for multi stage blade impellers as a function of the impeller clearance. Note that neighboring blades have been installed perpendicular to each other due to increased mechanical stability. For comparison reasons, the standard impeller from the first part of the paper [1] and a single blade impeller with  $h_{St}/d = 0.24$  are also presented in Figure 3. Two results are significant: The increase of the stirrer clearance  $s$  shows only a slight increase on  $Ne_{turb}$ ; a single stage stirrer with the same stirrer height as the cumulative stirrer height of a multi stage system always has a significant higher power uptake. The bottom clearance was kept constant for all investigations at  $h = 0.4T$  to avoid influence on the power consumption [12, 17]. The maximum stirrer clearance was defined by the point where still no surface aeration occurred.

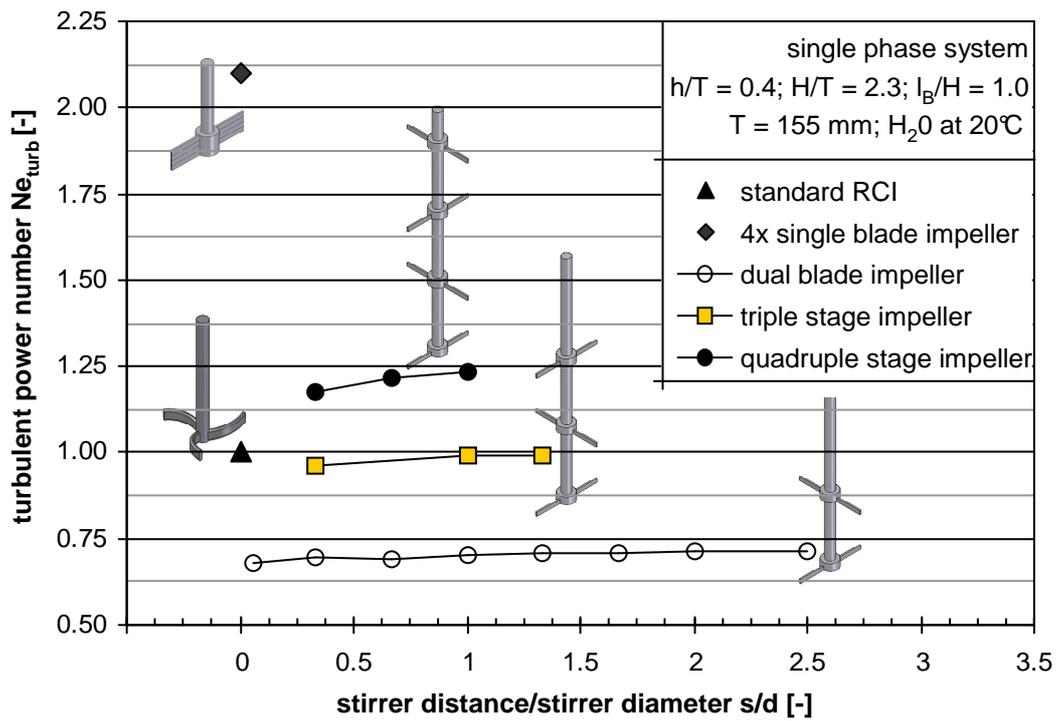


Figure 3 – Influence of the stirrer distances of crossed blade impellers on the turbulent power number.

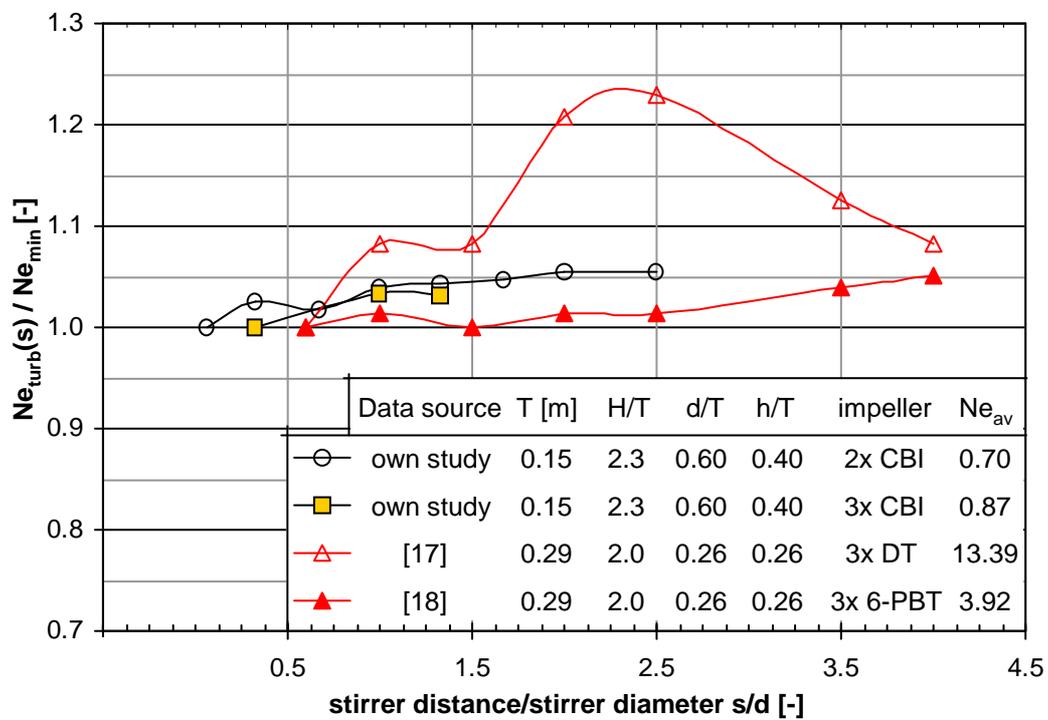


Figure 4 – Comparative experimental results on the influence of the impeller clearance from Armenante et al. [17, 18] and own results.

Every measurement was repeated five times; the standard deviations of such a set of experiments were always lower than two percent for the single and for the multiple impeller systems. A comparison with data from literature is plotted in Figure 4. Armenante et al. [17, 18] have studied a triple disk turbine (3xDT) and a triple 45° 6-blade pitched blade turbine (6-PBT). The absolute values of their determined  $Ne_{\text{turb}}$  are much higher than for the blade system. They are given as the average values ( $Ne_{\text{av}} = \sum_n Ne(s)/n$ ) in Figure 4. Therefore, all values have been divided by the turbulent power number resulting from the smallest impeller clearance. That was always the lowest value in every data set. The different developments over the stirrer distance are easy to compare. All four multiple stirrer systems are following more or less the same trend until a stirrer distance of 2.5d. The increase in the power consumption of approximately five percent is true for all systems beside the triple disk turbine. This stirrer configuration reaches increases for the power number of more than 20 percent. Interestingly such an intense rise has also been reported by Markopoulos et al. [35] for a dual Rushton turbine. The decrease of the power number for 3xDT is explained by Armenante et al. [17] with the limitations of the used vessel. The upper turbine had stayed constant at a fixed distance from the liquid surface to avoid aeration. The flow field of the second and the third turbine influence each other at the used H/T ratio. The full power was only drawn if each impeller was far away from the other [17]. To avoid such unspecific dependency in this study, the stirrer clearance  $s$  was always kept equal between all impellers over the shaft. Therefore, the maximum clearance, avoiding aeration, was 1.3d for the triple system and 1.0d for the quadruple system in the used experimental set-up with  $H/T = 2.3$ .

#### 4.1.2 Mixing time

Comparative mixing time experiments have been carried out to evaluate the flow field in a slim reactor. The H/T ratio was increased for different stirrer systems in the lab scale reactor. Results are given in Figure 5. Three different single stage impellers (RCI, BI with  $h_{\text{st}}/d = 0.12$ , BI with  $h_{\text{st}}/d = 0.24$ ) and a triple blade impeller with a cumulative stirrer height of 0.18d have been investigated. The linear increase of the liquid level caused an exponential increase of the dimensionless mixing time, due to the decrease of power input for a constant stirrer speed. This was expected and is already the basis for prediction correlations like the one presented by Vasconcelos et al. [16]. Comparing the experimental results with the recommended correlation from literature, one major result is obvious. The increase of  $\Theta_{95}$  with increasing H/T ratio is much stronger for the experiments than predicted. While the power number stays almost constant with increasing liquid level [1, 36], the reactor volume is dramatically increased. Therefore the power input decrease due to the increase of H/T at constant stirrer speed is not taken into account in the right scheme of magnitudes.

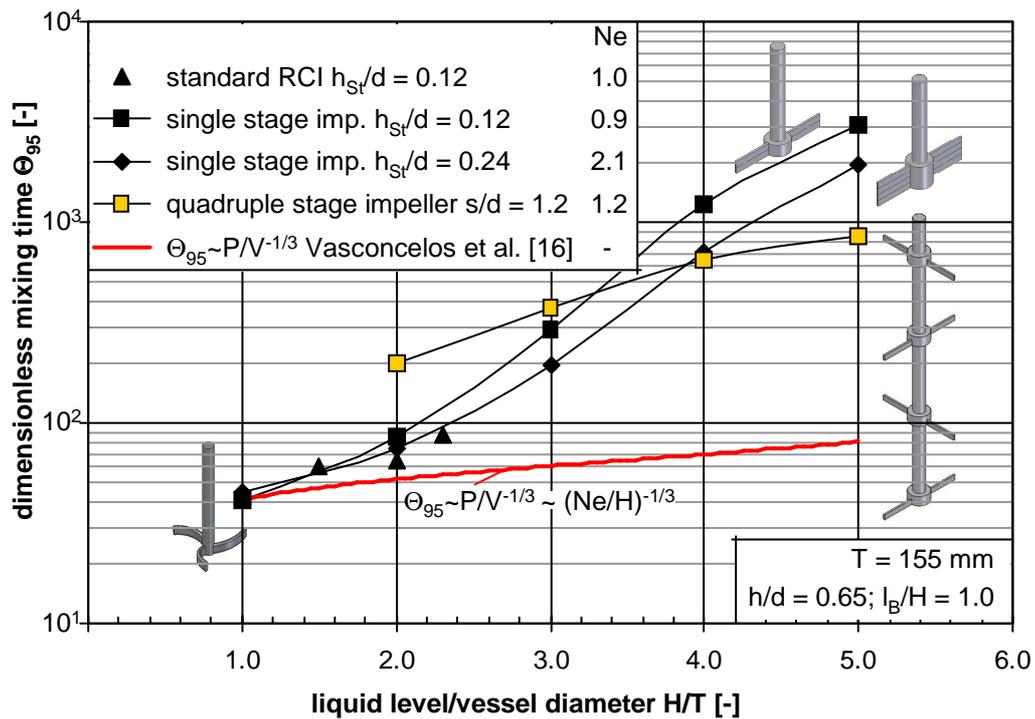


Figure 5 – Comparative results of the dimensionless mixing time for single and multi stage stirrer.

The comparison of the different experimental results shows that for the single stage stirrers those with the maximum stirrer height cause lower mixing times but consume more power (compare the power numbers in Figure 5 and see the dependencies of  $P/V$  in equation (4)). The multi stage stirrer (golden squares in Figure 5), which consumes almost as much power as both single blade impellers with  $h_{st} = 0.12d$  (RCI and BI in Figure 5), is worse than the single stage stirrer for  $H/T \leq 3.0$ . Mixing occurs not only in the vicinity of the stirrer but in the whole flow field by the turbulent fluctuation velocities. The cascaded circulation flow induced by the multiple impeller causes higher mixing times than the single stage impeller. After a critical liquid level, this compartmentalization [37] of the reactor leads to better mixing due to more homogeneous energy dissipation throughout the reactor and therefore a more homogeneous velocity field. The multi stage impellers become much more efficient in slim reactors with very high liquid levels ( $H/T > 3.5$ ).

#### 4.1.3 Minimum dispersion speed

The minimum agitation speed for the just drawdown condition of two immiscible fluids was determined experimentally in a slim reactor for an increasing liquid level up to a maximum  $H/T = 4.5$ . Four single stage blade impellers with different stirrer heights and a triple stage stirrer system have been compared. The results are given in Figure 6. Generally the increase in the liquid level demands an increase in the impeller speed for all investigated applications.

The increase of the stirrer height leads to a decrease of  $n_{\min}$  but also to an increase of power consumption. A critical liquid level can be determined for every used single stage impeller, where the just drawdown condition could not be achieved anymore. This is  $H/T = 2.0$  for the large blade impellers and even lower for the smaller ones. The single impellers have been installed close to the bottom of the vessel ( $h/d = 0.65$ ). The distance between the impeller and the surface becomes too long for the used radial impellers, so no complete dispersion is possible above this liquid level. The triple impeller system is decreasing this distance enormously. The circulation flow between the compartments of every blade impeller transports the dispersed phase slowly but steady down through the whole reactor.

The power consumption of all five stirrer applications is given in addition (see Table 5). The multi stage impeller is able to fulfill complete dispersion with almost the same power as the single BI ( $h_{St} = 0.12$ ) but for a reactor with more than twice the height. These results show clearly the need for multi stage stirrers in dispersion processes in slim reactors. They overcome the problems of missing axial flow close to the liquid surface, which occurs in single radial impeller systems. The circulation flows between the different cascades induced by every stirrer stage provides the axial mixing

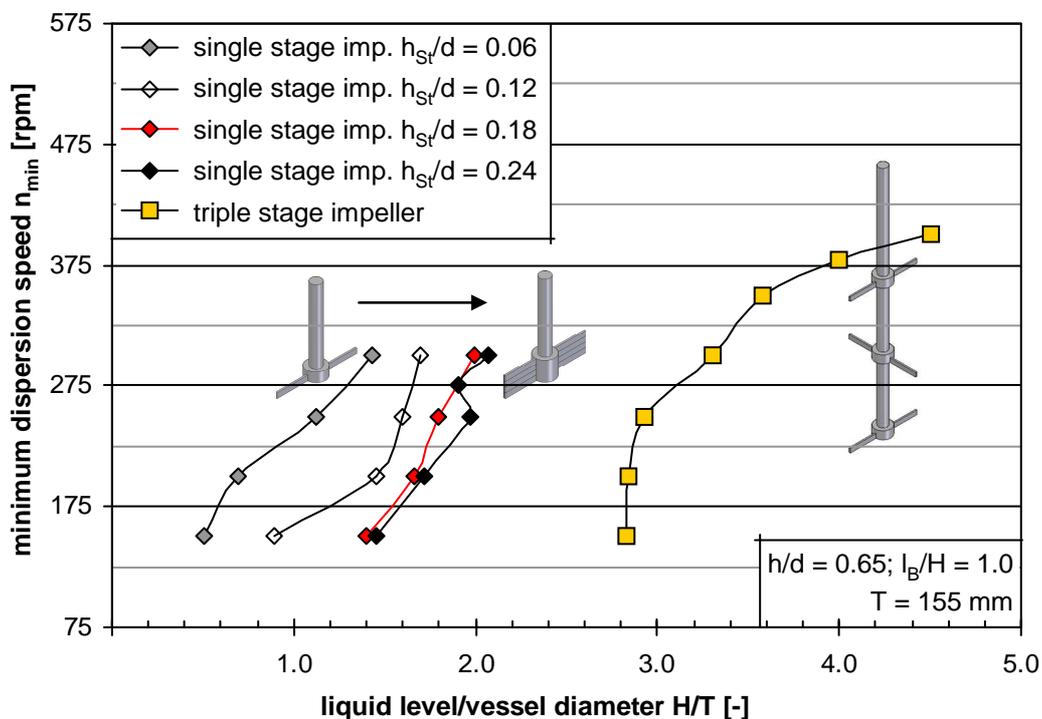


Figure 6 – Minimum impeller speed for various impeller systems at increasing  $H/T$  ratio.

**Table 5 – Power consumption analysis of different stirrer applications for a constant stirrer speed  $n = 300$  rpm and constant stirrer diameter  $d/T = 0.6$  in the lab scale vessel**

number and type of stirrers	$s/d$ [-]	$d/T$ [-]	$h_{Sv}/d$ [-]	$Ne$ [-]	$P(n = 300 \text{ rpm})$ [mW]
one BI	-	0.6	0.06	0.39	339
one BI	-	0.6	0.12	0.90	783
one BI	-	0.6	0.18	1.54	1339
one BI	-	0.6	0.24	2.10	1826
three CBI	1.2	0.6	0.18	1.00	870

#### 4.2 PBE parameter estimation for Sauter mean diameter prediction

Figure 7 shows three different simulation results compared to one experimental set. Both dispersion phenomena (breakage and coalescence) have been tested. After 60 min stirring with 410 rpm, the stirrer speed was reduced to 250 rpm. The experimental results for the 21 h of mixing are represented by the red circles. The drop size decreased within the first 60 min but stayed constant after the speed change. This distinctly shows the absence of coalescence under the chosen circumstances. The amount of introduced surfactant stabilizes the drops so that no size increase could be measured.

The simulation which was carried out with the parameter set from the first part of the publication [1] was not able to reproduce this system behavior. The influence of coalescence is especially overestimated, so that a strong increase in the drop size was simulated after the speed change. This is in agreement to the simulation results reported by Gäbler et al. [32]. Therefore, the coalescence parameter  $c_{2,coal}$  was fitted against the constant drop sizes after 60 min mixing. These simulation results show almost the same behavior for the higher stirrer speed but a constant  $d_{32}$  after the speed change. While the drop sizes are too large for both process steps, the most sensitive breakage parameter  $c_{1,break}$  was fitted against the experimental data. The simulation results are now in excellent agreement with the experimental data. The results for the different parameter sets are compared in the table within Figure 7. The necessity of fitting both the breakage and the coalescence kernel leads to the assumption that the influence of the dispersed phase fraction is underestimated in the breakage and the coalescence rate. This optimized parameter set is used for all following simulations in this work.

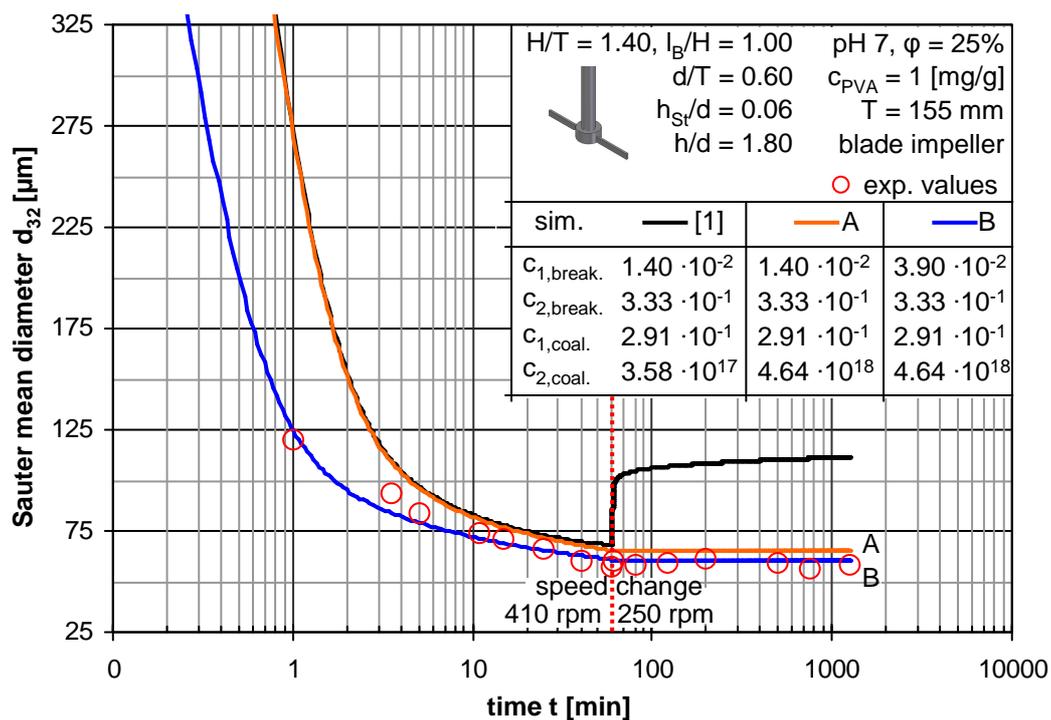


Figure 7 – Comparison of experiments and simulations of transient Sauter mean diameter for different numerical parameter.

### 4.3 Analysis of scale-up effects in stirred liquid-liquid dispersions

The simulations and experiments are used to determine suitable scale-up rules for the presented applications. The considered scale-up rules, equal power input per volume equal circulation time and equal impeller tip speed are adjusted by the stirrer speed  $n$  in the lab scale reactor while only one stirrer speed per parameter configuration is applied in the pilot plant scale vessel. All four configurations are compared with PBE-simulations. Those are carried out with a constant set of parameters, which are presented in Table 6.

Table 6 – Parameter overview of the used parameter set

	$C_{1,break.}$	$C_{2,break.}$	$C_{1,coal.}$	$C_{2,coal.}$
used parameter set	$3.90 \cdot 10^{-2}$	$3.33 \cdot 10^{-1}$	$2.91 \cdot 10^{-1}$	$4.64 \cdot 10^{18}$

#### 4.3.1 Effect of reactor diameter on drop sizes in liquid-liquid systems

In Figure 8 the effect of agitation speed (i.e., 200, 410, and 600rpm) on the Sauter mean diameter,  $d_{32}$ , is depicted for the used liquid-liquid dispersion. It is evident that the simulation results show a very good agreement with the experimental measurements for all three stirrer speeds. In all cases, the development of the liquid-liquid dispersion is dominated by a signifi-

cant droplet breakage, leading to a decrease in the mean size of the droplets. The results given in Figure 8 are consistent with those reported in the literature for dispersion [2, 5, 7, 10, 32, 38] and suspension polymerization [39, 40] processes. The increase in the agitation speed results in a subsequent increase of the droplet breakage rate, leading to an increase in the number of small droplets.

The results for the scale-up rule of constant power input for different vessel diameters under geometric similarity leads to excellent agreements. The  $d_{32}(40 \text{ min})$  is  $61 \mu\text{m}$  in the lab scale reactor, while it is  $62 \mu\text{m}$  in the pilot plant. Only in the first minute larger deviations are noticeable for the experimental results in the half logarithmic plot. The deviations of the experimental Sauter mean diameter between the lab and the pilot plant vessel are always below five percent after five minutes of stirring. In the initial phase the drops in the pilot plant vessel are larger due to the higher residence time of the dispersed phase in the bulk region in larger tanks. The simulations show almost no difference between the pilot plant and the lab scale reactor and are in marvelous agreements with the experimental results.

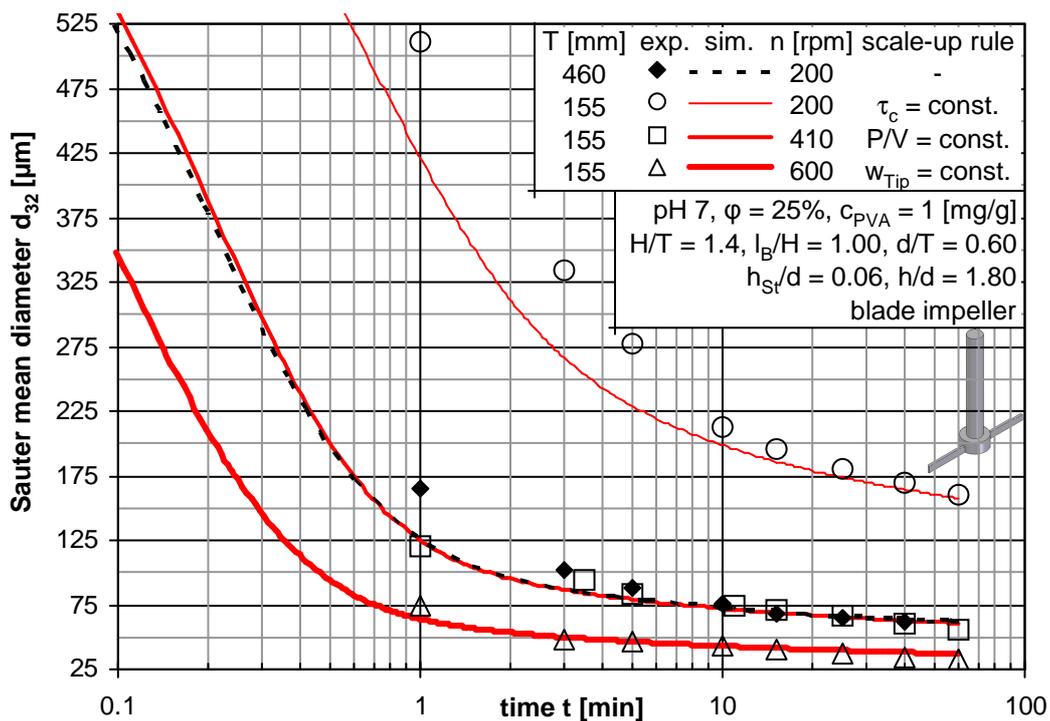


Figure 8 – Comparative results between simulation and experimental results of transient Sauter mean diameters.

The scale-up rule of constant circulation time leads to too small drops, while the constant stirrer tip speed leads to too large drops under scale-up conditions. The Sauter mean diameters

after 40 min mixing are now compared. For  $\tau_c = \text{const.}$ ,  $d_{32}(40 \text{ min})$  is 170  $\mu\text{m}$  in the lab scale reactor, while it is 62  $\mu\text{m}$  in the pilot plant. For  $w_{\text{Tip}} = \text{const.}$ ,  $d_{32}(40 \text{ min})$  is 34  $\mu\text{m}$  in the lab scale reactor, while it is 62  $\mu\text{m}$  in the pilot plant. Note that the results here are shown for a scale-down process (different scale-up rules are used for the smaller sized vessel, while only one set of experiments was carried out in the large vessel).

The minimum dispersion speed was barely fulfilled for the 200 rpm in the lab scale vessel. The simulated drop sizes are decreasing always faster over time than the experimental ones. The simulation starts with a fully dispersed system and a certain initial distribution over the whole reactor (see part one of the study for details [1]). This is a necessary assumption for the PBE, which is not in accordance to reality. The dispersion needs time to take place in the initial phase of the process. This is shown by the experimental results, which are slower over time than the simulation results, especially for the lower agitation speeds.

The filling level was doubled and the stirrer height was increased fourfold for the following experiments. Figure 9 shows the results for both reactor sizes. The constant power input per volume again leads to excellent accordance in both vessels. The minimum dispersion speed for the constant circulation time was not fulfilled with 200 rpm in the lab scale vessel because of the smaller bottom clearance than in the experiments shown before in Figure 8. The scale down process under the assumption of constant stirrer tip speed (5.78 m/s) leads to drops that are too small. The experiments could be reproduced with the PBE-simulations using the two zone model approach, at stirring times larger than three minutes. Again, the assumption of full dispersion already in the initial state of the simulation leads to all too low prediction values in the beginning of the process. This general deviation should always be considered when interpreting the results of PBE simulations.

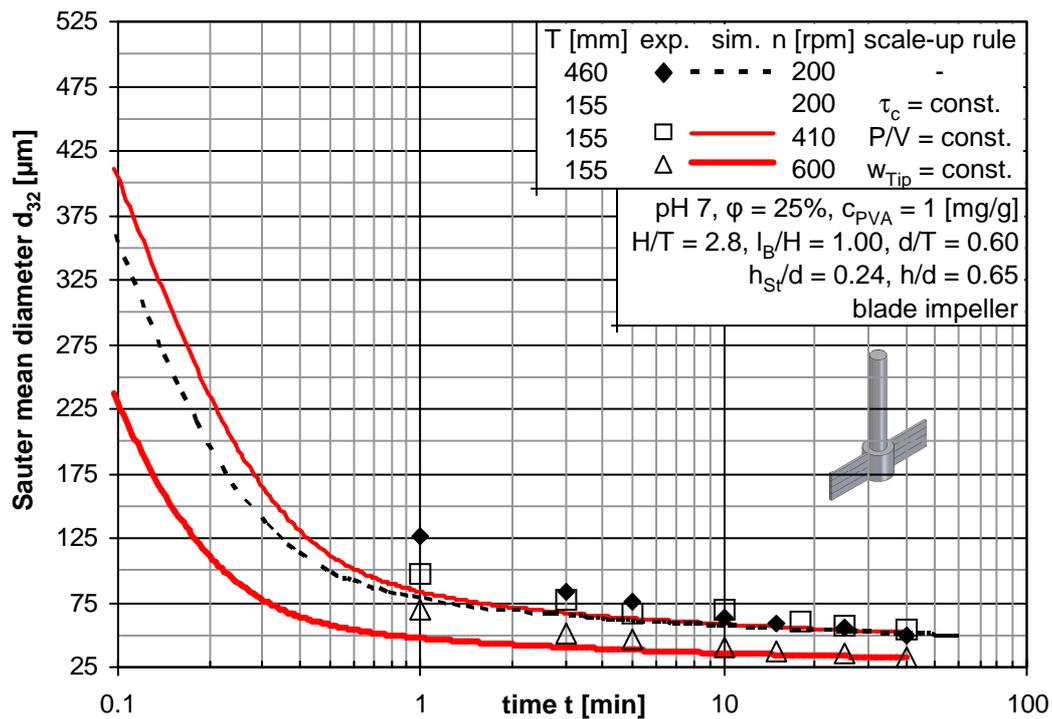
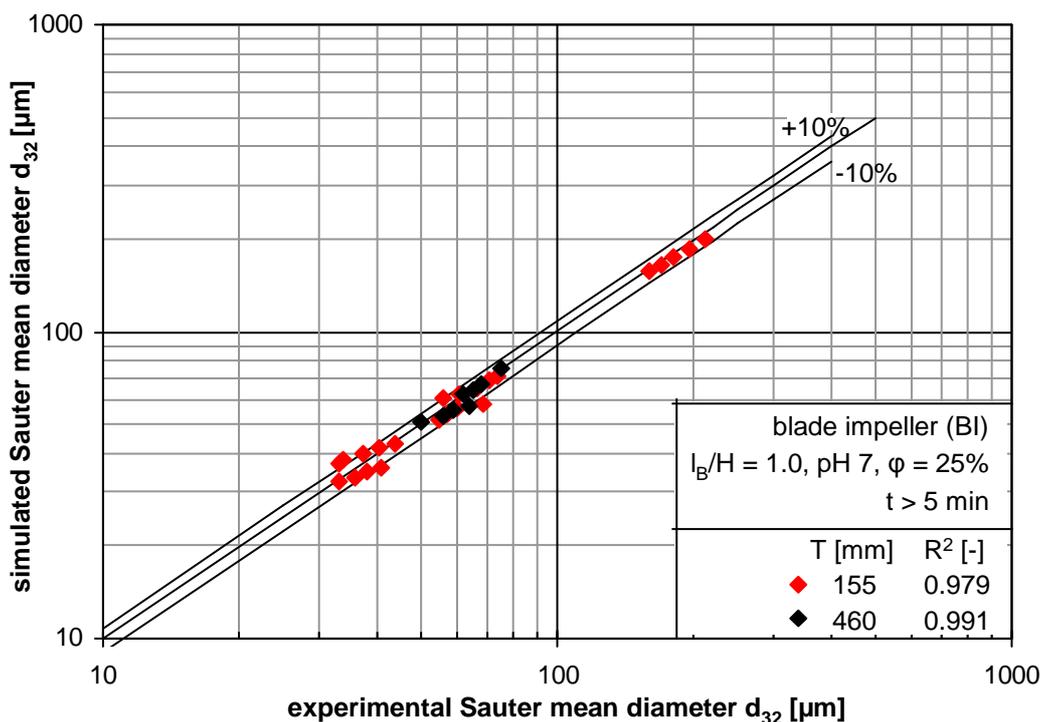


Figure 9 – Scale-up of liquid-liquid dispersions for constant power input and stirrer tip speed.

For concluding remarks on the scale-up of single stage impeller processes, all simulation results for stirring times larger than five minutes are summarized in a parity plot (see Figure 10). To differentiate the prediction results, they are split up based on the two used vessels (red diamonds –  $T = 155$  mm, black diamonds –  $T = 460$  mm). Both groups of values are astonishingly well predicted. The coefficient of determination was calculated for both. The results are given in Figure 10 and prove the quality of the prediction results. The two-zone PBE simulations are able to reflect the influence of various parameters on the transient behavior of drop sizes in stirred, slim reactors with high accuracy: stirrer speed, vessel diameter and liquid level. This complements the results from the first part of the paper and extends them on the influence of the vessel diameter. The next development stage in the modeling process is the extension from single to multi stage impellers, shown in Figure 11.



**Figure 10 – Deviation between experimental and predicted  $d_{32}$  with the two-zone PBE simulations at different stirrer speeds (200 - 600rpm), H/T ratios (1.4 & 2.8) and different BI's ( $h_{St} = 0.06$  and  $0.24$ ).**

No clear scale-up rule could be found out of the experimental results for the multiple impeller system. Under scale up conditions, the constant power input ( $368 \text{ W/m}^3$ ) leads to all too small drops, and the constant stirrer tip speed ( $5.78 \text{ m/s}$ ) to all too large drops. The experimental results for the constant circulation time (equal to constant stirrer speed – 200 rpm) show a high fluctuation over the time. The different impeller stages over the reactor height are building compartments. This leads to a decrease in the axial flow compared with a single stage impeller with an equal cumulative stirrer height. This compartmentalization delays the dispersion process especially for the lower stirrer speeds. The drop sizes were measured between the lowest and second lowest impeller stage. The complete dispersion, the state in which the complete organic phase is introduced into the aqueous phase, was achieved after 20 min stirring for 200 rpm and after 10 minutes for 410 rpm. The layer of n-butyl chloride was not completely withdrawn until those points in time. This leads to larger drops as expected by the simulations, because large drops are continuously unhinged out of the dispersed phase film at the top of the reactor. This explains also the increase of the drop size over time for 200 rpm. The consistent decrease of the Sauter mean diameter over time started after the complete dispersion was achieved. However, it seems that the steady state size range of the simulations and the experiments are in the same order of magnitude. The simulated size decrease is again much faster over time due to the assumption of complete dispersion at the beginning of the

process ( $t = 0$  min). Aside from these difficulties, all simulation results for the lab scale reactor are in very good agreement with the experimental results. The constant power input scale-up rule would be the presumption out of the PBE-simulations but could not be confirmed in the experiments.

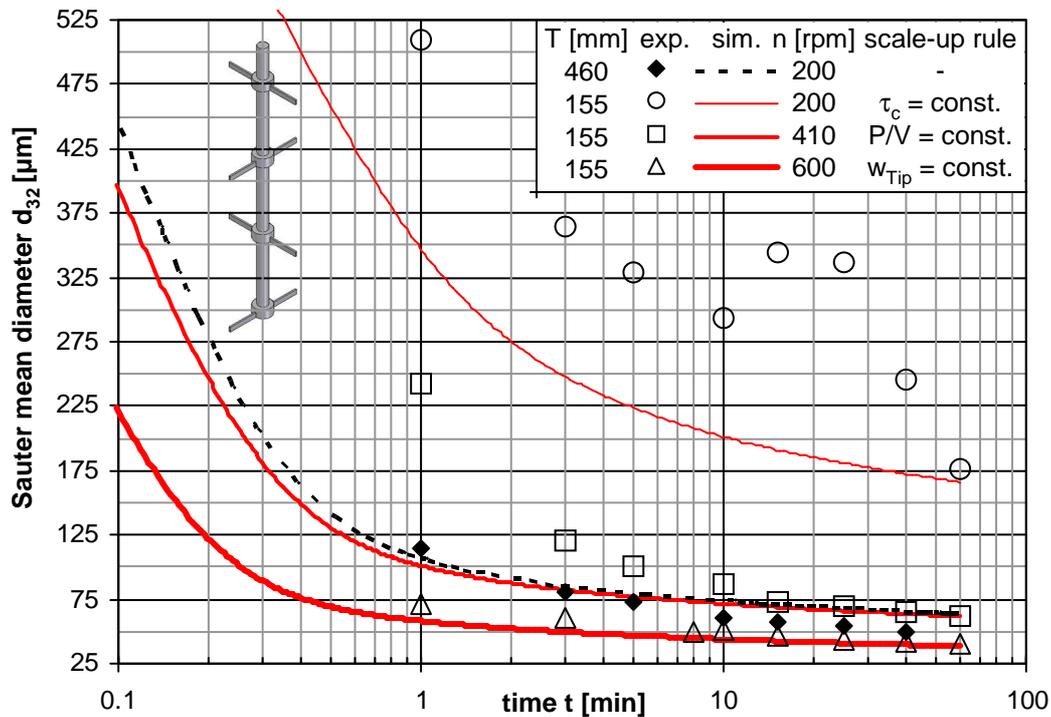


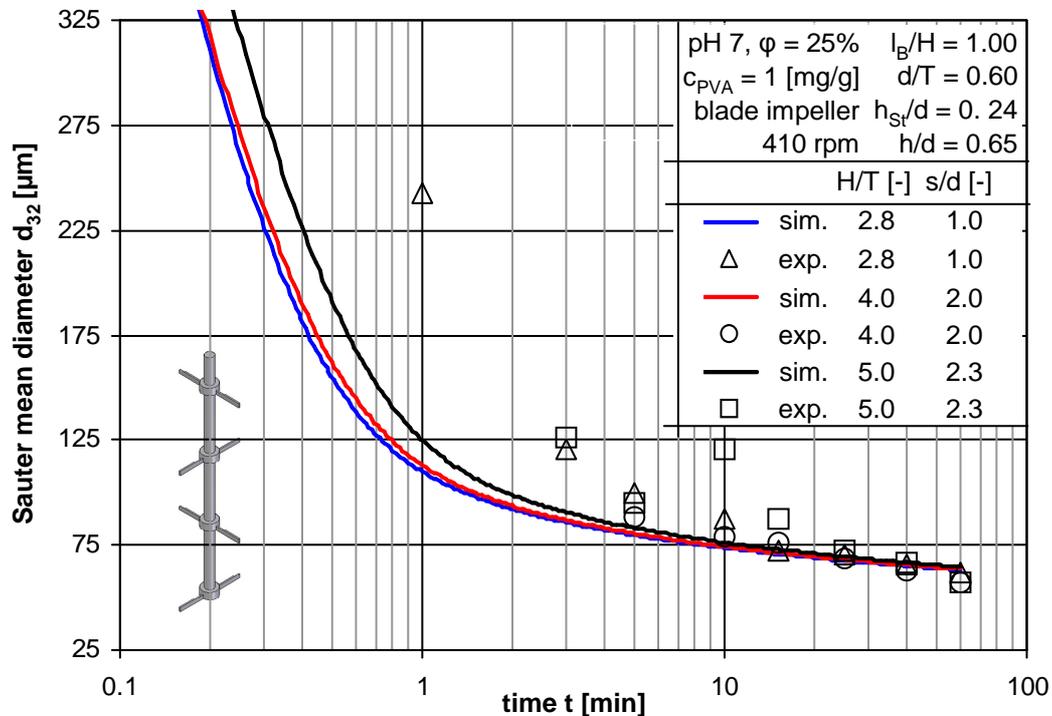
Figure 11 – Scale-up of liquid-liquid dispersions for multiple impeller system. The geometrical and chemical data are the same as in Figure 9.

#### 4.3.2 Effect of filling level on scale-up in multiple impeller systems

It appears that no work in the literature is available which delineates the effect of increasing filling level above an  $H/T$  ratio of 3.0 on the drop sizes. Experiments were conducted for three different  $H/T$  ratios (2.8, 4.0 and 5.0). The stirrer speed was kept constant at 410 rpm. The experiments were compared with each other but also with connected 2-zone PBE simulations. The results for both the experimental and simulated Sauter mean diameters are presented in Figure 12.

Again, the initial condition of complete dispersion in the simulations are leading to an under prediction, especially at the beginning of the stirring process. The simulations and the experiments are in good agreement after 5 min of mixing. The complete dispersion for the higher values was fulfilled after 15 min of mixing. The good agreements between the simulations and experiments after five minutes show that the complete dispersion was almost fulfilled. Although a continuous layer of n-butyl chloride was observed for over 10 minutes on the sur-

face, most of the disperse phase was already introduced into the aqueous phase at this point in time.



**Figure 12 – Influence of filling level on the transient Sauter mean diameter at constant stirrer speed for the quadruple stirrer system (exp. and simulation)**

For a more detailed comparison, the local drop size distributions are analyzed in Figure 13 with a simulation result. Two major effects can be seen: Firstly, the parallel recorded distributions from two different local positions (sampling points A & B) are almost equal, after 10 and also after 60 min. That underlines the assumption that after 10 min the complete dispersion is almost fulfilled. Secondly, the simulated distribution is much narrower than the experimental one. The parameter optimization of the PBE-models was always run for a Sauter mean diameter fit, which is much faster to achieve than for the whole distribution. Furthermore, the initial condition of complete dispersion leads to a more narrow distribution. All the drop building processes from the continuous layer on top of the reactor are neglected. That leads to a lower number of large particles in the distribution. This is balanced by a lower number of small particles to achieve the same Sauter mean diameter as the experimental results. However, these simulation results still give a reliable base for the scale-up of liquid-liquid processes. The major tendencies in the system are described with deviations smaller than five percent for the Sauter mean diameter.

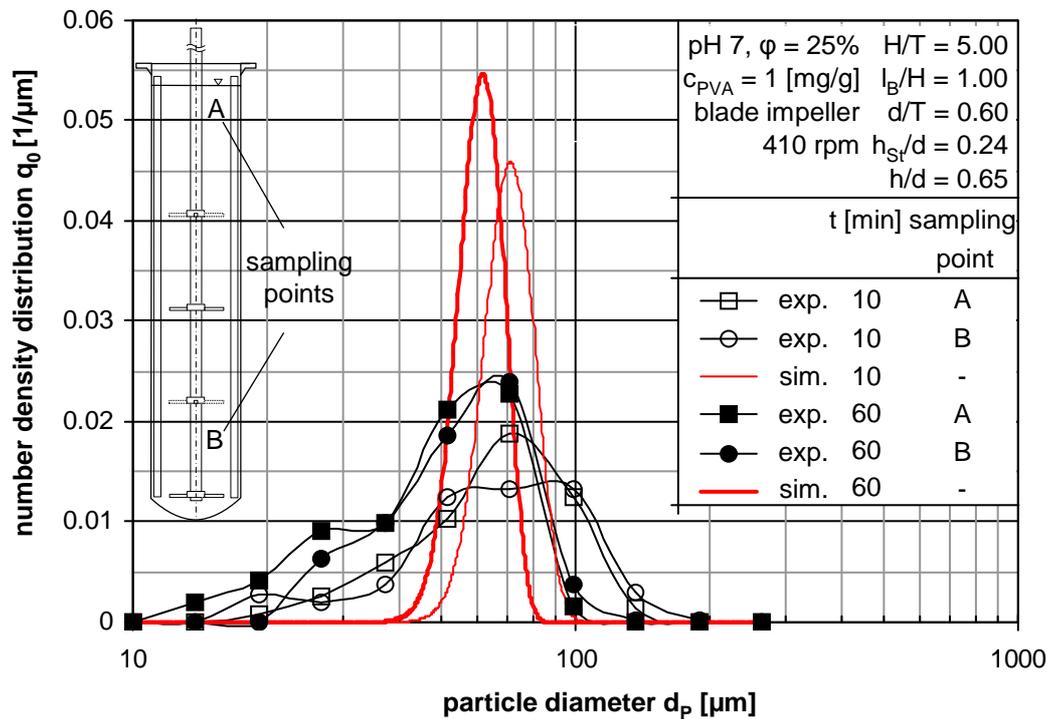


Figure 13 – Influence of the measurement point on transient drop size distribution.

A detailed evaluation of the prediction quality of the used model for the multi stage impeller is given by a parity plot (see Figure 14). All simulated Sauter mean diameters for process times longer than five minutes are shown over the corresponding experimental value. Values for shorter times were not taken into account due to the discussed problems occurring during the initialization of the dispersion process. The data are subdivided into three categories. Firstly, results for the lab scale reactor at the low stirrer speed of 200 rpm (red squares), secondly, the data from the lab scale vessel for the higher stirrer speeds of 410 and 600 rpm at different H/T ratios (red filled diamonds) and thirdly, the results of the pilot plant vessel at 200 rpm (black diamonds). The resulting coefficient of determination is only acceptable for the second data category, the variation of H/T ratio (2.8, 4.0 and 5.0) and the stirrer speed (410 and 600 rpm). The prediction for the low stirrer speed lab scale vessel (used for the scale-up rule of constant circulation time) cannot be described by the simulations. The fluid dynamics are overemphasized by the use of the model with the given boundaries. Although the experimental and simulation values converge after 40 min of stirring, the overall performance of the prediction is poor.

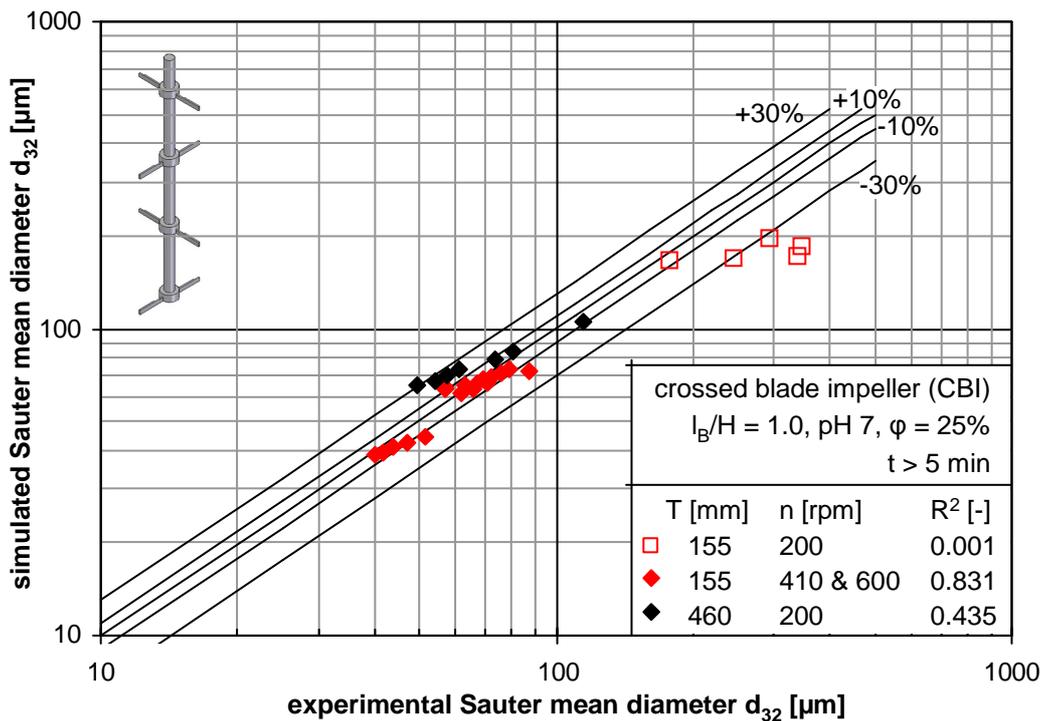


Figure 14 – Parity plot between experimental and predicted  $d_{32}$  with the two-zone PBE simulations at different  $H/T$  ratios (2.8 – 5.0) for a quadruple CBI ( $h_{St} = 0.24$ ) at varying stirrer distances  $s/d$  (1.0 – 2.3)

The agreement between experiments and simulations for the increasing vessel diameter is much better as for the last discussed case. The maximum differences are achieved for the smaller sizes. They are around thirty percent. As already discussed and presented in Figure 11, the simulations are following a different behavior than the experiments. No clear scale-up rule could be used to predict these changes. A more detailed analysis of the flow field induced by multiple impellers seems to be necessary to generate reliable input values for the PBE models. The average energy dissipation divided into two values for the cumulative stirrer and the bulk region is not suitable for this application. A straight numbering up of the stirrer zones according to the number of used stirrers and compartments did not improve the simulation results significantly. However, the calculation costs were increased dramatically. An increase in the number of zones from two to eight, displaying four stirrer regions in four compartments, only led to an increase of  $R^2$  by two percent but to an increase of the computation costs by 7500 percent. The standard two zone simulation takes 3 hours while the eight zone simulation took more than 11 days on a single standard CPU.

## 5 CONCLUSION

In conclusion, the prediction of drop sizes is of major interest for many industrial applications. The predictive capabilities of the used PBE model and the validity of the method were

demonstrated via the successful simulation of experimental measurements on the transient Sauter mean diameter for non-reactive liquid-liquid dispersions. This research is a straight continuation and extension of the first part of this paper [1]. The accuracy of the PBE simulations for single stage impellers are increased (deviation to experiments  $R^2 > 0.99$ ) and transferred to scale-up procedures of such applications. Therefore three different scale-up rules for liquid-liquid systems are tested: the constant circulation time, the constant power input and the constant stirrer tip speed. They are tested in two different sized, geometrically similar vessels. The scale-up rule for constant power input was successful in terms of achieving equal drop size in both vessels. This was shown not only by experiments but also by the PBE simulations (again deviation to experiments  $R^2 > 0.99$ ). While the aspect ratio  $H/T$  is further increased in this research, multi stage impellers are also tested against comparable single stage impellers in terms of power consumption, mixing time and minimum impeller speed. Especially for aspect ratios larger than three, multi stage impellers successfully compete with the single stage ones. The measured drop size distributions in slim reactors with multi stage impellers showed no dependency on the local position, although the dispersion process is tedious due to compartmentalization. The simulations are not able to reflect this initial phase of the dispersion process, but are in excellent (try to avoid using the word very if you can, it tends to be a weak word) agreement with the experiments after complete dispersion is fulfilled. Based on these experiences, the aspect ratio is increased up to five and the resulting drop size can be predicted with reasonable deviations (lower than 10 percent). The results of the scale-up of these multi stage impeller liquid-liquid systems do not lead to a clear conclusion. Although the simulations recommend the use of constant power input, the experiments could not support this, as any other traditional scale-up rule. However, it was possible to increase the reactor volume by a factor of two by increasing the aspect ratio up to five. Parallel the power consumption of the application was decreased by a factor of 1.7. The quadruple crossed blade impeller consumes much less power than the standard retreat curve or the parallel blade impeller with equal blade height. Overall, the results of power consumption, mixing time and dispersion behavior show a great potential of multi stage impellers for process optimization and intensification in slim reactors.

## SYMBOLS

- A - Hamakar constant [J]
- $c_i, C_i$  - numerical or correlation constants
- c - distance between liquid level and stirrer [m]
- d - stirrer diameter [m]
- $d_p$  - drop or particle diameter [m]

$d_{32}$	- Sauter mean diameter [m]
F	- coalescence rate [m/s]
h	- bottom clearance, stirrer height [m]
h	- collision frequency [m/s]
$h_{St}$	- cumulative stirrer height of the single or multiple impeller system [m]
$l_B$	- baffle length [m]
H	- liquid level [m]
n	- stirrer speed [rpm]
Ne	- Power or Newton number [-]
P	- power [W]
Re	- Reynolds number [-]
s	- stirrer distance, stirrer clearance [m]
t	- time [s]
T	- tank diameter [m]
V	- reactor volume [m <sup>3</sup> ]
$w_{Tip}$	- stirrer tip speed [m/s]
$\delta$	- relative deviation of the local concentration [-]
$\varepsilon$	- P/V - energy dissipation rate [W/m <sup>3</sup> ]
$\eta$	- dynamic viscosity [kg/(m·s)]
$\theta_{1-\delta}$	- mixing time of a certain deviation [s]
$\theta_{95}$	- mixing time [s]
$\Theta_{95}$	- dimensionless mixing time [-]
$\lambda$	- coalescence efficiency [-]
$\rho$	- density [kg/m <sup>3</sup> ]
$\tau_c$	- circulation time [s]
$\varphi$	- dispersed phase fraction [-]

#### ABBREVIATIONS

BI - blade impeller

- CBI - crossed blade impeller  
DSD - drop size distribution  
DT - disk turbine  
PBE - population balance equation  
PBI - parallel blade impeller  
PBT - pitched blade turbine  
PVA - poly vinyl alcohol  
PVC - poly vinyl chloride  
RCI - retreat curve impeller

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