

The effect of process conditions on the flocculation process occurring in an agitated vessel

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The effect of flocculation time, flocculant dosage, pollutant concentration, mixing intensity and tank size on flocculation performance was experimentally studied in a mechanically agitated vessel. The macroscopic approach was applied for flocculation modelling. The simple semi-empirical generalized correlation for flocculation kinetics proposed by Šulc (2003a) and the simple semi-empirical generalized correlation quantifying the effect of flocculation time and flocculant dosage on flocculation proposed by Šulc, Dítl (2007) were used for data treatment.

Keywords: flocculation, flocculation kinetics, mixing, Rushton turbine, flocculant dosage, scale up.

INTRODUCTION

Flocculation has very important application in solid – liquid separation processes, and in water supply and wastewater treatment. The primary purpose of flocculation is to transform smaller particles into larger aggregates – flocs that will eventually settle so that efficient separation is achieved. The first step in the flocculation process is a chemical interaction among the added flocculant, the water and the suspended colloidal solids. The second step is to aggregate the destabilized particles into flocs. The first phase is considered to be completed within a few seconds, whereas the second phase may continue for minutes or hours. During flocculation, the mixing of a suspension causes collisions between the particles and the formation of flocs. This brings about the growth of particle aggregates of such size and consistency that effective separation can take place. Increased mixing promotes floc collisions, and in this way their formation and growth are accelerated. However, if the agitation is too intensive, the turbulent shear forces can break the flocs down into smaller particles again. This breakage decreases the separation efficiency. Then, in order to achieve the required separation efficiency, the flocculation agent dosage has to be increased. Flocculation agents are expensive, and the dosage depends on flocculant type, preparation and mixing conditions.

Thomas et al.¹ classified published flocculation models according to modelling approach into the following groups: 1) modelling based on a microscopic approach, 2) modelling based on a macroscopic approach and 3) modelling based on a thermodynamic approach. Modelling based on a microscopic approach is based on the description of microscopic phenomena such as interparticle collisions, particle concentration and flocculation rate. These models can be further classified into models based on Smoluchowski² approach and models based on population balance. However, processes such as precipitation, destabilisation and surface adsorption can be affected by the surface chemistry. According to Thomas et al.¹ the models either ignore the surface chemical effects or dealt with them in a simplified form. The experimental results have often been obtained for idealised artificial particle suspensions such as suspensions of uniform latex spheres. In real systems the presence of

other soluble chemicals, e.g. such as alkalinity, hardness phosphate and humic compounds, complicates modelling. Unfortunately the interactions among these chemicals, the flocculant and primary particles are complex and poorly known. It could therefore be difficult to determine the mutual dependence between the process parameters such as mixing time and intensity, on the one hand, and flocculant dosage and flocculation kinetics, on the other. In addition, flocculation models adopting the Smoluchowski equation are limited by the constraints imposed by Smoluchowski's six principal assumptions and the population balance models have often been conditional on the floc size being less than Kolmogoroff length microscale. The macroscopic approach is a compromise between microscopic phenomena modelling and empirical measurements. For this approach, Thomas et al.¹ proposed that it is necessary to know the relation between macroscopic scales of flocculation, e.g. fractal dimension, and process parameters.

We adopted the macroscopic approach, which is a compromise between the microscopic phenomena modelling and the empirical measurements, and we focused our attention on finding the interrelation between macroscopic scale and process parameters. This paper presents a summary of our results obtained during a long-term research. The effect of flocculation time, flocculant dosage, impurity concentration, mixing intensity and tank size on flocculation performance will be presented here.

EXPERIMENTS

The flocculation experiments were conducted in a fully baffled cylindrical vessel of diameter $D = 150$ mm, filled to height $H = D$ with a model wastewater. The vessel was usually agitated by a Rushton turbine. A 6-pitched blade turbine was used in some of the experiments. The impeller of the diameter $d = 60$ mm was placed at an off-bottom clearance of $H_2/d = 0.85$ mm. The impeller speed was set up and the impeller power input value was calculated using the impeller power characteristics.

The flocs were separated by sedimentation. After the sedimentation running for 5 minutes (for sedimentation velocity $u_{sed} = 0.03$ m/min and given geometry) the clarified water sample was withdrawn. The sampling point was located at the level of the upper impeller edge. The

turbidity of the clarified water sample was measured using MultiLab5 (WTW, Germany). The built-in photometer measured the turbidity using the reference ray method. The aqueous suspension of amorphous SiO₂, clay and kaolin, respectively, was used as a model wastewater.

EXPERIMENTAL RESULTS

Flocculation kinetics

The turbidity is commonly used in practice as a measure of solid pollutant content. Thus flocculation efficiency is frequently defined as the rate of turbidity removal:

$$Z_e^*(t_F) = \frac{Z_e(t_F)}{Z_0} = \frac{Z_0 - Z_r(t_F)}{Z_0} = 1 - Z_r^*(t_F) \quad (1)$$

where Z_e^* is the degree of turbidity removal, Z_r^* is the degree of residual turbidity, Z_0 is the turbidity of the suspension before the beginning of flocculation, Z_e is the eliminated turbidity due to flocculation, Z_r is the residual turbidity of the clarified water after floc separation, and t_F is the flocculation time.

The effect of the flocculant dosage on the flocculation kinetics of kaolin slurry was investigated³ in a baffled tank agitated by a Rushton turbine at mixing intensity 168 W/m³ (corresponding impeller speed $N = 290$ rev/min) and kaolin concentration 0.58 g/l. The flocs were separated by 5-minute sedimentation. The tests were carried out on a model wastewater (a slurry of tap water and kaolin). The model wastewater was flocculated using Sokoflok 16A organic flocculant (solution 0.1% wt.; Sokoflok Ltd. Czech Republic). The experimental data are presented in Fig. 1. With increasing flocculation time, the degree of turbidity removal increases at the given flocculant dosage for $t_F \leq 10$ min. For $t_F > 10$ min, the degree of turbidity removal decreases slightly due to the floc breakage. Maximum turbidity removal 97.6% was observed at flocculation time 10 min and at flocculant dosage 2 ml/l. The simple semiempirical generalized correlation for flocculation kinetics proposed by Šulc⁴ and the simple semiempirical generalized correlation quantifying the effect of flocculation time and flocculant dosage on flocculation (see below) proposed by Šulc, Diti⁵ were used for data treatment.

The following generalized correlation for flocculation kinetics in an agitated tank, taking into account floc breakage was proposed⁴:

$$\Delta Z_r^* = A_{Z_r^*} \cdot (\Delta[N \cdot t_F]_{\log})^2 \quad (2)$$

where

$$\Delta Z_r^* = \frac{Z_r^* - Z_{r_{\min}}^*}{Z_{r_{\min}}^*} \quad (3)$$

$$\Delta[N \cdot t_F]_{\log}^* = \frac{\log(N \cdot t_F) - \log([N \cdot t_F]_{\min})}{\log([N \cdot t_F]_{\min})} \quad (4)$$

where $Z_{r_{\min}}^*$ is the minimum degree of residual turbidity achieved at time $[N \cdot t_F]_{\min}$, $[N \cdot t_F]_{\min}$ is the dimensionless flocculation time in which $Z_{r_{\min}}^*$ can be achieved, $A_{Z_r^*}$ is the residual turbidity shift coefficient, t_F is flocculation time, and N is impeller rotational speed. The residual turbidity shift coefficient parameter $A_{Z_r^*}$ is a measure of flocculation rate. In our case, a higher value of $A_{Z_r^*}$ means a faster decrease of turbidity up to a minimum that is characterized by $[N \cdot t_F]_{\min}$ and $Z_{r_{\min}}^*$. A higher

value of $A_{Z_r^*}$ indicates a faster increase in turbidity above the minimum due to floc breakage.

The generalized correlation parameters $Z_{r_{\min}}^*$ and $A_{Z_r^*}$ generally depend on the flocculation process conditions, i.e. mixing intensity, flocculant dosage. The correlation proposed here allows us to determine the flocculation time at which the minimum degree of residual turbidity and the maximum degree of turbidity removal, respectively, can be achieved. A comparison between the experimental data and the generalized correlation is shown in Fig. 2. The similar results were also obtained for amorphous SiO₂ and clay^{5,17,23}.

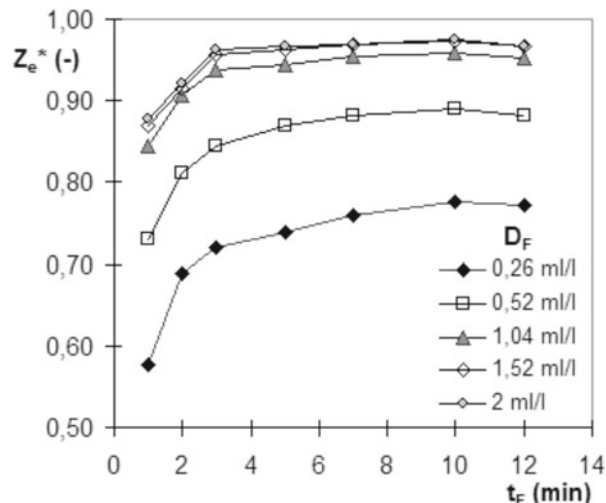


Figure 1. Experimental data $Z_e^* = f(t_F)$ for $D_F = \text{const.}$ – Šulc and Diti³, kaolin, $c_{K0} = 0.58$ g/l, $\epsilon_v = 168$ W/m³

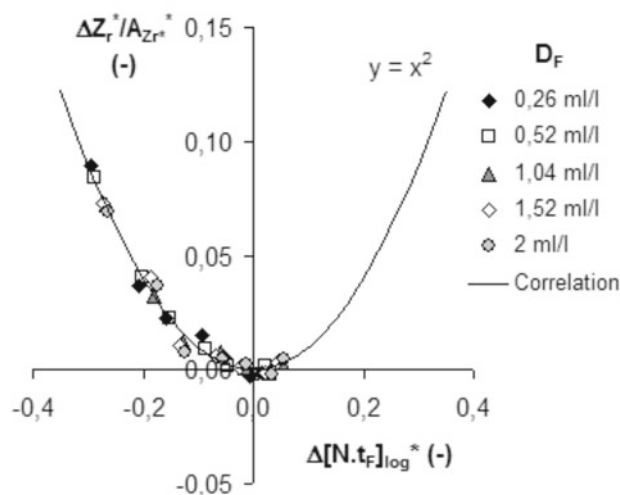


Figure 2. Experimental data vs. correlation (2) $\Delta Z_r^* = f(\Delta[N \cdot t_F]_{\log}^*)$ for $D_F = \text{const.}$ – Šulc and Diti³, kaolin, $c_{K0} = 0.58$ g/l, $\epsilon_v = 168$ W/m³

In Environmental Engineering, the flocculation performances occurring at various process conditions (impeller speed, mixing intensity, tank size) have been compared in the same flocculation time t_F . However, this comparison is correct only for the processes occurring in vessels of the same size, and for the same impeller speed. Both particle collisions and floc size reduction occur mainly in the impeller zone, where the turbulent energy dissipation rate is maximal. Therefore the effect of the number of

floc passages through the impeller zone on the flocculation course can be expected and should be taken into account. The number of passages through the impeller zone is proportional to the circulation time. Similarly, aggregation occurs during fluid circulation in a vessel, so it is also proportional to the circulation time. As the vessel size increases, the circulation time increases strongly. A similar effect of circulation time was also observed in a liquid – liquid dispersion process occurring in stirred vessels. The circulation time is indirectly proportional to the impeller rotational speed. This implies that dimensional flocculation time, which is proportional to the number of floc passages through an impeller zone, must be defined as the product of impeller speed N and flocculation time t_F . The dimensionless flocculation time defined as the product $N \cdot t_F$ has therefore been used as a scale of flocculation time. This phenomenon is crucial for process scale-up.

The proposed general correlation was tested on the experimental data published in Bernhardt, Schell⁶. Bernhardt, Schell⁶ investigated the effect of mixing intensity on flocculation kinetics of the model suspension of quartz particles (Sikron F100) flocculated by Fe^{3+} inorganic flocculant in a tank stirred by a two-paddle stirrer. The tests were carried out at a constant particle concentration. The flocs were separated by a filtration. The dependence of residual turbidity in a filtrate on the flocculation time at different mixing intensity was measured. The mixing intensity was characterized by the velocity gradient G , defined as $G = ((P/V)/\mu)^{1/2} = (\varepsilon_V/\mu)^{1/2}$. For $G = 16, 30$ and 40 s^{-1} (i.e. $\varepsilon_V = 0.26, 0.9$ and 1.6 W/m^3), no floc breakage was observed. In the remaining G -range of $50 \div 150 \text{ s}^{-1}$ (i.e. $\varepsilon_V = 2.5 \div 22.5 \text{ W/m}^3$) floc breakage was observed. No flocculation model was applied for data interpretation. A comparison between the experimental data and the generalized correlation is shown in Fig. 3. The relations $A_{Z_r^*}^* = 43.529 \cdot \varepsilon_V^{-0.766}$ ($R^2 = 0.9357$), $Z_{r^* \min} = 0.1386 \cdot \varepsilon_V^{0.425}$ ($R^2 = 0.9703$) and $[N \cdot t_F]_{\min}^* = 2182.9 \cdot \varepsilon_V^{0.825}$ ($R^2 = 0.9832$) were found for the model parameters.

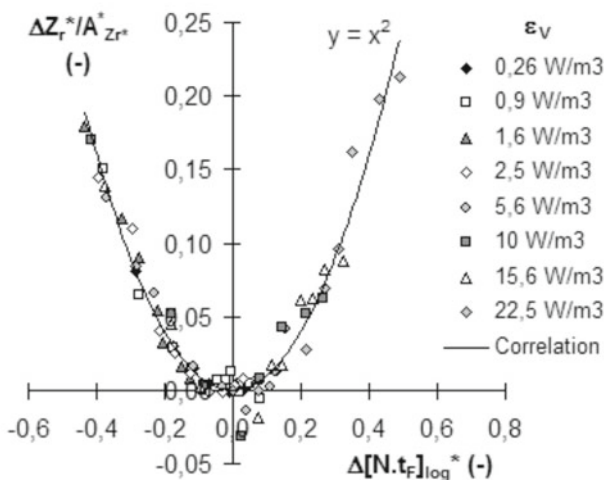


Figure 3. Experimental data vs. correlation (2) $\Delta Z_r^* = f(\Delta[N \cdot t_F]_{\log}^*)$ for $\varepsilon_V = \text{const.}$ – Bernhardt and Schell⁶, quartz, $c_{q0} = 0.025 \text{ g/l}$, $D_F = 2.48 \text{ ml/l}$

Camp number

The Camp number, defined as the product of velocity gradient G characterizing mixing intensity and floccula-

tion time t_F , i.e. $Ca = G \cdot t_F$ proposed by Camp⁷ postulates that same flocculation performance can be achieved at various velocity gradients at flocculation time satisfying relation $G \cdot t_F = \text{const.}$ Although the concepts of velocity gradient and Camp number have been criticized^{8–13}, they can be still used for flocculation process design^{14–16}.

The effect of mixing intensity on the flocculation kinetics of amorphous SiO_2 slurry was investigated¹⁷ in a baffled tank agitated by a Rushton turbine for various mixing intensities ranging from 40 to 520 W/m^3 at SiO_2 concentration 0.8 g/l and flocculant dosage $D_F = 2.48 \text{ ml/l}$. The floc breakage was observed in the whole measured mixing intensity range. The model wastewater was flocculated using Sokoflok 56A organic flocculant (solution 0.1% wt.; Sokoflok Ltd. Czech Republic). The experimental data is presented in Fig. 4, and a comparison of the experimental data and the generalized correlation (2) is presented in Fig 5. The parabolic curve according to the correlation (2) depicted in Fig. 4 does not mean that the floc formation and the floc breakage are related in symmetrical manner. The asymmetry between the floc formation and their breakage is masked by the statistical treatment leading to the generalization and nonlinearities of x - and y - axes⁴.

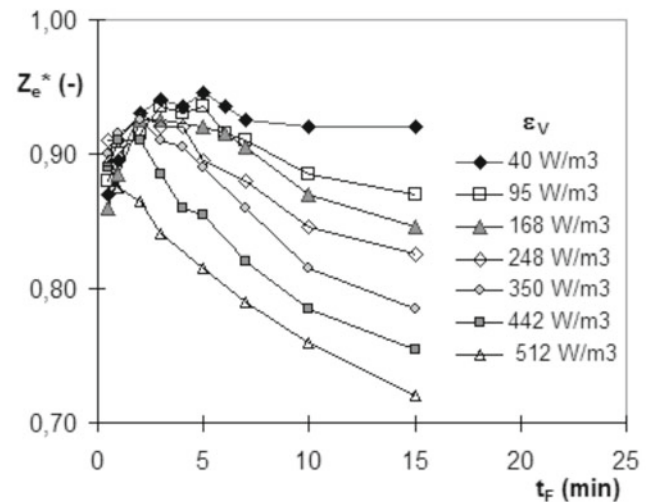


Figure 4. Experimental data $Z_e^* = f(t_F)$ for $\varepsilon_V = \text{const.}$ – Šulc and Ditl¹⁷, SiO_2 , $c_{S0} = 0.8 \text{ g/l}$, $D_F = 2.48 \text{ ml/l}$

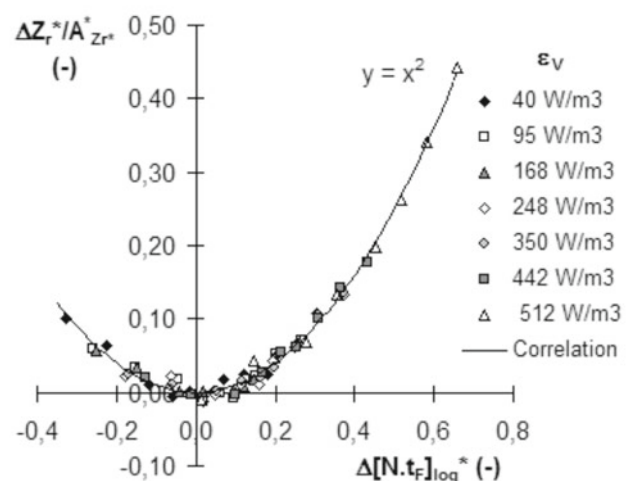


Figure 5. Experimental data vs. correlation (2) $\Delta Z_r^* = f(\Delta[N \cdot t_F]_{\log}^*)$ for $\varepsilon_V = \text{const.}$ – Šulc and Ditl¹⁷, SiO_2 , $c_{S0} = 0.8 \text{ g/l}$, $D_F = 2.48 \text{ ml/l}$

Table 1. Camp number calculated for given turbidity removal degree

Data: Šulc and Dittl ¹⁷ , SiO ₂ , c ₈₀ = 0.8 g/l, D _F = 2.48 ml/l				
ε_v	G	Ca = G.t _F (-)		
(W.m ⁻³)	(s ⁻¹)	Z _e [*] = 0.90	Z _e [*] = 0.91	Z _e [*] = 0.92
40	200	10 852	13 856	18 253
95	309	14 707	18 577	25 043
168	410	24 226	30 151	40 795
248	498	13 927	17 672	23 704
250	592	19 038	23 120	31 179
442	665	23 632	*	*
512	716	*	*	*
Data: Bernhardt and Schell ⁶ , quartz, c _{q0} = 0.025 g/l, D _F = 2.48 ml/l				
ε_v	G	Ca = G.t _F (-)		
(W.m ⁻³)	(s ⁻¹)	Z _e [*] = 0.60	Z _e [*] = 0.70	Z _e [*] = 0.80
1.6	40	11 700	19 625	45 341
2.5	50	13 792	21 655	48 196
5.63	75	13 041	26 257	*
10	100	18 592	*	*
15.6	125	*	*	*
22.5	150	*	*	*

Notice: * A given turbidity removal degree is not reachable at given conditions.

Table 1 illustrates the limited applicability of the generally used designing procedure based on the constant Camp number value. For a given turbidity removal degree, i.e. a given flocculation performance, the corresponding flocculation time was calculated for various mixing intensities using the correlation (2) for our data and also for the data published by Bernhardt, Schell⁶. Consequently, the Camp number values were calculated. From Table 1 it follows that the same flocculation efficiency (Z_e^{*} = 90% or 75%) can not be reached at the constant Camp number value.

Static mixer

It was found⁴ that the proposed generalized correlation (2) can be extended up to flocculation effected by a static mixer adopting dimensionless flocculation time equal to the number of elements i_S that a static mixer consists of:

$$t_F^* = \frac{t_M}{t_{M1}} = \frac{l_S}{l_E} = i_S \quad (5)$$

where t_M is the mean mixing time in a static mixer, t_{M1} is the mean mixing time in a static mixer element, l_S is static mixer length and l_E is element length. The dimensionless flocculation time defined in this way is analogous to the dimensionless time defined for an agitated vessel.

Thus, the generalized correlation is given as follows:

$$\Delta Z_r^* = A_{Z_r^*} \cdot (\Delta [i_S]_{\log}^*)^2 \quad (6)$$

where

$$\Delta Z_r^* = \frac{Z_r^* - Z_{r_{\min}}^*}{Z_{r_{\min}}^*} \quad (7)$$

where [i_S]_{min} corresponds to the number of static mixer elements in which Z_r^{*}_{min} can be achieved.

The proposed model was successfully tested¹⁸ on the data reported by Suharyono, Hogg¹⁹. Suharyono, Hogg¹⁹

performed a laboratory investigation of flocculation following the addition of an organic polymer flocculant to kaolin clay slurries flowing through static mixers (Kenics). The flocculation performance was evaluated according to the settling rate and the supernatant turbidity. The effect of the mixing time was controlled by the number of mixer elements in a pipe. The mixing intensity was varied according to the suspension flowrate. A comparison of the experimental data and the proposed generalized correlation (6) is depicted in Fig. 6. The specific power input was estimated using the Darcy-Weisbach equation.

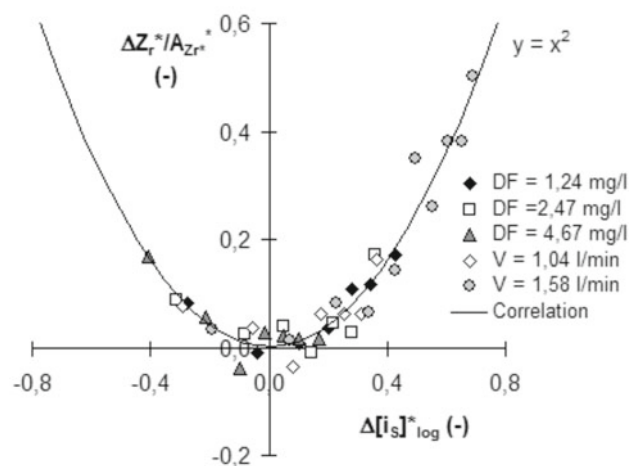


Figure 6. Experimental data vs. correlation (6) $\Delta Z_r^* = f(\Delta [i_S]_{\log}^*)$ – data Suharyono and Hogg¹⁹

Effect of the flocculant dosage

The effect of flocculant dosage on flocculation was investigated²⁰ in our laboratory using a baffled tank agitated by a Rushton turbine and a 6-pitched blade turbine at constant mixing intensity $\varepsilon_v = 40 \text{ W/m}^3$. The flocs were separated by sedimentation. The tests were carried out on a model wastewater (a suspension of tap

water + SiO₂; constant particle concentration 800 mg/l). The model wastewater was flocculated with Sokoflok 56A organic flocculant (solution 0.1% wt.; Sokoflok Ltd., Czech Republic) at pH 7.6.

A simple semi-empirical generalized correlation quantifying the effect of flocculant dosage on flocculation was proposed:

$$\Delta Z_r^* = A_{Z_r^*(D)}^* \cdot (\Delta [D_F]_{\log}^*)^2 \quad (9)$$

where

$$\Delta Z_r^* = \frac{Z_r^* - Z_{r_{\min(D)}}^*}{Z_{r_{\min(D)}}^*} \quad (10)$$

$$\Delta [D_F]_{\log}^* = \frac{\log(D_F) - \log(D_{F_{\min}})}{\log(D_{F_{\min}})} \quad (11)$$

where $Z_{r_{\min}}$ is the minimal degree of residual turbidity achieved at flocculant dosage $D_{F_{\min}}$, $D_{F_{\min}}$ is the flocculant dosage for which $Z_{r_{\min}}$ can be achieved, and $A_{Z_r^*(D)}^*$ is the residual turbidity shift coefficient.

The generalized correlation parameters $Z_{r_{\min}}^*$, $D_{F_{\min}}$ and $A_{Z_r^*(D)}^*$ generally depend on the flocculation process conditions, e.g. mixing intensity, flocculation time. The proposed model was successfully tested on the experimental data that was obtained. The experimental data for a Rushton turbine are presented in Fig. 7. A comparison of the experimental data and the generalized correlation for a Rushton turbine is depicted in Fig. 8.

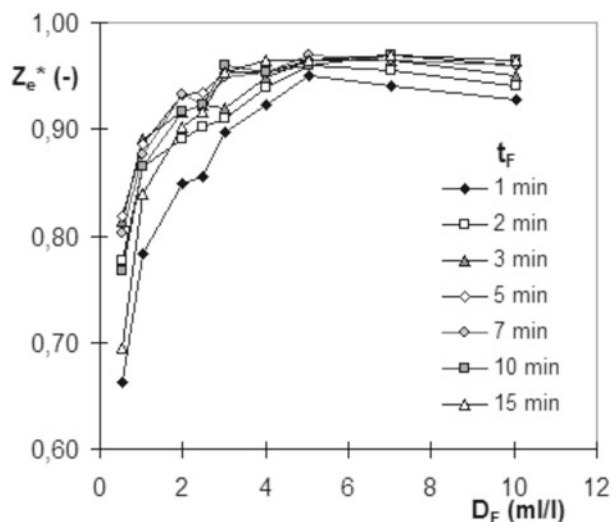


Figure 7. Experimental data $Z_r^* = f(D_F)$ for $t_F = \text{const.}$ – Šulc and Dítl²⁰, SiO₂, $c_{S0} = 0.8$ g/l, $\varepsilon_V = 40$ W/m³

Effect of flocculation time and flocculant dosage

On the basis of our earlier results, we proposed⁵ the following simple semi-empirical generalized correlation quantifying the effect of flocculation time and the flocculant dosage on flocculation:

$$\Delta Z_r^* = A_{12}^* \cdot (\Delta [N \cdot t_F]_{\log}^*)^2 + A_{22}^* \cdot \left(\Delta \left[\frac{D_F}{c_{p0}} \right]_{\log}^* \right)^2 + \quad (12)$$

$$+ B_{11}^* \cdot (\Delta [N \cdot t_F]_{\log}^*) \cdot \left(\Delta \left[\frac{D_F}{c_{p0}} \right]_{\log}^* \right)$$

$$\Delta Z_r^* = \frac{Z_r^* - Z_{r_{\min(t,D)}}^*}{Z_{r_{\min(t,D)}}^*} \quad (13)$$

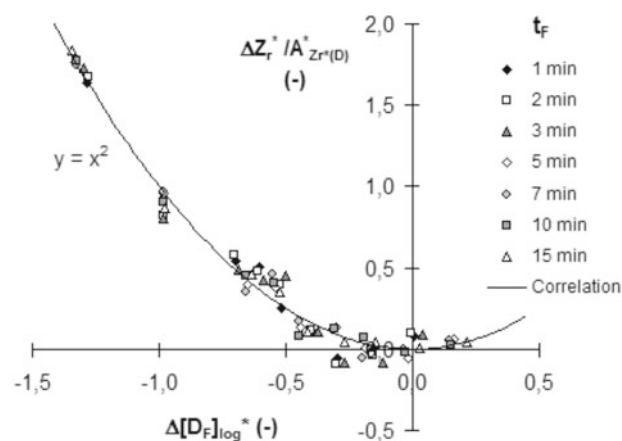


Figure 8. Experimental data vs. correlation (9) $\Delta Z_r^* = f(\Delta [D_F]_{\log}^*)$ – Šulc and Dítl²⁰, SiO₂, $c_{S0} = 0.8$ g/l, $\varepsilon_V = 40$ W/m³

The ratio D_F/c_{p0} represents the dimensionless flocculant dosage defined as the ratio of flocculant dosage D_F and the initial pollution concentration c_{p0} . Parameter $Z_{r_{\min(t,D)}}^*$ and the corresponding parameters $[N \cdot t_F]_{\min}$ and $[D_F/c_{p0}]_{\min}$ represent the optimal conditions under which minimum residual turbidity can be achieved. The coefficients A_{12}^* , A_{22}^* , B_{11}^* are residual turbidity shift coefficients due to flocculation time, flocculant dosage, and miscellaneous effects of both variables, respectively. The model parameters generally depend on the mixing intensity. The proposed generalized correlation was tested on our data³. A comparison of the experimental data and the generalized correlation (12) is depicted in Fig. 9.

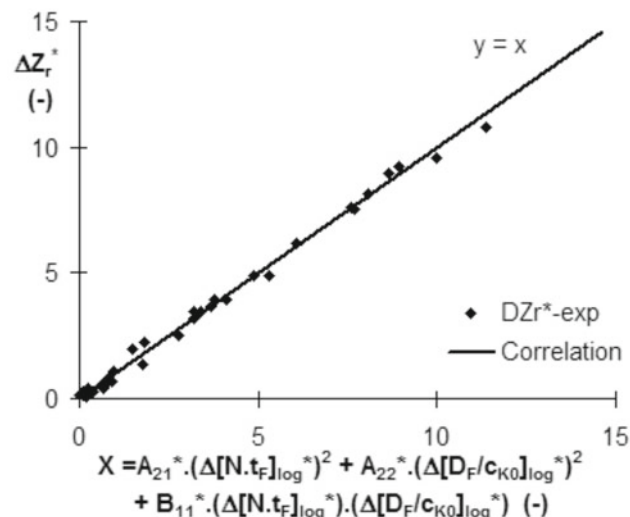


Figure 9. Experimental data vs. correlation (12) $\Delta Z_r^* = f(\Delta [N \cdot t_F]_{\log}^*, \Delta [D_F/c_{K0}]_{\log}^*)$ – Šulc and Dítl³, kaolin, $c_{K0} = 0.58$ g/l, $\varepsilon_V = 168$ W/m³, $t_F = 1-12$ min, $D_F = 0.26-2$ ml/l

An analytical description of the dependence of degree of residual turbidity Z_r^* on the flocculant dosage D_F and on flocculation time t_F enables us to calculate numerically the dependence of flocculant dosage D_F on flocculation time t_F for the desired flocculation efficiency, in this case for the desired degree of residual turbidity Z_r^* or degree of turbidity removal Z_e^* . For illustration, the graph of calculated isolines $D_F = f(t_F)$ for $Z_r^* = \text{const.}$ for a suspension of amorphous SiO₂ particles (constant particle concentration 800 mg/l) agitated by a Rushton

turbine at mixing intensity 40 W/m³ is presented in Fig. 10 for $Z_r^* = 0.15, 0.1, 0.08, 0.06, 0.05, 0.04, 0.035$, i.e. for $Z_e^* = 0.85, 0.9, 0.92, 0.94; 0.95, 0.96; 0.965$.

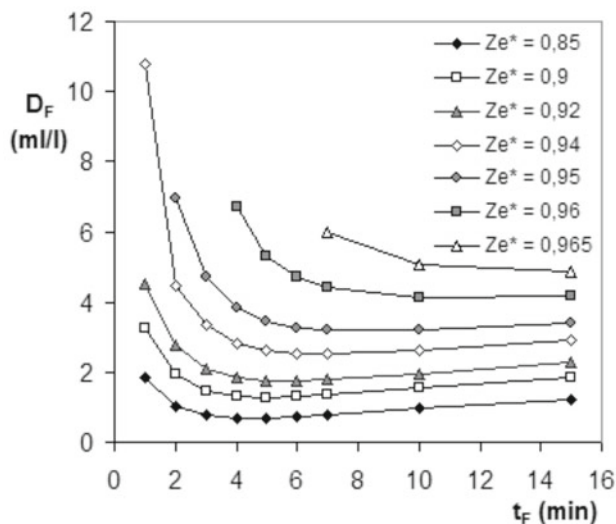


Figure 10. Isolines $D_F = f(t_F)$ for $Z_r^* = \text{const.}$ – data Šulc and Ditl²⁰, SiO₂, $c_{s0} = 0.8$ g/l, $\varepsilon_V = 40$ W/m³

The graph $D_F = f(t_F)$ for constant flocculation efficiency allows us to adjust the flocculant dosage according to the flocculation time, and vice versa, while the same flocculation efficiency is maintained. For example, flocculation occurs for 10 minutes and the flocculant dosage is 2 ml/l. In our case, the degree of residual turbidity 8%, i.e. degree of turbidity removal 92%, is achieved for the given conditions. According to the graph, the same effect can be achieved for the same dosage after just three minutes.

For the batch system, in this way, the process duty can be increased by the reduction of the flocculation time at the same flocculant dosage at the same flocculation efficiency. Similarly, the same effect can be expected for a continuous system.

Our model and the proposed data treatment is applicable for different systems and it enables for each system turbidity removal degree prediction, determination of dependence of flocculant dosage D_F on flocculation time t_F for the desired flocculation efficiency and scale-up for a particular system. The adjustable model parameters should be determined for each system separately.

Effect of pollutant concentration

The effect of clay concentration on flocculation performance was investigated⁵ for various flocculant dosages and for three clay slurry concentration values at mixing intensity $\varepsilon = 40$ W/m³. The flocculation experiments were conducted in a fully baffled cylindrical vessel of diameter $D = 150$ mm agitated by a Rushton turbine. A water – clay particle suspension was used as a model system. The solid fraction of the clay was 0.48, 0.58 and 0.92 g/l, respectively. The model wastewater was flocculated using the Sokoflok 16A organic polymer flocculant (0.1% wt. aqueous solution; Sokoflok Ltd. Czech Republic). The flocs were separated by the 5-minute sedimentation in the tank.

The fit tightness between the generalized correlation (12) and experimental data for the three different pollutant concentrations is depicted in Fig. 11. The systematic

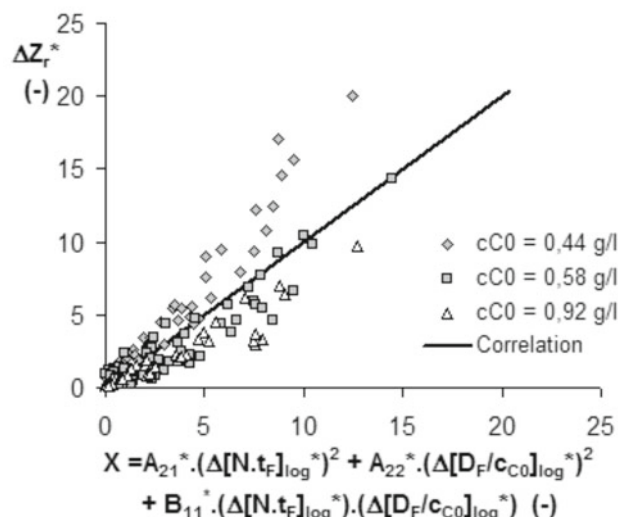


Figure 11. Experimental data vs. correlation (12) $\Delta Z_r^* = f(\Delta[N.t_F]_{\log^*}, \Delta[D_F/c_{C0}]_{\log^*})$ – Šulc and Ditl⁵, clay, $c_{C0} = 0.58$ g/l, $\varepsilon_V = 40$ W/m³, $t_F = 1-10$ min, $D_F = 0.26-3.04$ ml/l

deviations for the lowest concentration 0.44 g/l can be seen mainly. We assume that these deviations are caused by a nonlinear relation between the turbidity and the pollution concentration. Based on this hypothesis we introduced the correction based on the Lambert-Beer law. After this correction the nonlinear dependence of the turbidity vs. pollutant concentration is respected which is reflected in the excellent tightness of the measured and predicted data as seen in Fig. 12. The final results are prepared for publication.

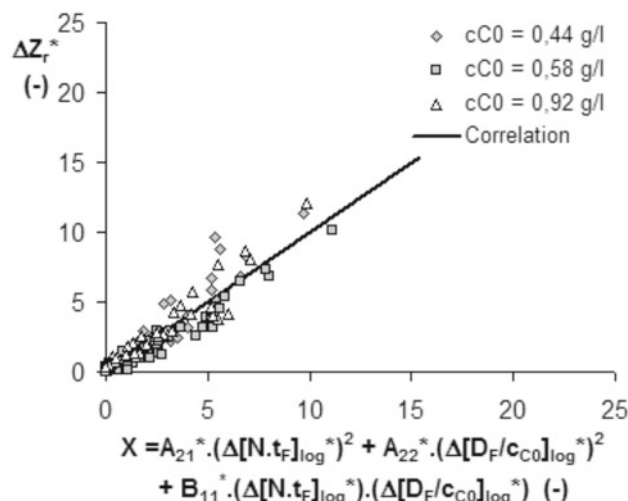


Figure 12. Experimental data vs. correlation (12) $\Delta Z_r^* = f(\Delta[N.t_F]_{\log^*}, \Delta[D_F/c_{C0}]_{\log^*})$ – data Šulc and Ditl⁵ – after correction

Scale up

Camp⁷ proposed as the main design parameters the average velocity gradient G and the product of velocity gradient and time, referred to as the Camp number, for which the flocculation performance should be the same, regardless of the tank size or impeller type. As mentioned above, although this approach has been discussed and criticized by many authors, it is still used in industrial practice for design and control. Oldshue and Mady²¹ investigated the effect of impeller speed on residual turbidity in two tanks 460 mm and 760 mm in

diameter stirred by a Rushton turbine and an A212 axial impeller. The flocs were separated by sedimentation. The average velocity gradient G , the flocculation time and the settling time were kept the same in both tanks. Ducoste and Clark²² investigated the effect of the impeller type and tank size on the flocculation process in three tanks 5 L, 28 L and 560 L in volume stirred by a Rushton turbine and an A 310 axial impeller. The experiments were carried out at a constant average velocity gradient and at a constant flocculation time in all three tanks. After flocculation, the sample was withdrawn and the final particle size distribution was determined. Both of these studies found that the flocculation performance decreases with the increasing scale.

Since flocculation is a very complex process consisting of many mechanisms affected by a wide range of variables, the scale-up criteria for each phenomenon need to be formulated, i.e. a multidimensional approach must be applied instead of the one-dimensional approach that has been used to date. If the scale-up criteria for all mechanisms are fulfilled, the same flocculation results might be expected. Among the participating mechanisms, we consider the following to be important for flocculation performance, so that their scale-up criteria need to be formulated: mixing, flocculation time, chemical environment and floc separation. On the basis of these findings, we proposed²³ the following scale up rules:

– $P/V = \text{const.}$ – the same specific power input into a tank D_2 in diameter as the specific power input into a tank D_1 in diameter to provide the same average mixing intensity.

– $t_F^* = N \cdot t_F = \text{const.}$ – the same dimensionless flocculation time in tank D_2 in diameter as the dimensionless flocculation time in the tank D_1 in diameter to provide the same number of passages of flocs through the impeller region.

– $D_F = \text{const.}$ – the same flocculant dosage in the tank D_2 in diameter as the flocculant dosage in the tank D_1 in diameter for a given flocculant type and wastewater quality to provide the same chemical environment.

Since the flocs will be separated by sedimentation, the sedimentation time was set up on the basis of the condition $u_{\text{sed}} = \text{const.}$ – the same sedimentation velocity in the tank D_2 in diameter as the sedimentation velocity in the D_1 tank in diameter to provide floc removal of same size. The same flocculation results were achieved fulfilling these conditions.

The scale-up criteria proposed here were tested in three baffled tanks 0.15 m-, 0.20 m- and 0.30 m in diameter agitated by a Rushton turbine. Clay slurry was used as a model wastewater. The experiments were carried out at the mixing intensity 40 W/m^3 and clay concentration 0.58 g/l . The model wastewater was flocculated using the organic flocculant Sokoflok 16A (solution 0.1 wt. %), flocculant dosage $D_F = 2.48 \text{ ml/l}$. The flocs were separated by sedimentation. The experimental data are presented in Fig. 13. It was found that the course of the flocculation for vessels 0.20 m and 0.30 m in diameter was practically the same. For the vessel of diameter $D = 0.15 \text{ m}$, the degree of turbidity removal was slightly lower, probably due to somewhat different hydrodynamic conditions in the impeller zone. A comparison of the experimental data and the generalized correlation (2)

is depicted in Fig. 14. The effect of the vessel size on the generalized correlation parameters was investigated statistically by hypothesis testing. According to the results of the test, all correlation parameters can be statistically taken as constant and independent of the vessel size. The maximum turbidity removal degree $Z_{e^* \text{ max}} = 96.4 \pm 1\%$ was found for the dimensionless flocculation time 1582 ± 230 .

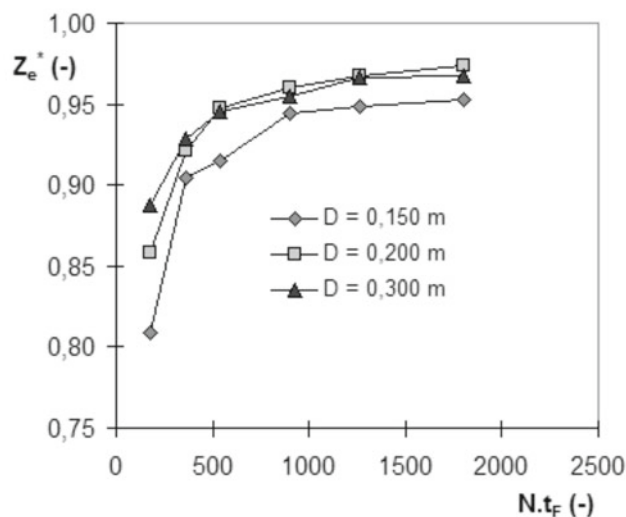


Figure 13. Experimental data $Z_r^* = f(N \cdot t_F)$ for $D_F = \text{const.}$ – Šulc and Ditl²³, clay, $c_{C0} = 0.58 \text{ g/l}$, $\varepsilon_V = 40 \text{ W/m}^3$, $D_F = 2.48 \text{ ml/l}$

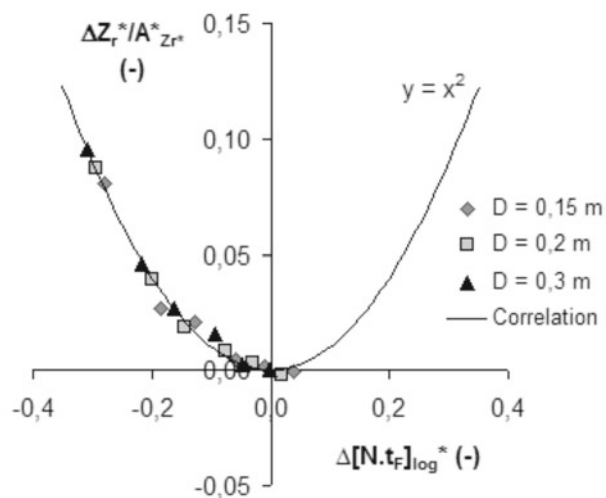


Figure 14. Experimental data vs. correlation (2) $\Delta Z_r^* = f(\Delta[N \cdot t_F]_{\log^*})$ for $D_F = \text{const.}$ – Šulc and Ditl²³, clay, $c_{C0} = 0.58 \text{ g/l}$, $\varepsilon_V = 40 \text{ W/m}^3$, $D_F = 2.48 \text{ ml/l}$

CONCLUSIONS

This paper has summarized our results obtained in the course of a long-term research on flocculation. The effect of flocculation time, flocculant dosage, impurity concentration, mixing intensity and the tank size on flocculation performance was experimentally studied in a mechanically agitated vessel. The macroscopic approach was applied for flocculation modelling.

The simple semiempirical generalized correlation⁴ for flocculation kinetics and the simple semiempirical generalized correlation⁵ quantifying the effect of flocculation time and flocculant dosage on flocculation were used for data treatment. Our model and the proposed

data treatment is applicable for different systems and it enables for each system turbidity removal degree prediction, determination of dependence of flocculant dosage D_F on flocculation time t_F for the desired flocculation efficiency and scale-up for a particular system. The adjustable model parameters should be determined for each system separately.

The proposed generalized correlation for flocculation kinetics was extended up to flocculation induced by a static mixer. The proposed model was successfully tested on data of Suharyono and Hogg¹⁹.

The experimental results obtained for the proposed scale up rules²³ have been presented. When the proposed scale-up rules were applied, the flocculation course for vessels 0.20 m and 0.30 m in diameter were found to be practically the same. For $D = 0.15$ m, the efficiency is slightly lower, probably due to various hydrodynamic conditions in the impeller region.

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Symbols

$A_{Z_r}^*$	residual turbidity shift coefficient; model parameter (2)
$A_{Z_r(D)}^*$	residual turbidity shift coefficient; model parameter (9)
A_{12}^*	generalized correlation parameter; model parameter (12)
A_{22}^*	generalized correlation parameter; model parameter (12)
B_{11}^*	generalized correlation parameter; model parameter (12)
c_{p0}	initial solid pollutant concentration, g/l
D	tank diameter, m
D_F	flocculant dosage, ml/l, mg/l
D_{Fmin}	generalized correlation parameter; model parameter (9), ml/l, mg/l
$[D_F/c_{p0}]_{max}$	generalized correlation parameter; model parameter (22)
$[D_F/c_{p0}]_{min}$	generalized correlation parameter; model parameter (12)
G	velocity gradient; $G = (\varepsilon_V/\mu)^{1/2} = (P/(V\mu))^{1/2} \text{ s}^{-1}$
$G.t_F$	Camp number
i_s	number of static mixer elements
l_s	static mixer length, m
l_e	length of static mixer element, m
N	impeller rotational speed, rpm
$[N.t_F]_{min}$	model parameter (2), (12)
P	impeller power input, W
t_F	flocculation time, minute
t_F^*	dimensionless flocculation time
V	tank volume, m ³
Z_0	turbidity before flocculation, FAU
Z_r	residual turbidity after flocculation, FAU
Z_c^*	turbidity removal degree
Z_r^*	residual turbidity degree
$Z_{r,min}^*$	model parameter (2), (6)
$Z_{r,min(D)}^*$	model parameter (9)
$Z_{r,min(t,D)}^*$	model parameter (12)

Greek letters

$\Delta[D_F]_{log}^*$	variable
$\Delta[D_F/c_{p0}]_{log}^*$	variable
$\Delta[i_s]_{log}^*$	variable
$\Delta[N.t_F]_{log}^*$	variable
ΔZ_r^*	variable
ε_V	specific impeller power input (per volume unit); $\varepsilon_V = P/V, \text{ W/m}^3$

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