

Factors influencing direct separation of destabilised impurities in a fluidised layer of granular material

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ABSTRACT

The results obtained from a study into the influences of certain technological factors on separation efficiencies of a fluidised layer of granular material (FLGM) are summarised. This study shows that the mechanism of interception of destabilised impurities on the surface of a granular medium corresponds to the mechanism of mutual particle aggregation of destabilised impurities. The efficiency of this process is influenced by the degree of destabilisation and degree of aggregation of the particles of impurities. The highest efficiencies are obtained when the particles of impurities enter a FLGM immediately after the completion of their destabilisation and their aggregation takes place entirely in the FLGM.

Key words | agitation, fluidised layer, water treatment

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INTRODUCTION

A fluidised layer of granular material (FLGM) has been applied in water treatment for the removal of phosphates (van Dijk & Braakensiek 1984), for water softening (Graveland *et al.* 1983; van der Veen & Graveland 1988, van der Veen 1985), and for the removal of heavy metals from effluents (Schöller *et al.* 1987). The FLGM was tested in pilot plant studies for treatment of waters containing high concentrations of mineral suspension (Tambo & Matsui 1989) as well as for treatment of highly coloured and turbid waters (Tambo & Wang 1993*a,b*). Applicability of the FLGM in pilot plant and full size operations was tested in the treatment of common quality surface waters (Mutl & Hereit 1981; Hereit & Mutl 1982; Hereit *et al.* 1982; Mutl 1990, 1993; Mutl *et al.* 1999, 2000).

The pilot and full size plant operations proved that a fluidised layer of granular material (FLGM) can be utilised in water treatment as the process unit for formation of suspensions separable by sedimentation, dissolved air flotation, filtration through a fluidised layer of flocculent suspension and deep bed filtration. Furthermore, the pilot plant testing proved that the FLGM can also be utilised for

direct separation of the destabilised particles of impurities, about 60% of which is intercepted in the form of a water-free coat on the surface of grains of the FLGM, with the remaining portion of impurities forming aggregates which are separable by a suitable separation method mentioned above.

The tests also proved that the mechanism enabling direct separation of the destabilised impurities in the FLGM is similar to that enabling formation of separable aggregates, i.e. the mechanism of mutual aggregation of the destabilised particles. The effect of certain factors affecting the character of the formed aggregates seems to be somehow different. From the technological point of view two groups of these factors are discernible:

- (a) factors determined by the properties of the FLGM;
- (b) factors determined by the properties of impurities, destabilisation reagent and water as the reaction medium.

Although the action of all of these factors is complex due to the interaction of the partial processes, their effect on

these partial processes can be optimised. From the results obtained during pilot and full size plant operations it is evident that the factors influencing conditions which affect the particles during their aggregation as well as the aggregates already formed, namely the magnitude of the velocity gradient, its distribution in the FLGM and duration of its action, belong to the most important factors of the first group. These factors are influenced by the type and material of the medium of the FLGM such as granularity of medium and height of the layer in a stable and a fluidised state. The most significant factors of the second group are those affecting the course and the velocity of the process of particle destabilisation and their aggregation, namely the type and dosage of destabilisation reagent, the reaction conditions and water temperature.

The objective of this research was to quantify the influence of these factors on separation efficiency of the FLGM. More specifically the research was aimed at the following:

1. the effect of the hydrodynamic characteristics of the FLGM on its technological properties;
2. the effect of the degree of destabilisation of impurities, reaction period and magnitude of surface area of grains in the FLGM on its separation efficiency.

PILOT PLANT

The effect of individual factors was investigated in pilot plant operations. The basic arrangement and the basic design parameters of the pilot plant used, enabling the FLGM reactor to operate in the separation mode, are described in previous work (Muhl & Polasek 1995; Mutl *et al.* 1999, 2000). The experimental work was carried out at Zelivka waterworks, treating water from the impounding Svihov reservoir. This reservoir is characterised by a relatively stable water quality over long periods of time, which ensures reproducibility of plant operations and hence the results obtained.

The properties of the medium material used in the FLGM affect its behaviour very significantly. The FLGM reactor was charged with silica sand of grain sizes varying

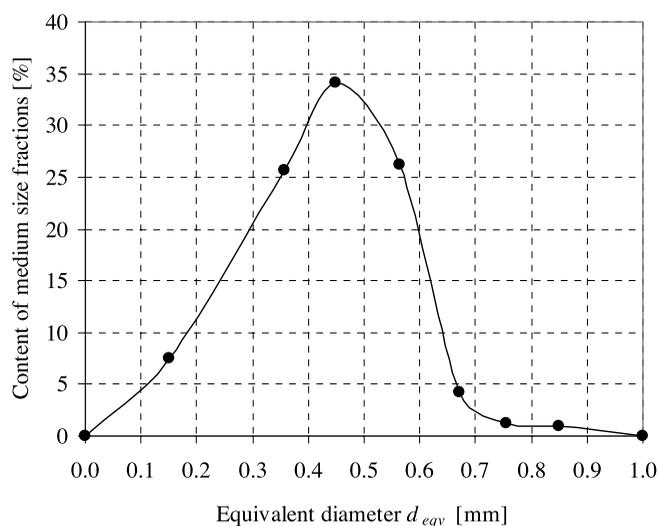


Figure 1 | Sieve analysis of the silica sand used to charge the FLGM reactor.

Table 1 | Basic characteristics of the partial size-fractions of granular medium

	d (mm)			
	0.315–0.40	0.40–0.50	0.50–0.63	0.63–0.71
d_{eqv} (mm)	0.337	0.426	0.587	0.674
S (mm ²)	0.357	0.570	1.082	1.426
ρ (g · cm ⁻³)	2.643	2.648	2.658	2.652
V (cm ³)	0.020	0.040	0.106	0.160

between $d_g = 0.315$ – 0.710 mm. Sand used in individual tests was pre-screened on laboratory sieves to different grain-size fractions. The sieve analysis of this sand is shown in Figure 1. The properties of medium grains are summarised in Table 1.

METHODOLOGY OF TESTING AND EVALUATION OF PLANT PERFORMANCE

The methodology of sampling, analytical processing of samples and evaluation of the results obtained can be

found in previous work (Hereit *et al.* 1977; Hereit & Mutl 1980; Polasek 1980; Polasek & Mutl 1995).

The basic criteria used to characterise the efficiency of the treatment process and the aggregation of the particles of impurities achieved are described below.

The principal criterion affecting the efficiency of treatment of surface waters is the degree of destabilisation of particles of impurities α_D . It is defined as a ratio of the number of particles destabilised by means of treatment reagent N_D to the total number of particles initially contained in the system N_0 and calculated from the following relationship:

$$\alpha_D = \frac{N_D}{N_0} = \frac{C_0F - C_EF}{C_0F} \quad (1)$$

The degree of destabilisation is influenced mainly by type of destabilisation reagent and its dosage. With increasing dosage its value increases and under ideal conditions its value reaches $\alpha_D = 1.0$.

Another principal criterion is the degree of aggregation α_A . It is defined as a ratio of the number of the particles that aggregated during the treatment process N_D to the total number of particles initially contained in the system N_0 and calculated from the following relationship

$$\alpha_A = \frac{N_A}{N_0} = \frac{C_0 - C_{EF}}{C_0} \quad (2)$$

The degree of aggregation is influenced mainly by the conditions of agitation, i.e. its intensity and duration. As the number of aggregated particles increases the degree of aggregation increases and under ideal conditions of aggregation its value reaches $\alpha_A = 1.0$.

Direct separation of the destabilised particles in the FLGM assumes that the mechanism of particle interception on the surface of grains of the FLGM is similar to that of mutual aggregation of the particles of impurities. It should be emphasised that the contacts between particle–grain, instead of particle–particle, are decisive with respect to separation efficiency of the FLGM.

The efficiency of the water treatment process was ascertained by determining actual separation efficiency φ . It is calculated according to the relationship:

$$\varphi = \frac{N_S}{N_0} = \frac{C_0 - C_E}{C_0} \quad (3)$$

The efficiency of the water treatment process was evaluated by:

- direct comparison of water quality determinants, namely content of the basic component of the hydrolysing reagent (Al) and chemical oxygen demand (COD_{Mn}), both measured in the raw water and the treated water;
- separation efficiencies φ_{Al} and φ_{COD} calculated from the data in (a) above.

The effect of agitation on efficiency of aggregation was evaluated by the degree of aggregation α_A determined in crucial stages of formation of suspension.

RESULTS AND DISCUSSION

Hydrodynamic characteristics of a fluidised layer

The principal hydrodynamic factors affecting aggregation and separation efficiencies of a fluidised layer are the magnitude and distribution of the velocity gradient and duration of its action in the FLGM.

Fluidisation of granular material in the FLGM reactor is dependent on an upflow velocity of water v_{up} . On exceeding the hydraulic resistance h of a stable layer, this layer expands and on exceeding the threshold velocity of fluidisation v_{tr} it becomes fully fluidised. The hydraulic resistance h at which the layer expands is defined by the equation:

$$h = L_0(1 - \omega_0) \left(\frac{\rho_g}{\rho_w} - 1 \right) \quad (4)$$

It follows from this equation that the hydraulic resistance h is dependent on initial height of a stable layer L_0 , its initial porosity ω_0 , densities of water ρ_w and medium ρ_g and thereby on mass of the medium m . From Table 2 and Figure 2 it is evident that the hydraulic resistance h increases proportionately with mass of the layer and that

Table 2 | Influence of the mass of the layer and water temperature on hydraulic resistance of the layer

L_0 cm	m (g)	h (cm) $T=5^\circ\text{C}$	$T=10^\circ\text{C}$	$T=15^\circ\text{C}$
90	5138	77	77	77
100	5709	86	86	86
110	6280	94	94	94
120	6851	103	103	103

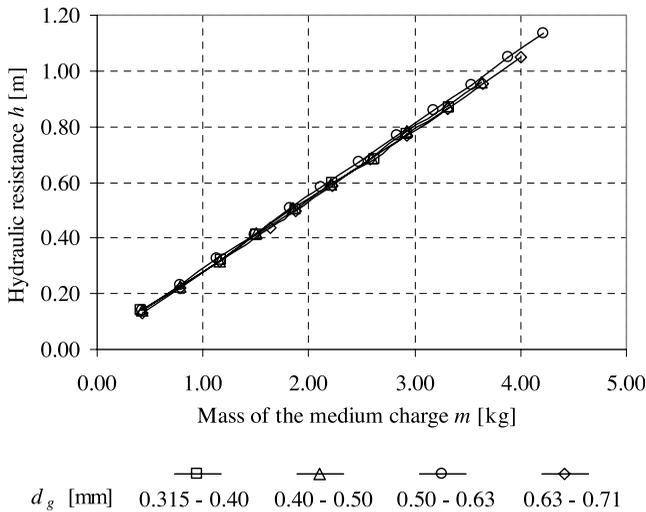


Figure 2 | Dependence of hydraulic resistance Δh on the mass of different medium size-fractions m charged to the FLGM reactor.

the influence of water temperature is negligible within the temperature range considered.

Transition of a stable layer to a fluidised layer is indicated by a change in the FLGM height from a height of stable layer L_0 to a height of fluidised layer L_E , its porosity from ω_0 to ω_E and by the mechanical energy W dissipated into the FLGM and expressed by a change in loss of head H which is expressed by the equation:

$$H = L_E \frac{(1 - \omega_E)(\rho_g - \rho_w)}{\rho_w} = L_E(1 - \omega_E) \left(\frac{\rho_g}{\rho_w} - 1 \right) \quad (5)$$

The expansion ratio E is expressed by the equation:

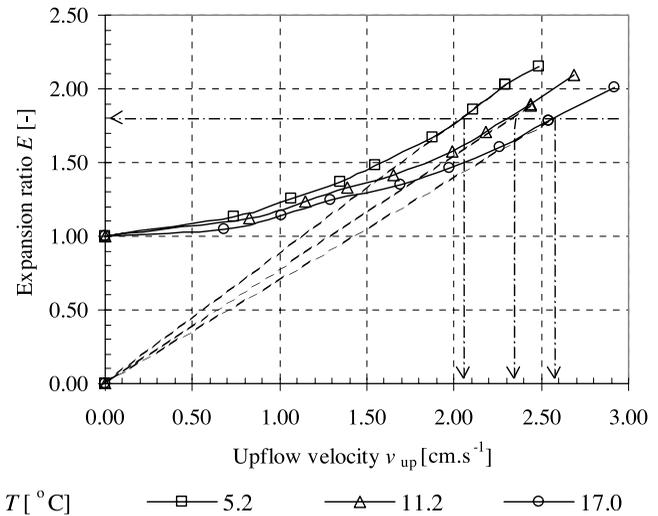


Figure 3 | Dependence of expansion ratio E on upflow velocity v_{up} in the FLGM reactor at different temperatures.

$$E = \frac{L_E}{L_0} \quad (6)$$

and the differential expansion e is expressed by the equation:

$$e = \frac{L_E - L_0}{L_0} = E - 1 \quad (7)$$

The changes in expansion ratio E and loss of head H , characterising changes in the FLGM height depending on changes in upflow velocity v_{up} are shown in Figure 3. It follows from this Figure that for a medium of grain size $d_g = 0.315 - 0.710$ mm, a fully fluidised state is reached at an expansion ratio $E > 1.70$ and that the upflow velocity v_{up} required for achieving it increases with water temperature.

The mean velocity gradient G acting on the aggregating particles in the FLGM is calculated from the equation:

$$G = \sqrt{\frac{W}{Vt\eta}} = \sqrt{\frac{Hg}{tv}} = \sqrt{\frac{g(1 - \omega_E)(\rho_g - \rho_w)v_{up}}{v(e + \omega_0)}} \quad (8)$$

Since the hydraulic resistance h , and hence the head loss H , are independent of flowrate, so is the velocity gradient G ; the value of G is influenced only by water temperature.

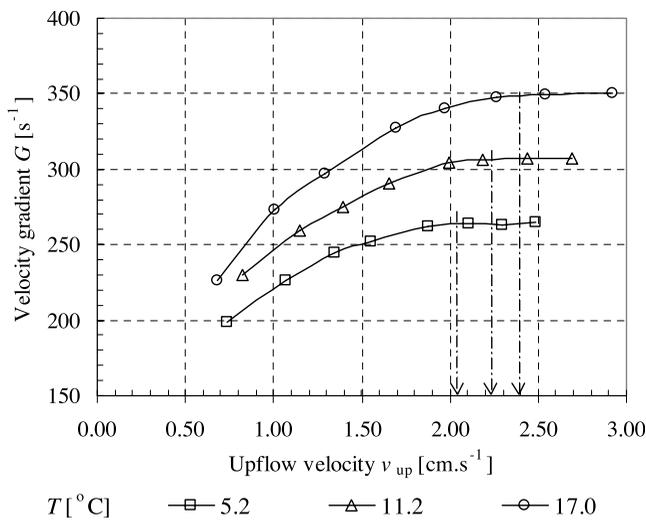


Figure 4 | Dependence of velocity gradient G on upflow velocity v_{up} in the FLGM reactor at different water temperatures.

The volumetric concentration of grains C_V in the FLGM is expressed by the equation:

$$C_V = \frac{V_N}{V_C} = \frac{V_N}{F L_E} = 1 - \omega_E \quad (9)$$

Because

$$L_O = L_E C_V = L_E (1 - \omega_E) = const. \quad (10)$$

the magnitude of dissipated energy W remains the same irrespective of the level of the FLGM expansion and this energy changes only with water temperature.

Since every grain in the FLGM is an independent agitation element, achieving full fluidisation of the medium is important from the point of view of the character of the formed aggregates. A uniform distribution of velocity gradient and its constant value is achieved when the FLGM is fully fluidised. The mean velocity gradient G generated by the FLGM is significantly influenced by water temperature, as follows from Figures 3 and 4. When the FLGM is charged with sand grains of $d_g = 0.315\text{--}0.710$ mm, is fully fluidised, and the water temperature fluctuates between $T = 5.2\text{--}17.0^\circ\text{C}$, the velocity gradient generated by the FLGM fluctuates with water temperature between $G = 345\text{--}260$ s⁻¹.

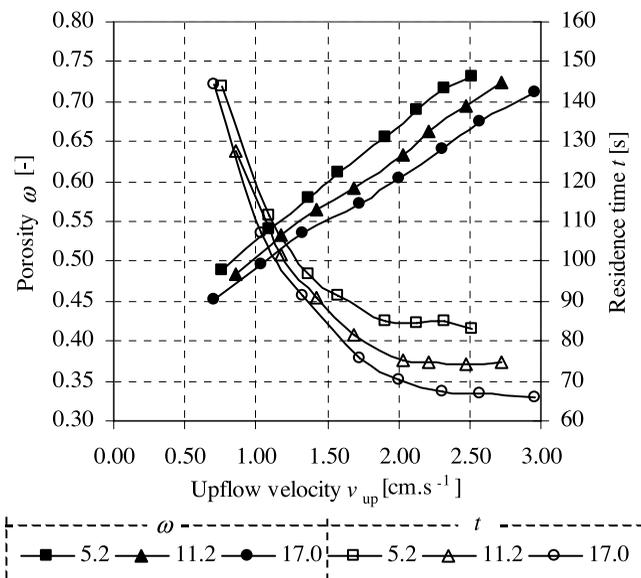


Figure 5 | Dependence of porosity ω and retention time t on upflow velocity v_{up} in the FLGM reactor at different water temperatures.

The period of action t of the velocity gradient G corresponds to the retention time of the water in the FLGM, i.e. to the volume of voids in the FLGM. The free volume of voids changes with changing porosity of the FLGM, which is dependent on the upflow velocity of water.

The retention time t in a layer of the medium is expressed by the equation:

$$t = \frac{V_p}{Q} = \frac{V_c \omega_E}{Q} = \frac{V_c \omega_E}{v_{up} F} = \frac{L_E \omega_E}{v_{up}} = \frac{L_O (e + \omega_0)}{v_{up}} \quad (11)$$

The dependence of retention time t and porosity of the media ω on the upflow velocity of the water v_{up} in the FLGM is evident from Figure 5. This Figure shows that retention period decreases with increasing upflow velocity and increasing porosity of the layer and after reaching a fully fluidised state it remains almost unchanged. This advantageous property of a fluidised layer makes it considerably different from agitation by hydraulic means in traditional flocculation chambers.

It follows from the foregoing that the magnitude of velocity gradient increases and the period of its action

Table 3 | Influence of dosage and water temperature on aggregation and separation efficiencies of the layer

Dosage (Al ₂ (SO ₄) ₃ mg · l ⁻¹)	T=5.8°C		T=10.7°C		T=15.5°C				
	φ _{Al}	φ _{COD}	α _{Al}	φ _{Al}	φ _{COD}	α _{Al}	φ _{Al}	φ _{COD}	α _{Al}
16.11	0.78	0.01	0.07	0.71	0.15	0.08	0.62	0.17	0.16
19.51	0.80	0.23	0.00	0.72	0.20	0.22	0.63	0.22	0.41
22.55	0.73	0.25	0.33	0.70	0.16	0.37	0.63	0.27	0.50
26.42	0.72	0.25	0.46	0.71	0.26	0.51	0.61	0.23	0.59
29.88	0.71	0.27	0.59	0.70	0.26	0.60	0.63	0.19	0.69

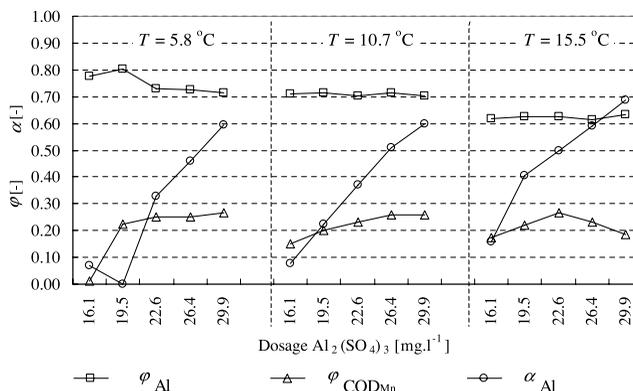
decreases with increasing water temperature. This effect influences very favourably the process of formation of flocculent suspension during periods of cold water. As a rule, the content of organic impurities in surface water decreases during these periods and the aggregates are formed mainly by mineral particles; they are fragile and less resistant to fragmentation so that the application of the low velocity gradient is advantageous. Since water contains lower concentrations of impurities during these cold periods, the longer action of the velocity gradient is favourable from the point of view of the kinetics of the process of aggregation. Similar adjustment of the conditions of agitation to the changes in water temperature for the plants agitated by mechanical means is difficult.

Influence of the degree of destabilisation of impurities on separation efficiency of the FLGM

The pilot plant tests were carried out during the period of the lowest water temperature; for investigations into the influence of changes of water temperature on the separation efficiency, the water was pre-heated to higher temperatures. For the purpose of verifying reproducibility of the results obtained, three series of pilot plant tests were carried out with water temperature varying between $T=5.0$ and 15.0°C , representing the usual water temperature variations at the Zelivka waterworks. The results obtained in all three series of tests were almost identical.

Therefore, as an example, only the results obtained from the first series of tests carried out with aluminium sulphate applied at dosages $D=16.0$ and $30\text{ mg} \cdot \text{l}^{-1}$ and water temperatures $T=5.8, 10.7$ and 15.5°C , are described below. These results are summarised in Table 3 and shown in Figure 6.

Figure 6 shows that the separation efficiency φ_{Al} changed depending on dosage and water temperature. The highest separation efficiency was achieved at a temperature $T=5.8^{\circ}\text{C}$ and a dosage $D=19.0\text{ mg} \cdot \text{l}^{-1}$. The average separation efficiency corresponded to $\varphi_{\text{Al}}=0.75$. At higher water temperatures the separation efficiency φ_{Al} decreased, but remained unchanged over the whole range

**Figure 6** | Dependence of separation efficiency φ and degree of aggregation α_A achieved by the FLGM at different alum dosages and water temperatures.

of dosages applied; at $T = 10.7^\circ\text{C}$ it dropped to $\varphi_{\text{Al}} = 0.71$ and at $T = 15.5^\circ\text{C}$ to $\varphi_{\text{Al}} = 0.62$.

The course of changes of residual COD_{Mn} value shows that at water temperatures $T = 5.8$ and 10.7°C the separation efficiency φ_{COD} increased with increasing dosage over the full range of dosages applied. At a temperature $T = 5.8^\circ\text{C}$ the rate of increase in φ_{COD} corresponded to that of residual Al. At a temperature $T = 15.5^\circ\text{C}$ and reagent dosages $D > 22.6 \text{ mg} \cdot \text{l}^{-1}$ a decrease in the φ_{COD} efficiency was evident. The maximum separation efficiencies φ_{COD} achieved at $T = 10.7$ and 15.5°C corresponded to $\varphi_{\text{COD}} = 0.26$.

The degree of aggregation α_{A} of the particles that were not intercepted by the FLGM increased with increasing reagent dosage and rising water temperature. At a temperature $T = 5.8^\circ\text{C}$ and a low dosage $D = 19.5 \text{ mg} \cdot \text{l}^{-1}$, the degree of aggregation dropped significantly from $\alpha_{\text{A}} = 0.07$ to $\alpha_{\text{A}} = 0.0$ and with increasing dosage it increased sharply from $\alpha_{\text{A}} = 0.0$ to $\alpha_{\text{A}} = 0.59$.

A drop of the degree of aggregation α_{A} to zero and the corresponding increase in separation efficiencies φ_{Al} and φ_{COD} is connected with the mechanism of particle interception. At low water temperatures the velocities of chemical reactions facilitating formation of hydroxopolymers are lower, and as a consequence the velocity of destabilisation reactions was also lower and they were slowing down after entering the FLGM. Therefore, at an optimum reagent dosage almost all particles of impurities, immediately after being destabilised, are intercepted on the surface of the medium grains and their mutual aggregation does not occur. In contrast, at higher dosages than the optimum one, the probability of particle collisions is higher due to an excessive quantity of hydroxopolymer particles present in the system; consequently the velocity of destabilisation reactions is also higher. As a result, the aggregation of destabilised particles is completed before they enter the FLGM. Higher temperatures affect the reaction velocity in a similar way—the reaction velocities are higher and the particles of hydroxopolymers, as well as the destabilised particles, move faster. Therefore, a drop in the value of degree of aggregation α_{Al} , accompanied by increased separation efficiencies φ_{Al} and φ_{COD} , occurred only at low water temperatures.

It is evident from the foregoing that the degree of destabilisation of impurities α_{D} influences the separation efficiency of the FLGM very significantly. The highest separation efficiencies were achieved at optimum reagent dosage $D = 19 \text{ mg Al}_2(\text{SO}_4)_3 \cdot \text{l}^{-1}$ and reaction pH around 6.0. The water temperature increase is accompanied by a significant drop in separation efficiency. In all cases, the degree of aggregation of particles not separated directly in the layer is high enough to enable these aggregates to be effectively separated, for instance, by deep bed filtration.

Influence of the degree of aggregation of destabilised particles on separation efficiency of the FLGM

The results from investigations into the influence of dosage of destabilisation reagent on separation efficiency of the FLGM proved that separation efficiency is influenced by the degree of destabilisation of particles of impurities α_{D} entering the FLGM. Decreasing separation efficiency with increasing water temperature also proved that separation efficiency is influenced by degree of aggregation of the destabilised particles entering the FLGM. This influence was investigated in the subsequent series of tests.

The basic arrangement of the FLGM reactor was used for these tests. The changes in the degree of aggregation α_{A} of the particles entering the FLGM were achieved by extending the feed pipe to the reactor with the reagent dosing point remaining at the inlet to the feed pipe. Agitation facilitating particle aggregation in the feed pipe was caused by means of flow in the feed pipe, and the period of particle aggregation corresponded to the retention time of water between the points of reagent addition and water discharge to the FLGM reactor. The results obtained from three series of tests were almost identical. Therefore, only the results from the tests taking place at optimum dosage of $D = 19.0 \text{ mg Al}_2(\text{SO}_4)_3 \cdot \text{l}^{-1}$ and water temperatures $T = 4.95, 9.05$ and 13.05°C are introduced below. The results obtained are summarised in Table 4 and illustrated in Figure 7.

It follows from Figure 7 that the degree of aggregation α_{A} determined at the inlet into the FLGM reactor

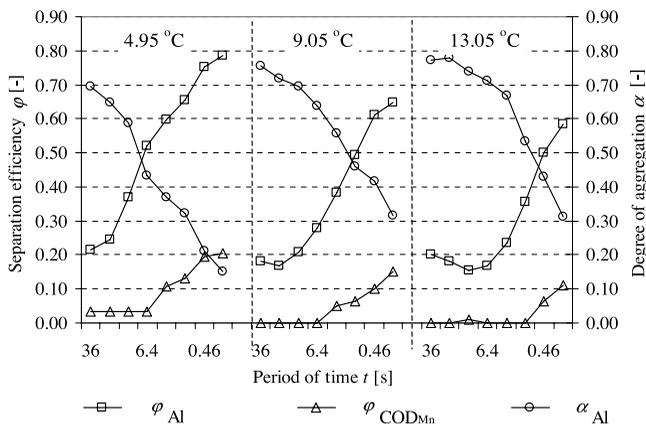
Table 4 | Influence of water temperature and duration of agitation upstream of the FLGM on separation efficiency of the FLGM

t (s)	$T=4.95^{\circ}\text{C}$			$T=9.05^{\circ}\text{C}$			$T=13.05^{\circ}\text{C}$		
	α_{Al}	φ_{Al}	φ_{COD}	α_{Al}	φ_{Al}	φ_{COD}	α_{Al}	φ_{Al}	φ_{COD}
36.00	0.49	0.28	0.00	0.69	0.19	0.02	0.58	0.43	0.05
21.00	0.52	0.25	0.02	0.66	0.19	0.04	0.62	0.36	0.13
12.50	0.53	0.26	0.05	0.68	0.19	0.07	0.52	0.43	0.01
6.40	0.55	0.26	0.05	0.63	0.19	0.05	0.63	0.34	0.24
3.60	0.47	0.31	0.09	0.56	0.25	0.11	0.60	0.29	0.05
1.37	0.41	0.43	0.010	0.58	0.27	0.11	0.61	0.30	0.08
0.46	0.28	0.58	0.14	0.56	0.32	0.12	0.60	0.28	0.06
0.23	0.27	0.67	0.21	0.52	0.41	0.15	0.58	0.30	0.01
0.11	0.25	0.72	0.25	0.42	0.52	0.23	0.54	0.35	0.00

increased with increasing length of the feed pipe. In contrast, the separation efficiencies φ_{Al} and φ_{COD} of the FLGM decreased as the period of aggregation increased and hence the value of degree of aggregation α_{A} of particles entering the FLGM increased. The highest separation efficiencies φ_{Al} and φ_{COD} were achieved when

the destabilisation reagent was dosed to the inlet into the FLGM reactor.

The influence of water temperature on separation efficiencies is also evident from Figure 7. With increasing temperature the aggregation efficiency α_{A} increased and consequently the separation efficiencies φ_{Al} and φ_{COD} decreased. The highest separation efficiencies were achieved at the lowest temperature and they decreased with increasing water temperature.

**Figure 7** | Influence of period of aggregation t prior to entering the FLGM reactor on the FLGM separation efficiency φ .

Significance of active surface area of medium grains

The size of grains of the FLGM, on the surface area of which the particles of impurities become attached, is also important from the separation efficiency point of view. This influence was investigated with increasing volume of four different size-fractions of the accurately sieved sand in the FLGM reactor. Since sand is anisometric in shape, grain size is defined, for the purpose of calculations, by an equivalent diameter calculated from the volume of an accurately defined number of grains determined by the pycnometric measurement. Basic data of the size-fractions of the sand used are summarised in Table 1.

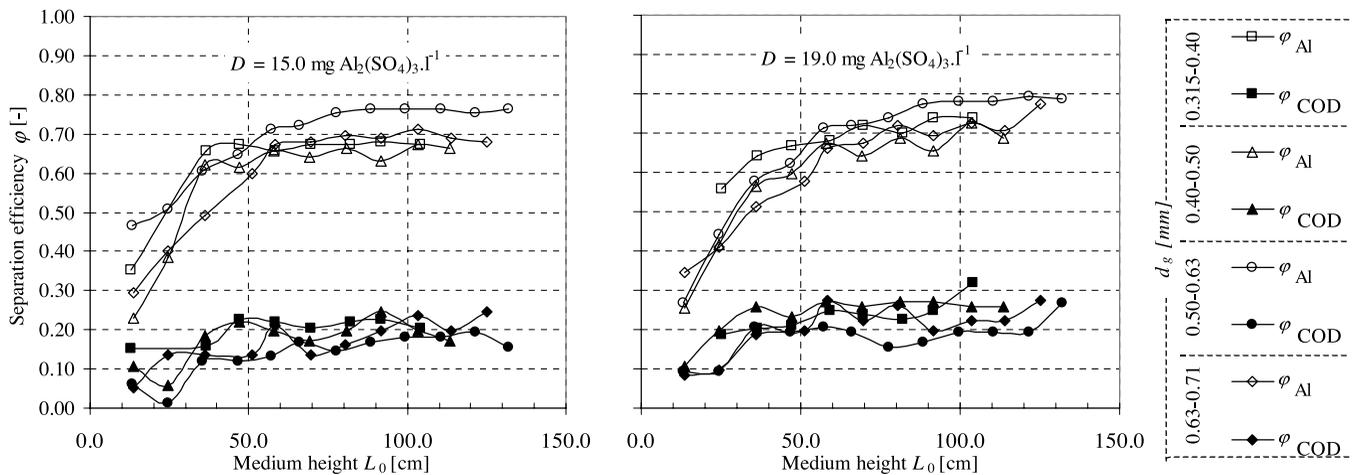


Figure 8 | Dependence of separation efficiency φ on medium height L_0 in FLGM reactor for different medium granulometries and alum dosages.

Since separation efficiencies φ_{Al} and φ_{COD} of the FLGM are also influenced by the degree of destabilisation α_D , the tests were carried out with two different dosages of aluminium sulphate:

- a lower dosage than optimum ($D = 15.0 \text{ mg} \cdot \text{l}^{-1}$) by which complete destabilisation of the particles of impurities is not achieved;
- an optimum dosage ($D = 19.0 \text{ mg} \cdot \text{l}^{-1}$) by which destabilisation of the particles of impurities takes place quantitatively and the chemical reactions are very fast.

The FLGM was operated at an upflow velocity $v_{\text{up}} = 13 \text{ mm} \cdot \text{s}^{-1}$ and water temperature varying between $T = 6.1$ and 6.3°C . The results obtained are illustrated in Figures 8 and 9.

The separation efficiencies φ_{Al} and φ_{COD} were lower at a suboptimum dosage when compared to those achieved at an optimum dosage. When these are compared with the results obtained with different individual medium size-fractions they did not differ much, as is shown in Figure 8. Furthermore, a comparison of the separation efficiencies achieved by the individual size-fractions shows that the same separation efficiencies were achieved when the total surface areas of grains in the FLGM were the same, regardless of the grain size.

The total surface area of grains enabling particle separation is determined by medium volume or its initial height. It follows from Figure 9 that besides the medium grain size a minimum height of medium required is also influenced by the degree of destabilisation α_D . At a lower α_D value the minimum medium height L_0 required was higher and increased with increasing medium coarseness. For clean medium grains its minimum height varied between $L_0 = 350$ – 600 mm . At an optimum reagent dosage it was slightly smaller $L_0 = 300$ – 550 mm .

With the separation mode of operation the basic characteristics of the FLGM change as coating of the grains progresses. Due to this, its operational parameters change, too. First of all, the media mass in the FLGM increases and, consequently, also its hydraulic resistance. Furthermore, due to interception of impurities on the grains, the size of grains also increases with the grains simultaneously changing shape. The initially anisometric sand grains gradually become spherical. Since the density of impurities is usually lower than that of medium grains, the density of the coated grains also gradually decreases.

The changes in head loss and medium depth in the FLGM, caused by the interception of impurities on the surface of medium grains, are evident from Figure 10. This figure shows that with increased medium mass in the

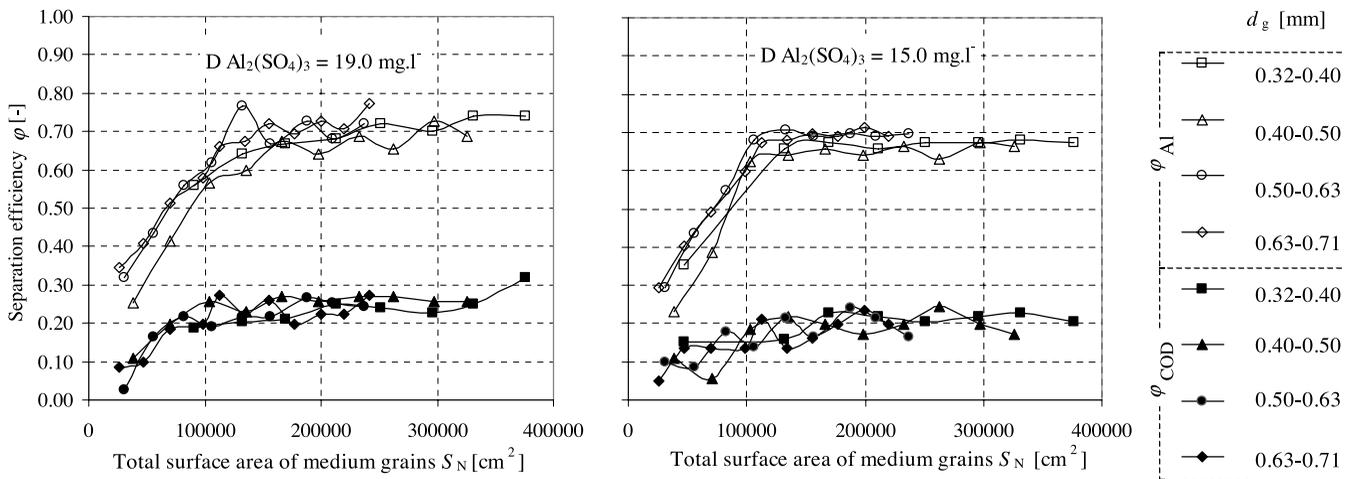


Figure 9 | Dependence of separation efficiency φ on medium grain surface area S_N for different grain-size fractions and two alum dosages.

FLGM by $\Delta m = 0.59 \text{ g} \cdot \text{h}^{-1}$ the pressure loss increased by $\Delta P = 6.4 \text{ kPa}$ and the height of the expanded medium layer by $\Delta L_E = 473 \text{ mm}$. This means that after 1250 hours of

operation its initial mass of medium increased by $\Delta m = 825.5 \text{ g}$.

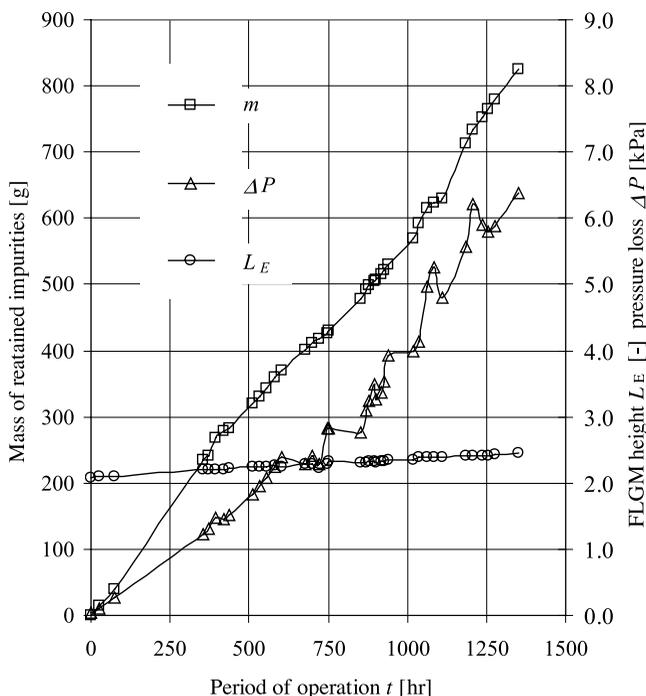


Figure 10 | Influence of period of operation t on pressure loss ΔP , FLGM height L_E and mass of retained impurities Δm .

CONCLUSIONS

1. With direct separation of impurities two processes take place simultaneously in the FLGM:
 - (a) adhesion of destabilised particles to the surface of medium grains;
 - (b) mutual combining of the remaining destabilised particles into flocculent aggregates.

The physical mechanism of both processes is identical. The prevailing process is dependent on the complex action of two basic parameters influencing the degree of destabilisation and degree of aggregation of particles entering the FLGM, namely, dosage of destabilisation reagent and water temperature.

2. The highest separation efficiencies are achieved with dosages of destabilisation reagent approaching the optimum dosage at which the highest degree of destabilisation is (achieved. With dosages larger than the optimum the separation efficiencies attainable by the FLGM decrease, because with increasing reagent dosage the velocity of particle

aggregation increases owing to the increasing number of particles of destabilisation reagent in the system. Comparison of these findings with the jar tests results shows that even at reagent dosages lower than the optimum one, higher separation efficiencies are readily achieved by the FLGM due to the high velocity gradient acting in the FLGM which makes forcible overcoming of the insufficiently suppressed stability barrier of the particles of impurities possible. Furthermore, the FLGM separation efficiency also increases with decreasing water temperature.

- The degree of aggregation of particles achieved prior to entering the FLGM is of great importance for the separation efficiency of the FLGM. At lower than optimum reagent dosages a drop in aggregation efficiency accompanied with a simultaneous increase in separation efficiency is evident. The highest separation efficiencies are achieved when the particles of impurities enter the FLGM immediately after being destabilised. With an increasing degree of aggregation of the particles achieved upstream of the FLGM its separation efficiency decreases.
- The test results proved that the FLGM separation efficiency is also dependent on the total surface area of all medium grains in the FLGM. Comparison of separation efficiencies achieved with the different medium size-fractions shows that achieving the same separation efficiencies requires the same total surface area of all grains in the FLGM.

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NOMENCLATURE

C_E concentration of determinant measured at the outlet from the unit operations ($\text{mg} \cdot \text{l}^{-1}$)

C_{EF} concentration of determinant remaining in the system at the point of testing after its separable portion was removed by centrifugation ($\text{mg} \cdot \text{l}^{-1}$)

C_O initial concentration of determinant measured in raw water ($\text{mg} \cdot \text{l}^{-1}$)

C_{OF} concentration of determinant remaining in raw water after its separable portion was removed by centrifugation ($\text{mg} \cdot \text{l}^{-1}$)

C_V volumetric concentration of medium grains in a stable layer ($\text{m}^3 \cdot \text{m}^{-3}$)

D reagent dosage ($\text{kg} \cdot \text{m}^{-3}$)

d_g diameter of grain (m)

d_{eqv} equivalent diameter (m)

E expansion ratio (–)

e differential expansion (–)

F cross-sectional area of the FLGM reactor (m^2)

G mean velocity gradient (s^{-1})

g gravity constant ($\text{m} \cdot \text{s}^{-2}$)

h hydraulic resistance of a stable layer (m)

H loss of head of a fluidised layer (m)

ΔH loss of head of a fluidised layer (m)

L_O height of a stable layer (m)

L_E height of a fluidised layer (m)

m mass of the medium (kg)

N_A total number of aggregated particles

N_D total number of destabilised particles

N_0 total number of all particles initially present in the system

N_S total number of particles separated (removed) from the system

ΔP pressure loss (Pa)

Q rate of flow ($\text{m}^3 \cdot \text{s}^{-1}$)

S surface of the particle (mm^2)

T water temperature ($^{\circ}\text{C}$)

t retention time in a fluidised layer (s)

V volume of a fluidised layer (m^3)

V_P volume of pores (m^3)

V_N volume of medium in FLGM (m^3)

V_C reactor volume occupied by medium (m^3)

v_{up} upflow velocity ($\text{m} \cdot \text{s}^{-1}$)

v_{tr} threshold velocity of fluidisation ($\text{m} \cdot \text{s}^{-1}$)

W mechanical energy dissipated into FLGM ($\text{kg} \cdot \text{m}^2 \cdot \text{s}^{-2}$)

α_A	degree of aggregation of the particles of impurities (—),
α_D	degree of destabilisation of the particles of impurities (—)
η	dynamic viscosity of water ($\text{kg} \cdot \text{m}^{-1} \cdot \text{s}^{-1}$)
ν	kinematic viscosity of water ($\text{m}^2 \cdot \text{s}^{-1}$)
ρ_g	density of medium grains ($\text{kg} \cdot \text{m}^{-3}$)
ρ_w	density of water ($\text{kg} \cdot \text{m}^{-3}$)
ω_O	porosity of a stable layer (—)
ω_E	porosity of a fluidised layer (—)
φ	separation efficiency (—)
φ_{Al}	separation efficiency expressed by Al (—)
φ_{COD}	separation efficiency expressed by COD _{Mn} value (—)

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