Aggregation capability of a fluidised layer of granular material during treatment of water with high DOC and low alkalinity

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ABSTRACT

The aim of this contribution is to investigate aggregation efficiencies when a fluidised layer of granular material (FLGM) is applied for a treatment of surface water with high DOC (Dissolved Organic Carbon) and low alkalinity in a pilot size operation. It was shown that the FLGM can be effective for agitation of a system using destabilised particles with the aim to form readily separable suspension. An aggregation efficiency of the FLGM reactor was evaluated by the degrees of destabilisation and aggregation of the particles of impurities. The maximal separation efficiencies of the pilot plant after filtration attained $\varphi_{Al} = 0.96$ and $\varphi_{DOC} = 0.73$. These values exceeded not negligibly the ones reported for the conventional aggregation processes.

Key words | aggregation, agitation, destabilisation, filtration, fluidised layer, water treatment

INTRODUCTION

Agitation represents one of the technological procedures applied in water treatment. It enables the aggregation of destabilised impurities and formation of separable aggregates (Tambo & Watanabe 1979; Francois 1988). Efficiency of separation processes and the properties of formed sludge strongly subjects to the properties of aggregates formed (their shape, size, geometric structure and density) generated by the conditions under which the aggregation is carried out (mean velocity gradient and time of its duration). The mean velocity gradient $\overline{G}$ is calculated from the relationship derived by Camp & Stein (1943)

$$\overline{G} = \sqrt{\frac{P}{V\eta}}$$

where $P$ represents power input to the water, $V$ is agitated volume of water, and $\eta$ stands for dynamic viscosity.

Equations (2) and (3) provide the calculation of the power input $P$ for the basins agitated by the hydraulic and mechanical means, respectively

$$P = \rho g \Delta H Q;$$

$$P = \sqrt{\frac{fApv^2}{2vV}} = Eu\rho n^3 d^5$$

where $\rho$ is liquid density, $g$ gravity constant, $\Delta H$ loss of head, $Q$ flow rate, $f$ resistance factor of the stirrer, $A_p$ area of stirrer blades perpendicular to movement, $v_r$ relative velocity of stirrer blades to liquid, $v$ kinematic viscosity, $Eu$ Euler number (power factor) adjusted for the conditions of mixing, $n$ a number of revolutions of the stirrer, and $d$ stirrer diameter.

Intensity of agitation and distribution of a velocity field in an agitated volume depend on an ultimate size of the aggregates required, which is determined by a method used for their separation. The rapid (high intensity) agitation, i.e. with velocity gradient $\overline{G} > 100 \text{ s}^{-1}$, results in a formation of micro-aggregates. The slow (low intensity) agitation, i.e. $\overline{G} < 60 \text{ s}^{-1}$, enables a formation of macro-aggregates from
the micro-aggregates formed during the preceding phase of rapid agitation. The slow agitation has its justification as the second phase only when formation of large and readily settleable aggregates is required and is only applied after completion of the formation of aggregates in the phase of rapid agitation. This has been verified in the following literature (Mutl et al. 2006; Pivokonska & Pivokonsky 2006).

The pilot plant operations proved that the FLGM can be utilised in water treatment as the process unit for formation of suspension separable by sedimentation, dissolved air flotation, filtration through a fluidised layer of flocculent suspension and deep bed filtration (Mutl et al. 1999, 2000). The FLGM was tested in the pilot plant studies for treatment of water containing high concentrations of mineral suspensions (Tambo & Matsui 1989) as well as for treatment of highly coloured and turbid water (Tambo & Wang 1993a, b). Applicability of the FLGM in the pilot plant and full size operations was tested in the treatment of common quality surface water (Mutl et al. 2000).

The magnitude, distribution of velocity gradient and duration of its action are the principal hydrodynamic factors affecting aggregation efficiencies of the FLGM. When liquid flows through a fixed layer of granular particles the hydraulic resistance of this layer increases with increasing flow velocity. When the flow velocity exceeds the critical one \( v_s \), the layer expands and fluidisation occurs. The upflow velocity \( v_{up} \) above the layer is given by

\[
v_{up} = \frac{Q}{F}
\]

where \( Q \) represents flow rate and \( F \) is a cross section of the FLGM reactor.

The hydraulic resistance \( h \) at which the layer expands is expressed by the relation

\[
h = L_0(1 - \omega_0) \cdot \left( \frac{\rho_k}{\rho_w} - 1 \right)
\]

where \( L_0 \) and \( \omega_0 \) are an initial height and porosity of stable layer, respectively; \( \rho_w \) and \( \rho_k \) is density of liquid and medium (granular material), respectively.

Transition from a stable layer to a fluidised one is indicated by a change in the FLGM height from that of stable layer \( L_0 \) to a height of fluidised layer \( L_E \), its porosity from \( \omega_0 \) to \( \omega_E \) and by the mechanical energy \( W \) dissipated into the FLGM and expressed by a change in loss of head \( H \) given by

\[
H = L_E \left( 1 - \frac{\omega_E}{\omega_0} \right) \frac{(\rho_k - \rho_w)}{\rho_w} = L_E \left( 1 - \omega_E \right) \left( \frac{\rho_k}{\rho_w} - 1 \right)
\]

(6)

The expansion \( e \) is defined by the relationship

\[
e = \frac{L_E - L_0}{L_0} = E - 1
\]

(7)

where the expansion ratio \( E \) is expressed by

\[
E = \frac{L_E}{L_0}
\]

(8)

The volume \( V_v \) of the voids in a layer is determined by

\[
V_v = V_t \omega_E
\]

(9)

where \( V_t \) is a total volume of the stable layer

Retention time \( T \) in the layer of the medium is expressed by the relationship

\[
T = \frac{V_v}{Q} = \frac{V_t \omega_E}{v_{up} F} = \frac{L_E \omega_E}{v_{up}} = \frac{L_0 (E + \omega_0)}{v_{up}}
\]

(10)

When liquid flows through the FLGM, each grain of the layer is a source of hydraulic shock and an origin of velocity gradient, the magnitude of which is given by

\[
\mathcal{G} = \sqrt{\frac{W}{V_H}} = \sqrt{\frac{\Delta H}{v_T}} = \sqrt{\frac{g(1 - \omega_0) \cdot (\rho_k - \rho_w) v_{up}}{v(T + \omega_0)}}
\]

(11)

Since each grain in the FLGM represents an independent agitation element, achievement of full fluidisation of the medium is important from the viewpoint of a character of the aggregates formed. A uniform distribution of velocity gradient and its constant value is attained when the FLGM is fully fluidised. The mean velocity gradient \( \mathcal{G} \) generated by the FLGM is significantly influenced by water temperature (Mutl et al. 1999).

In this paper there are summarised the results of operating the FLGM as an agitation element (high intensity agitation), aimed at the verification of its aggregation
capability during treatment of water with high DOC and low alkalinity in a pilot size operation.

METHODS

The experimental work was carried out at the Mezibori waterworks (near the town Litvinov, Czech Republic), treating water from the impounding Flaje reservoir. This reservoir is characterized by a relatively stable water quality and stable temperature over a long time period, which ensures reproducibility of the pilot plant operations and hence the results obtained.

The basic technological parameters (dosages of alkalisation and destabilisation reagents) were determined by the jar tests. A variable speed eight-paddle stirrer LMK 8-04 (EJP company, Czech Republic), and 2-litre jars were used for jar testing. The procedure consisted of a 1 minute rapid mixing \( \left( \bar{G} = 400 \text{s}^{-1} \right) \), 15 minutes of slow mixing \( \left( \bar{G} = 70 \text{s}^{-1} \right) \) and 60 minutes of settling.

The pilot plant tests were carried out with a plant consisting of a FLMG reactor (FLGM) and rapid gravity sand filter (RGF), see Figure 1. The FLGM reactor consisted of a 3,000 mm long Umaplex tube of an inner diameter \( d = 90 \text{mm} \) charged with silica sand of granularity \( d_s = 0.40–0.50 \text{mm} \). The sand depth corresponded to \( L_0 = 1,500 \text{mm} \). The FLMG expansion ratio \( E \) attained approximately 1.75. The mean velocity gradient produced was \( \bar{G} = 260 \text{s}^{-1} \) and a retention period in the fluidised layer corresponded to \( T = 170 \text{s} \). The velocity gradient was calculated using a head loss measured by head probe (P0) The rapid gravity filter consisted of a 2,500 mm long Umaplex tube of an inner diameter \( d = 125 \text{mm} \) with the probes (P1–P6) for measurement of head loss (14) along the depth and flow regulator (11) installed in the filtrate discharge pipe. The filter was charged with silica sand of a granularity \( d = 0.80–1.25 \text{mm} \) to a filter bed depth \( L_F = 1,100 \text{mm} \). During the pilot plant operation several filtration runs of the RG filter were evaluated at a filtration velocity \( v_f = 3.0, 6.0 \) and \( 9.0 \text{m} \cdot \text{h}^{-1} \). The backwashing of RGF is carried out using tap water (12).

The raw water is fed at maximum flow rate from raw water main (1) to a head tank. From the head tank raw water gravitates \( \text{via} \) a pipe fitted with flow regulator (3) and flow-meter (4) to a flow regulating system which is connected with the FLGM reactor. Flow regulating system enables stable flow rate (required to compensate the increasing head loss caused by formation of a cover layer on the grains of media) and automatic re-start of the plant in the case of accidental raw water cut off. The feed pipe, just before connecting to the FLGM reactor, is provided with the facilities for injection of an alkalisation reagent (lime water) (6), a destabilisation reagent (aluminium sulphate) (7) and a tube for sampling of the dosed water (8). Because the FLGM reactor was operated in aggregation mode in all cases of measurements, the facilities for injection of both reagent and its homogenisation with the raw water were placed at the inlet into a long pipe installed before a reactor. The water from the FLGM reactor gravitates through a pipe (10) with a tube for sampling of the flocculated suspension into the RGF. The filtration velocity is regulated using a flow regulator (11) installed in the filtrate discharge pipe. In the outlet from flow regulator there is placed a final sampling tube (15).

Data characterising the state of the FLGM reactor (head loss \( H_R \) and depths of layers \( L_0, L_E \) and filtration (head loss \( H_F \) and duration of filtration run \( T_F \) ) were used as the operational parameters. The duration of filtration runs \( T_F \) was limited either on exhausting the pressure head \( H_F \), or exceeding the maximum allowed Al concentration in tap water (\( C_{Al} = 0.20 \text{mg} \cdot \text{l}^{-1} \)). For the purpose of verifying reproducibility of the results obtained, three series of the pilot plant tests were carried out.

The efficiency of the water treatment process was evaluated by:

1. a direct comparison of water quality indicators: content of a metal component of destabilising reagent Al, chemical oxygen demand CODMn, dissolved organic carbon DOC, pH value, alkalinity and temperature;
2. determining separation efficiency \( \varphi \) of Al and DOC, according to the relationship

\[
\varphi = 1 - \left( \frac{C_F}{C_0} \right)
\]

where \( C_0 \) is the initial concentration of Al or DOC used and \( C_F \) is the final value in treated water.

The effect of agitation was evaluated by the degree of aggregation \( \alpha \) and determined in crucial stages of

\[
d = \frac{3.0 - 1.25}{0.80 - 1.25} \approx 1.75
\]
Figure 1 | Diagram of the pilot plant.
formation of suspension. It is calculated according to the relationship

$$\alpha_A = \frac{C_0 - C_F}{C_0}$$  \hspace{2cm} (13)

where $C_0$ is the concentration of monitored indicator (Al, DOC) at the point of testing and $C_F$ is the concentration of the monitored indicator determined in the sample after its treatment by centrifugation under the defined conditions.

Density and size of formed aggregates were determined by a test of aggregation, which enables separation of the formed aggregates into the four basic categories - non-aggregate particles ($NA$), primary aggregates ($PR$), micro-aggregates ($MI$) and macro-aggregates ($MA$). The particle distribution existing in the system at any given moment is evaluated by the test of aggregation that is based on sedimentometric analysis. The technologically significant categories of particles are determined according to the following relationships

$$MA = \frac{C_0 - C_5}{C_0}$$  \hspace{2cm} (14)

$$MI = \frac{C_5 - C_{60}}{C_0}$$  \hspace{2cm} (15)

$$PR = \frac{C_{60} - C_{60F}}{C_0}$$  \hspace{2cm} (16)

$$NA = \frac{C_{60F}}{C_0}$$  \hspace{2cm} (17)

and therefore

$$MA + MI + PR + NA = 1$$  \hspace{2cm} (18)

where $C_0$, $C_5$ and $C_{60}$ are the concentrations of the monitored parameter measured in the samples taken at the beginning of sedimentation, after 5, and after 60 minutes of sedimentation, respectively. $C_{60F}$ is a concentration of the monitored parameter determined in the sample after its treatment by centrifugation under the defined conditions (Hereit et al. 1980; Polasek & Mutl, 1996). All analytical measurements were conducted in triplicate and the errors were less than 3%.

### RESULTS

A pilot plant was installed in the Mezibori waterworks (planned output 400 l s$^{-1}$) treating water from the impounding reservoir Flaje. The perforated baffle-type flocculation chamber is used as an agitation element. The aggregates formed are separated using primary filtration. The water quality represents high concentration of organic matter of humic character ($DOC = 7.4 \text{ mg l}^{-1}$) and low alkalinity ($KNK_{4.5} = 0.16 \text{ mmol l}^{-1}$).

The efficiency of destabilisation of impurities in the treatment of humic water is, in general, reaction pH dependent. The jar tests were used for optimisation of aluminium sulphate dosage and reaction pH, see Figures 2 and 3. The highest destabilisation efficiencies represented by residual Al concentration value $\alpha_A(\text{Al}) = 0.94$ and DOC value $\alpha_A(\text{DOC}) = 0.76$, determined in the centrifuged samples taken after sedimentation, were reached at dosage varying over a narrow range from $D = 0.021$ to $0.026 \text{ mmol l}^{-1}$ and pH ranged between 5.8 and 6.2.

Figure 4 shows the influence of dosage of destabilising reagent on the particle size distribution. The minimal portion of non-aggregated particles ($p = 0.02$) was detected in the system for optimal values of destabilising reagent ($D = 0.021$ to $0.026 \text{ mmol l}^{-1}$) and reaction pH (around 6.0). The high portions of non-aggregated particles were measured for the lower and higher values of
destabilising reagent doses. In the jars with optimal destabilisation conditions, there were measured the high portions of primary- and micro-aggregates and relatively low portion of macro-aggregates.

The pilot plant testing was aimed at evaluating the aggregation efficiency of the FLGM (high intensity agitation). The pilot plant arrangement was supplemented with a long pipe installed upstream of the reactor, where the destabilizing particles aggregated into the primary-aggregates. The influence of varying dosage of aluminium sulphate on aggregation efficiency of the FLMG and at pH = 5.9 regulated by dosing of lime water, was established using pilot plant testing. The maximum efficiencies were achieved for the dosages varying over the same narrow range, which was achieved by the jar tests (\(D = 0.021\) to \(0.026\) mmol·l\(^{-1}\)). For an aluminium sulphate dosage \(D = 0.023\) mmol·l\(^{-1}\) (determined as the optimum dosage) the high degrees of aggregation \(\alpha_A(Al) = 0.92;\  \alpha_A(DOC) = 0.73\) were achieved using the FLMG reactor.

The efficiency of the FLMG on aggregation is demonstrated in Figure 5 depicting the portion of size fractions of the formed aggregates. A predominant portion of the formed aggregates is constituted by the primary-aggregates \((p_{PR} = 0.67)\), which are favourable for the one-step separation by filtration. The relatively high portion of micro-aggregates \((p_{MI} = 0.19)\) and low portion of non-aggregated particles \((p_{NA} = 0.08)\) and macro-aggregates \((p_{MA} = 0.06)\) were measured in the system using FLMG reactor as the agitation element. In contrast to it, the high portions of non-aggregated particles \((p_{NA} = 0.37\) and 0.34) and relatively high portion of macro-aggregates \((p_{MA} = 0.21\) and 0.16) were measured during agitation using the perforated baffles (low intensity agitation) in the full size agitation element installed at the waterworks and in the pilot plant, see Figure 5. The aggregates formed during slow agitation exhibit unsuitably wide size-distribution using the perforated baffles in full size or in a pilot plant operation.
During the pilot plant operation several filtration runs of the RG filter were evaluated at the filtration velocities $v_f = 3, 6$ and $9 \text{ m·h}^{-1}$. The results characterising the course of typical filtration runs for all filtration velocities are summarised in Table 1 and illustrated in Figures 6–8.

At a filtration velocity $v_f = 3 \text{ m·h}^{-1}$, the separation efficiency of the RG filter attained $\varphi_{Al} = 0.96$ and $\varphi_{DOC} = 0.73$, and the filtration run ended after 28 h operation also due to Al breakthrough at a very low filter head loss $\Delta H = 2.1 \text{ kPa}$.

For the parallel tests carried out with substantially higher velocities $v_f = 6$ and $9 \text{ m·h}^{-1}$, the total separation efficiencies of the pilot plant after filtration were $\varphi_{Al} = 0.95$ and 0.93, and $\varphi_{DOC} = 0.69$ and 0.67, respectively. The filtration runs during application of both highest filtration velocities were characterised by a short time of duration and its runs ended after 19 h ($v_f = 6 \text{ m·h}^{-1}$) and 14 h ($v_f = 9 \text{ m·h}^{-1}$) operation also due to Al breakthrough. The maximal separation efficiencies of the

### Table 1 | Comparison of operational performance and separation efficiencies of the pilot plant

<table>
<thead>
<tr>
<th>FLGM reactor</th>
<th>RG filter - outlet</th>
</tr>
</thead>
<tbody>
<tr>
<td>Inlet</td>
<td>Outlet</td>
</tr>
<tr>
<td>Al [mg·l$^{-1}$]</td>
<td>1.32</td>
</tr>
<tr>
<td>DOC [mg·l$^{-1}$]</td>
<td>7.40</td>
</tr>
<tr>
<td>$\alpha_A$ (Al) [-]</td>
<td>-</td>
</tr>
<tr>
<td>$\alpha_A$ (DOC) [-]</td>
<td>-</td>
</tr>
<tr>
<td>$\varphi_{Al}$ [-]</td>
<td>-</td>
</tr>
<tr>
<td>$\varphi_{DOC}$ [-]</td>
<td>-</td>
</tr>
<tr>
<td>$T_F$ [h]</td>
<td>-</td>
</tr>
<tr>
<td>$H$ [kPa]</td>
<td>-</td>
</tr>
</tbody>
</table>
The high portion of primary-aggregates (\(\varphi_{PR} = 0.67\)) and micro-aggregates (\(\varphi_{MI} = 0.19\)), and low portion of non-aggregated particles (\(\varphi_{NA} = 0.08\)) and macro-aggregates (\(\varphi_{MA} = 0.06\)) were measured in the system using the FLMG reactor as an agitation element.

- The filtration run ended after 28 h (\(v_f = 3\, \text{m·h}^{-1}\)), 19 h (\(v_f = 6\, \text{m·h}^{-1}\)) and 14 h (\(v_f = 9\, \text{m·h}^{-1}\)) operation also due to Al breakthrough at very low head loss \(\Delta H = 2.1\, \text{kPa}\) (\(v_f = 5\, \text{m·h}^{-1}\)), 2.7 kPa (\(v_f = 6\, \text{m·h}^{-1}\)), and 3.4 kPa (\(v_f = 9\, \text{m·h}^{-1}\)). The maximal separation efficiencies of the pilot plant after filtration attained \(\varphi_{Al} = 0.96\) and \(\varphi_{DOC} = 0.73\).

- The filtration run of the full size filters installed at the waterworks, when the plant is operated at a dosage \(D = 0.027\, \text{mmol·l}^{-1}\) and a filtration velocity 3 m·h\(^{-1}\) ended after 14 hours of operation due to Al breakthrough. The maximal separation efficiencies of the full size plant after filtration are \(\varphi_{Al} = 0.82\) and \(\varphi_{DOC} = 0.64\).

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