Application of a fluidised layer of granular material in the treatment of surface water. Part 2: separation efficiency of the layer

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ABSTRACT

Separation efficiencies attainable with a fluidised layer of granular material (FLGM) when applied for the treatment of different quality surface waters in pilot and full size operations were investigated. It was established that the FLGM can be effective for direct separation of the destabilised impurities and for agitation of systems containing destabilised particles aimed at the formation of readily separable suspension. The separation efficiency of FLGM was found to be, among other factors, influenced by reaction conditions influencing the process of destabilisation of impurities, such as dosage of destabilisation reagent, water temperature, reaction time and measure of FLGM expansion. Furthermore, it was established that incorporation of a FLGM reactor into the treatment process design significantly improved the operational performance of the works, especially in the process configuration of a single-stage separation arrangement. Replacement of silica sand used as the FLGM charge with thermally treated bentonite enabling easier processing of the coated grains as the raw material for production of Al_2O_3 was verified.

Key words | aggregation, destabilisation, fluidisation, water treatment

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INTRODUCTION

When verifying the FLGM effect with respect to particle aggregation at the Želivka waterworks (Mutl & Knesl 1999) it was discovered that, in addition to aggregation efficiency, the FLGM also manifests a separation capability. This capability has been investigated in a number of research projects aimed at uprating several waterworks treating water of different characteristics and degree of pollution. The results obtained from the pilot scale operations at the Meziboří and Kutná Hora waterworks together with the results obtained from the pilot and full size operation at the Vír waterworks, where one of the existing sludge blanket clarifiers was converted to a FLGM reactor, are presented in this paper. Furthermore, for information only, comparison of performance results of a FLGM reactor installed at the Waterworks Želivka operating in the aggregation 'A' mode (without interception of the destabilised particles of impurities in the FLGM) as well as in the separation 'S' mode (with interception of the

destabilised particles of impurities on to the grains of the FLGM) are also presented. This was carried out when the reactor was operating in the S mode with the unmodified A mode reactor arrangement in which only the reagent dosing point was changed.

The theoretical principles of the aggregation and separation processes taking place in the FLGM reactor together with the results of research into the influence of different physical-chemical factors on the course and efficiency of separation of impurities in a FLGM reactor installed at the Zelivka waterworks will be presented elsewhere.

EXPERIMENTAL PART

Operational performance of the plant with single and two-stage separation of suspension arrangement with the

Table 1 | Technological parameters of the tested raw waters

Waterworks		Meziboří		Kutná Hora		Vír		Želivka	
Reservoir		Fláje		Vrchlice		Vír		Švihov	
		min	max	min	max	min	max	min	max
Temperature	°C	2.0	12.0	3.5	12.7	3.0	12.0	2.4	6.1
pH	—	6.0	6.1	6.4	7.2	6.0	7.5	7.0	7.7
Alkalinity	mmol l^{-1}	0.08	0.12	0.40	1.80	0.25	0.60	0.75	0.97
Acidity	mmol l^{-1}	0.04	0.10					0.02	0.15
COD_{Mn}	mg O_2 l^{-1}	2.5	4.0	3.5	9.5	2.0	8.5	2.8	3.6
Mn	mg l^{-1}		0.00		0.00	0.00	0.85	0.00	0.06

FLGM reactor operating in the A and S mode was evaluated based on the results obtained from jar tests, pilot and full size plant operations.

The FLGM reactor was tested at those localities where the technological survey aimed at obtaining design data for maximising the capacity of the existing works for the purpose of increasing their capacity and improving the treated water quality. The technologically significant determinants of the raw waters tested are shown in Table 1.

Experimental facilities

Basic technological parameters were determined by jar tests, evaluated in accordance with the methodology described by Hereit & Mutl (1980) and Polasek & Mutl (1996). The testing procedures according to this methodology are superior to that commonly used by centrifugation of a portion of samples to be analysed. This facilitates determination of the quantity of the non-separable portion of impurities remaining in the water and hence indicates the attainable treatability of water.

The pilot plant tests were carried out with a plant consisting of a FLGM reactor (FLGM), sludge blanket clarifier (SBC) and rapid gravity sand filter (RGF). The plant arrangement is shown in Figure 1.

The reactor (FLGM) consisted of a 3000 mm long Perspex tube of an inner diameter $d = 80$ mm. During testing it was charged with silica sand of a granularity $d = 0.315$–0.80 mm to a depth so that at a no flow condition the sand depth corresponded to $L_0 = 1000$ or 1250 mm. The FLGM expansion ratio E was approximately 1.75. Other materials, such as thermally treated bentonite and aluminium sludge of the same granularity and layer depth were also tested.

The sludge blanket clarifier (SBC) consisted of a 2250 mm long Perspex tube of an inner diameter $d = 300$ mm. In the lowermost part there is a perforated baffle defining the zone for slow agitation (11), where water is agitated by means of an electrically driven paddle-type stirrer. The sludge blanket level is maintained constant by means of overflow of excess sludge from the level of the sludge blanket in the compartment (12) into a sludge compartment (13), which is caused by continuous water draw-off (14) from the sludge compartment.

The rapid gravity filter (RGF) consisted of a 2500 mm long Perspex tube of an inner diameter $d = 125$ mm. It is provided with outlets for measurement of head loss along the bed depth and a flow regulator (15) installed in the
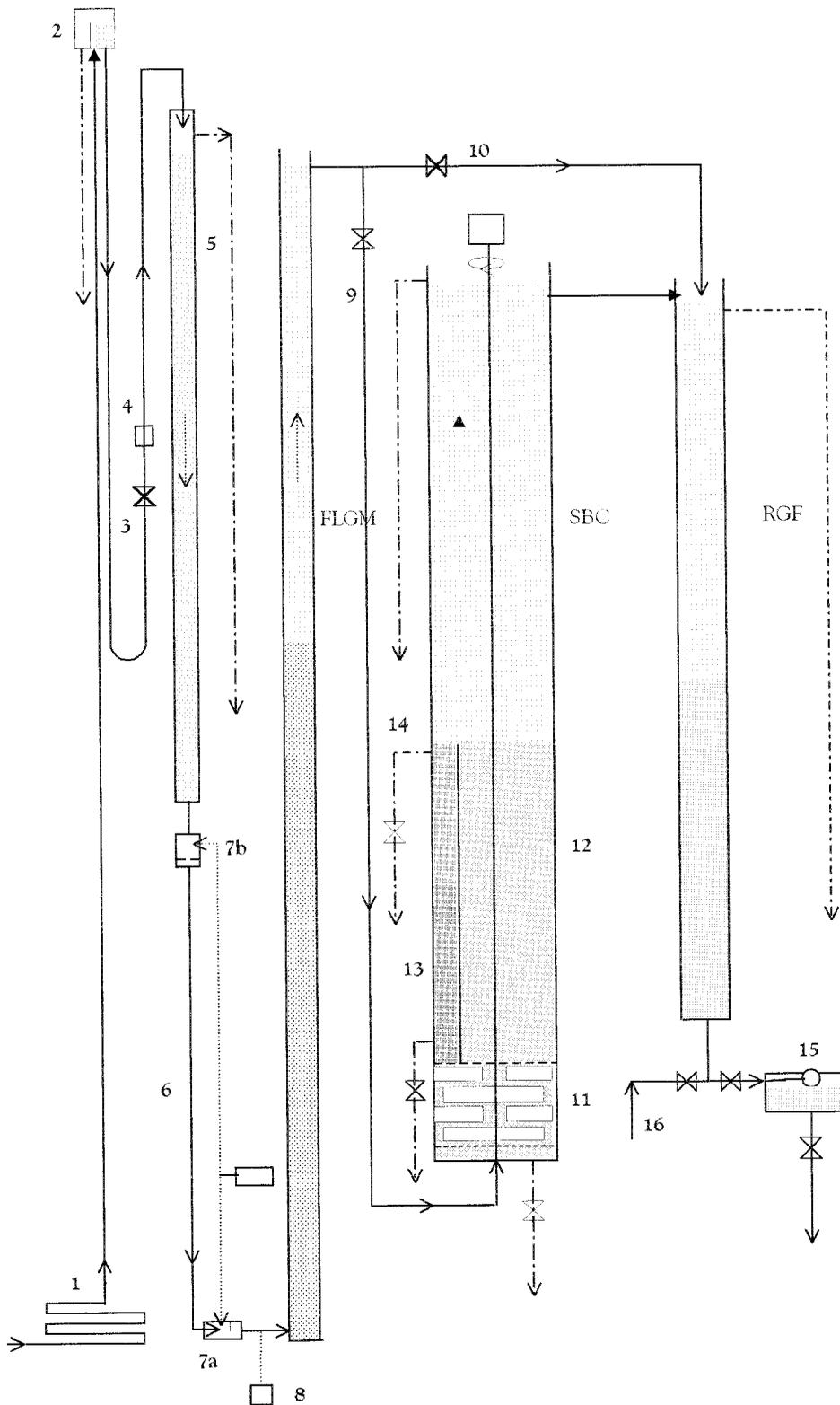
Figure 1 │ Flow diagram of the pilot plant with FLGM reactor, sludge blanket clarifier and rapid gravity filter.

filtrate discharge pipe. Filter backwashing is carried out by tap water (16). The filter was charged with silica sand of a granularity $d = 0.50$–1.25 mm to a filter bed depth $L_F = 1000$ mm.

The raw water is fed via a heating element (1), enabling temperature adjustment between 0.5°C and 25.0°C at maximum flow, to a header tank (2). From the head tank water gravitates via a pipe fitted with a regulation valve (3) and a rotameter (4) to a flow regulating system (5) and then via a pipe (6) into the FLGM reactor. The feed pipe, just before connecting to the reactor, is provided with a pressure sensor (8) and a tube for sampling of the dosed water. When the reactor is operating in the separation mode S, a facility for injection of a destabilisation reagent and its homogenisation with the raw water is placed close to the intake to the reactor (7a). When it is operating in the aggregation mode A, the feed pipe is replaced with a long pipe installed upstream to the reactor and the facility for injection of a destabilisation reagent is placed at the inlet into this pipe (7b). The flow regulating system allows for continuous plant operation, automatic adjustment of the flow rate (required to compensate for increasing head loss caused by formation of a cover layer on the grains of media) and automatic re-start of the plant in case of accidental raw water cut off.

Water from the FLGM reactor gravitates through pipe (9) into the sludge blanket clarifier (SBC) with a fully fluidised blanket of flocculated suspension or through a bypass (10) into the filter (RGF).

Aluminium sulphate of a quality available at the individual waterworks was used as the destabilisation reagent. The raw waters were pre-treated with sulphuric acid and lime (as lime water) respectively.

Methods of evaluation

The aggregation efficiency α and separation efficiencies φ_{Al} and φ_{COD} achieved were evaluated by the methodology described in Mutl (1983, 1984, 1985, 1990). The lowest efficiencies are achieved when α and φ values equal zero and highest efficiencies are reached when they approach 1.0. The basic determinants measured were: content of metal component of destabilising reagent Al, chemical

oxygen demand determined as oxidisability COD_{Mn} (Kubel), pH value and micro-organism count Bio. Other determinants, such as alkalinity, acidity, turbidity, colour etc. were measured, but are not stated and commented on in this paper.

Data characterising the state of the FLGM reactor, such as head loss H_R, depth of layer L_0, L_E and filtration, namely head loss H_F and duration of filtration run T_F, were used as the operational parameters. The length of filtration runs T_F was limited either on exhausting the pressure head available H_F, or exceeding the maximum permissible concentration of Al in the filtrate ($c = 0.20$ mg Al l^{-1}).

RESULTS AND DISCUSSION

Waterworks Meziboří

The Meziboří waterworks, design capacity $Q = 600$ l s^{-1}, treats water from the Fláje impounding reservoir. By its quality, the raw water represents a very soft, low salinity water with high leaching capability, polluted mainly by organic matter of humic character; mineral suspensions are rarely present.

The treatment process was designed for removal of humic substances and it includes:

- prealkalisation of the raw water with lime water;
- destabilisation of impurities with aluminium sulphate;
- formation of suspension in a swirling type mixer;
- separation of the formed suspension by direct filtration in a single stage by direct filtration; and
- hardening of the treated water with carbon dioxide.

The efficiency of destabilisation of impurities in the treatment of humic waters is, in general, reaction pH dependent. The typical jar test results obtained from optimisation of both aluminium sulphate dosage and reaction pH are shown in Figure 2. The efficiency of destabilisation process expressed by residual Al concentration and COD_{Mn} value, determined in centrifuged samples taken after sedimentation (AlF, COD_{Mn}F), does
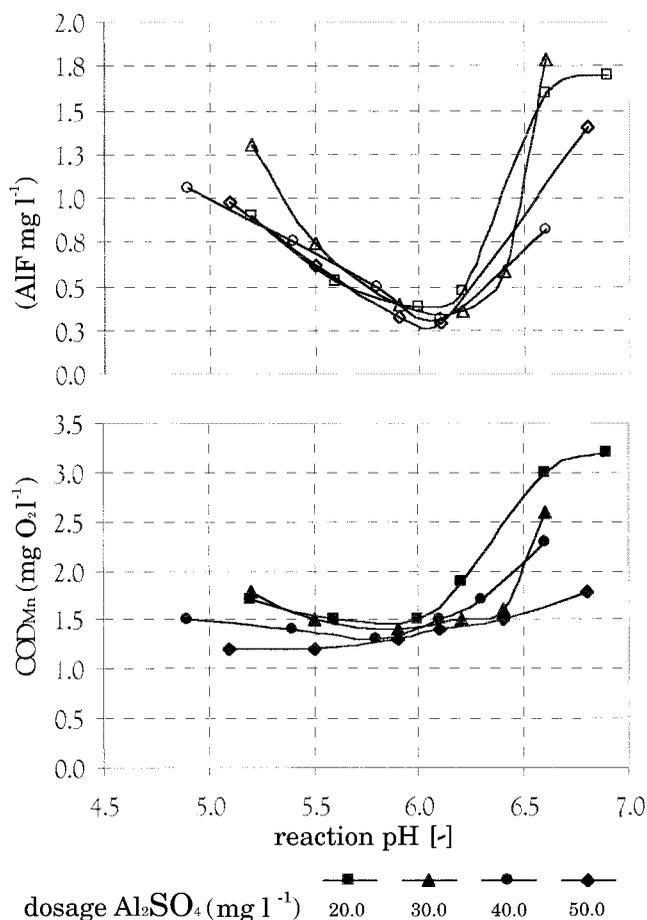
Figure 2 | Jar test optimisation of aluminium sulphate dosage and reaction pH: Waterworks Meziboří.

Figure 3 | Pilot plant optimisation of aluminium sulphate dosage with respect to separation efficiencies.

not vary much over the full range of dosages applied. The highest destabilisation efficiencies $\varphi_{Al} = 0.90$ and $\varphi_{COD} = 0.57$ were obtained in a relatively narrow pH range between pH 5.8 and 6.2.

The influence of varying dosages of the aluminium sulphate on the separation efficiency of the FLGM at pH = 5.8 regulated by dosing of lime, was established by pilot plant testing with the layer depth $L_0 = 1250$ mm. The results obtained are shown in Figure 3. The maximum efficiency was achieved at dosages varying over a relatively wide range from $D = 12.0$ to 25.0 mg l^{-1}. The separation efficiencies achieved are slightly lower than those achieved by jar tests and their value correspond to $\varphi_{Al} = 0.48$ and to $\varphi_{COD} = 0.32$.
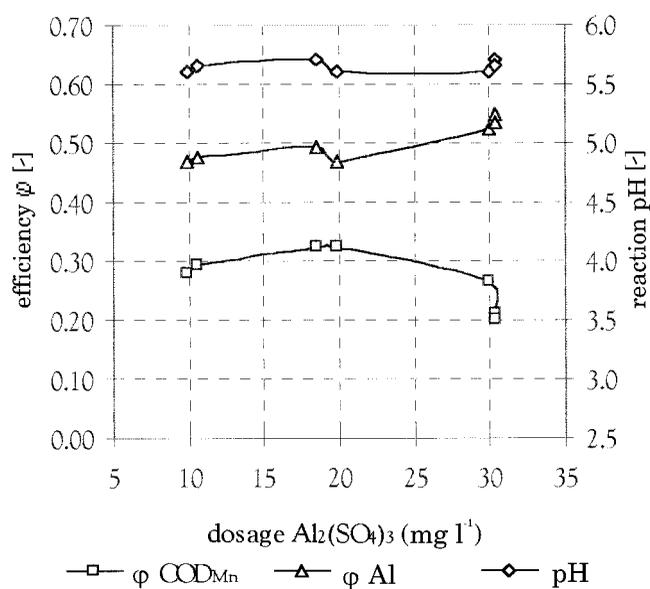
The results obtained from the optimisation tests were verified during pilot plant operation at organic pollution levels characterised by the average value $COD_{Mn} = 2.90$ mg O_2 l^{-1}. Aluminium sulphate was applied at a dosage $D = 22.0$ mg l^{-1} and reaction pH was adjusted by lime water to pH 5.8. The results are shown in Figure 4. As can be seen, the separation efficiencies attained by the FLGM remained relatively steady around $\varphi_{Al} = 0.60$ and $\varphi_{COD} = 0.40$ over the entire period of testing.

During the pilot plant operation several filtration runs of the RG filter were evaluated at a filtration velocity $v_f = 10.0$ m h^{-1}. At the time of testing the raw water was highly polluted by microorganisms which cannot be retained in the FLGM, $COD_{Mn} = 3.19$ mg O_2 l^{-1}, reaction $pH = 5.95$. The results characterising the course of a typical run are summarised in Table 2 and illustrated in Figure 5. At a dosage $D = 22.0$ mg l^{-1} (1.78 mg Al l^{-1}), the separation efficiencies of the FLGM averaged $\varphi_{Al} = 0.52$ and $\varphi_{COD} = 0.21$ and the total separation efficiencies achieved by the plant after filtration are $\varphi_{Al} = 0.85$ and $\varphi_{COD} = 0.53$. The filtration runs ended due to Al breakthrough after 16 hours operation at a very low filter head loss $\Delta H = 6.2$ kPa. In a parallel test carried out
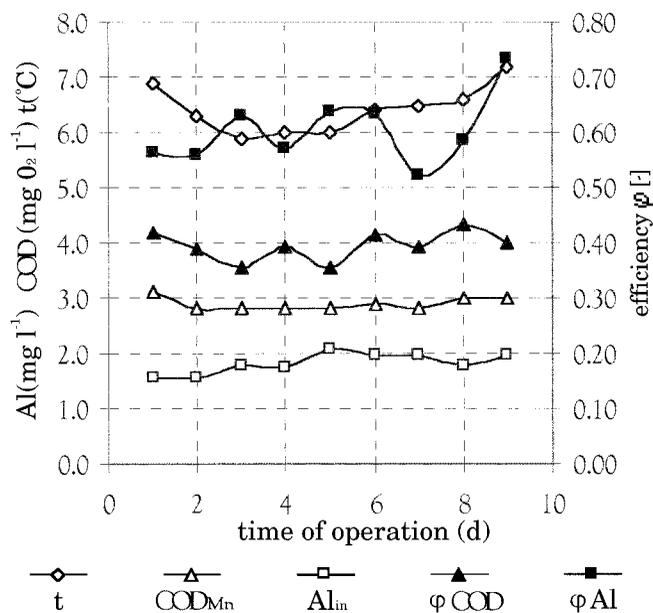
Figure 4 │ Evaluation of separation efficiencies of the pilot FLGM reactor.

with a substantially lower dosage, $D = 12.0$ mg l^{-1} (0.72 mg Al l^{-1}), the separation efficiencies of the FLGM are $\varphi_{Al} = 0.51$ and $\varphi_{COD} = 0.21$, and the total separation efficiencies of the plant after filtration are $\varphi_{Al} = 0.89$ and $\varphi_{COD} = 0.53$. These results are very similar to those obtained at higher dosage. The filtration run ended after 28 hours operation also due to Al breakthrough, but at a filter head loss $\Delta H = 11.1$ kPa. For comparison, the filtration run of the full size filters installed at the works, when the plant was operated at a dosage $D = 22.0$ mg l^{-1} and a significantly lower filtration velocity $v_f = 3.5$ m h^{-1}, ended after 21 hours of operation due to Al breakthrough.

The pilot plant tests carried out with this type of water confirmed that the FLGM is capable of separating the destabilised impurities with high efficiency. The pilot plant tests proved that the FLGM improves performance of filters because the filtration runs in the technological arrangement with FLGM were longer even when the RG

Table 2 │ The influence of dosage on operational performance and separation efficiencies of the FLGM reactor and RG filters

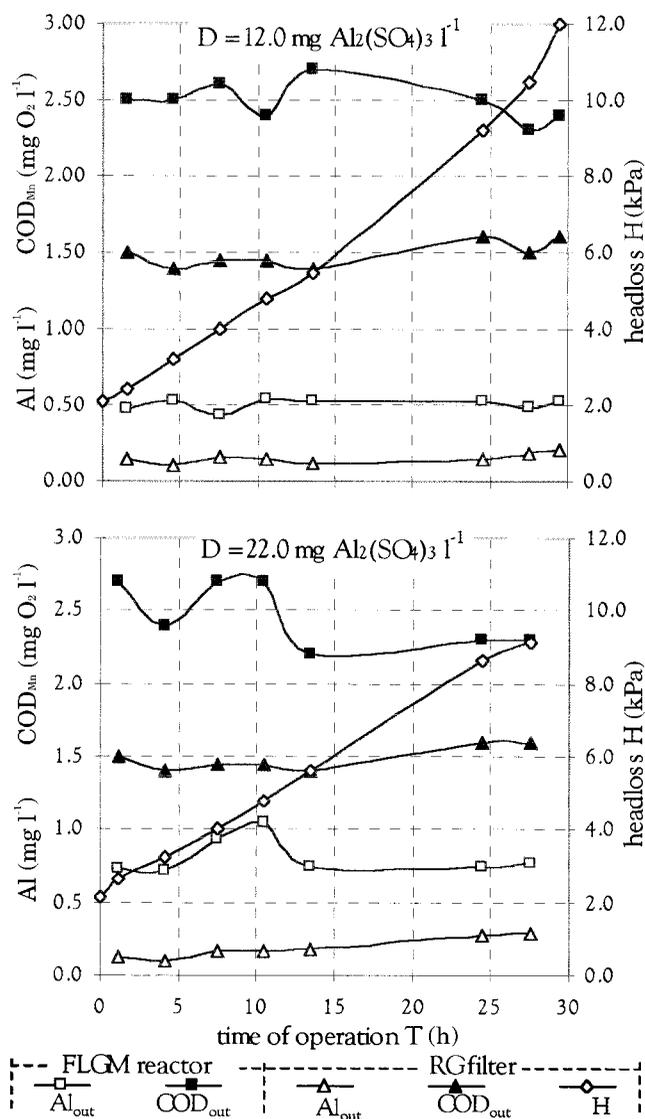| Dosage Al$_2$(SO$_4$)$_3$ 18 H$_2$O			mg l^{-1}	22.0	12.6
FLGM reactor	Inlet	pH	—	5.96	5.95
		Al	mg l^{-1}	1.71	1.02
		COD$_{Mn}$	mg O$_2$ l^{-1}	3.15	3.16
	Outlet	Al	mg l^{-1}	0.81	0.50
		COD$_{Mn}$	mg O$_2$ l^{-1}	2.47	2.49
		φ Al	—	0.52	0.51
		φ COD$_{Mn}$	—	0.21	0.21
RG filter	Outlet	Al	mg l^{-1}	0.19	0.15
		COD$_{Mn}$	mg O$_2$ l^{-1}	1.49	1.49
		φ Al	—	0.89	0.85
		φ COD$_{Mn}$	—	0.53	0.53
		H	kPa	16.0	28.0
		T	h	6.2	11.1

Figure 5 | Comparison of the course of filtration runs at different dosages.

filter was operated at a much higher filtration velocity. Furthermore, it was proved that with the use of the FLGM a reduced dosage of destabilisation reagent can be applied while maintaining a high efficiency of treatment.

Kutná Hora waterworks

The Kutná Hora waterworks, designed for a capacity $Q = 90.0 \, l \, s^{-1}$, treats water from the Vrchlice reservoir. This water is laden with organic matter of a type that is not

easily removable. The characteristic feature of this pollution is its widely fluctuating concentration over very short periods of time.

Treatment of this water is aimed at removing organic matter and, at times of heavy rains, also impurities of a mineral nature. The works treatment technology includes:

- pre-treatment of water by aeration;
- destabilisation of impurities with aluminium sulphate; and
- two-stage separation of formed suspension in rectangular sedimentation tanks followed by filtration in RG sand filters.

Both aluminium sulphate dosage at which maximum reduction in COD_{Mn} is achieved and the best attainable treatment process efficiency are reaction pH dependent. According to the waterworks experience the best results are usually obtained at a reaction pH value varying between 5.8 and 6.1.

The results of jar tests carried out during the period for average raw water quality, characterised by $COD_{Mn} = 4.5 \, mg \, O_2 \, l^{-1}$, are shown in Figure 6. The raw water was pre-treated with sulphuric acid in order to maintain the reaction pH in the above-mentioned range. The results proved that at a reaction pH of about 5.8, the maximum treatment process efficiency with respect to reduction of COD_{Mn} value is achieved with a dosage higher than $D = 60.0 \, mg \, l^{-1}$ and remains practically unchanged up to a dosage $D = 115 \, mg \, l^{-1}$. Because concentration of residual Al begins to increase from a dosage $D = 80.0 \, mg \, l^{-1}$, any dosage within the range $D = 60–80 \, mg \, l^{-1}$ is considered to be operationally applicable.

The course of the curves illustrating changes in residual Al concentration and COD_{Mn} value with reaction pH value shows that with respect to efficiencies of separation the optimum reaction pH is around 6.0; with decreasing pH value residual Al concentration increases and CODMn value decreases.

The waterworks design incorporates a two-stage separation of suspension arrangement. Therefore, the pilot plant arrangement incorporating the FLGM reactor, sludge blanket clarifier and RG sand filter was operated and continuously monitored over a period of more than four months.
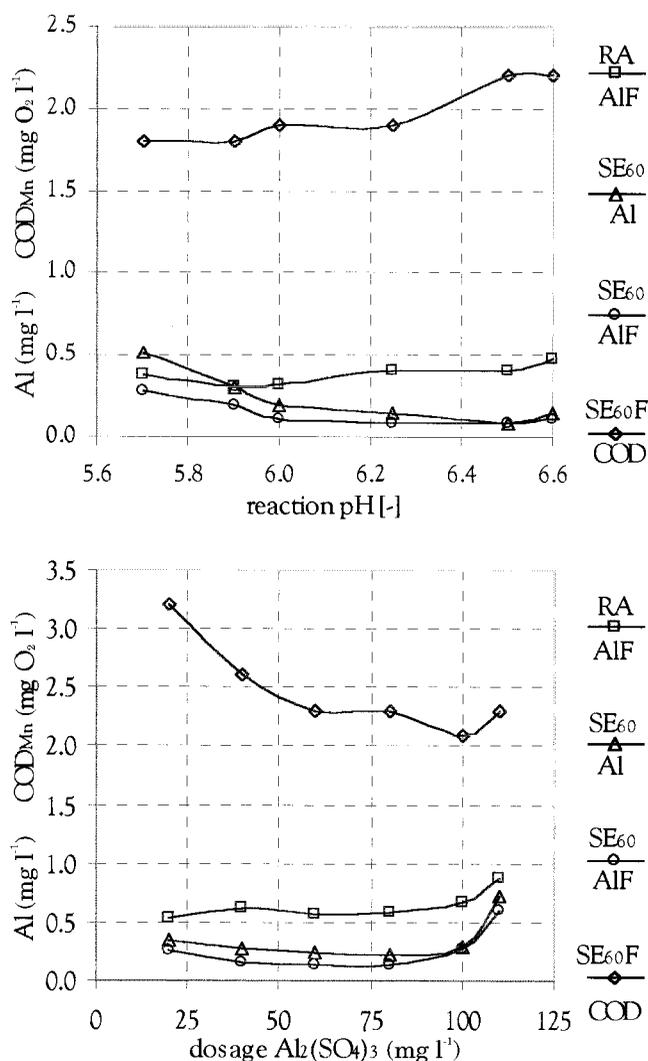
Figure 6 | Jar test optimisation of aluminium sulphate dosage and reaction pH: Waterworks Kutná Hora.

The first part of the pilot plant testing was aimed at evaluating both the aggregation efficiency and separation capability of the FLGM and the FLGM reactor was operated in A mode as well as in S mode. The results obtained are summarised in Table 3 and shown in Figure 7.

When the FLGM reactor was operated in A mode, the pilot plant arrangement was supplemented with a long pipe installed upstream of the reactor, where the destabilised particles aggregated to microparticles. At an aluminium sulphate dosage $D = 67.0$ mg l^{-1} (determined as the optimum dosage with respect to aggregation of particles taking place in the pilot plant) a high degree of aggregation ($\alpha = 0.77$) was achieved in water entering the FLGM and remained almost unchanged when passing through the reactor ($\alpha = 0.78$). The separation efficiency of the FLGM was zero.

When the FLGM reactor was operated in the S mode, the basic unchanged arrangement of the pilot plant was used. At a dosage $D = 62.0$ mg l^{-1} (determined as the optimum dosage with respect to separation efficiency attainable by the FLGM) the separation efficiencies achieved in the FLGM reactor were high, $\varphi_{Al} = 0.55$ and $\varphi_{COD} = 0.29$. The degree of aggregation of particles entering the reactor was nearly zero, while the degree of aggregation of particles leaving the reactor was high, $\alpha = 0.85$, higher than that reached when the reactor was operating in the A mode.

The separation efficiency determined at the outlet from the sludge blanket clarifier (taking into consideration also the separation efficiency achieved by the FLGM), when evaluated by residual Al concentration, was the same irrespective of the mode in which the FLGM was operated, $\varphi_{Al}A = \varphi_{Al}S = 0.72$. This means that with respect to Al removal the sludge blanket compensates for the missing separation function of the FLGM when it operates in A mode. When evaluated by COD_{Mn}, the separation efficiency of the clarifier was higher with the FLGM operating in S mode, than in A mode, $\varphi_{COD}S = 0.42$ and $\varphi_{COD}A = 0.33$. A higher separation efficiency of the sludge blanket with respect to COD_{Mn} was obtained with the FLGM operating in S mode. This was probably due to a greater binding power of the organic particles, just having being destabilised in the FLGM, to its grains. Therefore, the final quality of treated water produced with the FLGM reactor operating in S mode was a little better with respect to COD_{Mn}.

The benefit of the FLGM operating in S mode is also evident at filtration through RG filters. The filtration runs were approximately 30% longer in comparison to those obtained with the FLGM operating in A mode, and they ended due to utilisation of the pressure head available, ending filtration runs. In contrast, when the reactor was operating in A mode, the filtration runs ended due to

Table 3 | Comparison of operational performance and separation efficiencies of the plant with the FLGM reactor operating in A and S mode

Arrangement				A	S
FLGM reactor	Inlet	Al	mg l^{-1}	5.42	5.04
		AlF	mg l^{-1}	1.24	4.97
		COD_{Mn}	mg O$_2$ l^{-1}	5.51	4.94
		α Al	—	0.77	0.01
FLGM reactor	Outlet	Al	mg l^{-1}	5.42	2.27
		AlF	mg l^{-1}	1.21	0.33
		COD_{Mn}	mg O$_2$ l^{-1}	5.49	3.52
		φ Al	—	0.00	0.55
		φ COD_{Mn}	—	0.00	0.29
		α Al	—	0.78	0.85
Clarifier	Outlet	Al	mg l^{-1}	1.50	1.43
		COD_{Mn}	mg O$_2$ l^{-1}	3.69	2.83
		φ Al	—	0.72	0.72
		φ COD_{Mn}	—	0.33	0.42
RG filter	Outlet	Al	mg l^{-1}	0.09	0.08
		COD_{Mn}	mg O$_2$ l^{-1}	2.90	2.46
		φ Al	—	0.98	0.99
		φ COD_{Mn}	—	0.47	0.50
		T	h	30.0	39.5
		H	kPa	11.3	20.1

Al breakthrough to the filtrate long before the available pressure head could be utilised.

The second part of the pilot plant testing was aimed at comparing the course and the process efficiency of separation of suspension taking place in a single and two stage arrangement of the plant incorporating the FLGM reactor operating in S mode. The dosages of aluminium sulphate applied were as follows: $D = 50.0$ mg l^{-1} for the single-stage separation arrangement and $D = 70.0$ mg l^{-1} for the two-stage separation arrangement. The raw water was pre-treated with sulphuric acid in order to maintain reaction pH around 6.2. The results obtained are summarised in Table 4 and illustrated in Figure 8.

Despite the differences in dosages applied the separation efficiencies of the FLGM were nearly the same. The
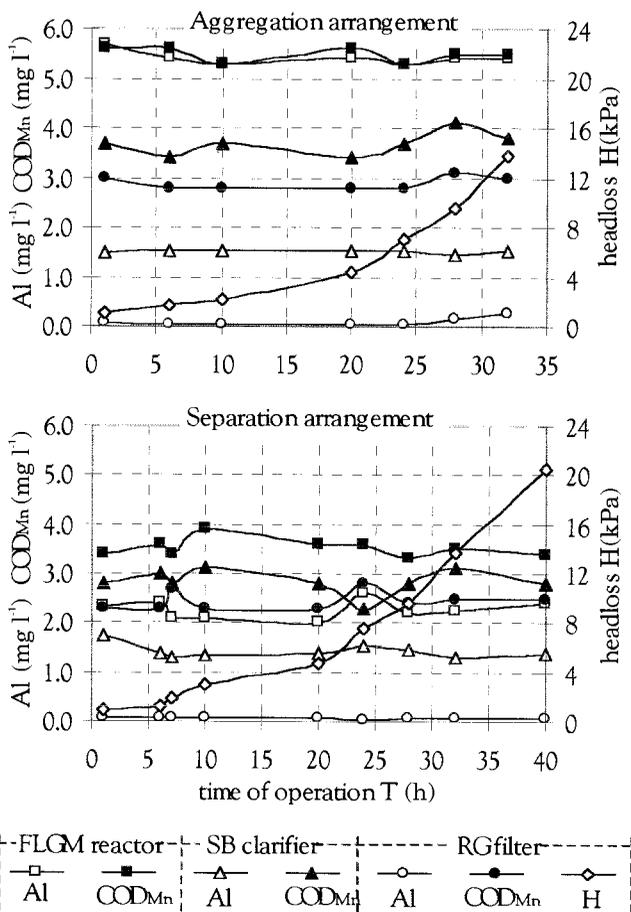
Figure 7 | Comparison of the performance of the pilot plant operating in the aggregation and separation arrangement.

degree of aggregation of suspension at the outlet from the reactor was high, implying that the suspension formed was in a form separable by sedimentation as well as filtration.

The filtrate quality was the same for both arrangements, although the dosage applied was considerably lower in the case of the single-stage separation arrangement. In both arrangements filtration runs ended due to Al breakthrough. The difference between the two was the duration of filtration runs; in the two-stage separation arrangement $T_F = 27$ hours and in the single-stage separation arrangement $T_F = 14.5$ hours. The course of head loss development shows that in the single-stage separation arrangement the head loss was increasing more rapidly due to higher content of suspension entering the filter.

The pilot plant operation proved the suitability of the FLGM for treatment of water from Vrchlice impounding reservoir, which is characterised by a high concentration of organic matter. Moreover, the results obtained show that the benefit of the FLGM operating in S mode is the improved performance efficiency of the final separation stage (filtration), when the single-stage separation arrangement is applied.

Vír waterworks

The Vír waterworks, designed for a capacity $Q = 140 \, \text{l s}^{-1}$, treats water from the Vír impounding reservoir. The raw water is treated by destabilisation of impurities with aluminium sulphate. Dispersion and homogenisation of aluminium sulphate with the raw water takes place in a stator type flash mixing chamber. Agitation facilitating aggregation of the destabilised particles of impurities takes place in two countercurrent-type mixing chambers. The formed suspension is separated in two stages. The first separation stage consists of two upflow-type sludge blanket clarifiers and the second separation stage comprises ten RG sand filters.

The raw water quality is relatively steady. The main objective of the treatment is removal of organic matter, which is difficult to remove. The works is operated as a two-stage separation plant only during the periods of poorer raw water quality, usually during spring and autumn, and for the rest of the year as a single-stage separation plant by way of direct filtration.

The purpose of the technological survey, within the framework of which the tests were carried out, was optimisation and intensification of existing water treatment processes resulting in the most effective upgrading and uprating of the waterworks.

The jar test results in Figure 9 show that the operational applicable dosage is between $D = 17.5$ and 20.0 mg l^{-1}. The degree of aggregation attained with this dosage is relatively high immediately after rapid agitation and does not change significantly during sedimentation. Effectiveness of the treatment process is satisfactory with respect to COD_{Mn} value; with respect to residual Al it is relatively low, because the residual Al content exceeds the

Table 4 | Comparison of operational performance and separation efficiencies of the plant with the FLGM reactor operating in single (I) and two (II) stage separation arrangement

Arrangement				I	II
FLGM reactor	Inlet	Al	mg l^{-1}	3.97	5.64
		COD$_{Mn}$	mg O$_2$ l^{-1}	4.6	4.6
FLGM reactor	Outlet	Al	mg l^{-1}	1.53	2.40
		AlF	mg l^{-1}	0.73	0.69
		COD$_{Mn}$	mg O$_2$ l^{-1}	3.5	3.5
		φ Al	—	0.61	0.57
		φ COD$_{Mn}$	—	0.24	0.21
		α Al	—	0.82	0.88
Clarifier	Outlet	Al	mg l^{-1}	—	1.57
		COD$_{Mn}$	mg O$_2$ l^{-1}	—	3.1
		φ Al	—	—	0.72
		φ COD$_{Mn}$	—	—	0.31
RG filter	Outlet	Al	mg l^{-1}	0.13	0.08
		COD$_{Mn}$	mg O$_2$ l^{-1}	2.6	2.5
		φ Al	—	0.95	0.97
		φ COD$_{Mn}$	—	0.43	0.43
		T	h	14.5	27.0
		H	kPa	15.0	12.6

permissible value $c = 0.20$ mg Al l^{-1}. As is evident from Figure 10, an excessive portion of non-aggregated particles remains in the system even after 60 minutes' sedimentation. Similar results were also obtained after the first separation stage in the full-size plant, where the satisfactory quality of the treated water was attained due to the aggregational ability of the filter bed. This implies that the particles of impurities were not sufficiently destabilised and the remaining barrier had to be overcome by the high velocity gradient in the filter bed.

Verification of the suitability of the FLGM as an effective separation element was carried out in two phases. In the first phase the separation efficiency was tested under the conditions of pilot plant operation and in the second phase under the conditions of full-size operation in a partly modified plant.

The performance results obtained with the pilot plant operation are shown in Figure 11. Evidently, the high separation efficiencies exceeding $\varphi_{Al} = 0.57$, $\varphi_{COD} = 0.27$ were achieved at dosages larger than $D = 17.5$ mg l^{-1}. The
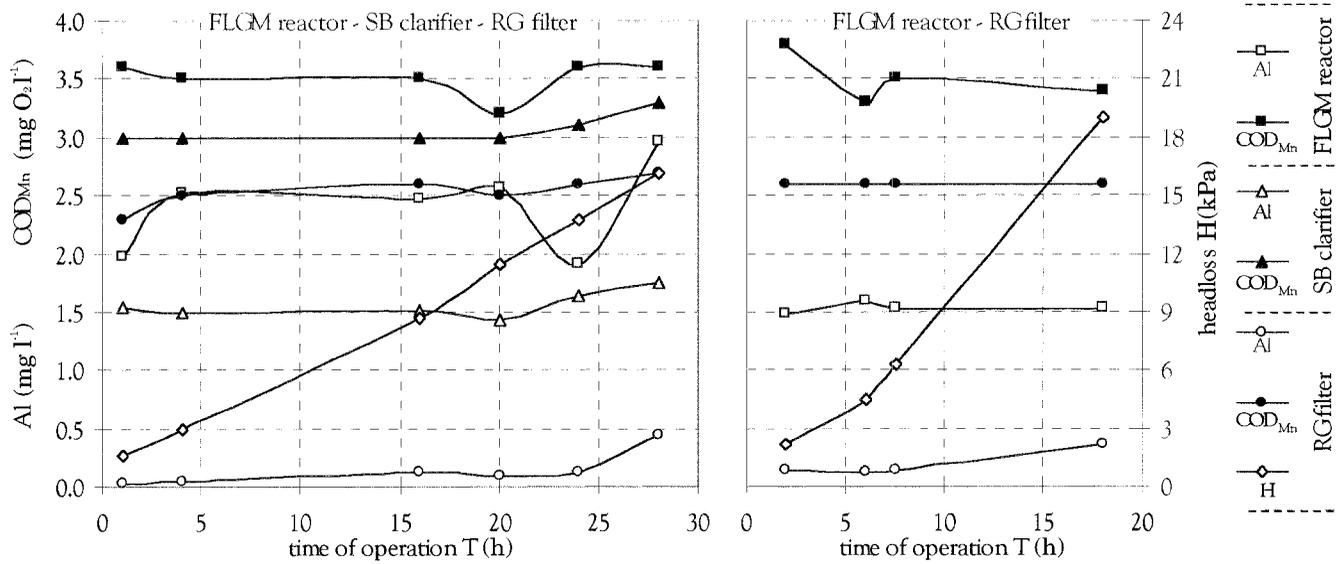
Figure 8 | Comparison of operational performance of the pilot plant arrangement with and without first separation stage.

degree of aggregation of the formed suspension leaving the FLGM increased rapidly up to a dosage $D = 20.0$ mg l^{-1} and thereafter remained almost unchanged. It is seen that, as a result of existing mutual interdependence between separation ability of destabilised impurities and their degree of aggregation, the course of the dependence of

degree of aggregation and of the efficiency of separation on the dosage are similar.

The course of a typical filtration run achieved with the FLGM operating in S mode is illustrated in Figure 12. The quality of water at the outlet from the RG filter was very steady and considerably better than could be expected from the jar test results and the works operational performance. The filter head loss reached after 24 hours
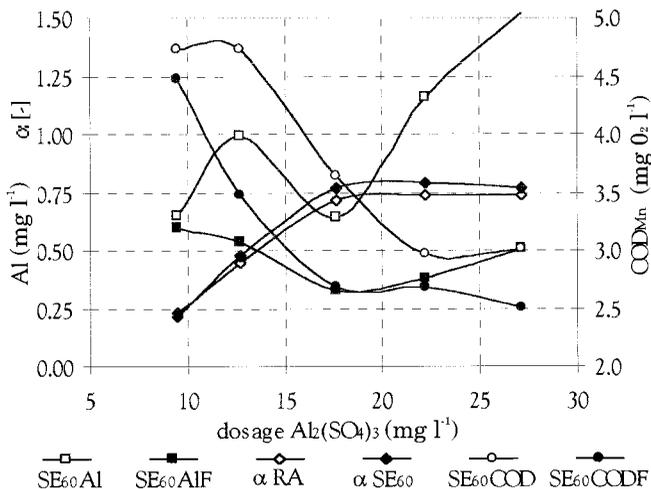
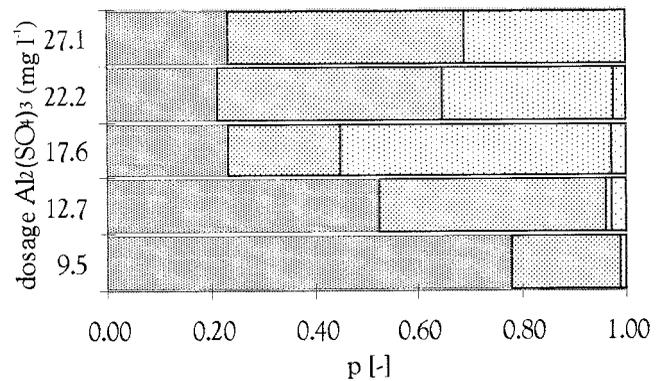**Figure 9** | Jar test optimisation of aluminium sulphate dosage: Waterworks Vir.
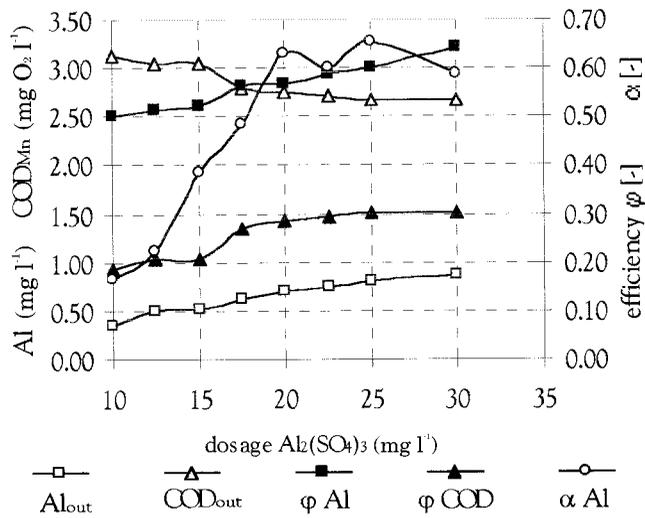

Figure 10 | Particle size distribution: jar test.

Figure 11 | Pilot plant optimisation of separation efficiencies.

Figure 13 | General arrangement of the converted FLGM reactor.

run at an average dosage of aluminium sulphate $D = 17.0$ mg l^{-1} and solids loading $M = 7.8$ mg Al m^{-2} h^{-1} was $\Delta H = 20.0$ kPa.

Determination of the separation efficiency of the FLGM under a full-size operation was carried out with one of the sludge blanket clarifiers converted to a FLGM reactor. The design data for this conversion were obtained from the results of the pilot plant operation. The arrangement of the converted FLGM reactor is shown in Figure 13.
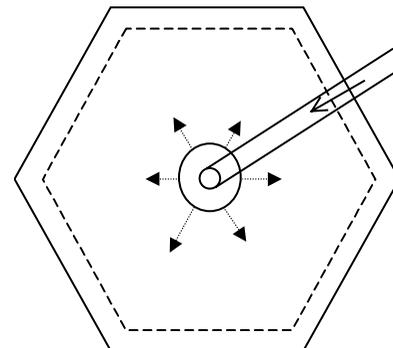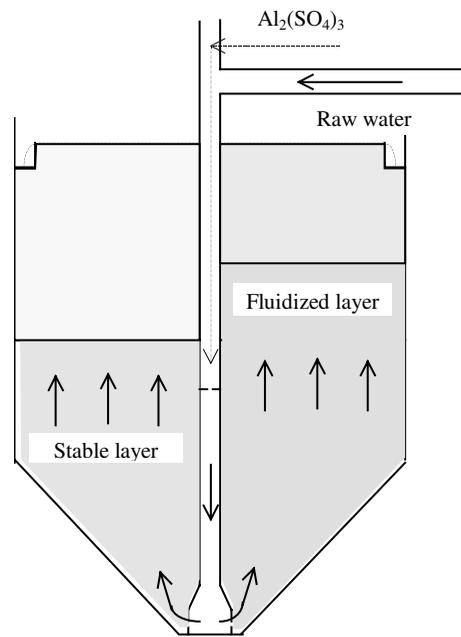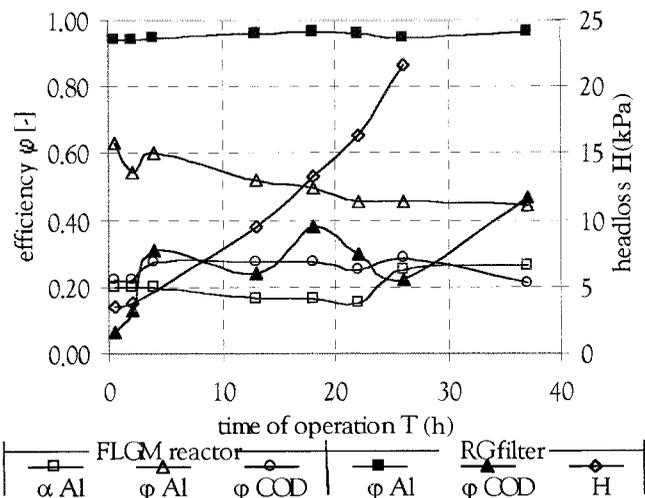
The rate of flow through the FLGM reactor varied between $Q_R = 90$ and 150 l s^{-1}. The reactor was charged with volume of sand $V_S = 3.1$ m^3, giving retention time in the FLGM of $T_{FLGM} = 100$–150 s depending on flowrate. The granularity of the sand used was $d = 0.315$–0.80 mm and its sieve analysis is shown in Figure 17. The upflow velocities in the reactor, above the FLGM, varied between $v = 10.0$ and 24.2 mm s^{-1} and the mean velocity gradient between $G = 280$ and 440 s^{-1}, depending on water temperature and clogging of the FLGM. Hydraulic

Figure 12 | Pilot plant operational performance during a filtration run.
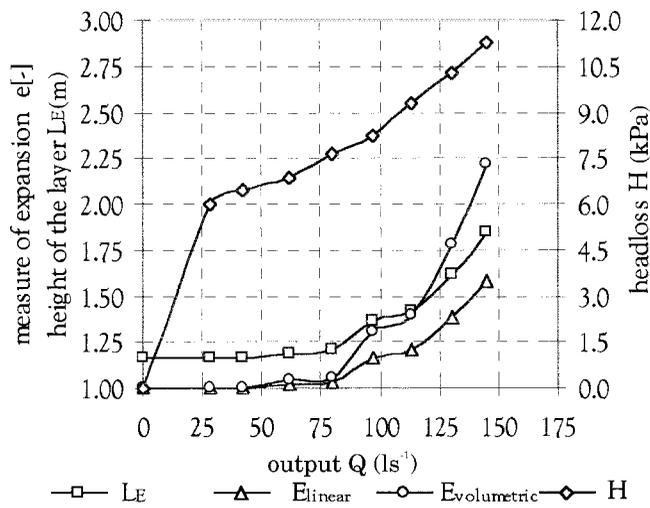
Figure 14 │ Hydraulic characteristics of the fluidised layer in the adapted reactor.

resistance of the clean FLGM was expected to be around $\Delta H = 20.0$ kPa at a flowrate $Q = 100 \text{ l s}^{-1}$. Both linear and volumetric characteristics of expansion of the FLGM are shown in Figure 14.

In the first phase of testing the FLGM reactor was kept in continuous operation by the waterworks staff and operated under the same conditions as the waterworks.

In this phase separation efficiency of the FLGM was evaluated. The results obtained are illustrated in Figure 15. The daily flowrates varied between $Q_R = 90$ and 140 l s^{-1} (an average flow $Q = 115 \text{ l s}^{-1}$) and aluminium sulphate dosage varied over a wide range between $D = 5.0$ and 34.0 mg l^{-1} (optimum dosage $D = 17.5 \text{ mg l}^{-1}$). These considerable fluctuations resulted from inaccuracies in modifications carried out on the existing dosing system. As a result, the FLGM separation efficiencies varied between $\varphi_{Al} = 0.20$ and 0.78 and $\varphi_{COD} = 0.02$ and 0.22.

In the second phase, testing aimed at evaluating the FLGM separation efficiencies was carried out. The operational performance of the reactor and the optimum dosage were continuously checked. The plant was operated at a dosage $D = 22 \text{ mg l}^{-1}$. The results obtained from one of the monitored filtration runs, when the dosage was stable, are illustrated in Figure 16. The criterion for ending the filtration run was the value of $COD_{Mn} = 3.0$ mg $O_2 \text{ l}^{-1}$ in the filtrate. It can be seen that this value was exceeded after 16 hours of filtration while residual Al was low, $c_{Al} = 0.06 \text{ mg l}^{-1}$, and head loss was also low, $\Delta H = 9.1$ kPa. The average FLGM separation efficiencies achieved by the FLGM were $\varphi_{Al} = 0.58$ and $\varphi_{COD} = 0.11$. The total separation efficiencies of the plant were $\varphi_{Al} = 0.97$ and $\varphi_{COD} = 0.32$.
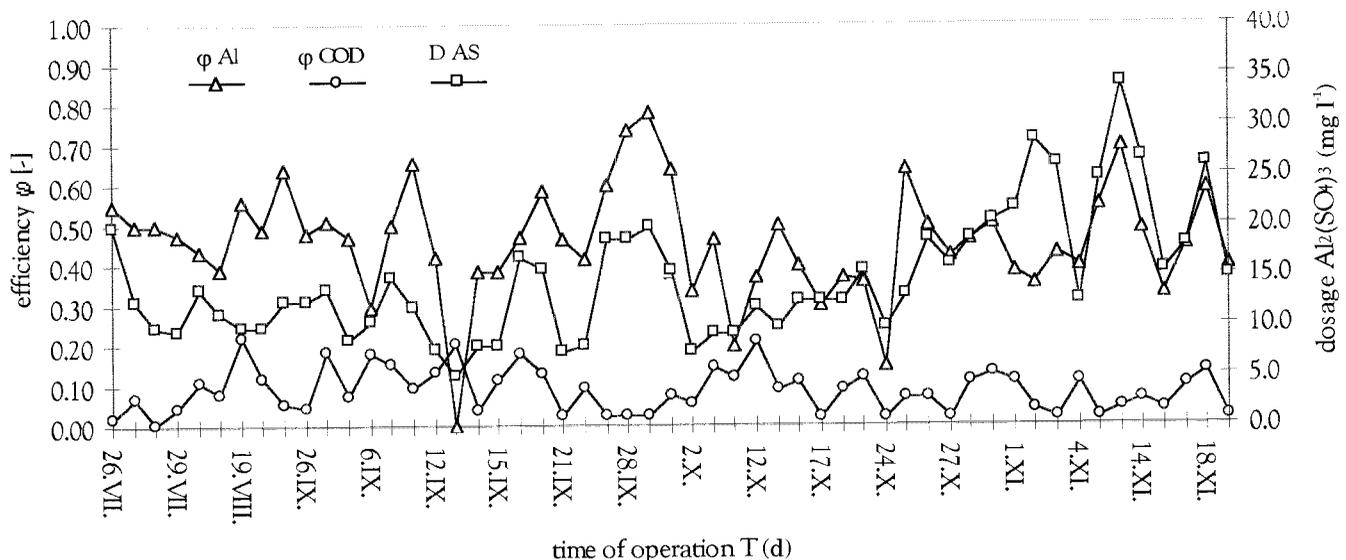

Figure 15 │ Separation efficiency of the converted full-scale FLGM reactor.
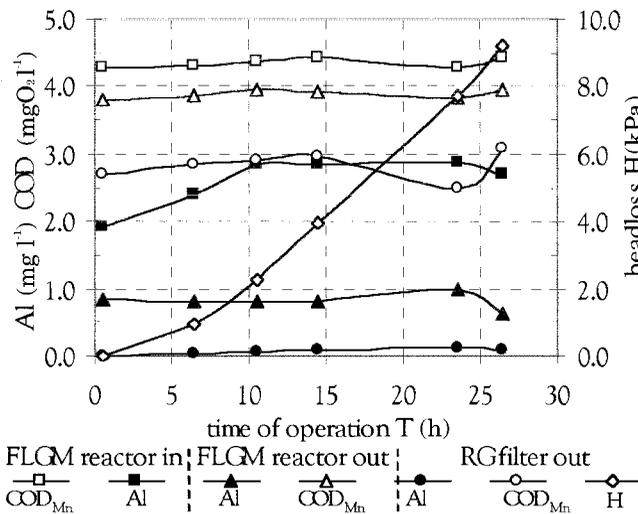
Figure 16 | Full-size plant operational performance during a filtration run.

raw water of 40 organisms ml^{-1}, the operation of the FLGM reactor was stopped. The rector was put back into service at the same flowrate after 15 hours interruption and 27 organisms ml^{-1} of live cells were counted at the outlet from the reactor; composition of the species was the same as in the raw water and the balance of the cells were mechanically destroyed. After the next 15 hours from re-starting, the biological picture remained almost unchanged. Similarly, the BOD_5 value also remained unchanged.

During a long-term operation of the full-size FLGM reactor, the changes in properties of the granular medium were evaluated. Samples of the medium were taken at the time of charging the reactor and then after 57, 121 and 147 days of operation. The changes in granularity and density of the medium were analysed and the effect of impurities attachment to the media grains was determined by a material balance. The results obtained are summarised in Table 5 and illustrated in Figure 17.

Changes in the size of grains measured after 57 days of operation show enlargement by $\Delta d = 0.13$ mm, corresponding to an average monthly enlargement of $\Delta d = 0.068$ mm. The weight of grains increased by $\Delta m = 21.1$ g l^{-1} d^{-1} and their volume increased by $\Delta V = 9.45$ cm^3 kg^{-1} d^{-1}, while the density of the coated

Possible biological activity of the FLGM that might be experienced during interruption of its operation was also investigated. At a microorganisms count (Leptothrix, colourless flagellates and diatoms, mainly: *Asterionella formosa*, *Fragilaria crotonensis*, *Nitzschia* sp. div.) in the

Table 5 | Changes in properties of the granular medium in the FLGM reactor of the full scale plant caused by separation of impurities

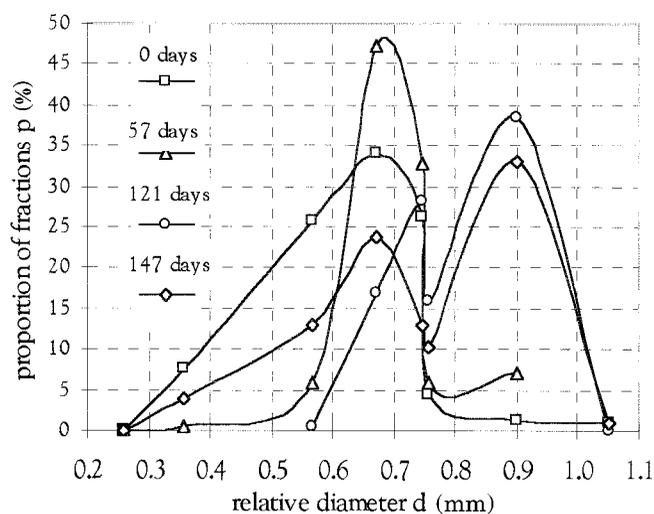| T (days)	ρ (g cm^{-3}) Coated	ρ (g cm^{-3}) Non-coated	ρ (g cm^{-3}) Coat	Δm (g d^{-1} l^{-1})	ΔV (cm^3 d^{-1} kg^{-1})
57	2.33	2.64	1.87	21.4	9.4
	2.33	2.64	1.85	20.7	9.5
Φ	2.332	2.639	1.856	21.06	9.45
121	2.21	2.63	1.84	16.9	4.5
	2.19	2.63	1.82	17.3	4.6
Φ	2.203	2.627	1.829	17.14	4.52
147	2.34	2.66	1.81	7.6	3.7
	2.20	2.68	1.85	17.2	3.7
Φ	2.270	2.671	1.831	12.39	3.72

Figure 17 | Changes of the size-distribution of grains of the FLGM due to separation of impurities in the full size reactor.

grain dropped by $\Delta\varrho = 0.00549$ g cm^{-3} d^{-1} ($\varrho = 2.3319$ g cm^{-3}), all in comparison with the 'virgin' sand.

Further measurement of the changes of the properties of the FLGM medium was carried out after 121 days' operation. The measurements of a shell thickness showed enlargement of the grain diameter by a total $\Delta d = 0.23$ mm. Other daily changes of the medium properties were smaller than those determined after the first 57 days of operation: the weight of grains increased only by $\Delta m = 17.1$ g l^{-1} d^{-1} and their volume only by $\Delta V = 4.52$ cm^3 kg^{-1} d^{-1}, while the grain density dropped by $\Delta\varrho = 0.00350$ g cm^{-3} d^{-1} ($\varrho = 2.2031$ g cm^{-3}), all in comparison with the 'virgin' sand. The deformation of curves characterising the size distribution of the grains is a result of uneven deposition of impurities on to the surface of grains, due to fluctuating flow resulting in frequently changing expansion and even sedimentation of a portion of the FLGM.

The measurement of coating thickness after 147 days of operation showed significant increase in non-uniformity of the medium granularity caused by uneven enlargement of grain diameters, which were found to vary between $\Delta d = 0.23$ and 0.36 mm. The shell thickness differs considerably from that established by the previous measurements. The weight of grains increased only

by $\Delta m = 11.1$ g l^{-1} d^{-1} and their volume only by $\Delta V = 4.24$ cm^3 kg^{-1} d^{-1}, while the grain density dropped by $\Delta r = 0.00273$ g cm^{-3} d^{-1} ($\rho = 2.2695$ g cm^{-3}).

It follows from the foregoing that the hydrodynamic conditions in the FLGM influence significantly its separation efficiency. During the first 95 days all grains were fully fluidised and therefore able to bind the destabilised impurities. Thereafter, due to disturbances of the hydraulic regime mentioned above the separation efficiency decreases, because larger grains sinking and depositing at the bottom of the reactor function only as the agitation elements. Reasonable consistency in the values of the coating density shows that the composition of the impurities did not change during the period of testing. Visual observation showed rounding of the grains.

Suitability of other materials that can be used to charge the FLGM reactor was tested. The reason for this was to replace silica sand with another suitable material in which the grains have only the composition of the shell, without containing any sand; this can make the recovery of Al$_2$O$_3$ from the grain shell cheaper. For this reason, the bentonite and the sludge produced by the waterworks were used as the materials for charging the FLGM reactor. Both bentonite and waterworks sludge were first thermally treated, then crushed, sieved to a granularity-corresponding to that of silica sand and then used to charge the reactor.

As an example, the results comparing the separation efficiencies attained with bentonite and silica sand are summarised in Table 6. In both series of tests the raw water quality was characterised by a relatively high content of organic matter, $COD_{Mn} = 6.24$ mg O$_2$ l^{-1}. In bothcases the water was treated with the same aluminium sulphate dosage $D = 20.0$ mg l^{-1}. The separation efficiencies attained by both materials were the same.

The application of different materials that can be used for charging the FLGM reactor, methods of their treatment, together with their utilisation for easy recovery of the metal oxide will be dealt with later.

Želivka waterworks

The raw water treated by the Želivka waterworks is characterised by relatively low content of organic matter,

Table 6 | Separation efficiencies achieved with silica sand and thermally treated bauxite used as the granular medium of the FLGM reactor

		Inlet	Outlet		φ	
T (h)	t (°C)	COD_{Mn} (mg l^{-1})	Al out (mg l^{-1})	COD_{Mn} (mg O$_2$ l^{-1})	Al	COD_{Mn}
Silica sand						
24	4.9	6,24	0.61	4.92	0.62	0.21
48	4.9	6,24	0.65	4.76	0.60	0.24
192	4.9	6,24	0.55	4.88	0.66	0.22
240	4.9	6,24	0.63	4.64	0.61	0.26
Φ	4.9	6,24	0.61	4.80	0.62	0.23
Termally treated bauxite						
24	4.9	6.24	0.63	4.92	0.61	0.21
48	4.9	6.24	0.65	4.92	0.60	0.21
192	4.9	6.24	0.51	5.03	0.69	0.19
240	4.9	6.24	0.63	4.88	0.61	0.22
Φ	4.9	6.24	0.61	4.94	0.63	0.21

which is difficult to remove. The treatment process arrangement of this waterworks is as described in Mutl and Knesl (1999). The waterworks incorporates two FLGM reactors operating in A mode. These reactors were used for verification of the pilot plant results obtained from the research into the FLGM and evaluation of the influence of design modifications of the FLGM reactors on their performance efficiencies.

The jar test results illustrating the treatability of water attainable by different aluminium sulphate dosages are shown in Figure 18. It is evident that the destabilisation efficiency increases with increasing dosage throughout the range of dosages tested. Determination of optimum dosage is not an easy task, as is evident from the curves illustrating residual Al concentration and COD_{Mn} value, both determined in samples taken after sedimentation. The changes in AlF values determined in the same samples after their centrifugation indicate that the impurities are sufficiently destabilised at an aluminium sulphate dosage $D = 15.0$ mg l^{-1}. At the works, the dosage applied is

determined in a similar way when microorganisms are present in raw water in quantities smaller than 100 individuals ml^{-1}; at quantities of microorganisms (as
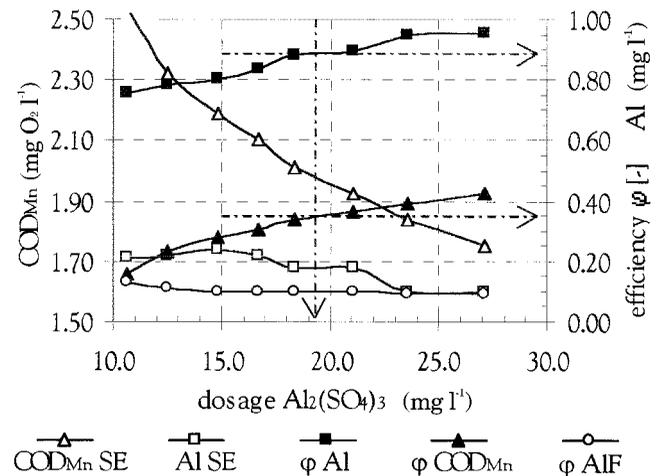
Figure 18 | Jar test optimisation of aluminium sulphate dosage: Waterworks Želivka.
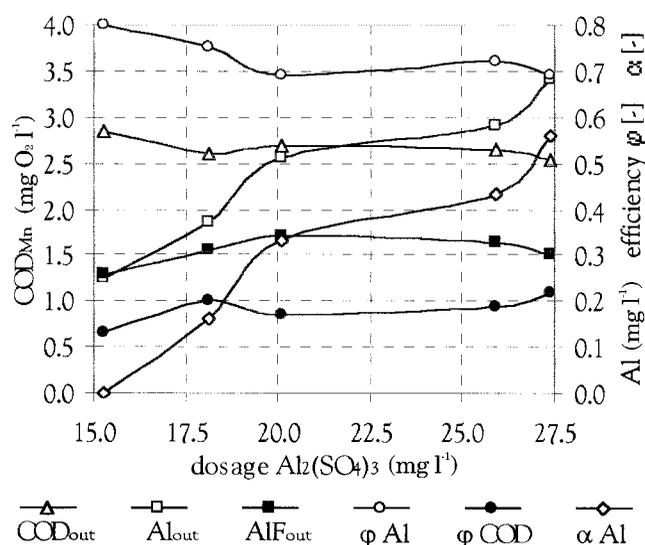
Figure 19 | Optimisation of dosage with respect to separation efficiencies by pilot FLGM reactor.

specified before) exceeding 100 individuals ml^{-1}, higher dosages are applied in order to ensure formation of larger aggregates, which are capable of intercepting micro-organisms by the mechanism of particle enmeshment (Mutl & Knesl 1999).

The effect of aluminium sulphate dosage with respect to the separation efficiency of the FLGM, achieved under the conditions of pilot plant operation, is evident from the results in Figure 19. As can be seen, high separation efficiencies evaluated by residual Al and COD_{Mn} value are achieved over the range of dosage. Residual Al concentration slightly increases with increasing dosage; comparison of curves illustrating changes in Al content in centrifuged and non-centrifuged samples shows that the increasing portion is in the form of aggregates. From the course of the curves illustrating changes in efficiencies of separation and aggregation it follows that a decrease in separation efficiency is accompanied by an increase of the degree of aggregation determined at the outlet of the FLGM. This finding is connected with the character and properties of the aggregates formed since the aggregates of lower development order have greater adhesion capability than the developed ones.

In view of the separation efficiencies achieved by the FLGM a dosage $D = 17.0$–18.0 mg l^{-1} was considered to be sufficiently effective. The separation efficiencies achieved are about $\varphi_{Al} = 0.70$ and $\varphi_{COD} = 0.19$ and are a little lower than those obtained by jar test $\varphi_{Al} = 0.87$ and $\varphi_{COD} = 0.33$. This dosage was applied during a long-term evaluation of separation efficiency of the FLGM. The results obtained are shown in Figure 20.

As can be seen, the final quality of water treated by the FLGM did not change during the entire period of operation. Comparison of this water quality with that produced by the works shows a slightly higher overall treatment efficiency achieved with the FLGM. The average separation efficiencies of the FLGM are $\varphi_{Al} = 0.67$ and $\varphi_{COD} = 0.25$.

When intercepted in the FLGM, the destabilised impurities are deposited on to the grain surfaces in the form of a coat. As a result, the expansion ratio and the hydraulic resistance of the expanded FLGM depth increases. When the depth of FLGM increased from its initial $L_E = 1750$ mm to $L'_E = 2450$ mm, suitability of the method proposed for regeneration of the FLGM was verified. The largest coated grains circulating at the bottom part of the reactor were siphoned-out in such a quantity that the resultant depth L'_E of the expanded FLGM was reduced by approximately 100 mm. At the same time the effect of the reduced FLGM depth on separation efficiency of the FLGM was evaluated. It was found that while the separation efficiencies did not change, the draw-off of the largest grains resulted in a drop of hydraulic resistance generated by the FLGM, corresponding to reduction of the FLGM mass. Coincidence of both parameters, L'_E and H, proved the validity of an assumption that hydraulic resistance can be used as a datum for regeneration of the FLGM.

The pilot plant testing was complemented with testing carried out in a full-size reactor designed to operate in A mode. This reactor was charged with the same sand that was used in the pilot plant to a depth $L_0 = 1000$ mm. In order to enable evaluation of efficiencies of separation, the only modification made to this reactor enabling it to operate in S mode was relocation of the aluminium sulphate dosing point close to the inlet of the FLGM, thus minimising aggregation of destabilised particles
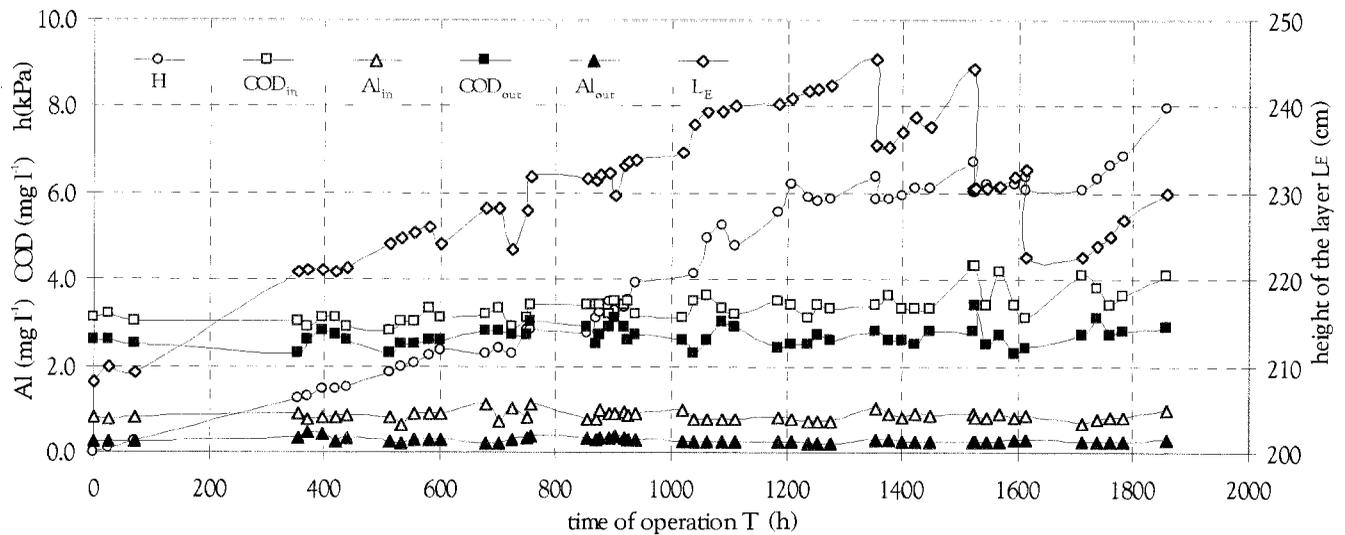
Figure 20 | Evaluation of direct separation capability of the FLGM under pilot plant operation.

upstream to the FLGM, i.e. ensuring that the degree of aggregation of the suspension entering the FLGM is as low as possible.

In the first series of tests evaluation of the FLGM reactor performance was aimed at optimising aluminium sulphate dosage with respect to maximisation of separation efficiency of the FLGM. The results of the operation carried out with approximately the same dosages of aluminium sulphate and the raw water quality as that treated by the pilot plant are summarised in Table 7 and shown in Figure 21. The efficiencies of separation of the FLGM slowly increased with increasing dosage. The best separation efficiencies achieved by the full size operation, namely $\varphi_{Al} = 0.25$ and $\varphi_{COD} = 0.08$, are considerably lower than the best separation efficiencies achieved during the pilot plant operation $\varphi_{Al} = 0.70$ and $\varphi_{COD} = 0.19$. The α values, characterising the degree of aggregation of destabilised particles at the inlet to and the outlet from the FLGM were high even at the low dosages and increased rapidly up to a dosage $D = 22.0$ mg l^{-1} and after exceeding this dosage increased very slowly. Due to a high degree of aggregation achieved by the system upstream to the FLGM, only a small portion of destabilised particles was still present in the non-aggregated form, i.e. in the form capable of entrapment in the FLGM.

Considerable differences, which are dosage dependent, were found with respect to the performance of filters. The duration of filtration runs decreased rapidly with dosage due to the increased quantity of suspension entering the filters. Significant shortening of filtration runs was experienced at dosages larger than $D = 25.0$ mg l^{-1}; the final head loss is approximately the same up to this dosage and then decreases with increasing dosage.

Based on the results obtained from the pilot plant operation, a much higher performance efficiency of the full-size FLGM reactor was expected. This proved not to be the case. As can be seen from comparison of the separation efficiencies achieved between the pilot and full-size FLGM operation demonstrated in Figure 22, the separation efficiencies achieved with the pilot plant are considerably higher. The reason for this is the long reaction time between the point of addition of aluminium sulphate to the water and point of water entry to the FLGM (long retention time in the reactor underfloor), which enables aggregation of destabilised particles to be enhanced considerably upstream of the FLGM.

While the pilot plant results show that separation efficiencies achieved by the FLGM are high and that its incorporation in the treatment process arrangement

Table 7 | The effect of dosage optimisation on operational performance of the full scale plant

Dosage $Al_2(SO_4)_3$			mg l^{-1}	12.0	16.2	23.2	25.8	32.0	34.2
FLGM reactor	Inlet	Al	mg l^{-1}	0.98	1.31	1.88	2.09	2.59	2.77
		AlF	mg l^{-1}	0.81	0.73	0.62	0.64	0.61	0.53
		COD$_{Mn}$	mg O$_2$ l^{-1}	3.05	2.96	3.02	3.02	3.04	3.18
		$\alpha\,Al$	—	0.17	0.43	0.67	0.69	0.76	0.81
FLGM reactor	Outlet	Al	mg l^{-1}	0.91	1.00	1.40	1.51	1.87	1.70
		AlF	mg l^{-1}	0.44	0.35	0.30	0.32	0.33	0.28
		$\alpha\,Al$	—	0.51	0.65	0.79	0.79	0.82	0.84
		COD	mg O$_2$ l^{-1}	2.85	2.78	2.82	2.74	2.86	2.60
		$\varphi\,Al$	—	0.07	0.23	0.26	0.27	0.27	0.39
		$\varphi\,$COD$_{Mn}$	—	0.06	0.06	0.07	0.09	0.06	0.18
RG filter 2	Outlet	Al	mg l^{-1}	0.07	0.05	0.05	0.10	0.18	0.18
		COD	mg O$_2$ l^{-1}	2.24	2.01	2.15	2.04	2.00	1.94
		$\varphi\,Al$	—	0.92	0.95	0.96	0.93	0.91	0.90
		$\varphi\,$COD$_{Mn}$	—	0.27	0.27	0.24	0.25	0.30	0.25
		H	kPa	24.0	24.0	23.7	23.2	13.7	12.9
		T	h	38.6	36.6	29.2	28.4	15.6	17.2
RG filter 12	Outlet	Al	mg l^{-1}	0.07	0.05	0.07	0.08	0.22	0.10
		COD	mg O$_2$ l^{-1}	2.34	2.02	2.13	2.07	1.96	1.78
		$\varphi\,Al$	—	0.93	0.95	0.95	0.95	0.87	0.96
		$\varphi\,$COD$_{Mn}$	—	0.22	0.27	0.24	0.24	0.32	0.31
		H	kPa	24.0	24.0	24.0	24.0	11.3	10.5
		T	h	40.6	37.6	31.0	30.2	14.0	15.6

significantly improves performance efficiency of the works, the full-scale plant results proved that the modifications carried out to convert the operation of the FLGM reactor from A mode to S mode to fully utilise its separation capability were not adequate.
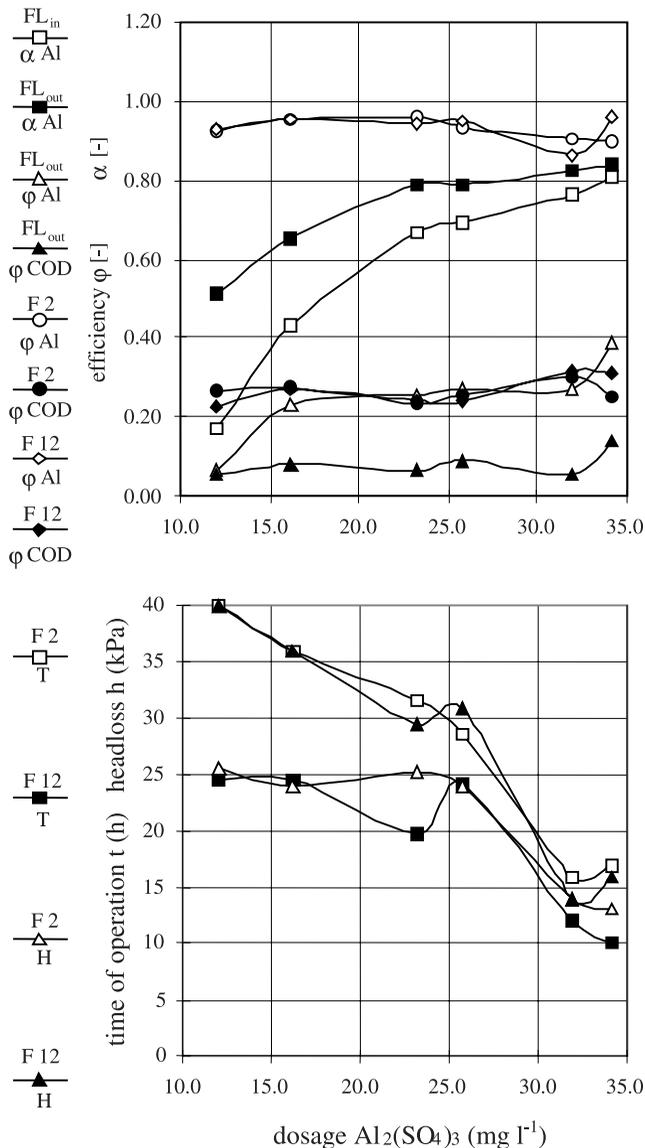
Figure 21 | Optimisation of dosage with respect to separation efficiencies by full size FLGM reactor.

Figure 22 | Comparison of the aggregation and separation efficiency in the pilot and full scale plant.

CONCLUSIONS

The results obtained from operation of the FLGM reactors at a number of different localities can be summarised in terms of the chemical technology point of view as follows:

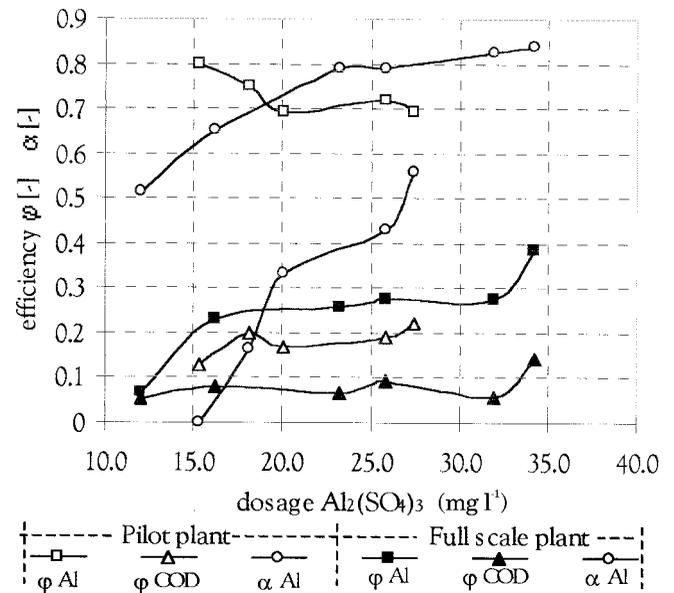1. The FLGM is capable of separating directly the destabilised impurities in the form of a waterfree coating developed on the surface of its grains. The separation efficiency is influenced by many factors, primarily by the degree of destabilisation of the particles of impurities and by the degree of their aggregation at the entry to the FLGM. Biological materials are not separable; due to motion of the grains they are only mechanically destroyed. The chemical composition of the shell is approximately the same as that of the waterworks sludge produced by conventional treatment processes.

2. The separation efficiency of the FLGM was more than 50% at all tested localities when evaluated by residual concentration of the metal component of the destabilisation reagent used, and around 30% when evaluated by COD_{Mn}. When only the separable portion of the particles is taken as a basis for evaluation, then the separation efficiency with respect to COD_{Mn} was around 50%. The destabilised impurities that are not intercepted in the FLGM are aggregated to readily separable aggregates.

3. The entrapment of impurities in the FLGM is particularly beneficial for treatment of water in a

single-stage arrangement. The influence of the aggregation capability of the FLGM on the character of the suspension remaining in the water after direct separation in the FLGM is of significance with respect to its separation in a two-stage arrangement.

4. The coated grains can be regenerated and after removal of the shell they can be returned to the reactor. A closed cycle for the replacement of the reactor charge can be constructed.

5. The use of a FLGM results in the reduction of the volume of sludge produced of more than 40%.

6. Thermally treated bauxite or waterworks sludge can be used as the reactor charge with the same efficiency as silica sand. The advantage of these two materials is that they can be used, as they are, as raw materials for producing aluminium oxide.

7. The separation efficiencies of a FLGM are influenced, among other factors, by reaction conditions influencing the efficiency of the process of destabilisation of impurities taking place in the FLGM, such as dosage of destabilisation reagent, water temperature, reaction time and measure of the FLGM expansion.

8. Modifications to the FLGM reactors used for verification of pilot plant results in a full-size plant were found to be inadequate with respect to the destabilisation process and therefore prevented total validation of the pilot plant performance results. The results made it possible to identify the important design parameters influencing the separation process in the FLGM and its efficiency in such a way that these could be optimised by subsequent research.

ACKNOWLEDGEMENTS

NOTATION

RA rapid agitation phase

SE_{60} 60 minutes sedimentation

Al_{out} residual aluminium measured at the outlet from evaluated treatment step

COD_{Mn} residual organic matter measured at the outlet from evaluated treatment step

φ_{Al} separation efficiency expressed by removed aluminium

φ_{COD} separation efficiency expressed by removed organic matter

AS aluminium sulphate

F means that the determinant was measured in a sample from which separable suspension was removed by centrifugation

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