

Application of a fluidised layer of granular material in the treatment of surface water. Part 1: aggregation efficiency of the layer

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ABSTRACT: Results obtained during the verification process of the new Zelivka Waterworks, which is equipped with a reactor containing a fluidised layer of silica sand are described in this paper. It was established that a fluidised layer of granular material represents an improved method for the application of shear to destabilised particles with respect to the kinetics of the aggregation process and of the character of formed aggregates. The separation of aggregates by filtration was enhanced compared to conventional aggregation processes.

INTRODUCTION

One of the basic prerequisites for facilitating the aggregation of impurities into separable aggregates is agitation, the conditions of which are determined by two parameters. One of these is the intensity of agitation: according to the magnitude of the mean velocity gradient, agitation can be either rapid or slow. The other parameter is the duration of agitation. The influence of both of these parameters on the kinetics of the treatment process has been studied by a number of authors and the recommended designs have varied widely [1–8].

A comparison of the properties of aggregates formed both in pilot and full scale plants showed that they differed considerably, although there are similarities between both processes from a technological point of view. One of the reasons for this is the fact that in the more commonly used agitation facilities, parameters defining the agitation intensity cannot be established precisely, due to the non-existence of sufficiently accurate methods for the calculation and determination of the velocity gradient. Another reason is the incorrect value of the agitation duration used, which in most calculations is taken as a theoretical and not an actual time. Finally, the third, and probably the most significant, cause is associated with the distribution of the velocity field in the agitated volume, because its uniformity is important with respect to the separation capability of the formed aggregates.

Differences in the properties of various aggregates, which affects the efficiency of the separation processes, led to a search for new methods of agitation, and resulted in the formation of more satisfactory aggregates. The use of a fluidised layer of granular material seemed to offer the greatest promise. Its application to the treatment of surface waters has been under study since 1979 [9,10] and it was installed as a technological element in a number of waterworks in the Czech Republic.

The results obtained from more than 3 years of operating a reactor with a fluidised layer of granular material, aimed at the verification of its aggregation capability, are summarised in this

paper. The results of the research on the separation ability of the fluidised layer will be published in part II of this paper.

FLUIDISED LAYER APPLIED AS AN AGITATION ELEMENT

When liquid flows through a fixed layer of granular particles in a direction opposite to that of the pull of gravity, the hydraulic resistance of this layer increases with increasing flow velocity. At a certain approach velocity (volumetric flow rate per unit face area) an equilibrium will be reached between the resistance H of the layer and its mass m . This balance is expressed by the equation:

$$\rho_f g h F = L_0 F (1 - \omega_0) (\rho_p - \rho_f) \quad (1)$$

The overpressure h is expressed by the equation:

$$h = L_0 (1 - \omega_0) \left(\frac{\rho_p}{\rho_f} - 1 \right) \quad (2)$$

When flow exceeds the critical velocity v_{cr} the layer expands and its fluidisation occurs. The upflow velocity v_{up} above the layer is

$$v_{up} = Q/F \quad (3)$$

This is much lower than the actual velocity v_{ac} in the layer, because the flow area is reduced by the particles forming the layer. The mutual relation between them is expressed by the equation:

$$v_{ac} = v_{up}/\omega_E \quad (4)$$

Transition from a stable layer to a fluid layer is indicated by a change in magnitude of mechanical energy dissipated in the layer:

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

Since the product of the concentration of particles and the depth of the layer is constant, and the loss of head is constant too, the magnitude of energy dissipated, regardless of the extent of expansion, remains the same. The extent of expansion is defined by the relationship:

$$\varepsilon = \frac{L_E - L_0}{L_0} \tag{6}$$

The volume V_v of voids in a layer, determining the retention period T of liquid in the layer, is calculated from the equation:

$$V_v = V_t \omega_E \tag{7}$$

At a rate of flow Q , the retention time in a layer T is:

$$T = \frac{V_v}{Q} = \frac{V_t \omega_E}{Q} = \frac{V_t \omega_E}{v_{up} F} = \frac{L_E \omega_E}{v_{up}} = \frac{L_0(\varepsilon + \omega_0)}{v_{up}} \tag{8}$$

When liquid flows through the layer, every grain of the layer

is a source of hydraulic shock and an origin of velocity gradient, the magnitude of which is given by the equation:

$$G = \sqrt{\frac{W}{V\eta}} = \sqrt{\frac{\Delta H}{vT}} = \sqrt{\frac{g(1 - \omega_E)(\rho_p - \rho_f)v_{up}}{\eta(\varepsilon + \omega_0)}} \tag{9}$$

Since the grains are uniformly distributed in the fluidised layer, the velocity gradient is also uniformly distributed.

METHODS OF EVALUATION

Evaluation of the efficiency of a fluidised layer was carried out on a full size plant at the Zelivka waterworks, where two different technological variants for the formation of suspension, separated by single-stage rapid gravity sand filters operating in parallel (Zelivka 1 and Zelivka 2) are installed. Both variants are shown schematically in Figs 1 and 2.

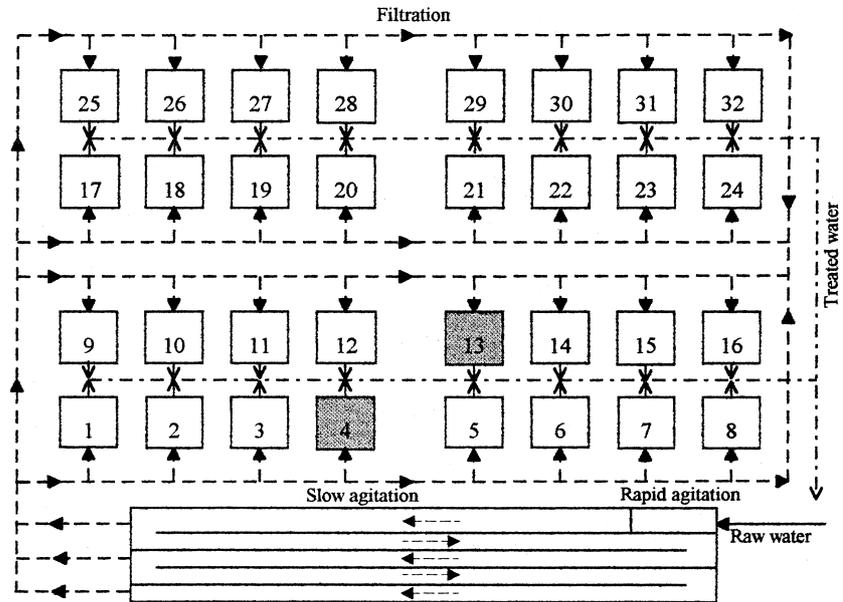


Fig. 1 Flow diagram of the Zelivka 1 waterworks.

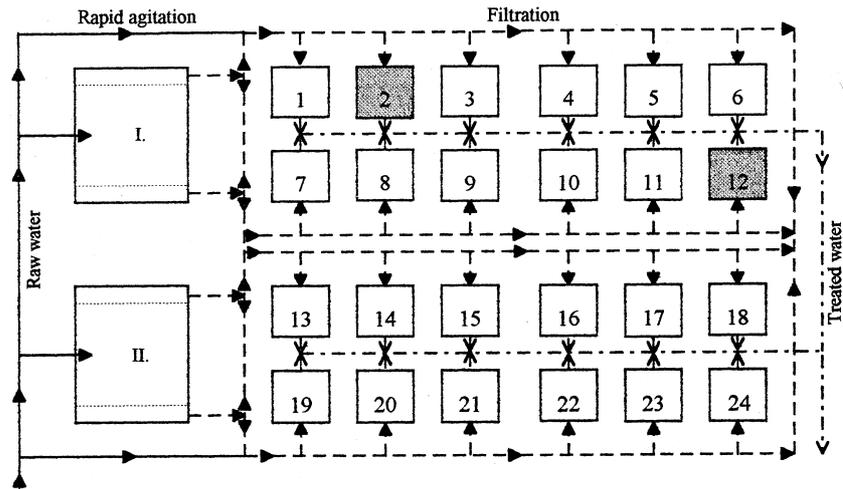


Fig. 2 Flow diagram of the Zelivka 2 waterworks.

The Zelivka 1 Waterworks was designed for a total capacity of $Q = 3.3 \text{ m}^3/\text{s}$. It incorporates rapid agitation (RA) in the tank ($7.5 \times 3.8 \times 4.0 \text{ m}$) by means of a blade stirrer designed to produce a mean velocity gradient $G_{RA} = 97/\text{s}$ for a retention time $T_{RA} = 35 \text{ s}$ and slow agitation (SA) in a meander-like chamber (1st chamber $25.4 \times 3.8 \times 4.0 \text{ m}$, 2nd to 5th chamber $32.5 \times 3.05 \times 4.0 \text{ m}$) producing a mean velocity gradient $G_{SA} = 29/\text{s}$, for a retention time $T_{SA} = 500 \text{ s}$.

The Zelivka 2 Waterworks, which was designed for a total capacity of $Q = 4.4 \text{ m}^3/\text{s}$, comprises two streams, each with a capacity of $Q_s = 2.2 \text{ m}^3/\text{s}$. Each stream is provided with a square reactor ($10.15 \times 10.15 \times 4.5 \text{ m}$) containing a layer of silica sand with a grain size $d_p = 0.3\text{--}0.8 \text{ mm}$. The stable depth of the layer is $L_o = 1000 \text{ mm}$, the mean velocity gradient produced is $G_{RA} = 285/\text{s}$, and the retention period in the fluidised layer corresponds to $T_{RA} = 56 \text{ s}$.

Separation of the formed suspension in both water works takes place in rapid gravity sand filters of the same arrangement, with the same granularity of filter media ($d = 1.1\text{--}1.6 \text{ mm}$) and the same backwashing process (air, air + water, water). The only difference between them are the depths of the filter bed (1600 mm in Zelivka 1 and 1400 mm in Zelivka 2) and the filtration velocities ($v_f = 3.83 \text{ m/h}$ with 32 filters in Zelivka 1 and $v_f = 6.67 \text{ m/h}$ with 24 filters in Zelivka 2).

Water containing flocculated aggregates is transported to the filters through a pipe, the diameter of which is the same in both waterworks (1000 mm) and its length to the filters under evaluation (Works 1–4, Works 2–2; Works 1–13, Works 2–12) is approximately the same.

The effect of agitation by a fluidised layer on the aggregation efficiency and separation properties of the suspension that is formed was evaluated in several stages. In the first stage, the aggregation in the reactors at Zelivka 2, both with and without a fluidised layer was compared. In the second stage, the effect of different technological arrangements of agitation, assuring the formation of a suspension in Zelivka 1 and Zelivka 2 were compared. In the third stage, the effect of aluminium sulphate and iron sulphate was compared. Finally, the effect of changes in the up-flow velocity in the layer was evaluated.

The efficiency of the water treatment process was ascertained by:

- 1 a direct comparison of water quality indicators: content of basic component of hydrolysing reagent (Me), chemical oxygen demand (COD_{Mn}), content of micro-organisms (Bio), determined in the raw water and in the filtrate;
- 2 determining separation efficiency ϕ , according to the relationship:

$$\phi = 1 - \left(\frac{C_f}{C_0} \right) \quad (10)$$

where C_0 is the initial concentration of indicators used and C_f is final value in the treated water.

The effect of agitation was evaluated by the degree of

aggregation α determined in the crucial stages of formation of suspension, calculated according to relationship:

$$\alpha = \frac{C_0 - C_f}{C_0} \quad (11)$$

where C_0 is the concentration of monitored indicator at the point of testing and C_f is the concentration of the monitored indicator determined in the same sample after its treatment by centrifugation under defined conditions.

The methods of evaluation have described previously [11–13].

The separation ability of the suspension was appraised by parameters measured during the operation of the filters, namely, the loss of head H (kPa), the duration of filtration run T (h) and quantity of suspension m (mg Me/L) removed by the filter bed during the period T . The evaluated filters are marked in the flow diagrams presented in Figs 1 and 2.

RESULTS AND DISCUSSION

Comparison of the aggregation efficiency of the reactor, with and without filling

When comparing reactor efficiencies both with and without the sand filling, the aggregation conditions differed by the value of the velocity gradient and its distribution in the agitated volume. For the reactor without sand filling, the mean velocity gradient, calculated on the basis of the measured loss of head according to eqn 9, corresponded to $G_{RA} = 95.3/\text{s}$, which is in agreement with the commonly accepted value of this design parameter, and the actual retention time in the reactor that has been determined chemically [14] corresponded to 60% of the theoretical retention time. For the reactor with sand filling, the velocity gradient was calculated to be $G_{RA} = 285/\text{s}$ and the uniformity of its distribution was ensured by the filling itself, so that the actual retention time corresponded to 93% of the theoretical time.

The operational conditions were the same for both reactors: flow rate $Q = 2.2 \text{ m}^3/\text{s}$, $D = 15.0 \text{ mg Al}_2(\text{SO}_4)_3 \cdot 18\text{H}_2\text{O}/\text{L}$, water temperature $t = 4.0^\circ\text{C}$.

The samples for evaluating the efficiency of the fluidised layer were taken at the inlet to the fluidised layer from an underfloor compartment and also from the reactor outlet on the upstream side to a discharge weir.

Rudimentary testing showed that the aggregation efficiency of the underfloor compartment was negligible; the value of the degree of aggregation ranged from $\alpha = 0.05$ to $\alpha = 0.10$. The determination of the α value at the inlet to the fluidised layer from the underfloor compartment was somewhat problematic; the measured values were probably higher than the actual values, because the samples are withdrawn through a pipe, having a diameter 20 mm and length of 5 m, in which additional aggregation takes place.

The indirect evaluation of the separation properties of the

Table 1 Influence of the length of the feed line on the aggregation efficiency

Sampling point	RA	F 1	F 2	F 3	F 4	F 5	F 6
α	0.45	0.52	0.56	0.59	0.62	0.72	0.75
Sampling point	RA	F 7	F 8	F 9	F 10	F 11	F 12
α	0.45	0.72	0.74	0.73	0.73	0.82	0.80

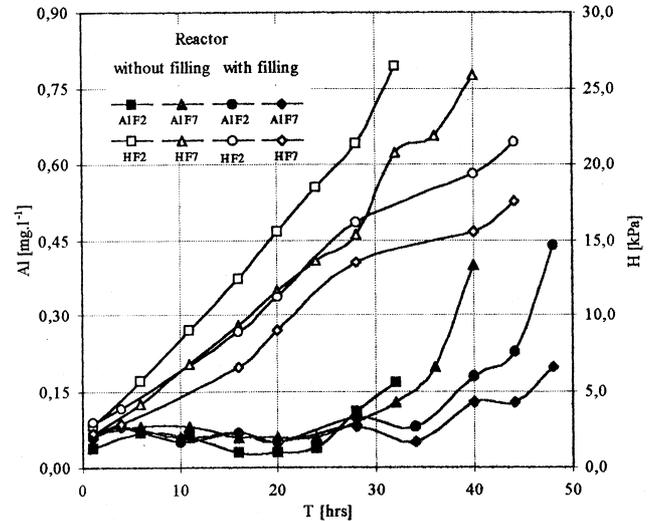
suspension was also accompanied by difficulties because aggregation is enhanced in the feed pipes to the filters. The aggregation efficiency of the feed pipe is evident from the difference in the α values determined at the outlet from the chamber for rapid agitation without the sand filling and inlet to the filters, as stated in Table 1. It is evident that the aggregation efficiency increases with the length of the feed pipe and the number of built-in fixtures. In stream 1 the aggregation efficiency is influenced mainly by the length of the feed pipe, and in stream 2 by the more built-in fixtures. The lowest α values were obtained from the filters which were closest to the reactor outlet.

Results comparing the aggregation efficiency and separation properties of the formed suspension obtained with the reactor both with and without the filling of granular material are summarised in Table 2. The course of filtration runs under comparison is graphically demonstrated in Fig. 3.

It follows from these results that the degree of aggregation at the outlet from the rapid agitation reactor without the filling is lower by $\Delta\alpha = 0.27$ than that at the outlet from the reactor with the filling. This difference is considerably reduced at the feed line to the filters, so that the difference at the inlet to filter no. 2 is lower by $\Delta\alpha = 0.11$ and to filter no. 7 by $\Delta\alpha = 0.03$. A comparison of the filter performances corresponding to the reactors with and without the filling, shows that the duration of filtration runs and the quantities of intercepted suspension from these filters indicates a considerably higher efficiency of filtration when the reactor with filling is used. With both alternatives, the filtration run ended with an Al breakthrough

Table 2 Performance efficiency of reactors with and without filling

	Filling	
	without	with
$D \text{ Al}_2(\text{SO}_4)_3 \cdot 18\text{H}_2\text{O}$ (mg/L)	15.0	15.0
Rapid agitation outlet α	0.45	0.72
Filter no.	2	12
inlet α	0.64	0.74
outlet α	0.75	0.77
outlet H (kPa)	23.5	21.5
T (h)	30.7	36.0
ϕ Al	0.96	0.92
ϕ COD_{Mn}	0.31	0.33
m (mg/L)	183.1	177.5


Fig. 3 Comparison of the filtration runs. Reactor with and without filling.

into the filtrate that was above the permissible limit of 0.20 mg Al/L. There was a higher head loss H across the filter when the reactor was operated without the filling than that obtained with the filling. The difference in head loss also corresponded to a difference in the quantity of suspension intercepted in filters, which in the case of the reactor with the filling was higher by 59 mg/L of the bed volume.

Since all other factors which could influence the duration of filtration runs and efficiency of filtration were the same, then the extended runs and higher filtration efficiency of the reactor with filling could only be due to the difference in the character of the formed suspension.

Comparison of the aggregation efficiency of mechanical agitation and hydraulic agitation in the fluidised layer

In the second stage of testing, the efficiency of mechanical agitation in Zelivka 1 was compared to that of the fluidised layer in Zelivka 2, under design flows for each plant. In both plants the optimum dosage of aluminium sulphate $D = 18.0 \text{ mg Al}_2(\text{SO}_4)_3 \cdot 18\text{H}_2\text{O/L}$ was applied. The results are summarised in Table 3 and graphically demonstrated in Fig. 4.

The aggregation efficiency of the agitation was found to be significantly higher with mechanical agitation than with the use of a fluidised layer. Due to the action of a velocity gradient in the feed lines to the filters, the efficiency of aggregation in both technological variants was substantially increased, so that the degree of aggregation α of suspension entering the filters was very high, varying from about $\alpha = 0.88$ for the mechanical agitation to about $\alpha = 0.80$ for the agitation in the fluidised layer. Although the degree of aggregation of suspension formed in the fluidised layer was lower, its separation ability with respect to separation by filtration was better. The filtration

Table 3 Comparison of aggregation efficiency and separation properties of suspension formed by traditional technology and in the fluidised layer under design conditions

Waterworks	1	2		
RA outlet α	0.72	0.72		
SA outlet α	0.86	–		
Filter no.	4	13	2	12
inlet α	0.89	0.88	0.80	0.81
H (kPa)	22.7	22.7	23.5	22.2
T (h)	36.2	36.2	25.6	32.6
ϕ Al	0.97	0.97	0.95	0.96
ϕ COD _{Mn}	0.35	0.35	0.33	0.33
ϕ Bio	0.92	0.95	0.91	0.95
m (mg/L)	117.9	118.1	162.1	214.8

runs ended with a breakthrough of aluminium into the filtrate that was greater than permissible levels (0.2 mg Al/L or 0.1 mg Fe/L); in Zelivka 1 it occurred at an unused pressure head. The duration of filtration runs in Zelivka 1 were longer, due to their lower filtration velocity, but the quantity of intercepted suspension in Zelivka 2, with the reactor with the fluidised layer, was greater.

Since it was difficult to include in the evaluation the influence of different filtration velocities as a factor influencing the course of retaining of suspension in the filter bed and thus to unequivocally establish the influence of the character of suspension, the performance efficiency was evaluated at the same filtration velocities. The plant output corresponded to the design capacity ($Q_1 = 3.3 \text{ m}^3/\text{s}$ and $Q_2 = 2.2 \text{ m}^3/\text{s}$), the dosage in both plants corresponded to $18.0 \text{ mg Al}_2(\text{SO}_4)_3 \cdot 18\text{H}_2\text{O}/\text{L}$, and the filtration velocity was set to be the same as in Zelivka 2

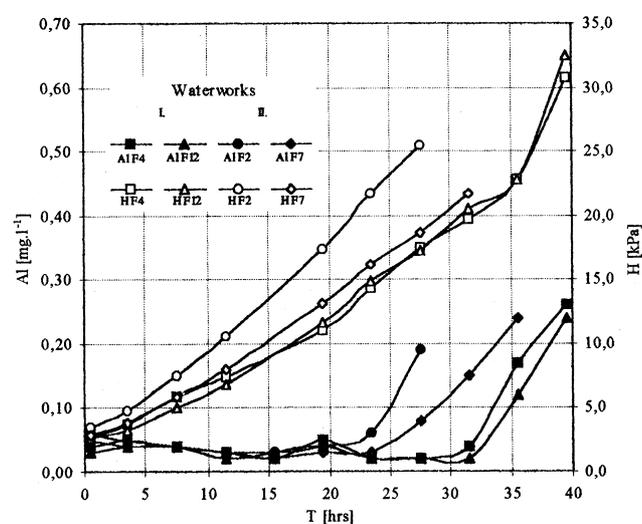
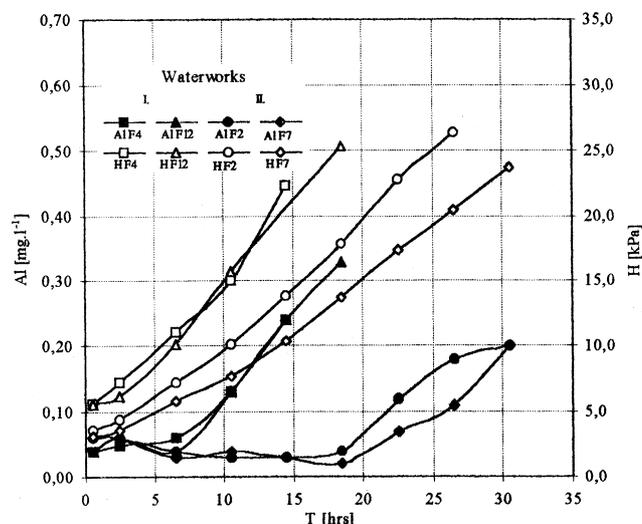
Table 4 Comparison of aggregation efficiency and separation properties of suspension formed by traditional technology and in the fluidised layer under the same filtration velocity

Waterworks	1	2		
RA outlet α	0.77	0.69		
SA outlet α	0.88	–		
Filter no.	4	13	2	12
inlet α	0.90	0.88	0.80	0.83
H (kPa)	16.9	17.4	23.5	23.5
T (h)	13.0	13.0	24.2	30.8
ϕ Al	0.93	0.94	0.94	0.94
ϕ COD _{Mn}	0.37	0.38	0.36	0.36
ϕ Bio	0.92	0.92	0.92	0.92
m (mg/L)	74.1	73.01	149.4	188.5

($v_f = 6.67 \text{ m/h}$). In Zelivka 1 the higher filtration velocity was achieved by varying the number of filters in operation. The results are summarised in Table 4 and plotted in Fig. 5.

It is evident that the degree of aggregation of suspension after slow agitation in Zelivka 1 was substantially higher than that of Zelivka 2 after rapid agitation. Although there was not a significant difference between the treatment efficiencies of either water works, the increased filtration velocity in Zelivka 1 manifested itself profoundly its performance efficiency. The filtration runs were terminated in both works because Al concentrations exceeded permissible levels in the filtrate. The filtration runs in Zelivka 2 were, however, 80–130% longer and the quantity of retained suspension about 100–200% greater than that in Zelivka 1, with the utilised pressure head being considerably lower in Zelivka 1.

When we evaluating the length of the filtration runs it was

**Fig. 4** Comparison of the filtration runs. Design parameters of compared waterworks.**Fig. 5** Comparison of the filtration runs using the same filtration velocity.

necessary to take into consideration the fact that the loading of filters in Zelivka 2—due to the direct separation of impurities that was taking place in the fluidised layer—was $\approx 15\%$ lower than that in Zelivka 1.

Influence of dosing rate and of type of destabilising reagent

The influence of dosing rate on the aggregation efficiency and on the separation properties of formed aggregates was evaluated using two different doses of aluminium sulphate and verified with two different dosages of iron sulphate. The results are summarised in Table 5.

As it is seen, the value of α at the outlet from the fluidised layer, was nearly the same for both the optimum dosage of the aluminium sulphate (19 mg/L) and for a dosage that was a little lower than optimum (18 mg/L) and it increased in the feed lines to the filters. The quantity of retained impurities was—due to the greater amount of formed suspension—greater at the optimum dosage. The filtration runs are maintained at the optimum dosage a little longer, because the fully destabilised particles are able to occupy the filter bed at more stable positions energetically and geometrically and therefore form more compact structures. Comparing the optimum dosage of iron sulphate (17.0 mg/L) to a dosage which was considerably higher than the optimum (22.0 mg/L) resulted in a marked increase in the aggregation efficiency. A substantial shortening of duration of the filtration run at the higher dosage was caused by the greater quantity of suspension produced.

A comparison of the influence of aluminium and iron sulphate at optimal dosages on the aggregation efficiency and on the separation properties of suspension is demonstrated in

Table 5 Influence of the dosing rate on the aggregation efficiency and separation properties of suspension

Reagent	$\text{Al}_2(\text{SO}_4)_3$		$\text{Fe}_2(\text{SO}_4)_3$	
Dosage (mg/L)	17.8	18.9	16.8	22.1
RA outlet α	0.69	0.72	0.53	0.75
Filter no. 2				
inlet α	0.80	0.80	0.70	0.83
outlet H (kPa)	23.5	23.5	23.5	23.5
T (h)	24.2	25.6	20.6	14.6
m (mg/L)	149.4	162.1	333.5	313.5
ϕ Me	0.94	0.95	0.97	0.97
ϕ COD _{Mn}	0.36	0.33	0.34	0.39
ϕ Bio	0.92	0.91	0.86	0.90
Filter no. 12				
inlet α	0.83	0.81	0.73	0.86
outlet H (kPa)	23.5	22.2	23.5	23.5
T (h)	30.8	32.6	25.8	15.6
m (mg/L)	188.5	214.8	417.3	335.0
ϕ Me	0.94	0.96	0.97	0.97
ϕ COD _{Mn}	0.36	0.33	0.34	0.37
ϕ Bio	0.92	0.95	0.85	0.87

Table 6 Influence of the type of destabilising reagent on the aggregation efficiency and separation properties of formed suspension

Reagent	$\text{Al}_2(\text{SO}_4)_3$	$\text{Fe}_2(\text{SO}_4)_3$
Dosage (mg/L)	18.9	16.8
RA outlet α	0.72	0.53
Filter no. 2		
inlet α	0.80	0.70
outlet H (kPa)	23.5	23.5
T (h)	25.6	20.6
m (mg/L)	162.1	333.5
ϕ Me	0.95	0.97
ϕ COD _{Mn}	0.33	0.34
ϕ Bio	0.91	0.86
Filter no. 12		
inlet α	0.81	0.73
outlet H (kPa)	22.2	23.5
T (h)	32.6	25.8
m (mg/L)	214.8	417.3
ϕ Me	0.96	0.97
ϕ COD _{Mn}	0.33	0.34
ϕ Bio	0.95	0.85

Table 6. It can be seen that the application of aluminium sulphate is advantageous, because at the same treatment efficiency the filtration runs are longer. This statement reflects the experience of the waterworks, particularly in the case where raw water contains large quantities of micro-organisms.

Influence of the throughput

The results summarised in Table 7 show that the aggregation efficiency changes with changing throughput. It increases with

Table 7 Influence of the throughput on the efficiency of aggregation and separation efficiency of filters

Throughput Q (m ³ /s)	2.2	1.65	1.10
Filtration velocity v_f (m/h)	6.67	5.01	3.34
RA inlet α	0.44	0.52	0.63
outlet α	0.72	0.75	0.76
Filter no. 2			
inlet α	0.75	0.77	0.78
outlet H (kPa)	19.6	16.4	12.9
T (h)	41.0	51.0	72.0
m (mg/L)	255.3	239.1	220.4
Filter no. 12			
inlet α	0.77	0.77	0.82
outlet H (kPa)	18.4	16.6	13.4
T (h)	48.0	60.0	85.0
m (mg/L)	249.8	223.7	215.8

decreasing throughput, in particular, upstream to the fluidised layer. The highest increase in the value of α was seen at the outlet from the rapid agitation; its subsequent changes in the course of the treatment process enhancement of aggregation were smaller and were found to be approximately the same. With increasing throughput, the duration of the filtration run shortens, whereas the shortening of the filtration run is not proportional to the increased loading; filtration runs ended with an AI breakthrough.

CONCLUSION

From these measurements it can be concluded that the fluidised layer of granular material represents a suitable technological addition, ensuring improved agitation in the aggregation with respect to the kinetics of the process, as well as to the character of formed aggregates from the point of view of their separation.

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BIBLIOGRAPHY

- 1 Amirtharajah A. Initial mixing. Presented at an AWWA Seminar, 'Coagulation and filtration. Back to basis', St Louis, Missouri, 1981.
- 2 Claus E. Mischungsprobleme in der Wasseraufbereitung. *Wasserw – Wassertechnik* 1967; **17**: 332–335.
- 3 Cohen JM, Hannah SA. *Water Quality and Treatment. Coagulation and Flocculation*. McGraw-Hill, New York, 1971.
- 4 Committee Report. *Water treatment plant design*. Report No. 10006, ASCE, AWWA and CSSE, New York, 1969.
- 5 Anonymous. Upgrading water treatment plants to improve water quality. Presented at the AWWA Seminar, Denver, Colorado, 1980.
- 6 Francois RJ. The influence of mixing parameters and water quality on the flocculation of kaolinite with aluminium sulphate. In: *Chemistry for Protection of the Environment II*. Elsevier, Amsterdam, Netherlands, 1986.
- 7 Montgomery JM. *Water Treatment Principles and Design*. Wiley-Interscience, New York, 1985.
- 8 Ives KJ. *The Scientific Basis of Flocculation*. Sijthoff & Nordhoff Intl Publ., Alphen aan der Rijn, Netherlands, 1978.
- 9 Mutl S. *Rapid Agitation in the Fluidized Layer* (in Czechoslovakian).

Report to the Ministry of Water Management. Hydroprojekt, Prague, Czech Republic, 1990.

- 10 Mutl S, Hereit F. *Intensification of the Waterworks Vir Applying the Fluidized Layer of the Fluidized Layer* (in Czechoslovakian), Report to the Ministry of Water Management. Hydroprojekt, Prague, Czech Republic, 1981.
- 11 Hereit F, Mutl S. Formation of separable suspensions and method of its assessment. *J Water SRT—Aqua* 1980; **5**: 95–99.
- 12 Mutl S. Evaluation of the course of the water treatment process by the characteristics of the formed suspension. Presented at the Symposium of The CAEM, 'Water treatment and modelling of the treatment processes'. Zivohost, Czech Republic, 1977.
- 13 Polasek P, Mutl S. *Guidelines to Coagulation and Flocculation for Surface Waters, Vol. I, Design principles for coagulation and flocculation systems*. P. Polasek & Associates, Marshalltown, South Africa, 1996.
- 14 Levenspiel O. *Theory and Calculation of Chemical Reactors*. Statni Nakladatelstvi Tecknicke Literatura, Prague, Czech Republic, 1967.

Appendix. Nomenclature

C	concentration (mg/L)
F	area, cross section of the reactor (m^2)
G	velocity gradient (/s)
g	gravity constant (m/s)
H	loss of head (kPa)
h	overpressure (m)
L_o	depth of the layer at no flow (m)
L_E	depth of the fluidised layer (m)
Q	rate of flow (m^3/s)
T	retention period of liquid in the layer (s)
V_t	total volume of the stable layer (m^3)
V_v	volume of voids in a layer (m^3)
v_{ac}	actual upflow velocity (m/s)
v_{cr}	critical upflow velocity (m/s)
v_{up}	upflow velocity (m/s)
W	dissipated energy ($kg.m^2/s$)
α	degree of aggregation (–)
ε	measure of expansion (–)
ρ_f	density of liquid (kg/m^3)
ρ_p	density of particles (kg/m^3)
ω_o	porosity of the layer at no flow (–)
ω_E	porosity of the fluidised layer (–)
η	dynamic viscosity ($kg/m.s$)
ν	kinematic viscosity (m^2/s)
ϕ	separation efficiency (–)