Environment Protection Engineering

Vol. 41 2015 No. 3

DOI: 10.5277/epe150308

MICHAŁ ZIELINA[1], WOJCIECH DĄBROWSKI[1], ROBERT IAN MACKIE[2]

# LABORATORY AND NUMERICAL EXPERIMENTS INTO EFFICIENT MANAGEMENT OF VDR FILTER PLANTS

One of the operation rules of variable declining rate (VDR) filters states that the longest filtration runs are achieved when the highest flow through the most recently backwashed filter and the highest available head loss coincide. This rule has previously been published by Dąbrowski, who used a simplified mathematical model of VDR filters developed by Di Bernardo. However, until now no experimental verification of this optimisation approach has been presented. A series of tests were carried out on a laboratory VDR filter plant, using a suspension collected from an industrial sedimentation tank and using alum as a coagulant. The results of the experiments were compared with computations carried out according to a combined unit bed element (UBE)-phenomenological model of deep bed filtration adapted to VDR operation by Mackie and Zhao. It was confirmed that the longest filtration runs did indeed correspond to the pairs of the highest $H$ and $q_1/q_{avr}$. The UBE-phenomenological model of deep bed filtration gave results of computations quite close to experimental results and improved the accuracy of calculations based on the theoretical model by Di Bernardo.

## 1. INTRODUCTION

Flow of water suspension through variable declining rate (VDR) filters is controlled by orifices usually installed at outflows from each of the filters in a plant. Turbulent head losses created by these orifices should be significant for a freshly backwashed filter and small in comparison with the laminar head loss of flow through clogged filter media just before a filter backwash. In other words, flow rates through clean filters of high media permeability are mainly controlled through turbulent head losses of flow through orifices but these head losses should be relatively small for clogged filters of high media resistance to flow, which mostly limits the value of flow

[1]Environmental Engineering Department, Cracow University of Technology, ul. Warszawska 24, 31-155 Cracow, Poland, corresponding author M. Zielina, e-mail: mziel@vistula.pk.edu.pl

[2]School of Engineering, Physics and Mathematics, University of Dundee, Dundee DD1 4HN, U.K.

rate, and thus the filtration velocity. A VDR plant will usually have the following features (cf. Fig. 1):

1) all filters are identical,

2) raw water inflow is located below the lowest water level above filter media,

3) head losses of flow through the pipe work are negligible in comparison with the head loss of flow through a filter with an orifice,

4) there are at least four filters in a plant.

Apart from these limitations, the filters are essentially the same as those operated under constant rate (CR) filtration or declining rate (DR) filtration.

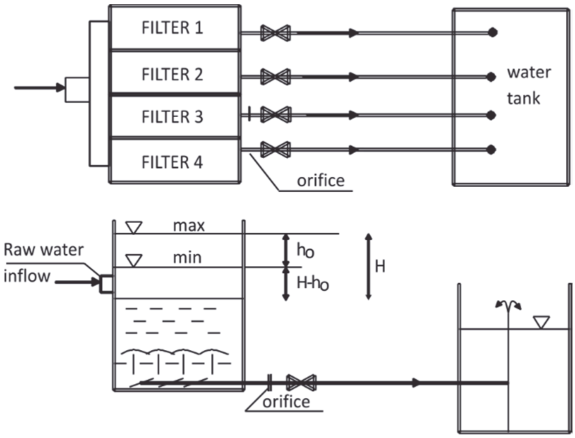


Fig. 1. An example of VDR filter construction

Fluctuations in time of the free surface water level above the filters and changes of the pattern of flow rate are presented in Fig. 2 developed from results of measurements made at a laboratory/pilot scale VDR filter plant constructed at the Cracow University of Technology.

For stable quality and temperature of raw water, filters are backwashed at regular periods of time selecting the most clogged unit each time when the water surface above filters reaches a given level. These values are used to calculate the total head loss, $H$, of flow through filter media, drainage and an orifice. This value is used during the designing process. Then the water level above all filters increases in time as deposit accumulates in the filter media pores. Requirement 2 results in the same water surface level above all filters in any moment of time. This level takes the lowest position just after connecting a freshly backwashed unit to work and the head loss of flow through all filters is lower by a value $h_0$ at this moment, thus it equals $H - h_0$. The flow rates and the water level patterns repeat each time after subsequent backwashes in a plant. It is important to note that changes of flow rates through each of the filters between subsequent backwashes in a plant are relatively small when compared with the changes just after the backwashes.

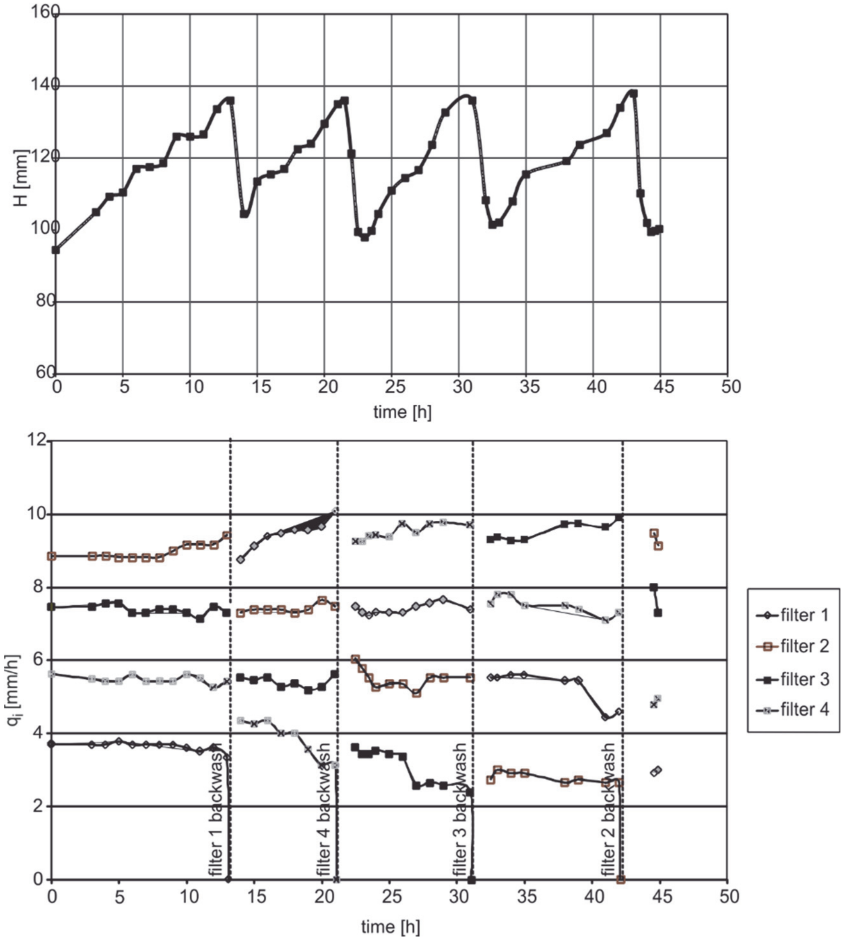


Fig. 2. A pattern of unconfined water surface level above a four filters plant in a laboratory/pilot plant scale and changes in flow rates according to measurements reported at the Cracow University of Technology

Based on the assumption that flow rates through filters are constant in time between backwashes and, moreover, neglecting water accumulation above filter media resulting from fluctuations of water surface levels, Di Bernardo [1] developed a useful mathematical model of a bank of VDR filters which enabled prediction of design parameters of orifices and the water table fluctuation height $h_0$ for the assumed total plant capacity $Q$, total head loss $H$ and an assumed ratio of the flow rate $q_1$ through the most recently backwashed filter to the average value of the flow rate per filter $q_{avr}$ for the plant. Thus although the length of a filter run and filtrate quality are dependent upon the raw water quality, if the backwashes are rigorously established for the same head loss H, the flow rates through filters essentially remain the same. This has been verified both numerically and experimentally [2–5]. The model did not include any equation describing the kinetics of filter media clogging, so the design method is universal. The description of this model has been simplified recently by cutting the number of equations and unknowns [6]. An optimisation procedure of VDR filter plant operation has been proposed [7, 8] which agree with results of computations performed on a unit bed element (UBE) mathematical model of (VDR) filters. However, this simple optimisation method of VDR plant operation needs more solid verification than just numerical results. Now empirical data confirming this operation procedure are reported both in a laboratory/pilot and with supporting evidence from a full scale installation.

## 2. PURPOSE OF THE STUDY

One rule for the efficient operation of VDR filters requires keeping as high as possible both the total head loss through filters before a subsequent backwash in a plant, and the ratio of the highest to average flow rates through filters. This rule is expected to provide the highest resistance to flow of the dirtiest filter just before a backwash, and perhaps the longest filter runs and the lowest losses of water for backwash purposes. It was developed [7, 8] by assuming the flow rates through filters followed a geometric progression with the ratio:

$$\frac{q_{i+1}}{q_i} = \frac{1-h_0}{H} \tag{1}$$

where: $q_{i+1}$ is the flow rate through filter $i + 1$, $q_i$ is the flow rate through filter $i$, $h_0$ is the height of water table fluctuation, $H$ is the head loss of the flow through a filter just before a subsequent backwash in a plant. Indexes $i$ denote the order of backwashes in a plant, thus the filter $i$ was backwashed just before the filter $i + 1$ and is capable to accept higher flow rate than the filter $i + 1$.

Equation (1) was also used to develop a strategy for adjusting filter operation parameters for keeping the same flow rate distribution among filters for various total head losses of flow through a plant, height of water table fluctuation, and resistance of orifice, used to maintain a required ratio of maximum to average filtration velocities in a plant [7, 8].

In the present study, the strategy of the plant operation and the optimisation approach are verified in both numerical and laboratory experiments. The UBE model [9] of deep bed filtration was combined with hydraulics of VDR filters, so that both hydraulic and removal kinetics were taken into account [10]. A laboratory set-up was constructed, consisting of four flocculators, four filtration columns operating in variable declining mode, to conduct tests on a coagulated water suspension under conditions typical of the operation of a surface water filter plant. Numerical and laboratory experiments made it possible to:

- verify both the investigated relationships between parameters of VDR filters and the optimisation approach to a plant management in two different circumstances: first for flow of non-coagulated fine suspension (numerical experiments), and second for clay suspension coagulated by small doses of alum (laboratory experiments),
- investigate how close the results of computations based on different models of VDR filters were to experimental results,
- check whether the UBE model was properly incorporating the equations governing the hydraulics of VDR filters.

## 3. EXPERIMENTS

### 3.1. EXPERIMENTAL SET-UP

The set-up presented schematically in Fig. 3 was constructed mostly in a basement (filtration columns) and on the first floor (flocculators and reservoirs) in a place where changes of air temperature were limited to 7 °C and did not affect the temperature of suspension by more than 2 °C. Direct filtration has been applied to test the efficiency of variable rate control operation system which is considered as an economical solution for treatment of high quality raw water. This system requires limited chemical doses and economical maintenance costs. Tanks with suspension (1), overflow tank (3) divided the suspension equally among the four flocculators. A vessel with 5 wt. % solution of aluminium sulfate and a dosing pump (6), flocculators (7), filtration columns (9) supplied with orifices controlling flow rates and arrangement of outflow (16) with flow meters (14), (17) were built as separate units and connected by flexible pipes to avoid any propagation of shakes from flocculators or tanks containing suspension to filtration columns.

Water suspension from one of two tanks (1) of the volume of 0.8 m$^3$ each was delivered with a circulation pump (2) Grundfos UPS 25-40 to a horizontal overflow tank (3) with four outflows (4) towards the flocculators (7). Using two suspension tanks (1) enabled continuous operation. Stirrers located inside these tanks protected suspension against deposition of large particles. To adjust identical inflow to each of four flocculators (7) the outflows from the open horizontal overflow tank (3) were equipped with orifices of the same opening diameters, located in caps movable up and down by a screw which enabled final adjustment of the flow rates. The flow through orifices was free in the sense that there was atmospheric pressure below each orifice. A constant level of water in the horizontal overflow tank (3) was maintained by an overflow discharging small excess of suspension back to the suspension tanks (1).

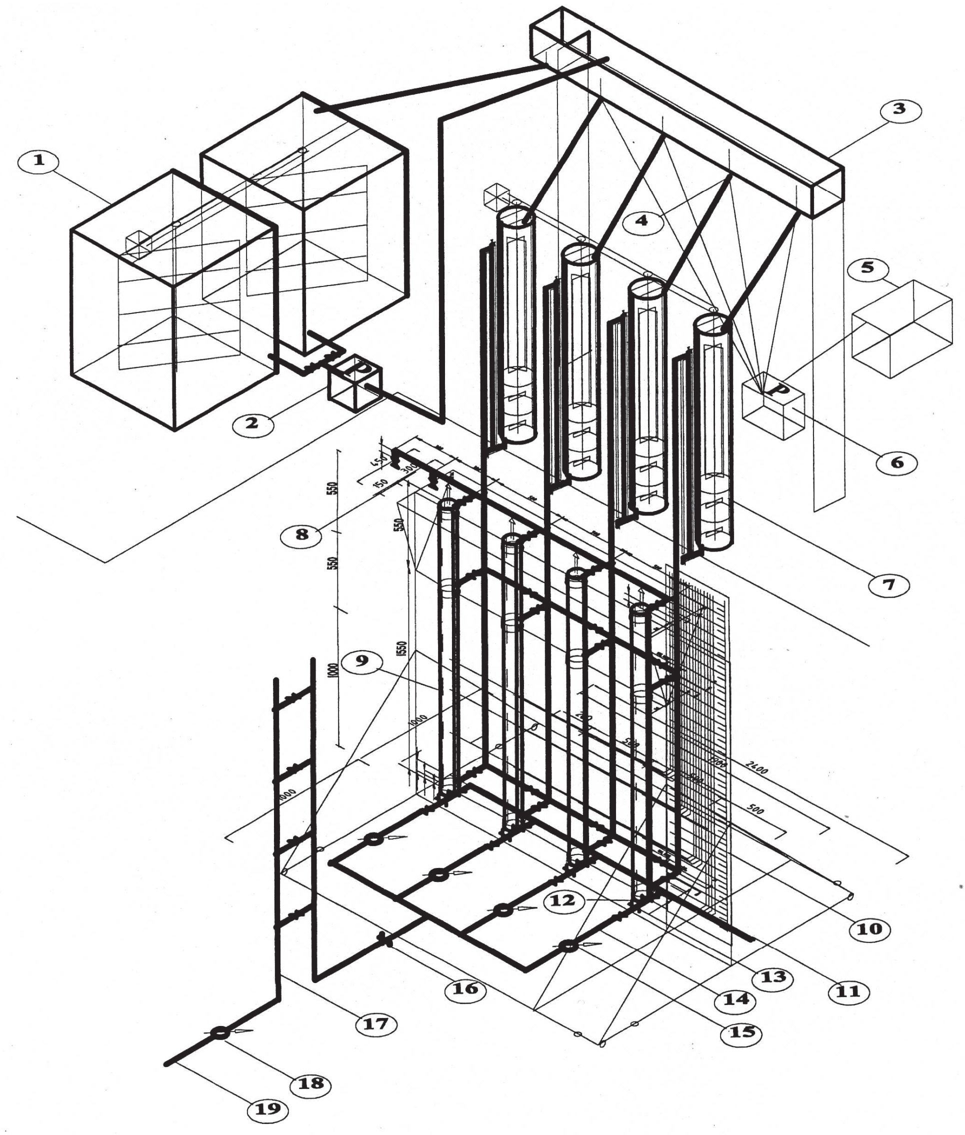


Fig. 3. Schematic of the laboratory set-up (for details, see the text)

Coagulant (alum) solution of 5 wt. % was dosed by a peristaltic dosing pump Zalimp-PP1B05 (6) from the vessel (5) to the flocculators (7). The mechanical flocculators (7) were divided into three compartments located inside in series vertically.

Paddles of all three compartments rotate around the same axis in each of flocculators. A chain transmitter connected all the flocculators with a small electric engine produced the same rotation speed for each of the sets of paddles. The number of paddles decreased moving from upper to lower departments of flocculators, to decrease the Camp–Stein gradient $G$ below 300 $s^{-1}$ in the lowest compartment. Higher Camp–Stein gradients than the conventional value 100 $s^{-1}$ were used to produce smaller flocs since there was no sedimentation tank in which the largest flocs could be settled. Smaller flocs penetrate deeper the filter media avoiding cake filtration at the surface of the sand bed.

The diameter of flocculators was the same as the diameter of a jar in a standard US jar test and the paddles in the lowest two compartments of the same shape and dimensions, thus the Camp–Stein gradient was estimated based on the curves known from the literature referring to the jar test experiments. The retention time in the flocculators was about 22 min. Outflows from flocculators (7) towards filtration columns (9) were constructed in a form of overflows thus the level of water in all flocculators was the same during all filter runs. Plexiglas columns of the internal diameter of 94 mm and 2 m high were filled up to 80 cm with quartz sand media of $d_{60}$ = 0.8 mm, $d_{10}$ = 0.5 mm and density of 2.56 g/cm$^3$. The average porosity for the whole media calculated from dry sand density and known mass of each of the fractions was 40.8%. The size distribution of dry fractions in the total mass of the media was the following: the fraction 0.4–0.5 mm – 10.0%, fraction 0.5–0.63 mm – 18.8%, fraction 0.63–0.8 mm – 31.2%, fraction 0.8–1.0 mm – 36.1%, fraction 1.0–1.25 mm – 3.9%. Below the 80 cm high filter media, 10 cm gravel support was constructed on a drainage consisting a horizontal plate containing 17 holes of the diameter of 6 mm, 5 cm below which another plate with symmetrically allocated holes of the diameters of 8 mm was located, to distribute backwash water more homogeneously. Each column had an overflow to remove water continuously during backwash. Backwash water was directly supplied from a drinking water tap. Backwash was provided by water alone, so some pulsation of flow rates was applied to avoid any mud ball problems. This was achieved without difficulties. Orifices delivering turbulent head losses required for VDR operation were located at the outflows from filters-between the drainages and flow meters Metron-JS0.6. These orifices, identical for all filters, were replaced for each series of experiments characterised by different ratio of maximum to average filtration velocities through the laboratory scale plant. The openings of these orifices were computed by the Di Bernardo method [1]. Each of the four filtration columns was equipped with seven connections to the piezometer board (10). One of these connections was located just above the drainage and one was located at the outflow below the orifice and flow meter, so they measured head of pressure once below and once after the turbulent head loss, which was found to be created almost exclusively by the orifice. Different settings of head were possible at outflow arrangement (16) by opening and turning off valves on different levels above the floor. Readings from flow-meters (14) were verified by the flow-meter (18) based on the mass balance.

Primary experiments were conducted to predict the efficiency of suspended solids removal. This has been evaluated indirectly by turbidity reduction up to 90%, which is an acceptable level for direct filtration.

### 3.2. PRELIMINARY EXPERIMENTS

In order to operate the set-up under conditions similar to those existing in water treatment plants, the suspension was prepared from natural wet clay, using an homogeniser and a suspension of the turbidity of 7.5 mg/dm$^3$ $SiO_2$ was produced in the tanks (1). This clay was collected from a treatment plant backwash water sedimentation tank. Jar tests indicated that a reasonable coagulant dose was 8 mg/dm$^3$ of alum. However, jar tests are used to predict doses optimal for sedimentation, and a dose of 6.5 mg/dm$^3$ was proved to produce similar quality filtrate at longer filter runs, and this dose was used for all experiments with the flocculated suspension. In addition to the coagulated experiments, one set of experiments was run with non-coagulated suspension. The properties to be verified experimentally should not depend on the kinetics of clogging and this additional set of experiments would increase the range of conditions under which the properties were tested.

It was important to characterise imperfections of the laboratory set-up. The flow rates of inflows to flocculators varied by ±1.5% from the average value. The coefficient $c_1$ is the ratio of head loss per hydraulic load. In several experiments, the coefficient $c_1$ representative for resistance of filter media to flow was measured for flow of suspension through clean sand bed at 18 °C, resulting in the following average values of $c_1$ (d$^{-1}$): 0.00236 for filter 1, 0.00219 d$^{-1}$ for filter 2, 0,00228 d$^{-1}$ for filter 3 and 0.00225 d$^{-1}$ for filter 4. The average value was 0.00236 d$^{-1}$. In spite of careful preparation of the filter beds, the divergences of $c_1$ were only slightly below ±4% from the average value. This coefficient depends on the way of turning off backwash water and to a lesser extent on the concentration of suspension. It was not expected to be of any importance at fairly low temperatures, but a small difference in head loss created by suspension and by water flows was observable at 17–19 °C . The measured head loss of flow through clean filter media was similar to the values predicted from the UBE model used here in numerical experiments and the results were not further than by ±7% of the measured values. The Kozeny–Carman equation was used in the UBE model to calculate head-loss of flow through clean filter media. One of the reasons for differences in resistances to flow of the examined filter media at the beginning of filter runs was a self-compaction of filter grains. It was observed that the height of the media decreased by up to 3% from the value immediately after backwash.

Suspension temperature remained in the range 17–19 °C. Orifices installed at the outflows from the filters were tested prior to installation. Measurements confirmed that the pressure losses caused by the drainage and flowmeters were negligible compared to the turbulent losses in the orifices. Since the flow through the orifices was fully turbulent, the exponent $n$ in the following equation was taken to be equal to 2, the value confirmed by tests:

$$\Delta h_{\mathrm{loss}\ (i)} = c_2 q_i^n \tag{2}$$

where: $\Delta h_{\mathrm{loss}\ (i)}$ is the turbulent head loss of flow through an orifice installed at the outflow of filter $i$, and $c_2$ is a coefficient describing turbulent head loss.

Three diameters of holes $d^{(j)}$ in orifices were used in the experiments, and three values of $c_2$ were experimentally obtained including the head loss in the drainage, orifice, and flow-meters: $c_2^{(1)} = 0.000007699$ m/(m/d)$^2$ for $d^{(1)} = 3.2$ mm, $c_2^{(2)} = 0.00001167$ m/(m/d)$^2$ for $d^{(1)} = 2.9$ mm, and $c_2^{(3)} = 0.0000109$ m/(m/d)$^2$ for $d^{(3)} = 3.0$ mm.

The inflows to individual flocculators were measured, adjusted, and measured again for each individual experiment. They differed by less than 4.5% from the average value. After passing all the flocculators the suspension was delivered by a common inflow to the filtration columns of low friction to flow ensuring the same water levels in all filtration columns at any moment, so the flow distribution to filters was ruled by hydraulics of VDR filters. The total plant capacity was found to vary by less than 2% from the average value. The accuracy of all flow meters was tested and the maximum error was below ±2% for the range of flow rates measured during the experiments. Calculations carried out based on the Navier–Stokes and continuity equations [11] revealed very small influence of wall effect for assumed conditions during laboratory experiments.

### 3.3. EXPERIMENTAL PRINCIPLES AND METHODS

Laboratory and parallel numerical experiments were carried out simultaneously. Results of calculations were used to design orifices and primarily operation parameters, and later the computations were repeated again for the conditions of the experiments, to make comparison of theoretical and laboratory results.

First different ratios $q_1/q_{\mathrm{avr}}$ of the maximum to average flow rates in the plant were assumed, and from them and the assumed value of $Q$, the required flow rate through the clean filter was calculated. Then the procedure developed by Di Bernardo [1] and then simplified by Dąbrowski [6] was used to calculate the parameters such as: $c_2$, $h_0$ for three different values of time. Observation of head loss caused by recirculation of filtrate through porous media filters shows that compressibility of accumulated deposits is an important factor of friction to flow when approach velocities are increased [12, 13]. However when the flow decreases, the decompression of sediment is much slower [12, 13] and consequently the models of VDR filters do not consider rheological properties of flocs trapped in filter pores. Similarly dislodgment of sediment is neglected [14] and in some UBE models of VDR filters reported here [2, 3, 15], the mass balance equation is simplified without visible impact on computing accuracy [16].

The orifices were prepared and the coefficient $c_2$ measured, giving always a little different value than that expected from calculations. This made it necessary to repeat the calculations for $h_0$ using the actual value of $c_2$ and for the same three values of $H$. In this way, the primarily conditions of filter run $c_2$, $H$, $h_0$ were predicted and used in laboratory runs. For each new $c_2$, or $H$ it was necessary to operate the laboratory treatment plant for a period of time long enough to observe the same flow rate pattern after subsequent backwashes in a plant, so a single experiment required from one to two weeks operation before the measurements were taken, depending on $H$ and $q_1/q_{avr}$ – higher values required longer periods before the plant achieved a stable state, this was also dependent upon the coagulant dose.

Due to imperfections of operation of the filter plant, which could be minimised but never totally eliminated, the real parameters of operation $H$, $h_0$ differed somewhat from the computed values and were predicted as an average value from a series of measurements. After that the Di Bernardo [1] set of equations [6] was applied again to make the computations for the actual circumstances of the experiment. Finally the UBE model [9] was used to make similar computations. So far their model has been calibrated only for fine suspensions, with particle sizes between 1 and 10 microns, which does not match the conditions of the laboratory experiments, especially for flocculated particles. This means that the computations done according to this model have to be treated as separate experiments for different circumstances, which serve here for verification of hydraulics relations between filters which were deduced independently on kinetics of deep filtration. The UBE model enables accounting the graduation of filter grains and non-homogeneity of suspension [10]. In the computations done by the UBE model, the same capacity of the plant, the same orifices and the same ratio of maximum to average flow rates were assumed in the data as for the experiments. The computations started from the same head loss $H - h_0$ for Di Bernardo and for the UBE model just after a backwash because both of them include governing flow through a filter at the very beginning of a filter run:

$$H - h_0 = c_1 q_1 + c_2 q_1^n \tag{3}$$

According to the Di Bernardo [1] model, if the backwash starts for the same $H$ the values of $h_0$, should not depend on the kinetics of filter clogging and result from hydraulic interaction of the filters [6, 8] . This was verified by laboratory and numerical results, based on the UBE model. Moreover, the approximated Eq. (1) was verified additionally by the Di Bernardo model.

## 4. RESULTS

### 4.1. VERIFICATION OF DESIGNING METHODS

First the approximated Eq. (1) was validated based on:

- laboratory experimental results,
- accurate solution to the set of equations describing the Di Bernardo [1] model in its shorter form presented by Dąbrowski [6],
- numerical experiments by the UBE model developed by Mackie and Zhao [9] and adapted by Mackie to VDR operation of a plant.

Experimentally determined flow rates through the plant, predicted from Di Bernardo, and from the UBE models, are shown for three different ratios of $q_1/q_{avr}$ in Figs. 4–6. In the Di Bernardo model all flow rates through filters between backwashes are assumed to be constant in time. The situation was somewhat different in both numerical and laboratory experiments, thus the values of flow rates in the middle of the period between subsequent backwashes in the plant were chosen for the discussion of the results presented here. Both the Di Bernardo and UBE approaches produced results close to those observed in the experiments. This confirms that some simplifications made by Di Bernardo [1] in his model were acceptable. The experimental work does not give an opportunity to verify the dynamic of clogging in the UBE model but it proves that it was correctly adapted to VDR operation by accounting for all equations describing cooperation of filters in a bank.

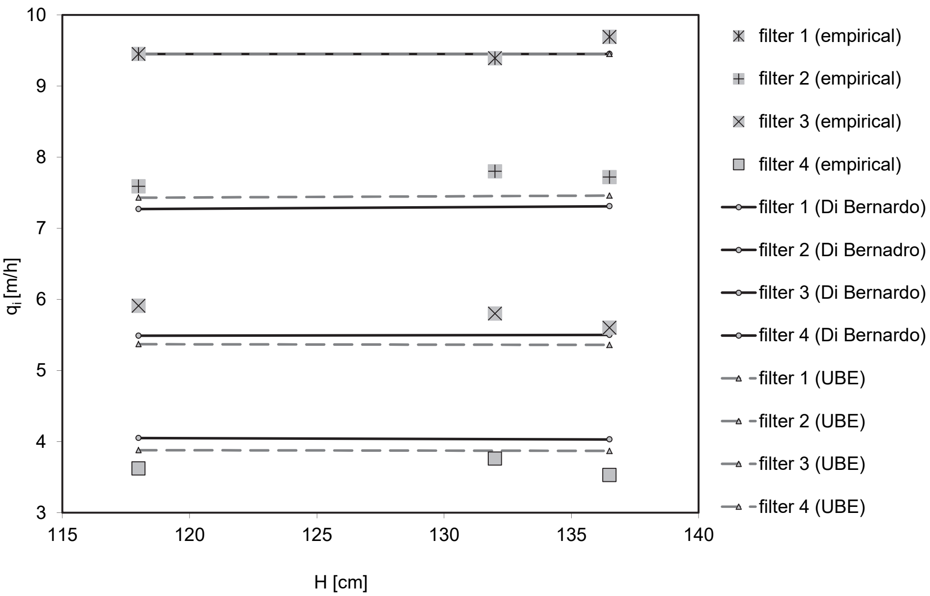


Fig. 4. Hydraulic loads in the middle of a period between backwashes for $q_1/q_{avr} = 1.44$

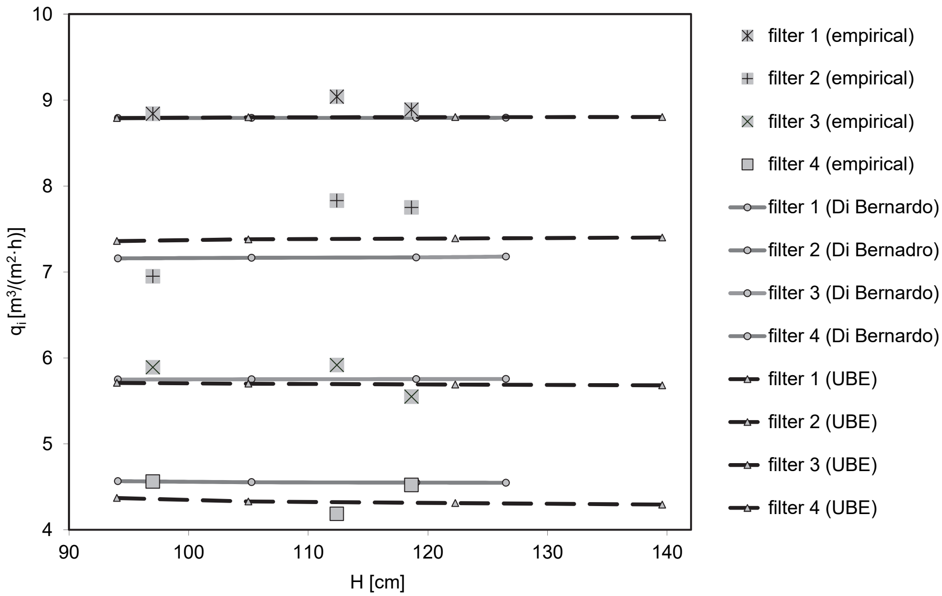


Fig. 5. Hydraulic loads in the middle of a period between backwashes for $q_1/q_{avr} = 1.34$

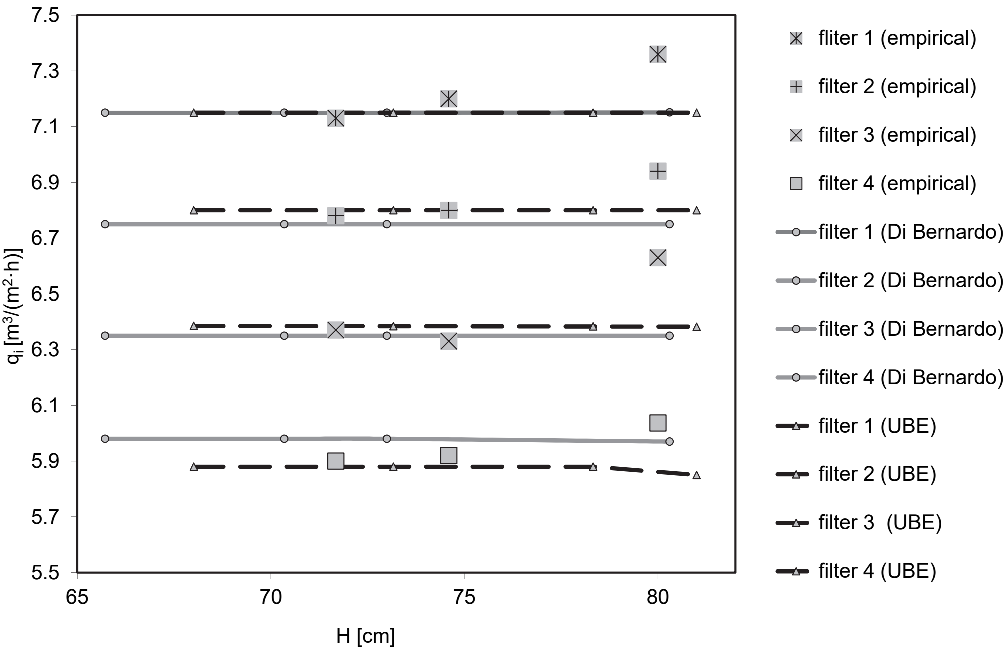


Fig. 6. Hydraulic loads in the middle of a period between backwashes for $q_1/q_{avr} = 1.09$

According to Eq. (1), the distribution of the flow rate between the filters is completely determined by $q_1$ and the ratio $h_0/H$. In laboratory and numerical experiments, $q_1$ was adjusted by the same head loss $H - h_0$ just after the backwash. In the laboratory set-up, this was done by a proper positioning of the level of outflow from the filter plant. However, later both $H$ and $h_0$ were created by numerical models and the laboratory filtration runs.

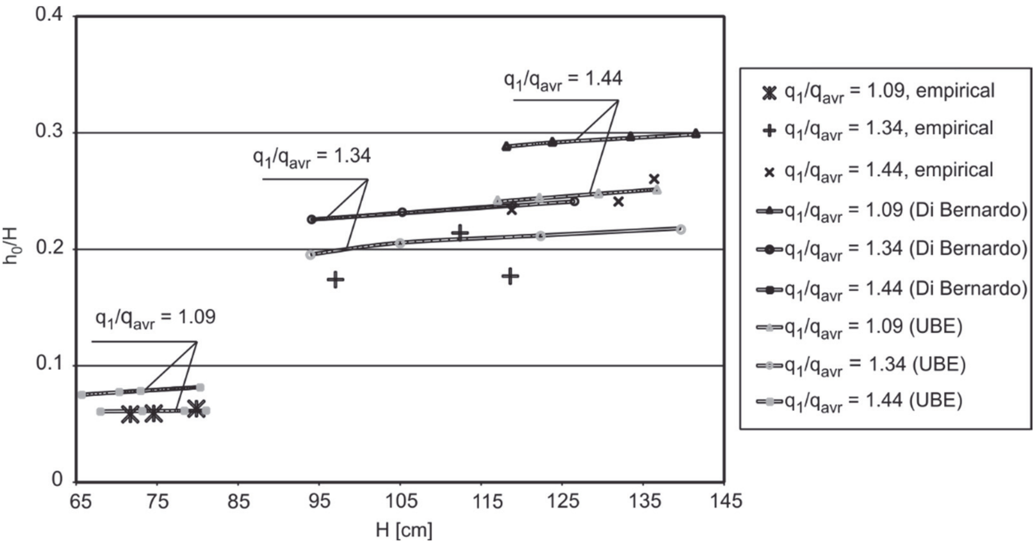


Fig. 7. Ratio of $h_0/H$ versus $H$ resulted in almost identical flow rates from Figs. 4–6

In Figure 7, the ratio of $h_0/H$ was collected for all three orifices predicted in both theoretical models in comparison with the experimental results. It can be seen that experimental points are more close to results of computations done by the UBE model. This may be simply explained by taking into consideration accumulation of water above filters and clogging of filter media during disconnection of one for backwash in the UBE model. The Di Bernardo model does not account for these phenomena. From that it may be concluded that perhaps there is some potential for applying of the UBE model by Mackie and Zhao [9] for improving design of filter plants according to the procedure developed by Di Bernardo, even if the fractions of suspensions are beyond the scope of the model. In all experiments the values of $h_0/H$ predicted from Di Bernardo set of equations were the highest.

### 4.2. TESTING OF OPTIMISING RULES

Based on simplified Eq. (1), it was deduced [6–8] that the highest resistance of the media to flow $c_{1z}$ defined by the equation:

$$H = c_{1z} q_z + c_2 q_z^n \tag{4}$$

is reached by the dirtiest filter $z$ just before its backwash, if simultaneously: the highest head loss $H$ and the highest ratio $q_1/q_{avr}$ were chosen as operation parameters. This has been verified here by laboratory and numerical tests. Measured and computed values of media resistance $c_{1z}$ are presented in Fig. 8, in function of both $H$ and $q_1/q_{avr}$. It is evident that both theoretical models and laboratory results are close to each other. All of them were increasing with increasing $H$, for the same $q_1/q_{avr}$ and vice versa.

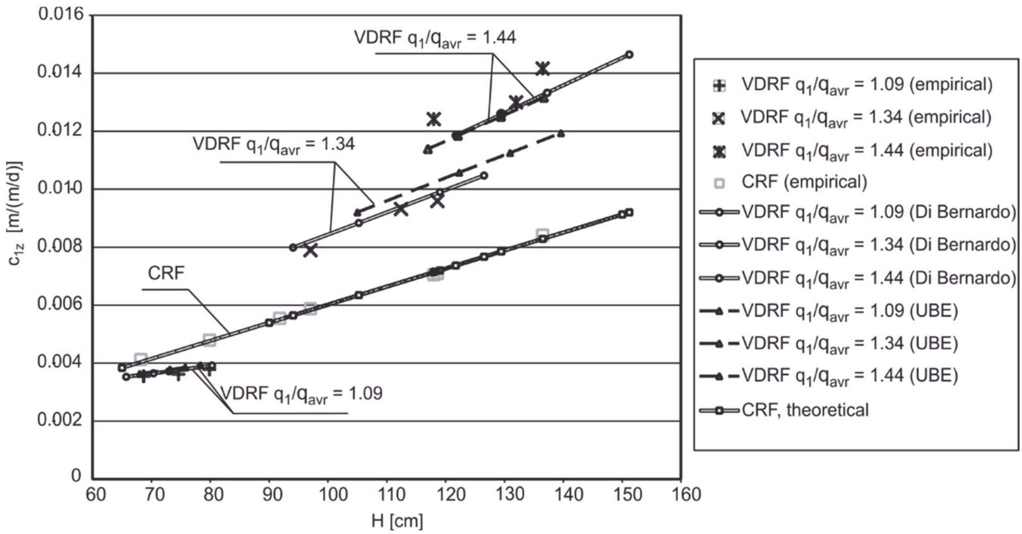


Fig. 8. Most clogged filter media resistance $c_{1z}$ just before its backwash, as a function of $H$ and $q_1/q_{avr}$

Dąbrowski [7] deduced from calculations based on the Di Bernardo model that higher filter media resistances $c_{1z}$, so likely longer filtration runs, were reached in the VDR system if $q_1/q_{avr} > 1.1$, while there was no benefit in applying VDR operation for lower ratio of $q_1/q_{avr}$. This was tested in Fig. 8 by comparing CR and VDR operations for various $H$ and various $q_1/q_{avr}$. For the same $H$, the filter media resistance $c_{1z}$ was always higher for $q_1/q_{avr} = 1.35$ and $q_1/q_{avr} = 1.44$ in the VDR mode but it was opposite to $q_1/q_{avr} = 1.09$.

Dąbrowski [7] assumed that the higher resistance to flow corresponds to longer filter runs, and supported by this assumption he formulated the objective function in his optimisation model describing the resistance of a filter media to flow just before backwash $c_{1z}$, to be maximized. The distribution of deposit in a filter may be significantly different for different filtration velocities, therefore experiments were conducted to test the applicability of Dąbrowski's assumption. The variation of the filter run period in function of the resistance to flow $c_{1z} = (H - c_2 q_z^{\alpha})/q$ is presented in Fig. 9 for the laboratory experiments described here.

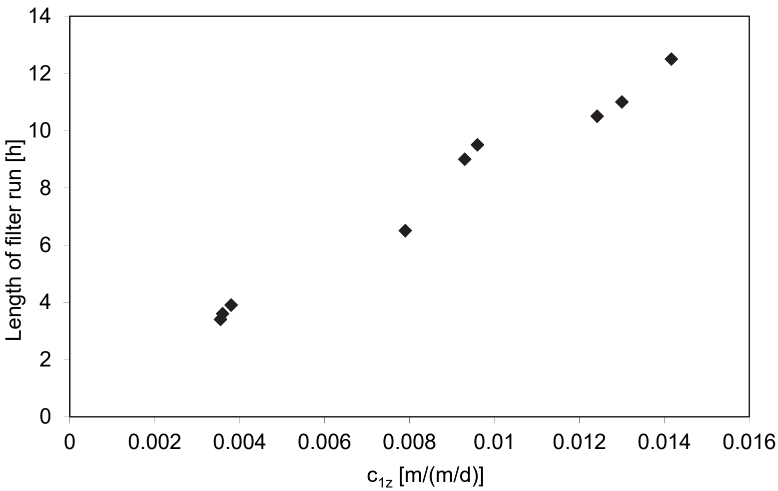


Fig. 9. A functional correlation between filter run periods
and the resistance of filter media just before its backwash

The flow rate $q_z$ through a filter is denoted and $c_2 q_z^{\alpha}$ describes the turbulent head loss in the orifice and transitive head loss in drainage. The laboratory results support the assumption that longer runs refer to higher resistance $c_{1z}$ of the filter media to flow just before a backwash for the conditions of the experiments.

## 5. FULL SCALE OPERATION

The rule of filter plant operation with simultaneously highest possible $q_1/q_{\mathrm{avr}}$ and $H$ was applied to a small treatment plant in Żywiec (South of Poland) in the early eighties of the previous century. The plant is supplied with raw surface water collected from a mountain river of low alkalinity, relatively low COD, $BOD_5$, and turbidity well below 5 NTU during dry weather flows. Due to a high quality of the raw water, it was coagulated only occasionally during the reported time of observations. The plant consisted of five identical coarse sand filters of a relatively low filter media depth. The product of mean grain diameter (by weight) and the filter media depth $h$ (both expressed in millimetres) was only slightly above 600, instead of 1200 which is recommended by more conservative US literature. Unfortunately, it was possible neither to replace the media by smaller grains, nor to increase the depth of the filter layers [17]. These two required improvements could not be applied because of a low elevation of the sedimentation tanks which had to be used during rainy weather. The following shortcomings of the previous plant management were recognized:

- Filters were operated with a constant flow rate without flow rate controllers.
- Each hour the filtration velocity was adjusted manually by valves, causing poor filtrate quality.
- Water levels above the filters were kept unnecessarily low, no matter whether the low through sedimentation tanks was required or not.
- A significant usage of backwash water was reported at the plant.

To improve the plant performance without substantial investments the following actions were taken:

- The filters were switched into the VDR system. The total plant capacity was adjusted by a butterfly valve installed especially for this purpose at the main collecting filtrate from all filtration units. The water distribution among the filters was realized by a properly chosen frequency of backwashing.
- Direct filtration was applied to make the operation of sedimentation tanks unnecessary during dry weather and thus enabled operation with a higher water level above the filters.
- The highest acceptable ratio of $q_1/q_{\text{avr}}$ was determined experimentally as equal to 1.5 which still gave an acceptable filtrate quality for the maximum plant capacity. This value of $q_1/q_{\text{avr}}$ was not exceeded during the VDR operation of the plant.
- The highest possible water table level above the filters (so also the highest $H$) was maintained during the plant operation.

The latter two points summarise briefly the rule of backwash water loss control. The changes of the plant management were made in the middle of September 1992. Results obtained are reported from the period September 1992 until the middle of May 1993. In May 1993, a reconstruction of the drainage system in one of the filters changed the hydraulics of the plant. The benefits resulting from the VDR operation remained essentially the same after the reconstruction of the drainage but they are not reported in Fig. 10. This figure clearly demonstrates that a substantial saving of backwash water arose from the properly managed VDR operation of the filter plant.

The evaluation of the filter plant performance during the period September 1992 –May 1993 is based both on the filtrate quality and on backwash water losses. Both raw water and filtrate turbidity values were measured starting in July 1992 (two months before the CRF was replaced by VDR filtration), so the previous measurements were perhaps less reliable. The plant efficiency, as described by the ratio of the filtrate turbidity to the turbidity of raw water, slightly improved in the VDR system. Perhaps this is not so much from the decrease in flow rates through clogged filters, but more likely because of smoother changes in filtration velocities in comparison with previous manual operation of valves to control flow rates through filters in the constant flow rate system. It is well known that chemical pre-treatment is of crucial importance to the filter performance, so it is not surprising that when the water was not coagulated poor quality filtration resulted as shown by high ratio of filtrate turbidity to turbidity of raw water. The next reasons of the poor filtrate quality were coarse filter media, low alkalinity of treated water and contamination by the highly dispersed particulates.

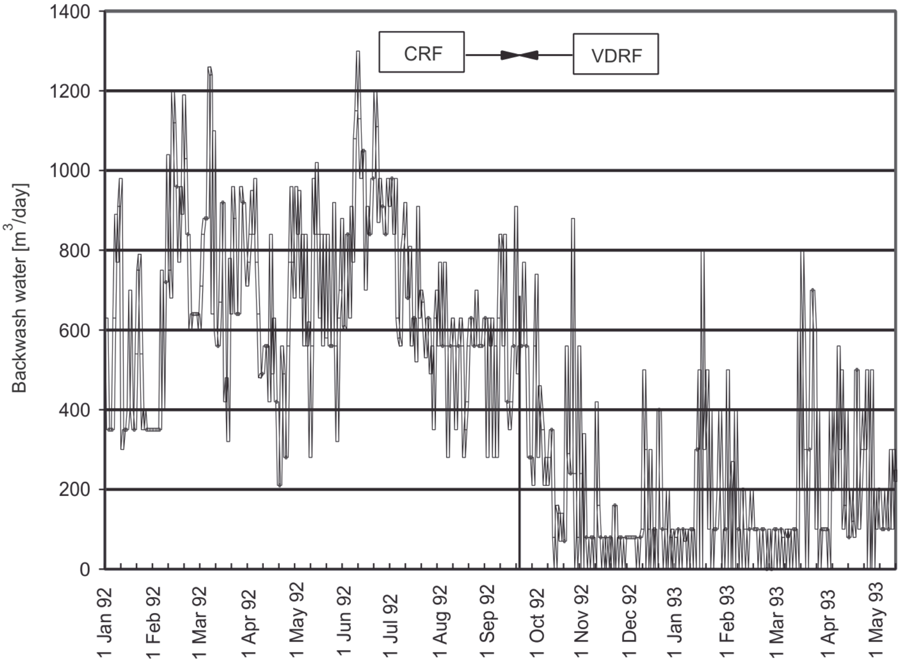


Fig. 10. Improvement in backwash water usage after changing CR
to VDR operation and applying the optimization approach

Backwash water performance strongly depends on its temperature. Because of this, the number of backwashes is used instead of total amount of backwash water in the plant to characterise the improved performance of VDR Filtration. Figure 10 shows both optimised and non-optimised VDR filtration (optimised VDR filtration means operation with both the highest possible water level above the filters and the highest acceptable ratio of $q_1/q_{avr}$ ). This period of optimised operation covered the last three months (October, November, and December) 1992. After the beginning of January 1993, the plant operated without strict adherence to the optimisation rules, following the same ratio of $q_1/q_{avr}$ but rather with lower and not well controlled water levels above the filters. Some differences between the number of backwashes in these two intervals of time are slightly visible in Fig. 10 however it is necessary to account for variable raw water turbidity, flow rates, and temperature. From the beginning of January 1993, the turbidity of raw water was so low that turbidity of the filtrate was sometimes about 0.1 NTU and the accuracy of the measurements was recognised to be not reliable enough to include all the results. In contrast, there were some small storm events during the first optimised period of VDR filtration. To take into consideration substantially different raw water quality, the operation of optimised and non-optimised VDR filtration was described by an inverse efficiency parameter $E$ defined as:

$$E = \frac{N}{V^2 C v} \tag{5}$$

where: $N$ is the number of backwashes per day, $V$ – daily water production (m/day), $C$ – turbidity of raw water expressed in NTU, $v$ – kinematics viscosity of water ($m^2/s$).

The coefficient $E$ was taken to be approximately proportional to the number of backwashes and approximately inversely proportional to the volume of suspended solids entering the filter media, to the filtration velocity and to water suspension viscosity. The volume of particles in water suspension inflowing to filters was very roughly assumed to be proportional to water turbidity [15, 18] and daily water production $V$. As the filtration velocity is really proportional to $V$, this parameter was raised to the power two in the denominator of Eq. (5). It is recognised that $E$ is a rather crude measure of backwash efficiency.

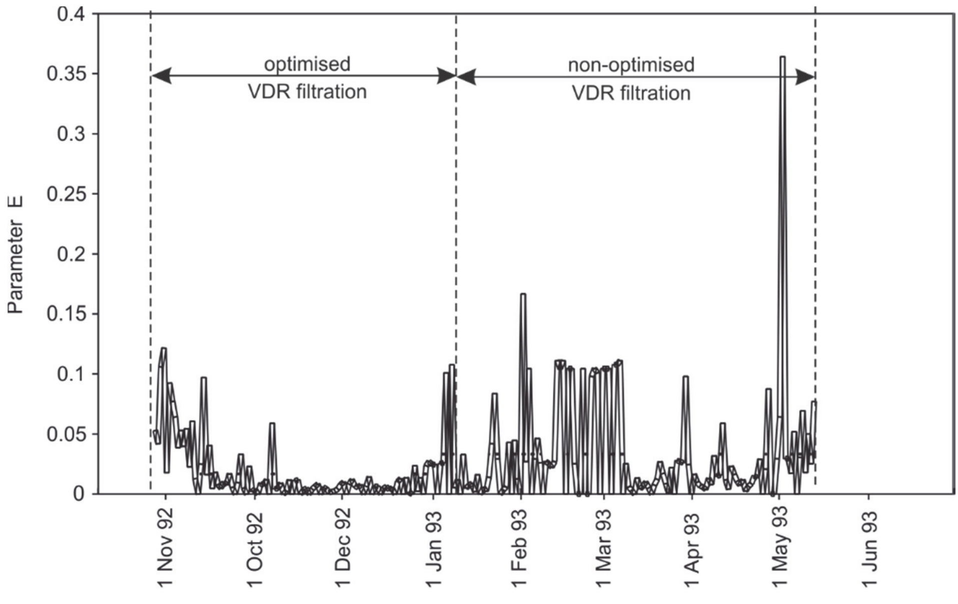


Fig. 11. Parameter $E$ characterizing inversely efficiency of backwash water usage for optimised and non-optimised VDR filter plant operation in Żywiec

As shown in Fig. 11, much lower values of the inverse efficiency coefficient $E$ demonstrate significant improvement in the plant performance during the first optimised period of the VDR filtration. An average value of the coefficient $E$ was equal to 0.017 for the optimised VDR filter plant operation and equal to 0.033 for the non-optimised one.

## 6. CONCLUSIONS

Equation (1), and its important consequences for proper VDR plant management were verified based on a simplified version of the Di Bernardo model, on a more complex UBE model, and, finally, on laboratory experiments. The tests were conducted for significantly different circumstances. The Di Bernardo model gave results representative for any kinetics of clogging but it included several significant simplifications. The UBE model was representative for fine suspension of the diameters from 0.5 μm to 10 μm. Coagulated and non-coagulated suspension of variety of fractions was used in laboratory tests. In spite of different conditions, the results confirmed the theoretical considerations presented in the previous papers [6–8]. The following conclusions were drawn from the tests.

- For the four filter laboratory scale plant tested, the rule of plant management based on simultaneously adjusting $q_1$ and the same ratio of $h_0/H$ was proved to result in the same flow rate distribution both in numerical and in experimental tests.
- For a definite ratio $q_1/q_{\text{avr}}$, the resistance of mostly clogged filter media before a backwash $c_{1z}$ resulted in higher values of the head loss $H$ of flow through the plant before a subsequent backwash. Similarly for the same $H$, higher $c_{1z}$ were received for higher $q_1/q_{\text{avr.}}$ This confirms the operation rule according to which the highest resistance of the filter media $c_{1z}$ (and likely the longest filtration run) is reached for simultaneously both the highest $H$ and the highest ratio of $q_1/q_{\text{avr}}$.
- For $q_1/q_{\text{avr}} = 1.09$ and for all investigated values of $H$, the filter resistance $c_{1z}$ was lower under VDR than under CR operation control system, while for $q_1/q_{\text{avr}} = 1.35$ and 1.44 the opposite was the case. This agrees with the theory presented in the previous papers [6–8], according to which there was no advantage of using VDR systems for $q_1/q_{\text{avr}} < 1.1$.
- In laboratory experiments carried out in a VDR filter plant higher resistances $c_{1z}$ of filter media to flow just before a backwash referred always to longer filtration runs.
- The UBE model used in the computations was adapted to the VDR operation system, and gave better approximation to the measured ratio of $h_0/H$ shown in Fig. 7 than the model by Di Bernardo, in spite that the fractions of suspension used in experiments were beyond the value the model was constructed for. This can be explained by the fact that the UBE model accounted for accumulation of water above filter media and for considering clogging of all servicing filters during the period when one of them is backwashing.
- Finally an example of applying the rules tested here was presented at full scale.

## REFERENCES

[1] DI BERNARDO L., *Designing declining-rate filters*, Filtr. Separat., 1987, 24 (5), 338.
[2] MACKIE R.I., DABROWSKI W., ZIELINA M., *Numerical study of a rational rule for the operation of variable declining rate filters in response to changes in raw water quality*, Environ. Prot. Eng., 2003, 29 (1), 45.
[3] ZIELINA M., DĄBROWSKI W., *Impact of raw water quality on operation of variable declining rate filter plants*, Environ. Prot. Eng., 2011, 37 (2), 133.
[4] HILMOE D.J., CLEASBY J.L., *Comparing constant-rate and declining-rate direct filtration of a surface water*, JAWWA, 1986, 78 (12), 26.
[5] CLEASBY J.L., DI BERNARDO L., *Hydraulics considerations in declining rate filtration*, J. Environ. Eng. Div., 1980, ASCE 106 (EE6), 1043.
[6] DĄBROWSKI W., *The progression of flow rates in variable declining rate filter systems*, Acta Hydroch. Hydrob., 2006, 34 (5), 442.
[7] DĄBROWSKI W., *Investigations into variable declining rate filters*, Cracow University of Technology, Cracow 1994.
[8] DĄBROWSKI W., *Rational operation of variable declining rate filters*, Environ. Prot. Eng., 2011, 37 (4), 35.
[9] MACKIE R.I., ZHAO Q., *A framework for modelling removal in the filtration of polydisperse suspension*, Water Res., 1999, 33 (3), 794.
[10] DĄBROWSKI W., MACKIE R.I., ZIELINA M., *Filtrate quality from different filter operations*, Acta Hydroch. Hydrob., 2003, 21 (1), 25.
[11] LIU S., MASLIYJAH J.H., *Single fluid flow in porous media*, Chem. Eng. Commun., 1996, 148 (1), 653.
[12] DĄBROWSKI W., *Hydraulic properties of deposit in coarse sand*, Arch. Hydroeng., 1993 (1/2), 135.
[13] DĄBROWSKI W., *Observations of deposit compressibility impact on head losses of flow through granular filter media*, Proceedings of the 2nd European Conference on Filtration and Separation, 12–13 October 2006, Compiegne, France, 191.
[14] DĄBROWSKI W., SPACZYŃSKA M., *Empirical investigations into the impact of deposit dislodgment on modeling of deep bed filtration*, Proceedings of the 2nd European Conference on Filtration and Separation, 12–13 October 2006, Compiegne, France, 104.
[15] ZIELINA M., *Experimental research into depth filtration of polydispersed suspension*, Environ. Prot. Eng., 2007, 33 (2), 249.
[16] DĄBROWSKI W., *Consequences of the mass balance simplification in modelling deep bed filtration*, Water Res., 1988, 22 (10), 1219.
[17] KŁOS M., TOKARCZYK J., *Modernization of coagulation and filtration systems in surface water treatment plants*, Ochr. Środ., 2005, 27 (3), 61.
[18] ZIELINA M., HEJDUK L., *Measurement of the efficiency of depth filters for water treatment*, Filtration, 2007, 7 (3), 225.