Practical Paper

Development of a model-based control system for membrane filtration process

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ABSTRACT

Reduction of operational cost and certification of safety for treated water are required for the membrane filtration process for drinking water. Therefore a model-based control system is developed to reduce operational cost. A process model is used for the model-based control system that calculates the transmembrane pressure (TMP) in the future. This allows the model-based control system to optimize the coagulant dosage for the pretreatment process and the backwash interval for the membrane filtration process automatically. The cost reduction effect is verified using a pilot-scale plant.

Key words | coagulation, control, membrane filtration process, process model

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INTRODUCTION

The number of water purification plants using a membrane filtration process is increasing. At the end of 2007, there were 623 plants in Japanese municipalities which used the membrane filtration process. For purification plants larger than an intermediate class, surface water is sometimes used as raw water to be treated. In such cases, a pretreatment process is usually required in order to reduce membrane fouling. In the conventional membrane filtration process, the operational conditions of the pretreatment process and filtration process have been decided empirically. For example, the terms of the filtration process and backwash process are controlled by an electrical timer in the control board and the coagulation dosage is set as proportional to the turbidity of the raw water. These control techniques may be sufficient when the raw water is relatively clean (lacking large amounts of foreign substances), but they may be insufficient for plants that treat surface water with rather large fluctuations in water quality because of weather conditions such as rainfall. On the other hand, the requirement for the reduction of operational costs is increasing, and the conventional control technique has not always been able to do this.

If there is a simulation system by which the fluctuation of transmembrane pressure (TMP) can be predicted, the operational cost could be evaluated for various operational conditions. By using this simulation system, the most desirable operational conditions could be obtained. In order to predict the TMP, a process model is needed to explain the accumulation and detachment of foulants in the filtration process and backwash process.

It has been reported that the increase of TMP in the filtration process can be explained by a blocking model

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Table 1 | Assumed equations of process model

Model	Formula	Eq. no.
Standard blocking model	$\Delta P = \frac{\mu \times J \times C_0}{(1-(V_{\mathrm{ir}}+V_{\mathrm{r}}))^2}$	(1)
Accumulation of difficult to detach foulants	$\frac{\mathrm{d}V_{\mathrm{ir}}}{\mathrm{d}t} = (C_1 \times \mathrm{Tu} + C_2 \times \mathrm{UV}_{260}) \times \exp(-C_3 \times \mathrm{PACl}) \times \exp(-C_4 \times \mathrm{temp}) \times (1.0 + C_5 \times \mathrm{cycle}) \times J$	(2)
Accumulation of easy to detach foulants	$\frac{\mathrm{d}V_{\mathrm{r}}}{\mathrm{d}t} = (C_6 \times \mathrm{Tu} + C_7 \times \mathrm{UV}_{260}) \times \exp(-C_8 \times \mathrm{PACl}) \times J$	(3)
Removal of easy to detach foulants	$\frac{\mathrm{d}V_{\mathrm{r}}}{\mathrm{d}t} = -C_9 \times V_{\mathrm{r}} \times Q_{\mathrm{b}} \times (1.0 + C_{10} \times \mathrm{temp}) \times (1.0 + C_{11} \times \mathrm{PACl}) \times (1.0 - C_{12} \times \mathrm{cycle})$	(4)

V_{ir}: index of difficult to detach foulants [-]; Tu: turbidity [NTU]; UV_{260}: ultraviolet absorbance [cm^{-1}]; PACl: PACl dose [$\mathrm{mg\,L}^{-1}$]; temp: water temperature [°C]; J: filtration flux [$\mathrm{m\,s}^{-1}$]; V_{r}: index of easy to detach foulants [-]; Q_{b}: flow rate for backwash [$\mathrm{m}^3\,\mathrm{s}^{-1}$]; Cycle: filtration process time [min]; ΔP: transmembrane pressure [Pa]; μ: viscosity of water [Pa s]; C_0: constant; $C_1 - C_{12}$: coefficients.

(Tambo 1994; Nishijima et al. 1998; Mingegishi et al. 2000; Kosvintsev et al. 2002; Susanto & Ulbricht 2008) or a cake filtration model (Fujita & Takizawa 1995; Bian et al. 2001; Lee et al. 2004; Yamamura et al. 2007). Both models include variables such as TMP and filtration flux, and the coefficients whose values differ depending on the raw water qualities. However, there are no reports in which the effect of each water quality item is shown as a mathematical expression. Also, there are no reports in which the effect of coagulant dosage on the reduction of membrane foulants is shown as a mathematical expression although there are some experimental results when the coagulation process is placed as the pretreatment for the membrane filtration process.

In this study, a process model is developed in which the accumulation and detachment amounts of membrane foulants can be calculated by the concentration of each raw water quality item and operational conditions. When the calculation procedure is established by which the operational cost can be calculated from the calculated TMP given by the process model, the operational cost can be obtained in advance from raw water qualities and operational conditions. In this study, coagulant dosage and backwash interval are selected as the items to be optimized.

MATERIAL AND METHODS

Process model for prediction of TMP

As a prerequisite for use in an automated control system, the main requirements for the process model are as follows.

- All input items should be measured automatically
- The number of coefficients (adjustable parameters) should be as small as possible
- Variation of TMP with time in the future should be calculable
- Both the filtration process and the backwash process should be provided in the model

To satisfy these requirements, equations shown in Table 1 were assumed. The foulants were classified into 'difficult to detach foulants' and 'easy to detach foulants'. Both difficult and easy to detach foulants were expressed by a standard blocking model. The indices of foulants in each of the equations were modelled as functions composed of raw water quality items. These functions express the relationship between increase and decrease of a property

Table 2 | Calibrated values of coefficients

C_1	3.3×10^{-8}
C_2	5.8×10^{-7}
C_3	3.2×10^{-2}
C_4	5.4×10^{-1}
C_5	1.3×10^{-6}
C_6	1.7×10^{-10}
C_7	2.1×10^{-6}
C_8	5.7×10^{-15}
C_9	8.0×10^{-1}
C_{10}	2.8×10^{-2}
C_{11}	7.4×10^{-2}
C_{12}	4.1×10^{-3}

simply based on experiments carried out previously. The values of the coefficients in these equations were calibrated by actual measurements of a pilot-scale facility. The calibrated coefficients are shown in Table 2. The calculated values of TMP showed good agreement with measurements and the mean error was 5.1–6.7 kPa.

Procedure for calculating operational cost

The formulae for computation of each evaluation index are shown in Table 3. The formulae for computation of each physical value are shown in Table 4. The TMP predicted by the process model shown above was used for the calculation of power consumption of pumps and the calculation of the interval of chemical cleaning. The chemical cleaning cost was calculated as one chemical cleaning unit cost that resulted when the TMP exceeded a set value. The membrane module replacement cost was calculated as one membrane replacement unit cost that resulted when pure water flux, which decreased at each chemical cleaning process, fell below a set value. Sometimes the membrane module is replaced regularly regardless of decline of its performance. Therefore the case in which the membrane module replacement cost was involved in the evaluation index and the case in which it was not involved were considered.

Table 3 | Equations for computing evaluation indices

	Eq. no.
Evaluation indices	
Operational cost [¥ m^{-3}] = power cost [¥ m^{-3}] + chemical cost [¥ m^{-3}] + sludge disposal cost [¥ m^{-3}] + chemical cleaning cost [¥ m^{-3}] + membrane replacement cost [¥ m^{-3}]	(5)
Power cost	
Power cost [¥ m^{-3}] = (power cost by filtration pump [¥] + power cost by backwash pump [¥] + power cost by dewatering equipment [¥])/effective quantity of water [m^3]	(6)
Power cost by filtration pump [¥] = unit price of power [¥ kWh^{-1}] × power consumption by filtration pump [kWh]	(7)
Power cost by backwash pump [¥] = unit price of power [¥ kWh^{-1}] × power consumption by backwash pump [kWh]	(8)
Power cost by dewatering equipment [¥] = unit price of power [¥ ton^{-1}] × (sludge quantity from suspended solids [ton] + sludge quantity from organic substances [ton] + sludge quantity from coagulant [ton]) × exp(−0.03 × standard value of coagulant dosage [mg L^{-1}])/exp(−0.03 × coagulant dosage [mg L^{-1}])	(9)
Chemical cost	
Chemical cost [¥ m^{-3}] = (cost for coagulant consumption [¥] + cost for NaClO consumption [¥])/effective quantity of water [m^3]	(10)
Cost for coagulant consumption [¥] = unit price of coagulant [¥ kg^{-1}] × dosage of coagulant [mg L^{-1}] × total quantity of filtrated water [m^3] × 10^{-3}	(11)
Cost for NaClO consumption [¥] = unit price of NaClO [¥ kg^{-1}] × dosage of NaClO [mg L^{-1}] × total quantity of backwash water [m^3] × 10^{-3}	(12)
Sludge disposal cost	
Sludge disposal cost [¥ m^{-3}] = (sludge quantity from suspended solids [ton] + sludge quantity from organic substances [ton] + sludge quantity from coagulant [ton]) × unit price for sludge disposal [¥ ton^{-1}]/effective quantity of water [m^3]	(13)
Chemical cleaning cost	
Chemical cleaning cost [¥ m^{-3}] = unit price for chemical cleaning [¥ $times^{-1}$ $Qty.^{-1}$] × times for chemical cleaning [times] × number of membrane modules [Qty.]/effective quantity of water [m^3]	(14)
Times for chemical cleaning [times] = calculation period [d]/predicted interval of chemical cleaning [d $times^{-1}$]	(15)
Membrane replacement cost	
Membrane replacement cost [¥ m^{-3}] = unit price of membrane replacement [¥ $times^{-1}$ $Qty.^{-1}$] × times for membrane replacement [times] × number of membrane modules [Qty.]/effective quantity of water [m^3]	(16)
Interval of membrane replacement [d $times^{-1}$] = interval of chemical cleaning [d $times^{-1}$] × log (decreasing rate of filtration performance [%]/100)/log (recovery rate by chemical cleaning [%]/100)	(17)

Table 4 | Equations for computing physical values

	Eq. no.
Effective quantity of water [m^3] = total quantity of filtrated water [m^3] − total quantity of backwash water [m^3]	(18)
Power consumption by filtration pump [kWh] = (0.163 × (filtration flux [m d^{-1}] membrane area [m^2]/24/60) × (mean TMP [kPa] + water head at the outlet [kPa] + pipe resistance [kPa])/9.81)/pump efficiency [-]) × (filtration process time[s]/3,600)	(19)
Power consumption by backwash pump [kWh] = (0.163 × flow rate for backwashing [m^3 min^{-1}] × (mean TMP [kPa]/9.81) × (backwash flux [m d^{-1}]/filtration flux [m d^{-1}])/pump efficiency [-]) × (backwash times [s]/3,600)	(20)
Power consumption by dewatering equipment [kWh] = power cost by dewatering equipment [¥]/unit price of power [¥ kWh^{-1}]	(21)
Sludge quantity from suspended solids [ton] = total quantity of filtrated water [m^3] × suspended solids in raw water [mg L^{-1}] × 10^{-6}	(22)
Sludge quantity from organic substances [ton] = total quantity of filtrated water [m^3] × (100/58) × TOC in raw water [mg L^{-1}] × 0.7(1.0 − exp(− 0.35 × dosage of coagulant [mg L^{-1} as AL])) × 10^{-6}	(23)
Sludge quantity from coagulant [ton] = total quantity of filtrated water [m^3] × dosage of coagulant [mg L^{-1}] × 10^{-6}	(24)
Mean TMP [kPa] = (TMP just after backwash [kPa] + TMP just before backwash [kPa])/2.0	(25)
Pipe resistance [kPa] = flow coefficient [-] × (10.666 × (number of membrane modules [Qty.] × membrane area [m^2] × filtration flux [m d^{-1}]/24/3,600)$^{1.85}$) × length of pipe [m]/(100$^{1.85}$ × pipe diameter [m]$^{4.87}$)	(26)

Effect of model-based control system

The process model and the calculation procedure for operational cost were implemented in control software with an optimization algorithm, and the verification experiments were carried out using the facility illustrated in Figure 1. The raw water was surface water from the Kuji River in Ibaraki Prefecture, Japan. It was first treated in the Mn removal column, then polyaluminium chloride (PACl) was added with rapid mixing. After that, it was filtered by the membrane. For the control evaluation, two systems were used which could treat the water at the same quality level. The external pressure type hollow fibre membrane modules were used. The membrane material was polyvinylidene difluoride (PVDF). The filtration area was 23 m^2 per module, and the nominal pore size was 0.1 μm. The turbidity was measured with a turbidity meter TR-502 (Kasahara Chemical Instruments). The kaolin turbidity standard was used so the unit for turbidity in this study was mg l^{-1}. The index of organic substances, UV$_{260}$, was measured by an ultraviolet absorptiometer DIAMON with a 1 cm quartz cell after removal of suspended solids by filtration through a paper filter of 1 μm pore size.

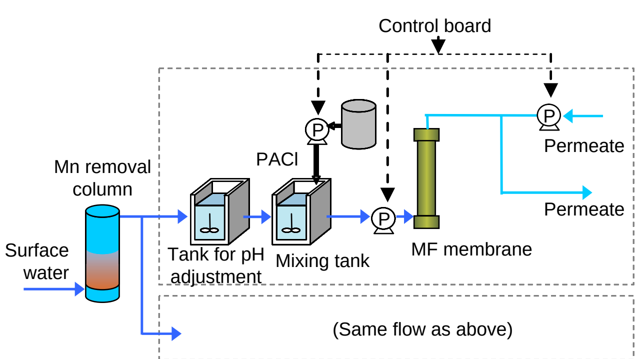

Figure 1 | Process flow of experimental facility.

One system was controlled by the model-based control method while the other was controlled by the conventional control method. This conventional control method was composed of coagulants dosage control, in which the dosage was proportional to turbidity of the raw water, and filtration cycle control, in which the filtration process time was set to a constant value (30 or 45 min).

Operational conditions

In each system, the generated water amount (filtered water amount minus water amount for backwash) was set to 2.5 m^3 h^{-1} and 1.25 m^3 h^{-1} and the recovery rate was set to 95%. The lower limit of filtration time was set to 20 min while the upper limit was set to 120 min. The lower limit of PACl dosage was set to 5 mg l^{-1} and the upper limit

Table 5 | Conditions for each run

Run No.	Generated water amount	Evaluation indices (in Table 3)	Upper limit of PACl dosage
Run-1		①②③④⑤	None
Run-2	2.5 $m^3 h^{-1}$		Half of the actual plant (rapid filtration process)
Run-3		①②③④	
Run-4	1.25 $m^3 h^{-1}$		None

to 60 mg l^{-1}. To evaluate the effect of restriction of PACl consumption, another case was also considered in which the upper limit of PACl was set to half that of an actual purification plant which treats water by a rapid sand filtration method.

In the verification, four kinds of experiment shown in Table 5 were carried out. When the TMP of either or both of the model-based control and the conventional control exceeded 150 kPa, operation of the experimental facility was suspended and chemical cleaning was carried out. Then, the next run was started. The operational cost was estimated from the data obtained until the chemical cleaning was done. The effect of model-based control was evaluated by comparing the results of the operational cost for the two controls.

RESULTS AND DISCUSSION

The transitions of raw water qualities are shown in Figure 2. The lowest value of the turbidity of the raw water was 2 mg l^{-1}, while the highest value was over 100 mg l^{-1}. The mean value of the turbidity of the raw water was 14.4 mg l^{-1}. The concentration of dissolved Mn was below 0.007 mg l^{-1}, and the UV_{260} was below 0.30 cm^{-1}.

The transitions of operational conditions and TMP are shown in Figure 3. In Run-1, the PACl dosage of

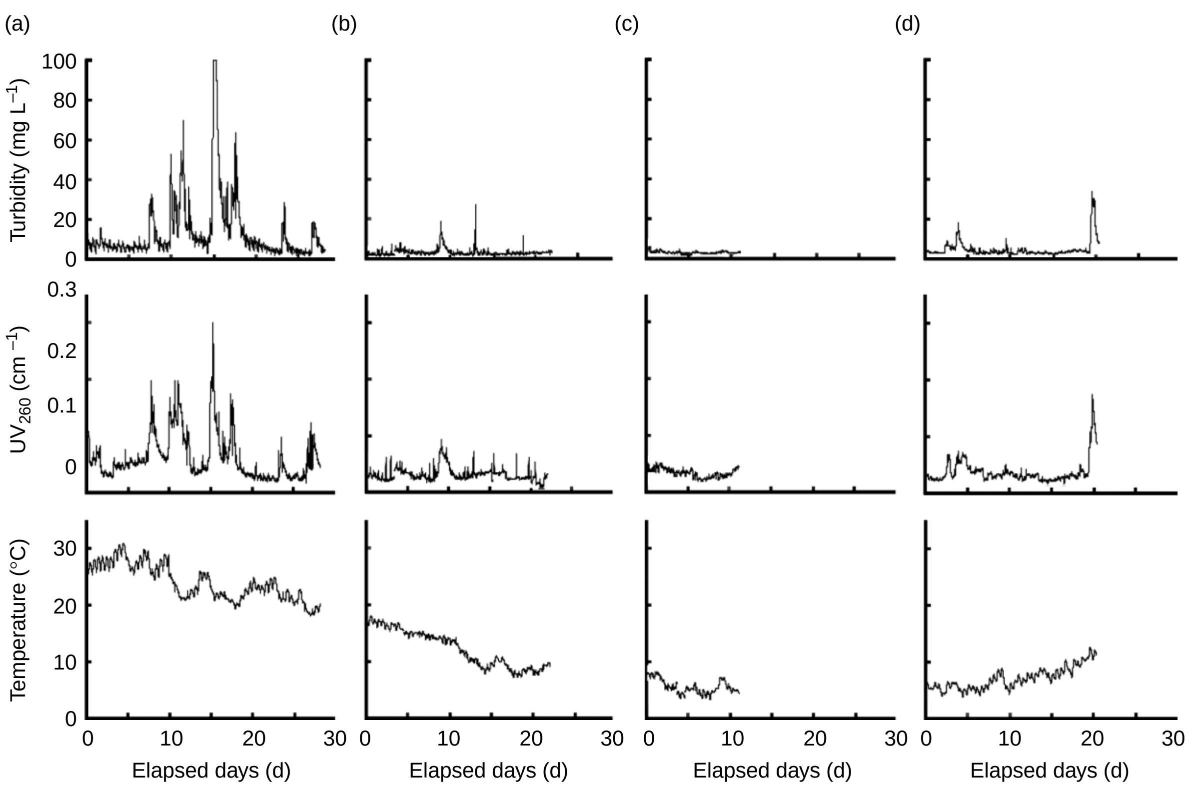

Figure 2 | Transitions of raw water qualities: (a) Run-1; (b) Run-2; (c) Run-3; (d) Run-4.

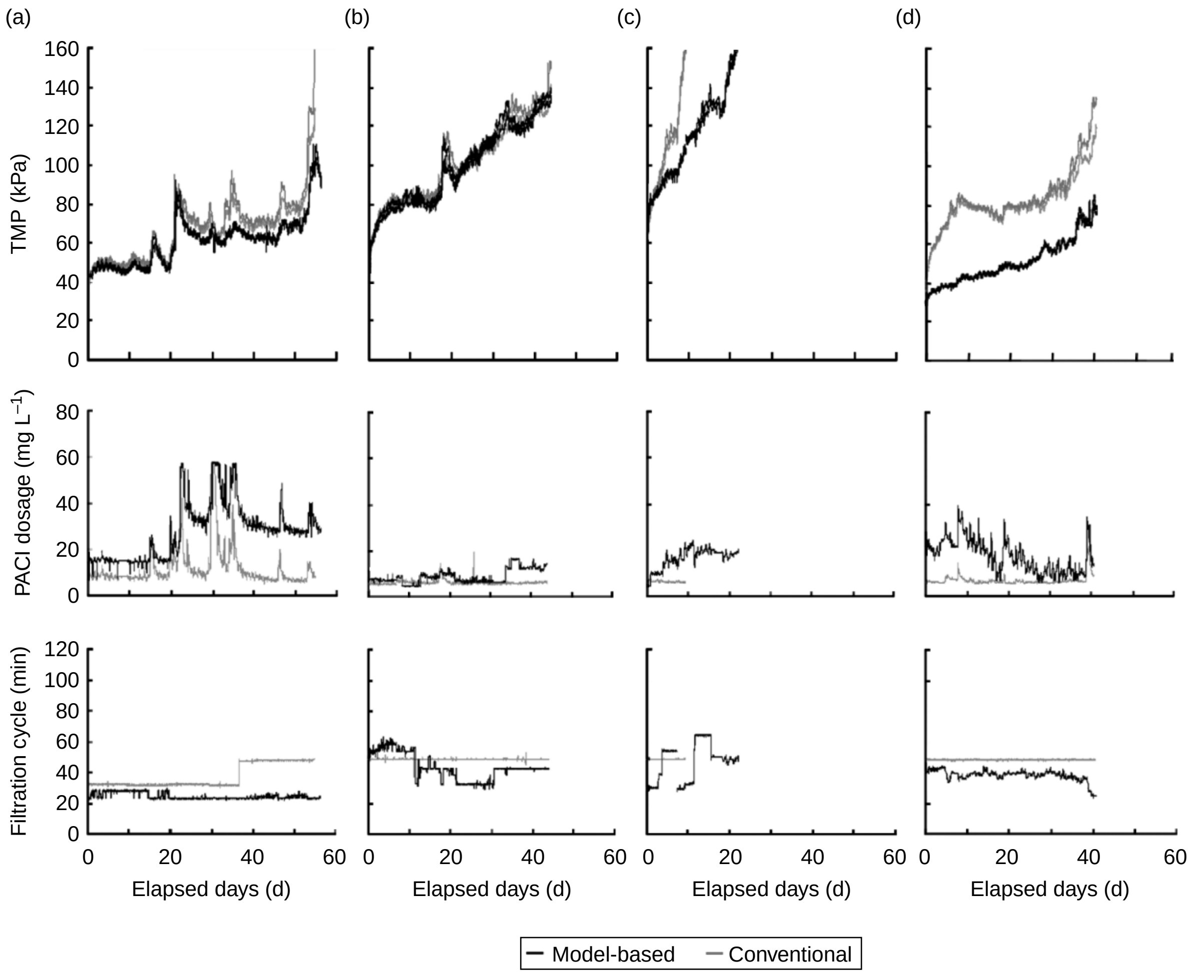

Figure 3 | Transitions of TMP and operational conditions: (a) Run-1; (b) Run-2; (c) Run-3; (d) Run-4.

model-based control fluctuated between two and three times higher than that of conventional control. The filtration process time of model-based control was about 25 min, which was shorter than that of conventional control. For the conventional control, the filtration process time was changed from 30 min to 45 min on the 37th day. In Run-1, TMP of the model-based control and the conventional control were almost the same for the first 23 days, but from 24 days the deviation increased. On the 55th day, the TMP of the conventional control was over 150 kPa, and the operation was changed to the chemical cleaning mode.

In Run-2, the PACl dosage of model-based control was lower than that of Run-1, because the upper limit of PACl dosage was set. On the 44th day, the TMP of the conventional control was over 150 kPa, and the operation was changed to the chemical cleaning mode.

In Run-3 and Run-4, the upper limit of PACl dosage was released. As a result, the PACl dosage of model-based control was about 2–3 times higher than that of the conventional control. The rate of increase of TMP was much higher than in Run-2, and between the 8th and 9th days, the TMP exceeded 150 kPa. Run-3 was continued until the TMP of the model-based control also exceeded 150 kPa. Finally, on the 20th day, the operation was changed to the chemical cleaning mode.

In Run-4, the filtration flux was decreased. The same as in Run-3, the PACl dosage of model-based control was higher and the filtration process time was shorter than the conventional control. Although the TMP of

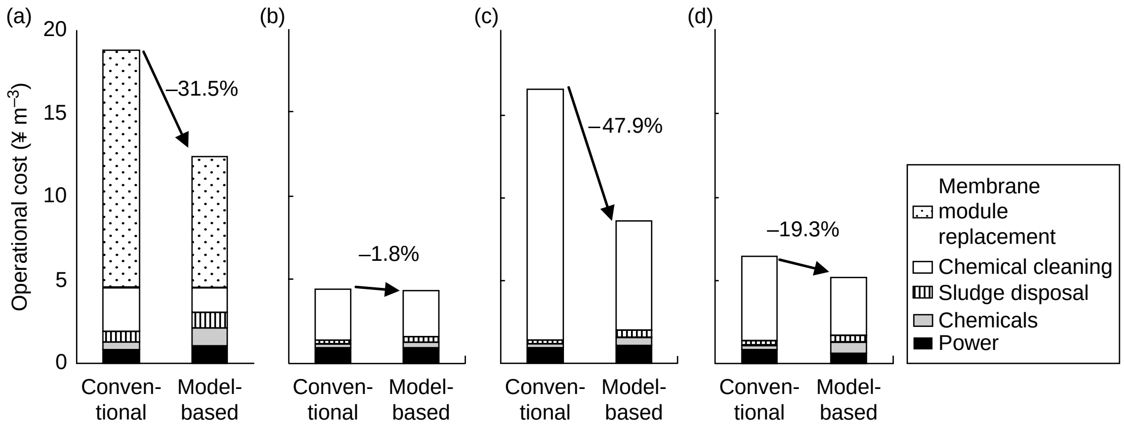

Figure 4 | Operational cost comparison: (a) Run-1; (b) Run-2; (c) Run-3; (d) Run-4.

the model-based control and the conventional control did not exceed 150 kPa, the operation was suspended because of the time limit of the experiment.

Based on the results shown in Figure 3, operational cost was estimated for each run. The interval of chemical cleaning of the model-based control was calculated by extrapolation of the TMPs shown in Figure 3 excluding Run-3. Figure 4 shows the operational cost results of each run.

In Run-1, the effect of cost reduction was 31.5%. The chemical cost and sludge disposal cost were higher in the model-based control, because the coagulant dosage of the model-based control was higher than the conventional one. The power cost was also higher because the sludge dewatering ability had fallen because of high coagulant dosage. However, the increase of TMP was restricted because substances such as dissolved organic matter, which may be the dominant difficult-to-detach foulants, were thought to be absorbed in large flocs generated by the high dosage coagulant. As a result, because the proportion of membrane module replacement cost and chemical cleaning cost was larger, the operational cost was reduced overall.

In Run-2, owing to the difference of the upper limit and the evaluation indices, the PACl dosage was almost the same. That is why the cost reduction effect was only 1.8%.

In contrast, the operational cost was cut almost in half in Run-3. The membrane module replacement cost was not included in evaluation indices in Run-3, and the effect of cost reduction was thought to be less than that of Run-1 at first. However, the accumulation rate of foulants on the membrane was faster because the water temperature was low, reaching a minimum of 5°C. As a result, TMP grew rapidly and chemical cleaning cost increased. Then, a larger effect on operational cost reduction was obtained than in Run-1. When the water temperature was low, the TMP increased because the viscosity of water increased. And the coagulation reaction of PACl was slow when the water temperature was low. It followed from this that the effect of pretreatment was reduced. In addition to these findings, the detachment effect of foulants on the membrane decreased when the water temperature was low. For these reasons, the TMPs in Run-3 were thought to increase rapidly.

In Run-4, the filtration flux was decreased. Therefore the effect of operational cost reduction was less than in Run-3 and the model-based control had a 19.3% operational cost reduction.

As shown above, although there were differences in the degree of the effect between 1.8% and 47.9%, the operational cost reduction by model-based control was verified for all runs. The extent of the reduction of operational cost and the optimum control conditions were found to differ due because of conditions of the upper limit of PACl dosage, filtration flux and evaluation indices.

The operational conditions obtained in this study cannot be used, for example, when the filtration flux, water resource or the membrane material is different. However, the model-based control method has a high flexibility because it consists mainly of the mathematical model and the procedure for calculating operational cost. Therefore it is possible to obtain the optimum operational conditions and control the facility just by changing the values in the software and carrying out the calibration. If the fouling phenomena do not differ much, this control method can be adapted to other membrane modules even though the manufacturer or the material is different.

CONCLUSIONS

In this study, a model-based control system was developed in which the coagulant dosage for the pretreatment process and the backwash interval for the membrane filtration process were automatically optimized based on the process model by which the future TMP could be calculated. From the results of a pilot-scale experiment, the operational cost was thought to be reduced by as little as 1.8 to as much as 47.9%. The extent of the reduction of operational cost and the optimum control conditions were found to differ depending on the conditions of the upper limit of PACl dosage, filtration flux and evaluation indices.

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