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CONTROL STRATEGY ROBUSTNESS WITH RESPECT TO HYDRAULIC MODEL SOPHISTICATION

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Abstract : The robustness of the control strategies developed in the COST 624 benchmark have been tested against the sophistication of the model describing the hydraulic behaviour of the biological reactor of a wastewater treatment plant by activated sludge. In such a large biological reactor of the channel type, hydrodynamics are intermediate between plug flow and well-mixed and are function of the liquid flow rate and the aeration intensity. No large effect of these hydrodynamics could be observed in the various tested scenarios, suggesting that the actual benchmark is indeed a good tool for a first assessment of the efficiency of control schemes. *Copyright IFAC 2002*.

Keywords : benchmarking, wastewater treatment, robustness, hydrodynamics

1. INTRODUCTION

Wastewater treatment plants are complex non-linear systems, subject to large perturbations and where different physical (such as settling) and biological phenomena are taking place.

Many control strategies have been proposed in the literature for wastewater treatment plants but their evaluation and comparison are difficult. This is partly due to the variability of the influent, to the complexity of the physical and biochemical phenomena and to the large range of time constants (from a few minutes to several days) inherent in the activated sludge process. Also complicating the evaluation is the lack of standard evaluation criteria. A benchmark, i.e. a simulation environment defining a plant layout, a simulation model, influent loads, test procedures and evaluation criteria has been proposed within the framework of COST Actions 682 and 624 (Pons et al., 1999; Alex et al., 1999). Although realistically chosen, the benchmark plant layout does not correspond exactly to any specific plant. It is therefore legitimate to evaluate the robustness of the control strategies validated on the benchmark with respect to design and operational parameters. Vanrolleghem and Gillot (2001) have investigated the influence of changes on the influent composition and flow rate and of the temperature. Another issue is the complexity of the hydrodynamics. The channel reactor is one of the most widespread reactors in wastewater treatment plants of large capacity. It is aerated from its floor by an air diffusion system. Due to the large gas velocities, the bubbles induce an upward motion of the liquid near the fixed walls, thus creating vertical recirculation cells. The relatively slow horizontal motion due to the incoming water flow is superposed to this vertical motion. Globally the resulting hydrodynamics are intermediate between well-mixed reactor and plug flow. For sake of simplicity the benchmark biological reactor, of total volume 5999 m^3, has been divided into five well-mixed compartments, two of them being anoxic and the three others aerated. In this contribution we propose to investigate the effect of the hydrodynamics on the basic control strategy of the benchmark, which is based on two PI control loops:

one controls the nitrate level in the anoxic section and the other controls the dissolved oxygen concentration at the end of the aerated section.

2. PLANT DESCRIPTION

Figure 1 summarises the benchmark plant layout. Full details are available on the COST 624 webpage (http://www.ensic.inpl-nancy.fr/COSTWWTP). The biological reactor has five well-mixed units for a total volume of 5999 m^3, with an anoxic section which occupies 1/3 of this volume. The clarifier has a volume of 6000 m^3 and a depth of 4m. The IAWQ Activated Sludge Model N° 1 (Henze et al., 1987) was chosen to simulate the biological process. Thirteen state variables describe the fate of biodegrable and non biodegradable, soluble and insoluble, carbon and nitrogen-based pollution as well as bacteria (heterotrophs and autotrophs). The double-exponential settling velocity model proposed by Takács *et al.* (1991) was selected to describe the behaviour of the clarifier. Dissolved oxygen concentration in the 3rd aerated unit and nitrate concentration in the second anoxic unit are controlled with PI controllers by means of the oxygen transfer coefficient, K_La, and the internal recycle flow rates (Q_a) respectively.

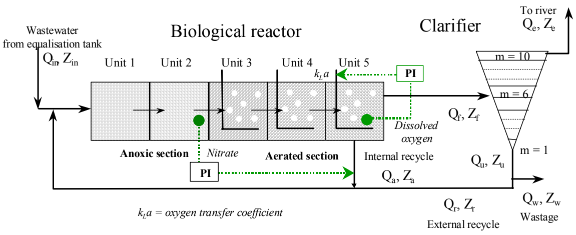


Fig. 1: Benchmark plant flowsheet

In the present implementation PI controllers are of the discrete type. Let Δt be the time interval between two actions of a controller, *y(k)* the measurement at time $k\Delta t$, and y^{set} the setpoint . The action to be applied, *u(k),* is calculated as follows :

$$u(k) = Du + u(k-1) \qquad (1)$$

with $$Du = K \left\{ [e(k) - e(k-1)] + \frac{\Delta t}{T_i} e(k) \right\} \qquad (2)$$

under the following constrains :

$|Du| \le Du_{\max}$ (limit on u variation between two successive actions)

$u_{\min} \le u(k) \le u_{\max}$ (permissible values of u)

e(k) and *e(k-1)* are respectively the errors at time $k\Delta t$ and $(k-1)\ \Delta t$:

$$e(k) = y^{set} - y(k) \qquad (3)$$

K_c and τ_i are respectively the proportional and integral constants of the PI controller.

A dry weather file, available on the COST website describes two weeks of variations of the influent flow rate and composition, without any rainfall. The weekend effect is taken into account.

The results discussed thereafter have been obtained using a FORTRAN code and the set of differential equations is integrated using a 4th-order Runge-Kutta algorithm and a constant integration step size (0.002 hr).

3. HYDRODYNAMICAL MODEL

The model is based on an existing biological reactor, whose size is similar to the size of the aerated zone of the benchmark plant (4000 m^3). The largest aerated biological reactor of the Nancy-Maxéville (F) wastewater treatment has a volume of 3300 m^3 and is aerated by gas diffusers located on the floor. Residence Time Distributions experiments have been run previously on that reactor by injecting an inert tracer (lithium chloride) (Potier et al., 1998). From these experiments performed under different liquid flow rates, it appears that this 100m long and 4m deep channel reactor can be modelled either by a plug flow model with axial dispersion, or a series of J completely mixed reactors, or J_a completely mixed reactor with back-mixing, q_b. It has been shown also that the axial dispersion coefficient D, the number of well mixed reactors J or the back-mixing flow q_b are function, for a fixed geometry of the reactor, of the liquid and gas flow rates (Potier et al., 2001). For the simulation structure it is easier to deal with a fixed number of reactors and a varying back-mixing flow rate.

Figure 2 summarises the model composed of J_{anox} units for the anoxic section and the J_{aera} units for aerated section. A wall between the two sections prevents back-mixing between them. In the aerated section,

$$q_{b,aera} = \alpha\left(Q, v_{aerated}\right) \cdot \varphi(K_L a) \cdot Q \qquad (4)$$

where $Q = Q_0 + Q_a + Q_r$ (Q_0 is the incoming flow rate, Q_a the internal recycle flow rate and Q_r the external recycle flow rate), K_La the local oxygen transfer coefficient, $v_{aerated}$ the volume of the aerated section and α and φ two functions. The variation of α for $v_{aerated}$ = 3999 m^3 is given in Fig. 3. φ is a function which translates the effect of aeration (through K_La) on back-mixing. A linear relationship has been assumed here. For back-mixing between unit i+1 and i:

$$\varphi_{\ i+1 \to i} = 1. + \frac{K_L a(i) + K_L a(i+1) - K_L a_{\max}}{K_L a_{\max}} \qquad (5)$$

where K_La_{max} is the maximal value of the oxygen transfer rate (set to10 hr^{-1}).

In the anoxic section, where the sludge suspension is only due to specially designed propellers,

$$q_{b,anox} = 0.1 \cdot \alpha\left(Q, v_{aerated}\right) \cdot Q \qquad (6)$$

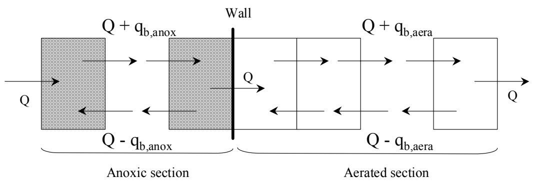


Fig. 2: Hydrodynamical model with a series of well-mixed tanks with back-mixing

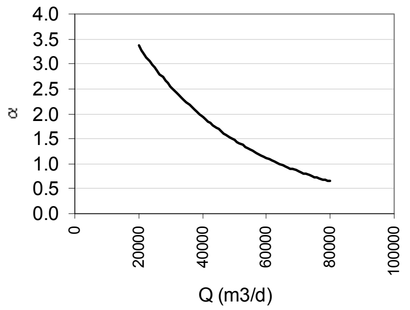


Fig. 3. Variation of α in function of the flowrate Q

In the original benchmark, the dissolved oxygen probe is located in the last one of the three aerated compartments and is used to manipulate the oxygen transfer coefficient in this compartment. Here, it is assumed that one could manipulate the oxygen transfer coefficient per zone equal to one third of the aerated section. Physically this is related to the fact that the channel reactors are usually folded. Here two cases will be tested. In the first case (Control_1), a single probe is located in the middle of the third zone. Therefore, when the basic control proposed in the benchmark is used, the control loop manipulates the oxygen transfer coefficient in units $1+2\cdot J_{aera}/3$ to J_{aera} (i.e. in the last third of the aerated section) and the oxygen transfer coefficients in the other units remain constant (10 h^{-1}). In the second case (Control_3), a probe is located in the final unit of each zone, three PI controllers are used and the oxygen transfer coefficient in each zone is manipulated. This control strategy was suggested by Vanrolleghem and Gillot (2001) as efficient with a reasonable investment cost. The nitrate sensor is always located in the last anoxic compartment.

4. PERFORMANCE ASSESSMENT

Various criteria have been defined within the benchmark to assess the general performance of the plant. For sake of simplicity, we will focus here on the pollution index (*E.Q.*), which is proportional to the fines to be paid in case of high pollution level in the receiving body, the aeration energy (*AE*) and the pumping energy (*PE*):

$$E.Q. = \frac{1}{T\cdot 1000}\int_{t=7days}^{t=14days}\cdot\begin{pmatrix} B_{SS}\cdot SS_e(t)+B_{COD}\cdot COD_e + \\ B_{NKj}\cdot S_{NKj,e}(t)+B_{NO}\cdot S_{NO,e}(t)+ \\ B_{BOD5}\cdot BOD_{5,e}(t)\end{pmatrix}.Q_e(t)\ dt \tag{7}$$

It should be noticed that *E.Q.* decreases when the effluent water contains less pollution.

$$AE = \frac{86400}{T}\int_{t=7\ days}^{t=14\ days}\sum_{iz=1}^{iz=3}\left(0.288\cdot(K_La)_{iz}^2+9.68\cdot(K_La)_{iz}\right)\cdot dt \tag{8}$$

where iz is the zone number in the aerated section.

$$PE = \frac{0.04}{T}\int_{t=7\ days}^{t=14\ days}\left(Q_a(t)+Q_r(t)+Q_w(t)\right)dt \tag{9}$$

Furthermore constraints with respect to the effluent quality are defined as follows: total nitrogen $S_{Nkj,e}$ < 18 mgN/L, COD_e < 100 mg/L, ammonia $S_{NH,e}$ < 4 mgN/L, suspended solids SS_e < 30 mg/L, $BOD_{5,e}$ < 10 mg/L. All the criteria are computed during the second week of a four-week dynamic test period (two weeks of dry weather followed by the two weeks of the weather file to be tested), that follows 50 days of stabilisation under constant influent conditions.

5. STEADY STATE

The effect of the modified hydrodynamical model on the steady-state concentrations, at the end of the 50 days stabilisation period, has first been examined. Table 1 summarises the results obtained on some effluent state variables, with no back-mixing ($q_{b,aera}$ = $q_{b,anox}$ = 0), in open loop, in function of J_{aera} and J_{anox} and for $\varphi(K_La) = 1$. The case where J_{aera} = 3, J_{anox} = 2 without back-mixing is the reference benchmark situation. In open loop the oxygen transfer coefficient is equal to 10 hr^{-1} in zones 1 and 2, and to 3.5 hr^{-1} in zone 3. Some variables, such as the total solid concentration in the effluent is not affected by the hydraulic model complexity. With respect to the original benchmark (J_{anox} = 2 and J_{aera} = 3), the increase of the number of units (the hydrodynamics are more of the "plug-flow" type) increases the global efficiency of soluble carbon and ammonia removal (6% decrease of $S_{S,e}$ and 33% for $S_{NH,e}$). In Table 2 the steady-state values obtained for J_{anox} = 6 and J_{aera} =12 with and without back-mixing are compared. With back-mixing the behaviour is less of the "plug-flow" type and the global efficiency decreases slightly but is still higher than in the original case (4% decrease of $S_{S,e}$ and 29% for $S_{NH,e}$). The effect on the effluent nitrate concentration is very small (< 1%).

The PI controller settings are given in Table 3 and the influence of the hydrodynamical model on the Control_1 closed-loop behaviour at the end of the 50 days stabilisation periods is summarised in Table 4. The effect is similar to what was observed in open loop, with an increase in ammonia removal.

Table 1: Influence of the number of units in the aerated and anoxic section on the steady-state effluent soluble substrate ($S_{S,e}$), nitrate ($S_{NO,e}$) and ammonia ($S_{NH,e}$) concentrations, without back-mixing (open loop)

J_{anox}	J_{aera}	$S_{S,e}$ (mg/L)	$S_{NO,e}$ (mg/L)	$S_{NH,e}$ (mg/L)
2	3	0.887	10.4	1.71
2	6	0.857	10.9	1.33
2	9	0.847	11.1	1.21
4	9	0.842	10.9	1.19
6	12	0.835	10.9	1.13

Table 2: Influence of back-mixing on the steady-state effluent soluble substrate ($S_{S,e}$), nitrate ($S_{NO,e}$) and ammonia ($S_{NH,e}$) concentrations, for $J_{anox} = 6$, $J_{aera} = 12$ (open loop)

$q_{b,aera}$	φ	$S_{S,e}$ (mg/L)	$S_{NO,e}$ (mg/L)	$S_{NH,e}$ (mg/L)
0	-	0.835	10.9	1.13
Eq. 4	1	0.843	10.9	1.17
Eq. 4	Eq. 5	0.849	10.8	1.22

Table 3: PI controllers settings

	Oxygen controller	Nitrate controller
K_c	0.9 $h^{-1}.(mg/L)^{-1}$	315 $(m^3/hr)(mg/L)^{-1}$
τ_i (hr)	0.05	0.3
Δt (hr)	0.02	0.17
y^{set} (mg/L)	2	1
u_{min}	0 hr^{-1}	0 m^3/hr
u_{max}	10 hr^{-1}	3843 m^3/hr
Du_{max}	0.5 hr^{-1}	∞

Table 4. Influence of back-mixing on the steady-state effluent soluble substrate ($S_{S,e}$), nitrate ($S_{NO,e}$) and ammonia ($S_{NH,e}$) concentrations, in Control_1 closed-loop

J_{anox}	J_{aera}	$q_{b,aera}$	φ	$S_{S,e}$ (mg/L)	$S_{NO,e}$ (mg/L)	$S_{NH,e}$ (mg/L)
2	3	No	No	0.806	13.5	0.667
6	12	No	No	0.721	13.9	0.210
6	12	Yes	No	0.758	13.4	0.367
6	12	Yes	Yes	0.781	13.3	0.431

6. DYNAMIC BEHAVIOUR

In Control_1 closed-loop, with the dry weather file, the effluent quality is slightly decreased when back-mixing is considered (4%). The operation costs are slightly increased from the aeration point of view (2%). They are decreased from the pumping point of view (12%) but this item represents only 17% of the operation costs. Fig. 4 presents the variations of the oxygen transfer rate in the aerated section in function of the hydrodynamical model. The amplitude of variations of K_La is smaller in the reference case, but it is very difficult to discriminate between the different scenarios when the number of units increases.

Table 5. Influence of back-mixing on the plant performance in Control_1 closed-loop

J_{anox}	J_{aera}	$q_{b,aera}$	φ	E.Q. (kg/d)	AE (kWh/d)	PE (kWh/d)
2	3	No	No	7540	7235	1517
6	12	No	No	7890	7378	1267
6	12	Yes	No	7830	7332	1307
6	12	Yes	Yes	7810	7353	1327

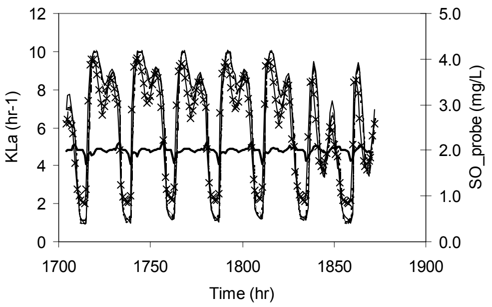


Fig. 4. Variations of the oxygen transfer coefficient (K_La) in the third aerated zone for $J_{anox} = 6$, $J_{aera} = 12$ without back-mixing (—), with back-mixing and $\varphi = 0$, with back-mixing and $\varphi \neq 0$. Comparison with K_La of the original case ($J_{anox} = 2$, $J_{aera} = 3$) (—×—) Dissolved oxygen at the probe level (▬).

As suggested by Vanrolleghem and Gillot (2001), two additional control loops on dissolved oxygen were implemented in the aerated section, with one control loop for each zone (Control_3). The set points were fixed at 2 mg/L in each zone. Table 6 summarises the results obtained for the different configurations with K_c(nitrate) = 31.5 $(m^3/hr)(mg/L)^{-1}$. The last line of Table 6 refers to the implementation of an equalisation tank in front of the biological reactor. This tank, of volume 4000 m^3, has been previously shown to dampen the large flow rate variations of the influent and to improve the treated water quality in the original benchmark (Pons and Corriou, 2001a, 2001b). The set point of the equalisation tank flow rate to the biological reactor was fixed at 750 m^3/h.

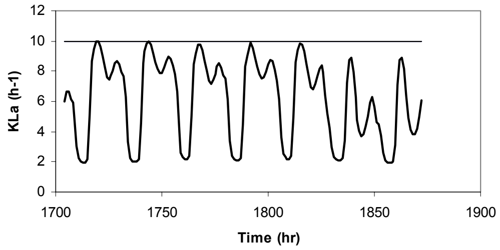
With the original benchmark as well as with the more sophisticated hydrodynamical model, the Control_3 strategy is more efficient in terms of water quality than the Control_1 strategy. The aeration energy is slightly reduced but the pumping demand is higher. The combination of the equalisation tank and the Control_3 strategy improves significantly the water quality and decreases slightly the aeration energy. Pumping energy will of course be much higher in the case of the equalisation tank (pumping from the tank to the biological reactor has not been included here).

Table 6. Comparison of the control strategies performances in closed-loop.

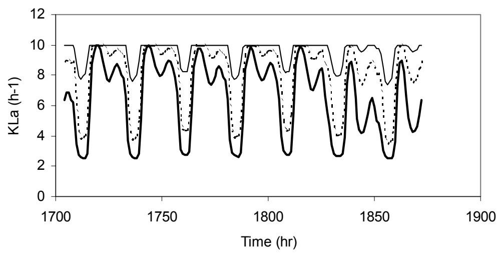
J_{anox}	J_{aera}	$q_{b,aera}$	φ	Control	*E.Q.* (kg/d)	*AE* (kWh/d)	*PE*
2	3	No	No	1	7990	7330	1249
2	3	No	No	3	7950	7170	1260
6	12	Yes	Yes	1	7820	7329	1326
6	12	Yes	Yes	3	7760	6097	1343
6	12	Yes	Yes	3 + Equal.	6980	5896	1483

Figure 5 compares the effect of the control strategies on the manipulated variables, i.e. the oxygen transfer coefficients in the three zones. With Control_1, the manipulated variables in zones 1 and 2 of course is always saturated. With Control_3 the effort is better distributed between the three zones. However with Control_3 and the equalisation tank, the manipulated variable in the first zone saturates almost for the full evaluation period, when the oxygen transfer coefficient in the last zone is smaller in average than in the two other cases. The manipulated variables variations are smaller in the latter case, which reduces the compressor wear.

(a)



(b)



(c)

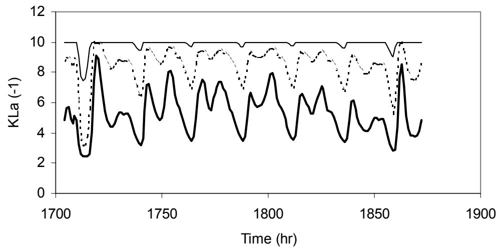


Fig. 5. (a) Control_1; (b) Control_3; (c) Control_3 + equalisation tank. Variations of the oxygen transfer coefficient (K_La) in the three aerated zones for J_{anox} = 6, J_{aera} = 12 with back-mixing: zone 1 (—), zone 2 (- - -) and zone 3 (▬).

7. CONCLUSIONS

The effect of the sophistication of the description of the hydrodynamics on the design of control strategies for wastewater treatment plant has been investigated using the COST 624 benchmark. The 5000 m^3 biological reactor has been supposed to be ~~of~~ the channel type, that is very common in full-scale plants. The hydrodynamics are intermediate between plug-flow and well-mixed, and their characteristics are function of the flow rate and the aeration. The sophistication of the biological reactor hydrodynamical model affects only some of the state variables at steady-state, and more especially the ammonia concentration in the plant effluent. Under dynamic conditions a similar effect is observed.

However, the simplicity of the reference benchmark hydrodynamical model for its biological reactor has not been challenged in any of the tested control strategies. The actual benchmark seems therefore a good starting point to evaluate the performance of control strategies, before of course taking into account the real behaviour of a plant for a more refined assessment. It should also be noted that the reduced dimension of the benchmark model makes the simulation time much shorter than with a complex model, allowing to test easily many control schemes.

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