

Control of a Class of Sulfate Reducing Chemostat Via Feedback Polynomial Injection

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ABSTRACT

The main goal of this work is to present a new class of feedback controller which contains on its structure a polynomial form of the named control error, the proposed controller is applied to a class of sulfate-reducing chemostat in order to control the sulfate concentration, which would be useful for several biotechnological issues, as heavy metal removal in wastewater. The closed-loop behavior of the chemostat is theoretically analyzed and a practical convergence to the selected optimum trajectory is proved. The proposed methodology is applied to an experimentally corroborated kinetic model of a sulfate-reducing bacterium and further numerical experiments show the satisfactory closed-loop performance of the process in comparison with other controllers.

Keywords: Sulfate reducing chemostat, polynomial feedback, closed-loop stability.

RESUMEN

El principal objetivo de este trabajo es el presentar una nueva clase de controlador de tipo retroalimentado, el cual contiene en su estructura una forma polinomial del llamado error de control, el controlador propuesto es aplicado a un quimiostato sulfato-reductor, el cual pudiera ser usado para varios fines biotecnológicos, como la remoción de metales pesados en aguas residuales. El comportamiento a lazo cerrado del quimiostato considerado es teóricamente analizado y se prueba convergencia práctica a la trayectoria óptima seleccionada. La metodología propuesta es aplicada a un modelo cinético de una bacteria sulfato-reductora experimentalmente validado y experimentos numéricos complementarios muestran un comportamiento a lazo cerrado satisfactorio en comparación con otros controladores.

1. Introduction

Sulfate reducing bacteria (SRB) are dispersed in the nature, are able to produce energy through electron transfer-coupled phosphorylation, use the sulfate as the terminal electron acceptor for respiration and produce sulfide as compound end. SBR play an important role in global cycling of numerous elements in anaerobic environments, particularly in sulfate rich environments, the global sulfur and carbon cycles are most relevant (Widdel, 1988). Beyond their role in ecology, SRB have an economic impact because they are associated with pitting corrosion of ferrous metals due to production of sulfide, which is highly reactive, corrosive and toxic, particularly in the petroleum industry (Neria-Gonzalez et al.; 2006).

On other hand, SRB have become especially important in biotechnological processes due to its ability to degrade organic material and remove heavy metals (Jorgensen, 1982). Many processes of material degradation and heavy metal removal are carried on anaerobic bioreactors, which are large fermentation tanks provided with mechanical mixing, heating, gas collection, sludge addition and withdrawal ports, and supernatant outlets. The anaerobic digestion is affected by many factors including temperature, hydraulic time, pH, and chemical composition of wastewater. In a practical framework, the bioreactors are operated on the basis of laboratory analysis of samples, which lead to empirical methodologies for operation policies. In general the above is related to the rising energy costs, increased global competition in terms of

both price and quality, and the need to make products in an environmentally benign manner. Many of the products obtained by the biological route either cannot be produced, or are very difficult to obtain, by conventional manufacturing methods.

Generally speaking, the following features distinguish the biochemical reactors from the chemical reactors:

1. Several of the crucial variables can not be directly measured quickly or easily. Quite often a time delay, which is larger than the system time-constants, is associated with their measurement. Therefore, a mathematical model must be used in place of feedback information.
2. As already pointed out, the fed-batch reactor operation does not have a true steady state. In this case, evaluation of the state variables will locate the position of the system on a trajectory through the operational cycle. Since these state variables can not be measured online, the estimation of state becomes an important element of optimization and control of the reactor.
3. Linear system analysis is mostly not applicable in case of bioreactors, especially for evaluation of long term response, since biochemical reaction systems usually are nonlinear. Hence, numerical solution of differential equations is required.

Biological reactors typically are governed by highly nonlinear behavior occurring on both a macroscopic reactor scale and a microscopic cellular scale. Reactions taking place at these scales also occur at different rates so that the bioreactor system is multi-scale both spatially and temporally. By increasing our understanding of these systems, it may be possible to avoid some of the empiricism associated with the operation of continuous and (fed-) batch bioreactors. Considerable benefit, in terms of reduced product variability and optimal resource utilization could be achieved. Process control mainly focuses on controlling variables such as pressure, level, flow, temperature, pH, level in the process industries; however, the methodologies and principles are the same as in all control fields. The early successful application control strategy in process control is in evolution of the PID controller and Ziegler-Nichols tuning method (Ziegler and Nichols, 1942). So far a high percentage of the controllers implemented

in the process industries are PID-type (Chidambaram and See, 2002); however, as (i) the industrial demands (ii) the computational capabilities of controllers and (iii) complexity of systems under control increase, the challenge is to implement advanced control algorithms (Spear, 2005).

Since achievable controller performance in a model-based control scheme is dependent on the quality of the process model (Aguilar, et al, 2005), a controller based on a model that captures events occurring at both the general considerations for control of bioreactors.

On the other hand, the difficulty of implementing a feedback control is threefold. Firstly, the response of sensors tends to be slower than many of the processes they monitor. Secondly, sensors are generally not available for measurement of substrate with rapid dynamics for feedback application; and thirdly, there is also generally not available a sensor for the biomass concentration without which the state of the system can not be estimated (Fordyce, et al, 1987)

Given the above objectives, broadly speaking, there are two ways to design an appropriate control system. The most frequently used method is to pre-select a controller structure and then to tune the parameters of this controller so that the desired closed-loop response is obtained. This is referred to as a parameter optimized control system; the most well-known example is probably the Proportional Integral Derivative (PID) controller (Bastin and Dochain, 1990). The other approach is the use of structure optimal control systems where both the structure and parameters of the controller are adapted to those of the process model (Betancour, et al, 2006). In practice, however, the use of the latter method is severely restricted because exact dynamic term cancellation is required in order to produce the optimal controller structure. This is usually not possible for various reasons: the lack of an appropriate process model, non-linearities and physical constraints on the process variables.

From the above, in this paper a class of nonlinear single-input single-output controller with polynomial output structure in order to provide stabilization to a class of continuous sulfate reducing bioreactor is proposed.

2. Experimental

In this paper, a sulfate-reducing bacterium (6SR), isolated from a biofilm sample obtained from the inner surface of pipelines that transport oil crude, is considered. The molecular identification of 6SR isolated showed a phylogenetic relation with *Desulfovibrio alaskensis* so the strain is named *Desulfovibrio alaskensis* 6SR (Neria-Gonzalez et al.; 2006). *Desulfovibrio alaskensis* 6SR was cultured and conserved in Ravot medium under an atmosphere of N₂-CO₂ (80:20, v/v).

2.1 Growth kinetic

A sample of congenital water medium (CW) was obtained from an oil pipeline located in the Mexican Southeast region. Chemical determination of water: chlorides 64 000 ppm, sulfur 178 ppm, sulfate 350 to 400 ppm, pH 8.84. A 1000 mL aliquot of congenital water was saturated with N₂ for an hour and was enriched with sodium lactate 6 mL, yeast extract 0.5 g, and reducing solution 5 mL (acid ascorbic 1 g L⁻¹, and sodium thioglicolate, 1 g L⁻¹). The pH was adjusted to 7 with KOH 1N. The CW medium was distributed in serum bottles of 60 mL using the Hungate's technique (Hungate, 1969) and they were autoclaved at 120 °C for 15 minutes.

The original cultures of *D. alaskensis* 6SR in medium Ravot were used to inoculate 45 mL of CW medium. The culture was incubated for 20 days at 37°C. This was used to inoculate three bottles with CW medium for different periods of time: zero, 24 and 36 hours, respectively, and were incubated under same conditions. The bacterial growth was followed through Optical Density (OD) measurements, consume of sulfate and production of sulfide. Samples from the cultures were taken anaerobically each hour. Sulfate in the medium was measured by the turbidimetric method based on the precipitation of barium (Kolmert A et al., 2000). Also, the production of sulfide was measured by a turbidimetric method (Cord-Ruwisch, 1985). The OD reading for cell growth was transformed to dry weight (mg/mL) through a standard curve of growth. The data were analyzed and only the points that adjusted a straight line (exponential phase) were used to determine the growth kinetic parameters according to the Monod model.

3. Process modeling

Sulfate reduction can be applied beneficially to biotechnology, i.e. for the removal of heavy metals from groundwater and waste water. These applications take advantage of differences in the chemical properties of metal sulfates and sulphides. Metal sulfates (cadmium, cobalt, copper, iron, nickel and zinc) are highly soluble but the corresponding metal sulphides have low solubility. Thus, by sulfate reduction, metals can be precipitated, recovered and reused. This concept has been applied to immobilize metals from surface water and process water from mining industries. Organic waste materials (for example, straw) are often used to immobilize heavy metals in lake sediments. From the above, the necessity to improve the operating performance of these processes seems very important and the system modeling, simulation, optimization and control appear to be tools for a complete process analysis.

For biological systems, the unstructured kinetic models are the simplest of all modeling philosophies used to describe the biological model. They consider the cell mass as a single chemical species and do not consider any intracellular reactions occurring within the cell. Unstructured models typically describe the growth phenomena based on a single limiting substrate and consider only substrate uptake, biomass growth, and product formation in the modeling framework. Thus the biological component of the system depends directly on the macroscopic reactor variables. These models give an adequate representation of the biological growth phenomena in relatively simple cases, when the cell response time to environmental changes is either negligibly small or much longer than the batch time (Schugerl, 1987).

The most commonly used unstructured model in the literature is the Monod model (Alford, 2006). And it is one of the earliest attempts at modeling biological systems. In this model the growth kinetics are expressed in terms of the specific growth rate. The parameters are obtained experimentally and they do not have a direct physical interpretation (Schugerl, 1987). A drawback of this model is that it does not capture the initial lag phase of growth which is observed in most batch cultures. There are variants to the Monod model that have also been used, such as

the models by Messier, Contois etc. reported in the open literature (Bailey and Ollis, 1986). These models differ in their substrate dependence and some include terms to account for saturation due to high substrate concentration and inhibition due to product or a competing inhibitor; however, these models do not differ significantly from the Monod model in the fact that they are empirical and represent all of the cellular processes with just a single equation for the specific growth rate.

Nevertheless, for control purposes, a reduced order model which can describe the dynamic behavior of the main state variables is adequate. The kinetic parameters can be determined via standard methodology (Bailey and Ollis, 1986) in a batch culture (Neria-González and Aguilar-Lopez, 2007).

By employing a Monod model for the specific growth rate, the following is obtained:

$$\mu(S) = \mu_{\max} \frac{S}{K_S + S} \quad (1)$$

or

$$\frac{1}{\mu} = \frac{1}{\mu_{\max}} + \frac{K_S}{\mu_{\max} S} \quad (2)$$

The corresponding kinetic parameters can be fitted by plotting μ^{-1} versus S^{-1} (Lineweaver-Burk plot). For our particular study case, the maximum specific growth rate is calculated as $\mu_{\max} = 0.035 \text{ h}^{-1}$ and the affinity constant $k_s = 0.9 \text{ g/L}$ with a correlation factor of $R^2 = 0.9786$.

Therefore:

$$\mu(S) = \frac{0.035S}{0.9 + S} \quad (3)$$

and $Y_{S/X} = 0.25$ and $Y_{P/X} = 0.263$.

The following mathematical model for the continuous bioreactor is hence proposed, based

on classical mass balances for biomass, sulfate (substrate) and sulfide (product) concentrations: Sulfate (S).-

$$\frac{dS}{dt} = D(S_{\text{in}} - S) - \mu(S) \frac{X}{Y_{S/X}} \quad (4)$$

Biomass (X).-

$$\frac{dX}{dt} = -DX + \mu(S)X \quad (5)$$

Sulfide (P).-

$$\frac{dP}{dt} = -DP + \mu(S) \frac{X}{Y_{P/X}} \quad (6)$$

Here D is the dilution rate (control input), μ is the specific growth rate, $Y_{S/X}$ is the sulfate coefficient yield and $Y_{P/X}$ is the sulfide coefficient yield.

In a previous paper (Neria-González and Aguilar-López, 2007), some operating regions related with specific process applications (for example, heavy metal removal) were determined. From the above, two open-loop steady states were selected as mentioned below; the first equilibrium point for heavy metal removal regimen is $S_{eq} = 2.25 \text{ g/L}$; $X_{eq} = 0.68 \text{ g/L}$; $P_{eq} = 2.61 \text{ g/L}$. The second equilibrium point: $S_{eq} = 0.36 \text{ g/L}$; $X_{eq} = 1.16 \text{ g/L}$; $P_{eq} = 4.4 \text{ g/L}$, is related to high sulfate consumption.

Now, evaluation of the the stability conditions of the above equilibrium points is necessary, this is done via the first Lyapunov criteria, where the corresponding eigenvalues for the first equilibrium point are $\lambda_1 = -1/40$; $\lambda_2 = -1/40$; $\lambda_3 = 43/280$, note that the real part of the eigenvalue corresponding to sulfide concentrations is positive; therefore, the equilibrium point is unstable.

On the other side, the corresponding eigenvalues for the second equilibrium point are $\lambda_1 = -1/100$; $\lambda_2 = -1/100$; $\lambda_3 = 59/350$; as above, the the equilibrium point is unstable too.

From the above results, it can be concluded that the open-loop operation is locally unstable in such a way that any disturbance arriving to the bioreactor can lead to the process to other, possibly undesirable, operating point; this justifies the application of a control law to force the bioreactor to remain on the selected operating points.

4. Proposed controller

A general model of the bioreactor can be described by Equations (7a) and (7b), which is related to an affine control law system for the considered process, as follows:

$$\dot{X} = f(X, U) = \mathfrak{F}(X) + \ell(X)U \quad (7a)$$

$$Y = h(X) = CX \quad (7b)$$

Where:

$X = [S, X, P]^T \in \mathfrak{R}_+^3$ is the corresponding state vector. Now consider the set $\Phi \subset \mathfrak{R}^3$ as the corresponding physically realizable domain such that $\Phi = \{(S, X, P) \in \mathfrak{R}_+^3 / 0 \leq S \leq S_m; 0 \leq X \leq X_{max}; 0 \leq P \leq P_{max}\}$; $\mathfrak{F}(X) \in C^\infty(X)$, i.e. this is a smooth function; $\mathfrak{F}(0) = 0$ and $\mathfrak{F}(X) \leq \Gamma \quad \forall X \in \mathfrak{R}_+^3$, where $\Gamma < \infty$, this means that \mathfrak{F} is bounded.

Defining the vector error as $e = X - X_{sp}$, where X_{sp} is the considered reference trajectory; therefore, $\dot{e} = \dot{X} - \dot{X}_{sp}$; where it is assumed that $\dot{X}_{sp} = \mathfrak{F}(X_{sp})$; this means that the proposed trajectory (sulfate concentration) obeys the mass conservation

principle despite the corresponding mass balance equation under the action of the considered control law. From the above, the corresponding error dynamic equation considering (7a) is as follows:

$$\dot{e} = \mathfrak{F}(X) - \mathfrak{F}(X_{sp}) + \ell(X)U \quad (8)$$

Now, let us consider the following hypothesis:

H1. For the realized control input vector $U(X(t))$, ($\|U(X(t))\| \leq \bar{U}_{max}$), the nominal closed-loop nonlinear system is quadratically stable; therefore, there exists a Lyapunov function $\Psi \geq 0$ that satisfies the following:

where $\Delta\mathfrak{F}(o) = \mathfrak{F}(X) - \mathfrak{F}(X_{sp})$, such that the uncontrolled states of Equation (7) are stable on the quadratic domain given by (9).

H2. The nonlinear vector function $\Delta\mathfrak{F}(o)$ is Lipschitz bounded i. e.: $|\Delta\mathfrak{F}(o)| \leq \Lambda|e|$.

H3. It is considered that $|\ell(X)| \leq \ell^+$, where ℓ^+ is an upper bound.

Proposition 1. For system (7) the following control U is considered:

$$U = g_1 C(x - x_{sp}) + g_2 C^w (x - x_{sp})^w \quad (10)$$

then U is able to provide semi-global stabilization to system (7).

Sketch of proof of Proposition 1.

Taking into account the above assumptions and applying the triangle inequality, the following expression can be obtained:

$$\frac{\partial \Psi}{\partial X} [\Delta\mathfrak{F}(o) + \ell(X)U] \leq -\varphi_1 \|X\|^2, \quad \left\| \frac{\partial \Psi}{\partial X} \right\| \leq \varphi_2 \|X\|, \quad \varphi_1, \varphi_2 > 0 \quad (9)$$

$$\left| \dot{e} \right| \leq \Lambda|e| + g_1 C \ell^+ |e| + g_2 C^w \ell^+ |e^w| = (\Lambda + g_1 C \ell^+) |e| + g_2 C^w \ell^+ |e^w| \quad (11)$$

$$\text{sign}\left(\dot{e}_i\right) \dot{e}_i \leq (\Lambda + g_1 C \ell^+)_i \text{sign}(e_i) e_i + (g_2 C^w \ell^+)_i \text{sign}(e_i^w) e_i^w \quad (12)$$

Considering a scalar element i of the above vector differential inequality and employing the equality: $|e| = \text{sign}(e)e$, the following equation is obtained:

Therefore, after algebraic manipulations, the following differential inequality is generated:

$$\dot{e}_i - \lambda_{1i} e_i \leq \lambda_{2i} e_i^w \quad (13)$$

Where:

$$\lambda_{1i} = \frac{(\Lambda + g_1 C \ell^+)_i \text{sign}(e_i)}{\text{sign}\left(\dot{e}_i\right)} \quad (14)$$

$$\text{and } \lambda_{2i} = \frac{(g_2 C^w \ell^+)_i \text{sign}(e_i^w)}{\text{sign}\left(\dot{e}_i\right)}$$

To solve the above inequality, consider the change of variable:

$$\mathcal{G}_i = e_i^{1-w}, \quad w > 1 \quad (15)$$

Thus

$$\text{if } e_i = 0 \Rightarrow \mathcal{G}_i = 0$$

and

$$\text{if } e_i \neq 0 \Rightarrow \mathcal{G}_i \neq 0, \quad \text{i.e.} \quad (16)$$

$$\left\{ \begin{array}{l} e_i > 0 \\ e_i < 0, w \text{ odd} \end{array} \right\} \Rightarrow \mathcal{G}_i > 0$$

Here after we consider

$$e_i \neq 0 \quad \text{with } w \in \mathbb{Z}^+, w \text{ odd}, w > 1.$$

Considering that inequality (12) is a Bernoulli type differential inequality and taking into account the change of variable (15), the next first order ordinary differential inequality is generated:

$$\dot{\mathcal{G}}_i - (1-w)\lambda_{1i}\mathcal{G}_i \geq (1-w)\lambda_{2i} \quad (17)$$

By solving the above:

$$\mathcal{G}_i \geq \mathcal{G}_{0i} \exp(-\{w-1\}\lambda_{1i}t) + \frac{\lambda_{2i}}{\lambda_{1i}} (1 - \exp(-\{w-1\}\lambda_{1i}t)) \quad (18)$$

For $t \rightarrow t_\infty$

$$\mathcal{G}_i \geq \frac{\lambda_{2i}}{\lambda_{1i}} \quad (19)$$

To get back to the original variable (regulation error):

$$e_i \leq \left(\frac{\lambda_{1i}}{\lambda_{2i}} \right)^{\frac{1}{w-1}} = \left(\frac{(\Lambda + g_1 C \ell^+)_i \text{sign}(e_i)}{(g_2 C^w \ell^+)_i \text{sign}(e_i^q)} \right)^{\frac{1}{w-1}}, \quad w > 1 \quad (20)$$

Defining the maximum error as $\bar{e} = \max_i e_i$. It can be concluded that $\bar{e} \in B_{\bar{e}}(0)$. 2

Remarks:

Note that the estimation error can be diminished as desired, considering g_2 large enough or considering $g_1 \approx -\Lambda(C\ell^+)_i^{-1}$, both considerations can be used as tuning rules guidelines; besides, as w increases, the regulation error is diminished as shown in Equation (20).

5. Numerical experiments and results

Firstly, the performance of the experimental kinetic model is shown in Figure 1, where an adequate performance can be observed. On the other hand, several numerical simulations were carried out in order to show the performance of the proposed methodology. The mathematical model of the considered continuous bioreactor was simulated via Math Lab tools (23S ODE library) ®; the corresponding initial conditions for the sulfate, biomass and sulfide concentrations are 6.0 g L^{-1} , 1.0 g L^{-1} and 0.1 g L^{-1} , respectively. The bioreactor is operated in an open-loop mode from the zero time at 50 h when the controller is tuning up, the original set point is defined at 2.25 g L^{-1} ; at $t = 75 \text{ h}$, the set point changes to 0.36 g L^{-1} and, finally, at $t = 125 \text{ h}$, the set point is 1.5 g L^{-1} of sulfate. For comparison purposes, several controllers were implemented; an ideal Input/Output linearizing controller (Kalafatis, et al, 2005), linear PI controller and the proposed methodology. The linear PI controller was tuned by Internal Model Control (IMC) guidelines (Rivera et al, 1986); the corresponding tuning is done via a step disturbance of 5% in the nominal value of the control input ($D = 0.025 \text{ h}^{-1}$), the steady-state gain is calculated as $K = 1400 \text{ g h L}^{-1}$, the characteristic time is $\tau = 170 \text{ h}$, the time delay is $\theta = 6 \text{ h}$ and the corresponding proportional gain is $K_p = 4.857 \text{ h}^{-1}$ and integral time $\tau_i = 170 \text{ h}$ for the closed-loop time constant $\lambda = 35 \text{ h}$ and the control's gain for the linearizing I/O controller is $\tau = 1.0 \text{ h}^{-1}$, for the proposed controller the gains are selected as; $g_1 = -2.857 \text{ h}^{-1}$ and $g_2 = 0.5 \text{ g}^{-w+1} \text{ L}^{-w+1} \text{ h}^{-1}$, considering $w = 3$.

The "Integral Time-Weighted Squared Error" (ITSE) defined by (18) suggested by Ogunnaike and Ray (1994) is considered to measure the impact of the control error. ITSE provides the advantage of heavy penalization of large errors at long time; therefore, an adequate measure of resilience of the considered controllers.

$$\text{ITSE} = \int_0^{\infty} t \varepsilon^2 dt \quad (21)$$

As can be observed in Figure 2, the proposed methodology acts almost immediately, leading the sulfate trajectory to the corresponding set points without overshoots and settling times, with an adequate effort of the control as shown in Figure 3. Note that all the controllers act over reachable operating regions, which is feasible for a real implementation. Figure 4 shows the closed-loop behavior of the uncontrolled states (biomass and sulfide concentrations); as can be observed, they are stable trajectories; therefore, it can be concluded that the bioreactor is stabilizable.

Figure 5 is related to a 3D phase-portrait, including all the trajectories of the state variables. Note that the trajectories remain in a bounded region. Figure 6 shows a comparison of the performance of the proposed controller against other control structures. As can be observed, the linear PI controller provides the worst performance, even more, the performance index (ITSE) does not reach a steady state value. The ideal linearizing I/O controller provides a satisfactory performance as observed because this controller cancels the nonlinearities of the system and imposes a closed-loop linear behavior, leading the corresponding trajectory to the selected set points in an asymptotic and exponential way; however, note that the performance of the proposed controller is the best in comparison with the other control schemes, without the need for a perfect plant model. Figure 6 is related to the closed-loop behavior of the corresponding concentration trajectories, under different initial conditions, it can be observed that all of them reach the corresponding closed-loop steady-state. Furthermore, additional numerical simulations were done in order to show the effect of the polynomial degree on the proposed control performance. As proved in Section 3, when the order of the polynomial contribution increases, the corresponding closed-loop performance is improved. Figure 7 is related to the sulfate closed-loop trajectory under several polynomial degrees and, finally, Figure 8 shows the corresponding performance index, which is in agreement with the expected result.

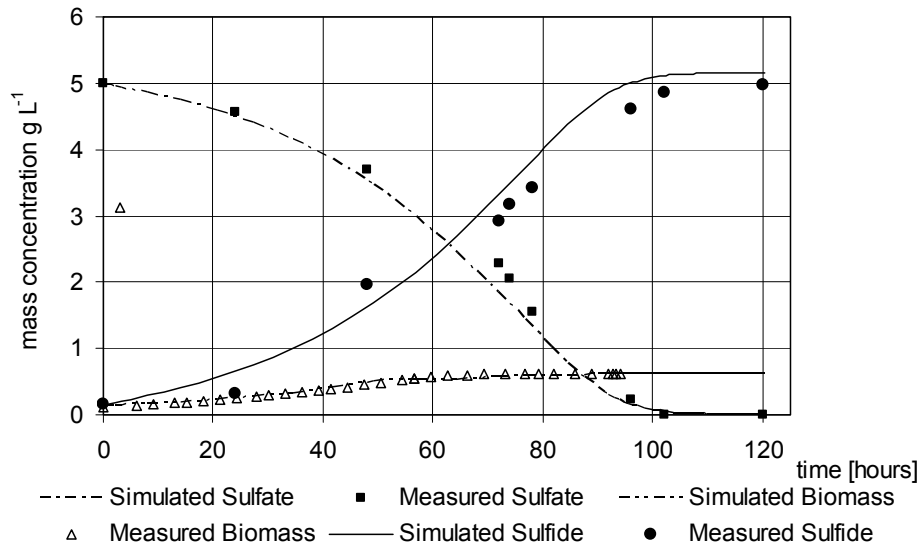


Figure 1. Experimental validation of the kinetic model.

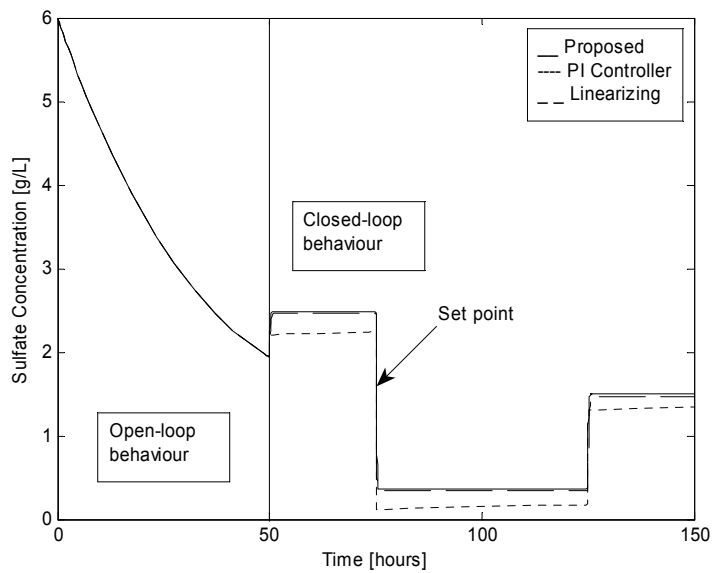


Figure 2. Closed-loop trajectory of the sulfate concentration.

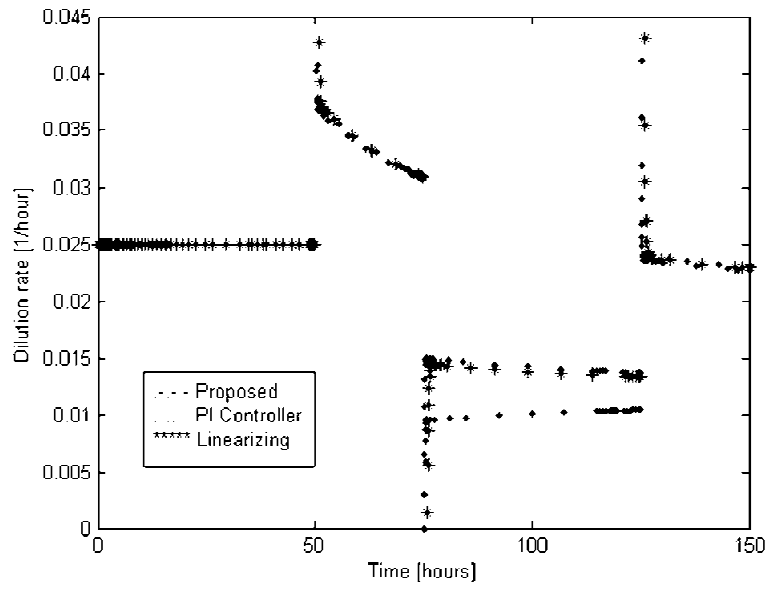


Figure 3. Control input efforts.

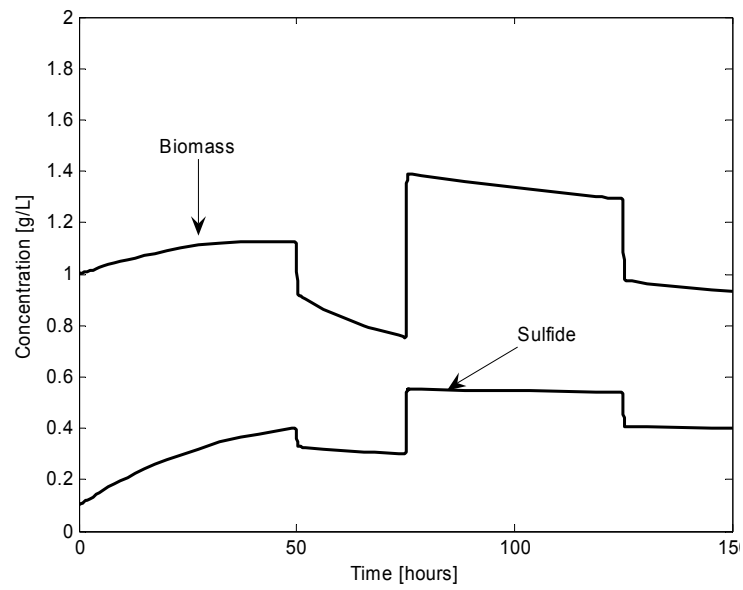


Figure 4. Closed-loop performance of the uncontrolled mass concentrations.

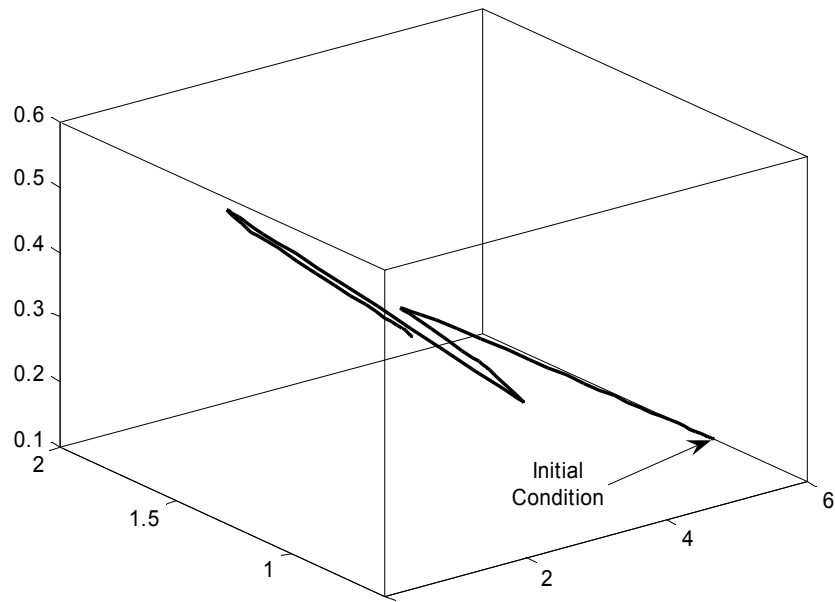


Figure 5. Closed-loop phase portrait.

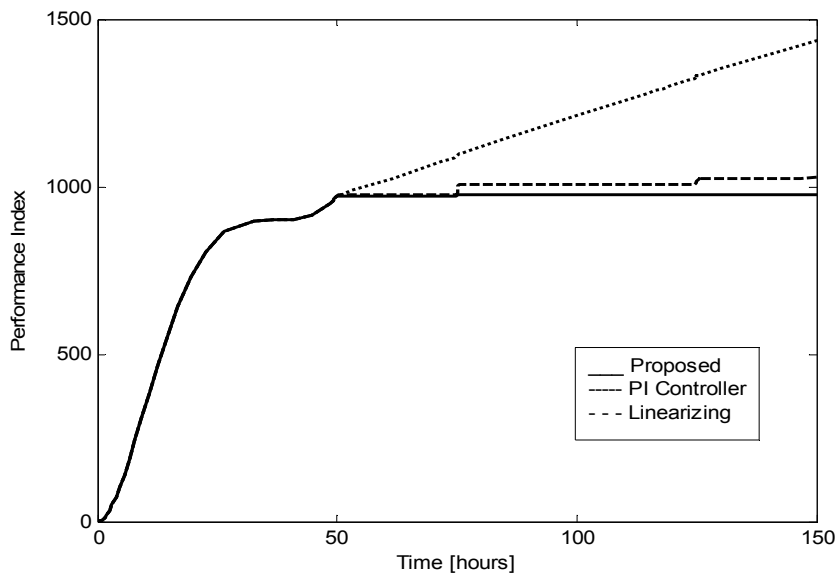


Figure 6. Performance index (ITSE) for the considered controllers.

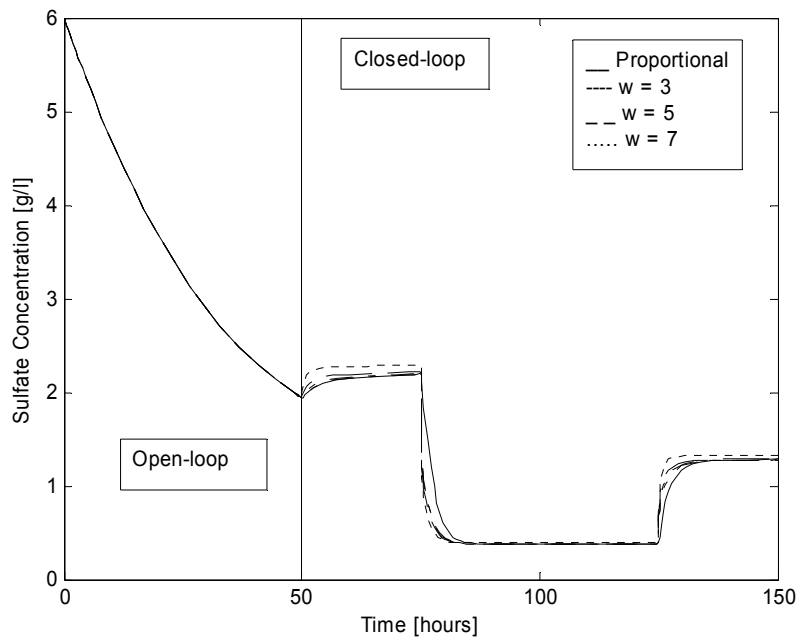


Figure 7. Closed-loop performance of the sulfate concentration for several values of the polynomial degree of the proposed controller.

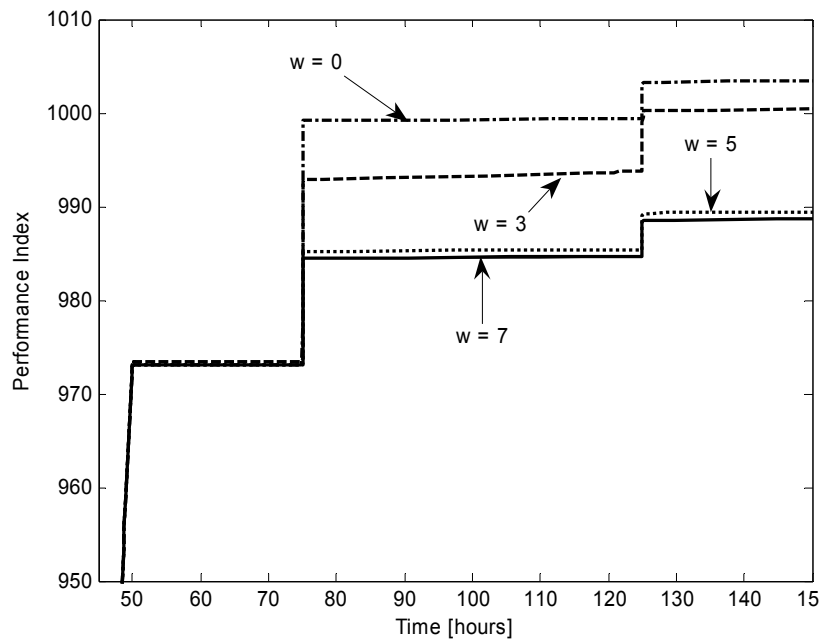


Figure 8. Performance index for several values of the polynomial degree of the proposed controller.

5. Concluding remarks

In this paper, a class of nonlinear controller with polynomial output injection which provides semi-global stabilization under the considered assumptions is proposed. The proposed controller is able to provide adequate performance for regulation and tracking purposes in a better way than a linearizing I/O control law does and it is able to avoid overshoots and setting time, in comparison with the linear PI control. As predicted, when the order of the polynomial increases, a better performance is achieved. No perfect knowledge of the model is required to design the controller which is a valuable feature process application.

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