AN ANALYSIS OF THE IMPACT OF CONTROLLED VARIABLES SELECTION ON THE OPERATION OF ANAEROBIC DIGESTION PROCESSES

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ABSTRACT Different adaptive nonlinear controller designs are evaluated for the operation of the anaerobic digestion process. The selected controlled variables are the Chemical Oxygen Demand (COD), propionate concentration, and dissolved hydrogen concentration. The manipulated variable is the dilution rate. Because the nonlinear control algorithms are designed based on the mass balance equations, distinct assumptions have to be made depending on the choice of the controlled variable. Although the obtained responses are model-specific, the methodology presented can be readily applied to an actual lab scale or industrial scale bioreactor. The study of the time response of the volatile fatty acids and dissolved hydrogen concentration, and of the methane flowrate obtained by controlling one variable at a time reveals the intrinsic behaviour of the balances between the assumed bacterial species and can be useful to investigate for instance the inhibition induced by a given product present in the reactor. The developed methodology demonstrates the usefulness of advanced control techniques to get a better understanding of the process under study even if the controller designs are based on incomplete knowledge.

KEYWORDS Anaerobic Digestion, Adaptive Control, Nonlinear Systems

1. INTRODUCTION

It is generally recognized that the operation of bioreactors for the anaerobic treatment of wastewater is particularly sensitive to input variations such as organic load, influent flowrate, and the presence of toxic material. Procedures for the design, start-up and operation of anaerobic treatment systems to improve their stability are available in the literature (e.g. Harper and Poland, 1986, Hickey et al., 1991, Weiland and Rozzi, 1991). The application of process control techniques may provide an additional way to achieve stabilization while maintaining the highest possible conversion of wastes.

Several simulation and experimental studies have appeared to help understanding what are the consequences of the controlled and manipulated variables selection on the operation of anaerobic bioreactors. Only a few of these are reported here to illustrate the diversity of approaches. Dalla Torre and Stephanopoulos (1986) have proposed several feeding procedures to minimize start-up time. Denac et al. (1988) have suggested a pH control strategy based on the alkaline consumption which appeared to be the most satisfactory performance indicator for their experiments. In Renard et al. (1988, 1991), the control of the COD and the propionate concentration has been studied and applied on a pilot-scale reactor. It was shown that adaptive nonlinear controllers could maintain the controlled variables at their set-point. Various nonlinear adaptive control algorithms based on different modelling assumptions are presented and evaluated in Bastin and Dochain (1990).

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As pointed out by Weiland and Rozzi (1991), it is difficult to compare the various approaches based on the results obtained from different reactors, treating different wastes, with different control strategies and algorithms. It would appear useful to compare the performance of a given control algorithm based on the same model structure for selected controlled and manipulated variables.

This paper presents the design and evaluation of control algorithms for dissolved hydrogen concentration, propionate concentration, and COD by manipulation of the dilution rate. All the controllers use the same information, i.e. the measurement of the controlled variable concentration, the influent substrate concentration, and the methane flowrate. A description of the modelling approach is given first. The necessary steps for the synthesis of the nonlinear adaptive control laws are then presented. The performance of each controller is finally compared by simulation for load changes in the influent substrate concentration.

2. MODEL DESCRIPTION

The five bacterial populations model structure proposed by Mosey (1983) has been used by several researchers (e.g. Jones and Hall, 1989, Costello et al., 1991) to represent adequately the dynamic behaviour of anaerobic bioreactors after step changes in influent concentrations, and hydraulic residence times. These bacterial populations are : the acid formers, propionate-utilizing acetogens, butyrate-utilizing acetogens, acetoclastic methanogens, and hydrogen-utilizing methanogens. The model presented here includes four bacterial populations where the butyrate-utilizing and propionate-utilizing acetogens are merged into the Obligate Hydrogen Producing Acetogens (OHPA).

2.1 Reaction pathways

The following reaction pathways are assumed to describe the anaerobic digestion process :

Acid Formers Bacteria (X_1)

$$k_1S_1 \rightarrow X_1 + k_3S_2 + k_6S_3 + k_8S_4 + k_{11}S_5 \qquad (1)$$

OHPA (X_2)

$$k_2S_2 \rightarrow X_2 + k_6S_3 + k_9S_4 + k_{12}S_5 \qquad (2)$$

Acetoclastic Methanogenic Bacteria (X_3)

$$k_4S_3 \rightarrow X_3 + k_{13}S_5 + k_{14}P_1 \qquad (3)$$

Hydrogen-utilizing Methanogenic Bacteria (X_4)

$$k_7S_4 + k_{10}S_5 \rightarrow X_4 + k_{15}P_1 \qquad (4)$$

where S_1, S_2, S_3, S_4, S_5, P_1 are, respectively, glucose, propionate, acetate, dissolved hydrogen, inorganic carbon, and methane. The yield coefficients are represented by k_i (i=1 to 15).

2.2 Dynamic Model

The dynamic mass balance equations on each bacterial population, substrate and product are readily obtained from equations (1) to (4) and can be formulated under the following general form (Bastin and Dochain, 1990):

$$\frac{d\xi}{dt} = K\varphi - D\xi - Q + F$$

$$\xi^T = \left[\, X_1\ S_1\ X_2\ S_2\ X_3\ S_3\ X_4\ S_4\ S_5\ P_1 \,\right]$$

$$K^T = \begin{bmatrix} 1 & -k_1 & 0 & k_3 & 0 & k_5 & 0 & k_8 & k_{11} & 0 \\ 0 & 0 & 1 & -k_2 & 0 & k_6 & 0 & k_9 & k_{12} & 0 \\ 0 & 0 & 0 & 0 & 1 & -k_4 & 0 & 0 & k_{13} & k_{14} \\ 0 & 0 & 0 & 0 & 0 & 0 & 1 & -k_7 & -k_{10} & k_{15} \end{bmatrix}$$

$$\varphi^T = \left[\, \varphi_1\ \varphi_2\ \varphi_3\ \varphi_4 \,\right]$$

$$F^T = \left[\, 0\ DS_{in_1}\ 0\ DS_{in_2}\ 0\ DS_{in_3}\ 0\ 0\ DS_{in_5}\ 0 \,\right]$$

$$Q^T = \left[\, 0\ 0\ 0\ 0\ 0\ Q_{H_2}\ Q_{CO_2}\ Q_{CH_4} \,\right]$$

3. ADAPTIVE NONLINEAR CONTROLLER ALGORITHMS

The controller algorithm for each variable (dissolved hydrogen, propionate, and chemical oxygen demand) was obtained in three steps : model reduction, feedback linearization, and parameter adaptation.

3.1 Model Reduction

Dissolved Hydrogen Control (S_4)

Assuming that the first three pathways are much faster than the hydrogen-utilizing methanogenic pathway and that methane is a low solubility product, we obtain the following output dynamical equation :

$$\dot{S}_4 = -DS_4 - Q_{H_2} - \theta_{14}Q_{CH_4} + \theta_{24}DS_{in_1}$$

where

$$\theta_{14} = \frac{k_7}{k_{15}}$$

$$\theta_{24} = \left[\frac{k_7k_{14}(k_2k_5 + k_3k_6)}{k_1k_2k_4k_{15}} + \frac{k_8}{k_1} + \frac{k_9k_3}{k_1k_2}\right]$$

It should be noted here that the assumptions necessary to perform model order reduction can be mathematically formalized by using singular perturbation techniques (VanBreusegem and Bastin, 1991). Note also that in practice, the model reduction can be performed here in a quite direct way by simply setting S_1, S_2, S_3 and P_1, and their time derivatives to zero. The assumptions considered for model reduction are used solely for controller design and may not actually reflect the actual reaction kinetics.

Propionate Control (S_2)

The output dynamical for propionate is obtained assuming that the pathways 1, 3, and 4 are faster than pathway 2 and that methane is a low solubility product :

$$\dot{S}_2 = -DS_2 - \theta_{12}Q_{CH_4} + \theta_{22}DS_{in_1}$$

where

$$\theta_{12} = \left(\frac{k_2k_4k_7}{k_4k_9k_{15} + k_{14}k_7k_6}\right)$$

$$\theta_{22} = \frac{k_3}{k_1} + \left(\frac{k_{15}k_4k_2}{k_4k_9k_{15} + k_{14}k_7k_6}\right)\left(\frac{k_8}{k_1} + \frac{k_{14}k_7k_5}{k_1k_{15}k_4}\right)$$

Chemical Oxygen Demand (COD)

Let us define the Chemical Oxygen Demand as :

$$COD = c_1S_1 + c_2S_2 + c_3S_3$$

where c_1, c_2, c_3 are conversion factors from g/L to (g COD)/L. Multiplying the mass balance equations on S_1, S_2, S_3, by c_1, c_2, and c_3 respectively and adding the three equations we obtain :

$$\frac{dCOD}{dt} = (-c_1k_1+c_2k_3+c_3k_5)\varphi_1 + (-c_2k_2 + c_3k_6)\varphi_2 + (-c_3k_3)\varphi_3 + D(c_1S_{in_1} - COD)$$

The output dynamical equation for COD is obtained by assuming that pathways 2, 3, and 4 are faster than pathway 1 and that here again methane is low solubility product :

$$\frac{dCOD}{dt} = -\theta_{11}Q_{CH_4} + D(c_1S_{in_1} - COD)$$

where

$$\theta_{11} = \frac{k_1k_2k_4k_7}{k_7k_{14}(k_2k_5 + k_3k_6) + k_4k_{15}(k_2k_8 + k_3k_9)}$$

3.2 Feedback Linearizing Controllers

The output dynamical equation for each controlled variable can be represented by the following general expression :

$$\frac{dy_j}{dt} = -Dy_j - \theta_{1j}Q_{CH_4} + \theta_{2j}DS_{in_1}$$

If we want to impose first-order stable linear closed-loop dynamics :

$$\frac{dy_j}{dt} = C_1(y_{sp_j} - y_j) \qquad (C_1 > 0)$$

The following control law is obtained :

$$D = \frac{C_1(y_{j_{sp}} - y_j) + \theta_{1j}Q_{CH_4}}{\theta_{2j}S_{in_1} - y_j} \qquad (5)$$

The first term on the right-hand side of Equation 5 can be viewed as a proportional control term where the gain depends on the desired closed-loop time constant, the influent substrate concentration, and the actual level of the controlled output. The second term represents a nonlinear feedforward component on the actual methane gas production. It can be shown (Perrier and Dochain, 1992) that Eq. 5, coupled with the adaptation mechanism described in the next section, includes the key features (proportional and integral action) of a classical PI controller.

3.3 Adaptation mechanism

The parameters θ_{11} for COD control or θ_{2j} (j=2,4) for propionate and hydrogen control respectively were updated based on a Lyapunov approach (Bastin and Dochain, 1990) where the parameter rate of change is proportional to the control error $\tilde{y} = y_{sp} - y$.

$$\frac{d\theta_{11}}{dt} = C_2Q_{CH_4}(y_{1_{sp}} - y_1)$$

$$\frac{d\theta_{2j}}{dt} = -C_2DS_{in_1}(y_{j_{sp}} - y_j) \qquad (C_2 > 0)$$

Defining the system formed of the control error and the parameter error $\tilde{\theta}$, its dynamics is given by :

$$\frac{d}{dt}\begin{bmatrix}\tilde{y}_j\\ \tilde{\theta}_{1j}\end{bmatrix} = \begin{bmatrix}-C_1 & Q_{CH_4}\\ -C_2Q_{CH_4} & 0\end{bmatrix}\begin{bmatrix}\tilde{y}_j\\ \tilde{\theta}_{1j}\end{bmatrix} + \begin{bmatrix}0\\1\end{bmatrix}\frac{d}{dt}\theta_{1j}$$

$$\frac{d}{dt}\begin{bmatrix}\tilde{y}_j\\ \tilde{\theta}_{2j}\end{bmatrix} = \begin{bmatrix}-C_1 & -DS_{in_1}\\ C_2DS_{in_1} & 0\end{bmatrix}\begin{bmatrix}\tilde{y}_j\\ \tilde{\theta}_{2j}\end{bmatrix} + \begin{bmatrix}0\\1\end{bmatrix}\frac{d}{dt}\theta_{2j}$$

The last two equations can be used to select values of the tuning parameters C_1 and C_2.

4. RESULTS AND DISCUSSION

The performance of each controller has been individually evaluated by simulation for step changes in the organic load (S_{in1}). The anaerobic process has been simulated by integration of the mass balance equations with the following yield coefficients, specific growth rate expressions, and initial conditions :

Yield coefficients

k_1=3.2, k_2=1.5, k_3=0.7, k_4=12, k_5=0.27, k_6=0.53, k_7=1.15, k_8=0.6, k_9=0.1, k_{14}=0.3, k_{15}=0.08

Specific growth rate expressions

$$\mu_1 = \frac{\mu_{max_1} S_1}{1 + (S_2+S_3)/K_{I_1}} \qquad \mu_2 = \frac{\mu_{max_1} S_2}{K_{S_2}(1 + S_3/K_{I_2})}$$

$$\mu_3 = \frac{\mu_{max_3} S_3}{K_{S_3} + S_3 + S_3^2/K_{I_3}} \qquad \mu_4 = \frac{\mu_{max_4} S_4}{K_{S_4} + S_4 + S_4^2/K_{I_4}}$$

$$\begin{array}{lll} \mu_{max_1} = 0.2 & K_{I_1} = 10 & \\ \mu_{max_2} = 0.5 & K_{S_2} = 0.4 & K_{I_2} = 0.5 \\ \mu_{max_3} = 0.4 & K_{S_3} = 0.5 & K_{I_3} = 4 \\ \mu_{max_4} = 0.5 & K_{S_4} = 4 & K_{I_4} = 3 \end{array}$$

Initial steady-state conditions

$S_1(0)$=0.79 g/L, $S_2(0)$=0.28 g/L,
$S_3(0)$=0.31 g/L, $S_4(0)$=3.0 μM.

$X_1(0)$=7.6 g/L, $X_2(0)$=3.3 g/L,
$X_3(0)$=0.29 g/L, $X_4(0)$=1.6 g/L.

The performance of each controller was evaluated for an influent substrate concentration (S_{in1}) change from 25 to 35 g/L applied at t=2 days. The closed-loop time constant for each case was chosen to be equal to 1 day (C_1=1). The parameter estimation dynamics is imposed by C_2. C_2 was chosen to be inversely proportional to the square of the regressor (Perrier and Dochain, 1992), i.e. $C_2=C_2'/4Q^2_{CH4}$ for θ_{11} and $C_2=C_2'/4\{(D+D_{min})^2 S^2_{in}\}$ for θ_{2j}. C_2' is set equal to 1 for each case.

The concentration of each substrate and bacterial population is shown in Figure 1a and 1b, respectively, for dissolved hydrogen control. The controller proves to be able to maintain the hydrogen concentration at the desired set-point. The concentration of bacterial populations slowly reaches new levels while the substrates go back to their initial values. Good regulatory performance is also achieved by using the COD controller as shown in Figure 2a and 2b. However, it can be seen in Figure 2a that the dissolved hydrogen concentration reaches a new level (S_4=4.2 μM > $(K_{S4}K_{I4})^{½}$=3.5 μM) which corresponds to an inhibitory value in the Haldane model of μ_4. Results obtained using the propionate controller are very similar to those of the COD controller and are not shown here for limited space reasons (see Perrier and Dochain, 1992, for further details).

It is also interesting to check how the controllers would perform when the step change in inlet substrate concentration is not monitored. As shown in Figures 3a and 3b the substrates and biomass concentration behaviour is only slightly worse than for the case where the change is included in the hydrogen controller. The adaptation of parameter θ_{24} (Figure 3c) is able to account for the change. The COD controller is not able to reject the disturbance as efficiently because the dilution rate is not reduced sufficiently rapidly as can be seen from Figure 4a. As a consequence, hydrogen will tend to accumulate and will inhibit the methane production (Figure 4b). Again, similar results are obtained by using the propionate controller.

5. CONCLUSIONS

A general nonlinear controller structure has been proposed for dissolved hydrogen, propionate and COD. The control law has been derived by assuming a model structure based on four bacterial populations. In addition to the measurement of the chosen controlled variable, the controller requires the knowledge of the methane flowrate and the influent substrate concentration.

Simulation results have shown that each controller was able to maintain the controlled variable at the desired setpoint after a step change in influent substrate concentration. At this point, it has been found out that after an unmeasured organic load change, only the hydrogen controller was able to recover quickly.

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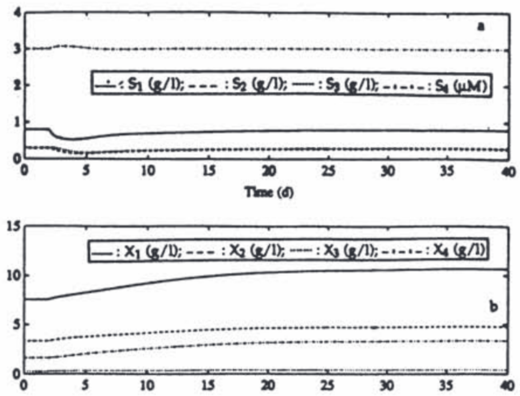

Fig.1. Adaptive linearizing control of H_2

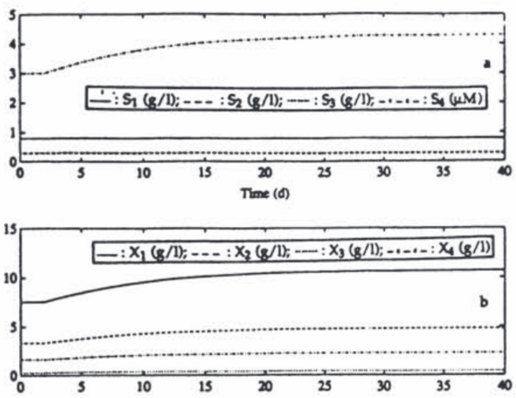

Fig.2. Adaptive linearizing control of COD

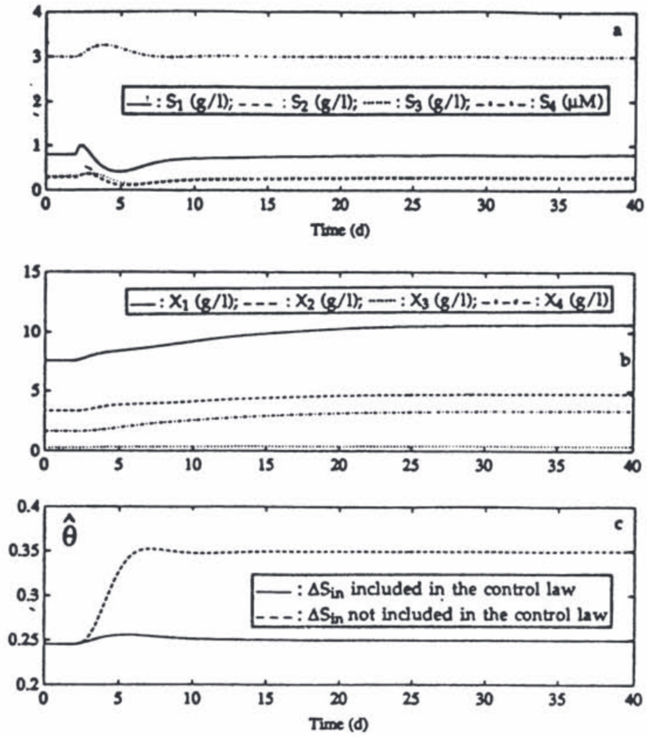

Fig.3. Adaptive linearizing control of H_2
(when ΔS_{in} is not included)

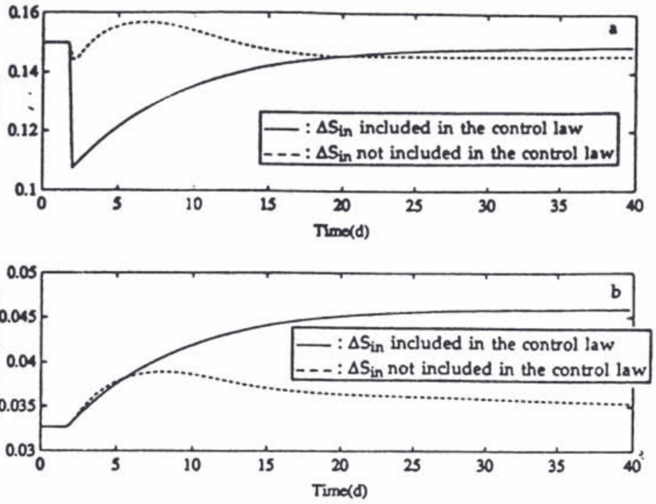

Fig.4. Comparison of the COD control with
or without including ΔS_{in}