Adaptive Control of Anaerobic Digestion Processes — A Pilot-Scale Application

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A simple adaptive control algorithm, for which theoretical stability and convergence properties had been previously demonstrated, has been successfully implemented on a biomethanation pilot reactor. The methane digester, operated in the CSTR mode was submitted to a shock load, and successfully computer controlled during the subsequent transitory state.

INTRODUCTION

The need for computer control in biotechnological processes in order to improve product quality or to optimize production efficiency is becoming more and more obvious. For the last two decades, it has been the subject of much research work and has resulted in numerous scientific publications. For a survey on the topic, see ref. 1, while typical control examples can be found in refs. 2 and 3.

The computer control of variables such as temperature or pH has become a routine. But real-life control applications of the biological variables such as biomass substrate, S, active biomass, X, or product, P, are developing slowly. There are two main reasons for this. First, basic biological processes underlying biotechnologies, and their dynamics, only begin to be well understood. They involve living organisms, the dynamical behaviour of which is strongly nonlinear and nonstationary. Secondly, there is a lack, in most cases, of cheap sensors capable of providing reliable on-line measurements of the biological and biochemical parameters required to implement high performances computer control strategies.

The available strategies for computer control of the biological variable such as S, P, or X can be roughly classified into three categories. Some consider extensive biological models of the process including, for the fermentation parameters such as the specific growth rate, $\mu(t)$, structured relations with fixed coefficients. A second category of control schemes is based on linear "black-box" models of the system. However, because of the nonlinearity and the nonstationarity of the underlying process, it is not possible to emphasize any stability or convergence property for these control algorithms, and their performances are difficult to evaluate.

Our control scheme belongs to a third category, and has been described earlier.[4] It refers to the most recent research trends in adaptive control theory.[5,6] It is based on a simple nonlinear mathematical model of the fermentation process, and does not require any analytical expression for the specific growth rate, $\mu(t)$. Moreover, it has been proven stable and convergent.[4] It will be explained below in simple mathematical terms. In this article, we present the real-life validation of our control algorithm on an anaerobic digestion process conducted in the CSTR mode and which was being submitted to shock loads.

MATERIALS AND METHODS

The Biological Process of Methanogenesis

It has long been known that organic compounds can be metabolized in anaerobiosis down to a mixture of methane and carbon dioxide.[7] It is commonly considered as a two-step biological process: fermentation (or acidogenesis) and methanogenesis. In the first step, organic compounds are fermented usually to volatile fatty acids by a group of acidogenic bacteria. In the second step, the methanogenic archae-bacteria convert the products of acidogenesis, into methane, CH_4, and carbon dioxide, CO_2. Recently, the bio-

logical process has appeared more complicated.[8] However, the two-step picture will suffice for our purpose.

If, as it is often assumed, the methanogenic step is considered to be rate-limiting, the following state space representation based on mass balances can be used to describe the dynamical behavior of a continuously fed, completely mixed anaerobic digestion reactor.[9] The net accumulation of active biomass in the reactor is

$$\frac{dX(t)}{dt} = \mu(t)X(t) - D(t)X(t) \qquad (1)$$

The net accumulation of substrate biomass in the reactor is

$$\frac{dS(t)}{dt} = -k_1\mu(t)X(t) + D(t)S_{in}(t) - D(t)S(t) \qquad (2)$$

The rate of methane gas (product) production is

$$Q(t) = k_2\mu(t)X(t) \qquad (3)$$

where

$X(t)$ is the methanogenic bacterial (active biomass) concentration;

$S(t)$ is the substrate biomass concentration in the effluent (g COD/L);

$S_{in}(t)$ is the substrate biomass concentration in the influent (g COD/L);

$Q(t)$ is the methane gas (product) production rate (L CH$_4$/h);

$D(t)$ is the dilution rate (day^{-1});

$\mu(t)$ is the specific growth rate (day^{-1}); and

k_1 and k_2 are the yield coefficients.

The specific growth rate, $\mu(t)$, is known to be a complex function of many physicochemical and biological environmental factors, such as the substrate biomass concentration, S, the active biomass concentration, X, and the pH or the temperature, and many different analytical expressions have been suggested to account for these factors in anaerobic waste treatment processes.[7,10-14]

It is worth remembering that our adaptive control algorithm[4] remains valid for any specific growth rate, $\mu(t)$, provided that it fulfills the following two mild and realistic assumptions. First, the specific growth rate $\mu(t)$ is positive and bounded, for all values of t: $0 \leqslant \mu(t) < \mu^*$, where μ^* is the maximum specific growth rate. Secondly, there is no growth without substrate: $\mu(t) = 0$ when $S(t) = 0$.

Design and Operation of the Methane Digester

The anaerobic digestion process was conducted in a pilot-scale methane reactor of 60 L working volume. The reactor is constructed from standard glass parts used for the construction of chemical reactors (Corning, Stone, England).

The loading substrate consists of spent liquor from citric acid fermentation, made available by La Critique Belge (Tienen, Belgium) as a 50% (volatile solids) syrup. The syrup is diluted to the appropriate concentration by a solution of NaHCO$_3$ (Solvay, pure) in softed water in order to avoid precipitation of CaCO$_3$ (softener YM 515, Belgian Water Systems). This reconstitutes a typical fermentation industry wastewater, the composition of which will be quantitatively characterized as COD.

The methane reactor is loaded continuously. It is mechanically stirred with a two-bladed propeller double helix so as to operate in the CSTR mode. Moreover, the mixed liquor is continuously recirculated from the bottom to the top of the reactor by a volumetric membrane pump, type B.04.017N (CfG Prominent, Heidelberg, Germany), with a flow rate of 10 L/h, to avoid solids accumulation at the bottom of the reactor. The reactor is maintained at the constant temperature of 35°C by an external electrical heating coil Stabilag (Etirex, Soissons, France). Figure 1 shows the draft of the reactor and a schematic view of the whole data acquisition and monitoring system. The microcomputer is a Z80 (Furrer-Gloor, Zurich, Switzerland).

Volumetric membrane pumps LMI, type A523 (Mecaflo, Acton, MA), are used to load the digester. Their electronic systems have been modified to accept external electric impulses. The microcomputer is able to send given numbers of impulses, each of which corresponds to one stroke of the pump, i.e., one filling and emptying of the membrane chamber. The volume of this chamber can be mechanically adjusted for calibration purpose. A special chamber for viscous liquids was chosen for the syrup.

Monitoring of the methane reactor is carried out by a dedicated data acquisition system of on-line sensors which provide the following measurements: the gas flow rate, the percentage (v/v) methane and carbon dioxide in the gas, pH, and temperature (for full details, see ref. 15).

The gas flow rate is measured with a home-modified type 1 volumetric gas meter (Contigea Schlumberger, Dordrecht, Holland). Eight reed-relays are equidistantly stuck on the gas meter dial. Each time the pointer, on which a permanent magnet has been fixed, passes before a reed relay, an impulse signal is sent to the microcomputer and the production of one eighth L gas is recorded. Percent methane and carbon dioxide are made available through gas analyzers type Ultramat 1 (Siemens, Karlsruhe, W. Germany) by specific infrared absorption. The pH is measured in the recirculation circuit, with a combined 405.DXK.K8 equipped with a jellied electrolyte–Xerolyt system (Ingold, Zurich, Switzerland), especially worked out by Ingold for steady measurements in anaerobic, low redox, S^{2-} containing mixed liquors. Temperature is measured by an adjustable glass thermometer Jumo DBP, type MS 121 S (Juchweim, Fulda, W. Germany).

As on-line measurements of COD of both the influent and effluent biomass substrate concentrations were found unreliable for the purpose of these experiments, COD measurements were carried out manually every two hours, namely the time necessary to perform the analysis, and manually stored in the microcomputer.
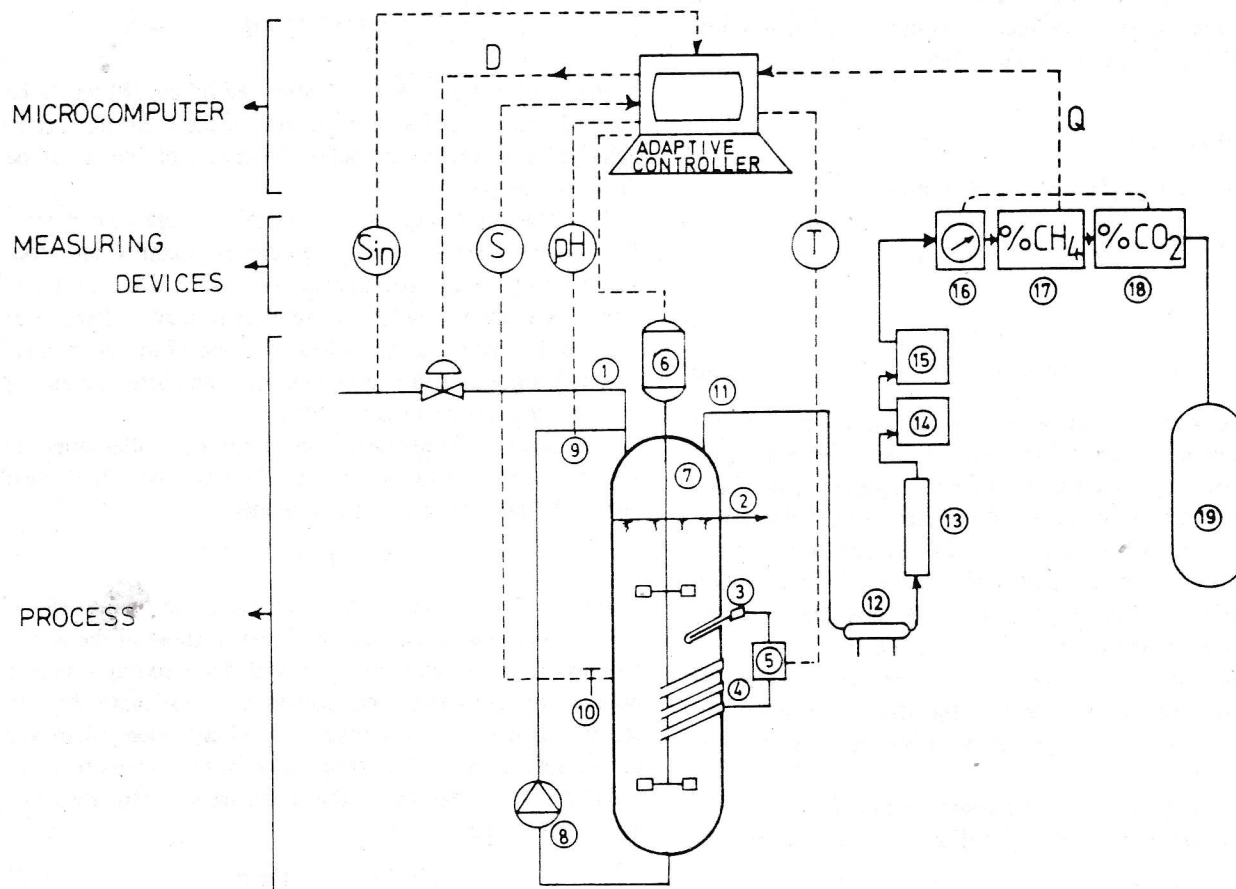
Figure 1. Flow diagram of the methane reactor and the data acquisition and monitoring system: (1) feed inlet, (2) effluent outlet, (3) contact thermometer, (4) electrical heating coil, (5) regulation, (6) stirring system, (7) propeller shaft, (8) recirculation pump, (9) pH electrode, (10) mixed liquor sampler, (11) gas outlet, (12) condenser, (13) H₂S filter, (14) bubbler, (15) water trap, (16) volumetric gas meter, (17) CH₄ gas analyzer, (18) CO₂ gas analyzer, and (19) gas holder.

Methods

The chemical oxygen demand (COD) was determined by the sulfuric-acid–dichromate method,[16] modified according to Leithe.[17] The reflux heating time is reduced from 2 h to 10 min. The boiling temperature of the reacting mixture is correspondingly increased by an increase of the concentration of sulfuric acid from 50 to 57% (v/v). The COD analyses were performed on soluble samples obtained after centrifugation at $1.2 \times 10^4 g$ for 20 min in a SS-3 automatic centrifuge (Sorvall, Newtown, MA), and subsequent filtration of the decanted supernatant on a membrane filter of 0.45-μm pore diameter (Sartorius, Göttingen, Germany), according to the ISO norm DIS 6060.2. Since the COD analysis requires diverse laboratory manipulations, each of them possibly eliciting measurement errors, samples were analyzed in triplicate. The dispersion of the COD values was kept in this way, in all cases below 2%, usually below 0.5%. Volatile fatty acids were determined after acidification with one volume of metaphosphoric acid [25% (v/v) in water] for five volumes of sample by gas liquid chromatography.[18]

RESULTS

Control Scheme

Statement of the Control Objective

If we consider a process similar to that used for waste-water treatment in food industry, then the control objective is to regulate the output pollution level, S, in the effluent (as COD), at a prescribed level S^*, despite fluctuations of the influent pollution level, S_{in}, by acting on the hydraulic loading rate, and hence on the dilution rate, D.

Outline of the Control Scheme

The working of the control law can simply be described as follows (Fig. 1). Every two hours, measurements of the influent and effluent waste concentration, $S_{in}(t)$ and $S(t)$ (available through COD analysis), and of the methane gas production rate, $Q(t)$ (available as direct measurement), are considered by the controller in order to compute a new value

of the control input, namely the dilution rate, $D(t)$, which will be imposed to the bioreactor as hydraulic loading rate during the following two-hour period.

The Control Law

Introducing eq. (3) into eq. (2) yields

$$\frac{dS((t)}{dt} = -KQ(t) + D(t)S_{in}(t) - D(t)S(t) \tag{4}$$

with

$$K = k_1/k_2 \tag{5}$$

As can be seen, we now have a dynamical equation in the substrate concentration, $S(t)$, in which the variable $X(t)$, the measurement of which is difficult to apprehend, has been replaced by a measurable on-line variable, $Q(t)$. Moreover the specific growth rate $\mu(t)$ has disappeared from the dynamical equation in $S(t)$. Note also that the yield coefficient, K, is a conversion factor of the biomass substrate, $S(t)$, into the produced methane gas. Equation (4) is the basis for the derivation of the control law. For this purpose, it also expresses conveniently the control objective: to maintain $S(t)$ as constant as possible, irrespective of variations of $S_{in}(t)$ by acting on $D(t)$.

Let us decide to handle the control input, $D(t)$, using as a tool in automatic control the following nonlinear expression:

$$D(t) = \frac{C_1(t)[S^* - S(t)] + KQ(t)}{S_{in}(t) - S(t)}, \qquad C_1(t) > 0 \tag{6a}$$

where $C_1(t)$ is a design parameter, the choice of which will be discussed below.

The control law[4] has been deduced by external linearization techniques,[19] which consist in introducing nonlinearities of the system into the control scheme in order to obtain a linear kinetics for the closed-loop system (i.e. the process plus the controller). In fact, introducing eq. (6a) into eq. (4) yields the following closed-loop dynamical equation:

$$\frac{dS(t)}{dt} = C_1(t)[S^* - S(t)] \tag{7}$$

i.e. a first-order linear kinetics which, moreover, is unconditionally stable: the substrate biomass concentration, $S(t)$, converges to the prescribed level S^* with a rate equal to $C_1(t)$, irrespective of any fluctuation of the influent substrate biomass concentration, $S_{in}(t)$, for instance.

Since the yield coefficient, K, is assumed to be *a priori* unknown, it will be replaced in eq. (6a) by an on-line estimate $\hat{K}(t)$:

$$D(t) = \frac{C_1(t)[S^* - S(t)] + \hat{K}(t)Q(t)}{S_{in}(t) - S(t)} \tag{6b}$$

Parameter $\hat{K}(t)$ is estimated on-line as a function of time, i.e., updated by the following equation, to be considered once again as a tool in automatic control:

$$\frac{d\hat{K}(t)}{dt} = C_2 Q(t)[S^* - S(t)]; \qquad C_2 > 0 \tag{8}$$

In this control study, $\hat{K}(t)$ as estimated by eq. (8) has to be basically considered as a parameter of the controller. Here again, C_2 is a design parameter, the choice of which will be discussed below.

The structure of eq. (8) is classical in adaptive control.[20] It has been chosen so as to guarantee stability and convergence of the control algorithm.[4] It should not be interpreted as a kinetic law for the yield coefficient K. Parameter $\hat{K}(t)$ can be viewed as an estimated value of the yield coefficient K only in stationary conditions, and after the end of the convergence process for $\hat{K}(t)$.

The practical implementation of the controller imposes bounds on the control input: the dilution rate, $D(t)$, must clearly be positive and upper bounded:

$$0 \le D(t) \le D_{max} \tag{9}$$

In practice, a reasonable choice for D_{max} will be the critical value of the dilution rate at which the washout of the active biomass from the completely mixed, once-through reactor occurs. Furthermore, the updating of estimate $\hat{K}(t)$ is stopped, and $\hat{K}(t)$ used at its unchanged last value, whenever the computed value of $D(t)$ by eq. (6b) reaches its bounds. This ensures, in particular, that estimate $\hat{K}(t)$ remains positive at all times:

$$\hat{K}(t) > 0; \qquad \text{for all } t \tag{10}$$

Since measurements of S are not instantaneous, a discretized version of the control algorithm represented by eqs. (6b), (8), and (9) must be considered in the real-life implementation. Here, we have chosen to use a simple first-order Euler approximation for $d\hat{K}/dt$:

$$\frac{d\hat{K}}{dt} \rightarrow \frac{\hat{K}_{t+1} - \hat{K}_t}{T} \tag{11}$$

where t is the time index ($t = 0, 1, 2, 3 \ldots$) and T is the sampling period, here equal to two hours, i.e., the duration of the COD analysis.

Hence, eq. (8) becomes:

$$\hat{K}_{t+1} = \hat{K}_t + TC_2 Q_{t+1}[S^* - S_t] \tag{12}$$

where $\hat{K}_{t+1}$ is the value of $\hat{K}$ estimated at present time from the value of $\hat{K}_t$, namely, the value of $\hat{K}$ estimated two hours earlier. The first value of $\hat{K}_t$ at $t = 0$ is arbitrarily chosen by the user and its practical choice will be discussed below. Parameter S_t is the value of COD in the effluent measured two hours earlier, and only available two hours later, and Q_{t+1} is the methane gas production rate over the preceding two hours.

Similarly, eq. (6b) becomes:

$$D_{t+1} = \frac{C_{1,t+1}[S^* - S_t] + \hat{K}_{t+1} Q_{t+1}}{S_{in,t} - S_t} \tag{13}$$

It is the introduction of the on-line estimate, $\hat{K}(t)$, which gives its adaptive feature to the control law (6b), (8), and

(9). The updating eq. (8) has two advantages: apart from the fact that no precise *a priori* value of K must be known, it allows also to track long-term variations of this parameter (e.g. due to "nonobservable physiological or genetic events"[21]).

Besides, it is worth noting that, for the control algorithm constituted by eqs. (6b), (8), and (9), theoretical stability and convergence properties have been demonstrated.[4] These properties guarantee, first of all, that the overall system (process plus controller) represented by eqs. (4), (5), (6b), (8), and (9) is stable. In particular, this means that the controller is able to stabilize an unstable reactor which would be led, without the controller, to a washout.[22] Secondly, it guarantees that the regulation error $[S^* - S(t)]$ tends asymptotically to a value close to zero. This value is even equal to zero in the free-disturbance case, i.e., when for instance, unnoisy measurements of $S(t)$ are available.

Experimental Validation of the Control Scheme

Two different adaptive control experiments were carried out. The first one consisted to apply a square influent substrate concentration charge. The second one consisted to apply a step of the influent substrate concentration.

First Experiment

During the three weeks before the experiment, the reactor is loaded at the volumetric loading rate, B_V, of 2 g COD/L day. The mean retention time, θ, is 15 days. These values correspond to a substrate concentration, S_{in}, of 30 g COD/L and a dilution rate, D, of 0.067 day^{-1}. The process remains in steady state and is characterized by a gas production rate, $r_{V\,gas}$, of 1 L gas/L reactor day (L gas/L day) with 52% CH$_4$ and a conversion, $Y_{CODr/CODo}$ (r for removed and o for influent), of 70%. The pH is 7.5. The total volatile fatty acid concentration, TVA, is 3.5 g COD/L, i.e., an acetate concentration, VA_2, of 2.6 g COD/L and a propionate concentration, VA_3, of 0.7 g COD/L.

The prescribed level, S^*, of 8.9 g COD/L is chosen from the mean value of the effluent COD measurements over the preceding days. The following procedure is applied to choose the initial value of $\hat{K}$. Since the process is in steady state, the time derivative, dS/dt, in eq. (4) is equal to zero. Parameter K is then the only unknown in eq. (4) and therefore, $\hat{K}_0$ can be set to:

$$\hat{K}_0 = \frac{D(S_{in} - \overline{S})}{\overline{Q}} \qquad (14)$$

where $\overline{S}$ and $\overline{Q}$ are the mean values of S and Q over the preceding days, i.e. 8.9 g COD/L and 1.3 L CH$_4$/h (0.5 L CH$_4$/L day) respectively. Here, $\hat{K}_0$ has been set to 1.09 (g COD h/L CH$_4$ L day). The upper bound on the dilution rate, D_{max}, and the design parameters $C_1(t)$ and C_2 have been chosen as follows (Table I):

$$D_{max} = 0.4 \text{ day}^{-1};$$

$$C_1(t) = C_1 Q(t) \quad \text{with } C_1 = 5 \text{ h/L CH}_4 \text{ day}; \quad \text{and}$$
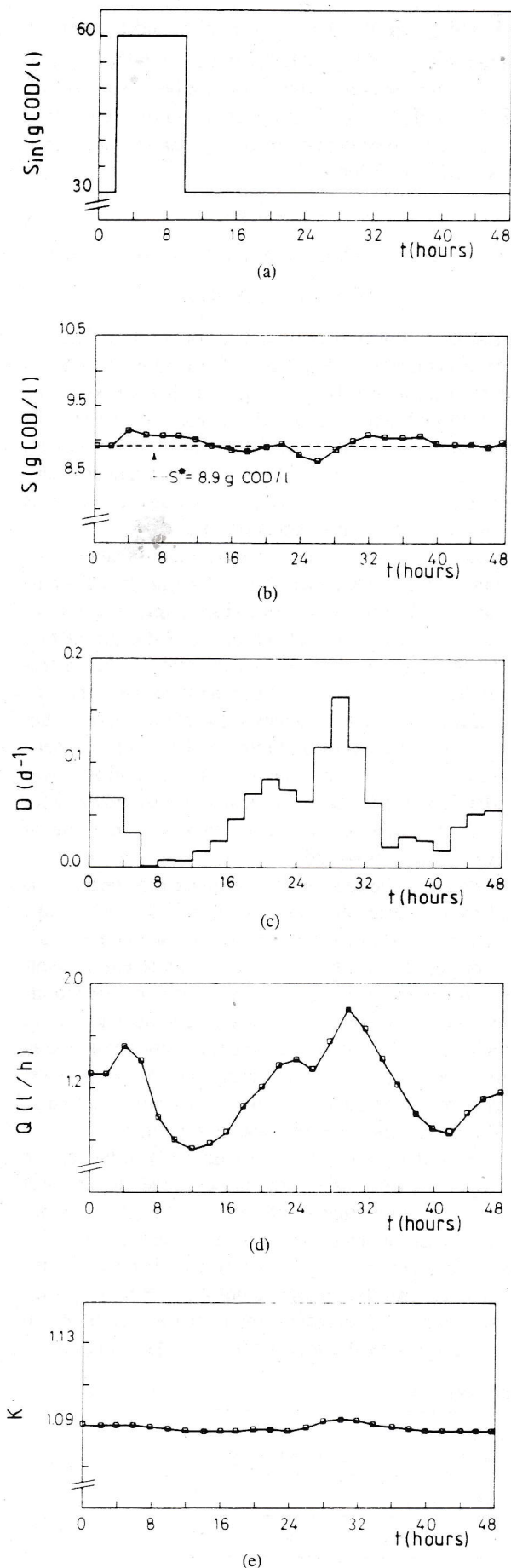
$$C_2 = 0.1 \text{ h}^2/\text{L}^2 \text{ CH}_4 \text{ day}^2.$$

In this adaptive control experiment, a square (8 h) influent substrate concentration, S_{in}, from 30 to 60 g COD/L, is applied to the process [Fig. 2(a)]. For this purpose, the reactor is loaded with syrup, which has been previously diluted at the concentration of 60 g COD/L. The vessel, which contains the loading solution, is maintained at a temperature of 4°C. The loading pump is calibrated at the value of 1.23 mL/stroke, with a precision of 2%.
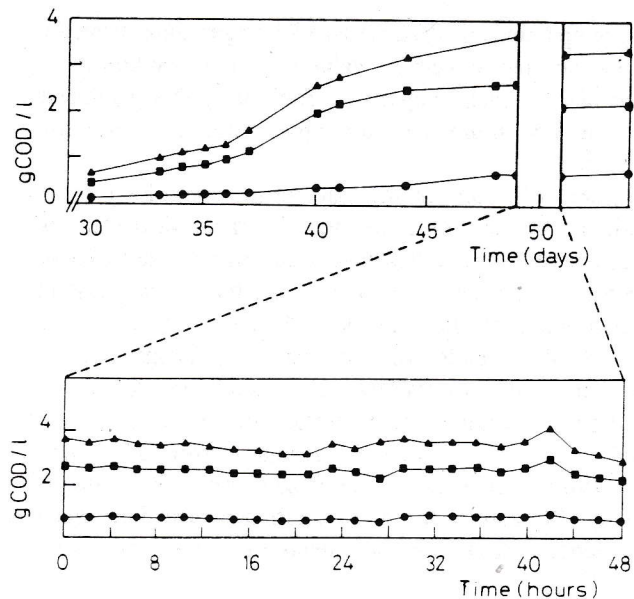
Every two hours, a sample of mixed liquor is taken from the middle of the reactor (see Fig. 1) and analyzed for its COD content. The result is introduced into the microcomputer. The mean value of the methane gas production rate, $Q(t)$, over the two hours is available on the microcomputer. The COD result and the mean value $Q(t)$ are used by the controller to calculate a new value of the dilution rate, $D(t)$. The results are shown on Figure 2. The control experiment lasted 48 h. Following this period, the controller was switched off and the COD of the effluent measured daily for some days. No significative deviation from the value of 8.9 g COD/L was observed.

The stability of the reactor over a longer time period was assessed by the concentration in total volatile fatty acids, and more particularly in acetate and propionate in the mixed liquor, before, during and after the control experiment. Figure 3 shows these evolutions. All three concentrations increased progressively in the days preceding the control experiment. However, as well during the control experiment and thus with the controller as during the following days, and thus without the controller, no significant variation in any of the three concentrations was observed.

As expressed by eq. (6b), the calculated value of the control input, D, fluctuates with respect to the variations of S, S_{in}, Q, and $\hat{K}$. In particular, D varies as the inverse of S_{in}. Therefore, when an increase of S_{in} is detected (i.e. after 2 h), D is reduced in a way which directly depends on the variation of S_{in} (and the feeding is not just simply stopped). Conversely, when S_{in} decreases, the increase of D is modulated with respect to the amplitude of S_{in} [see Fig. 2(c)].

Table I. Values chosen for the parameters of both control experiments.

Controller parameters	S^* (g COD/L)	$\hat{K}_0$ (g h/L CH$_4$ L day)	D_{max} (day^{-1})	$C_1(t)$ (day^{-1})	C_2 (h^2/L^2 CH$_4$ day^2)
First experiment	8.9	1.09	0.4	5	0.1
Second experiment	3.4	0.94	0.4	5 $Q(t)^a$	5

[a] See the Discussion section.

Figure 3. Stability of the reactor before, during, and after the first experiment: (▲) concentration in total volatile acids, (■) concentration in acetate, and (●) concentration in propionate.

Under the influence of the controller, the variations of the controlled variable, namely the effluent COD, S, are small [Fig. 2(b)], while large oscillations of the dilution rate, D [Fig. 2(c)], and of the methane gas production rate, Q [Fig. 2(d)] appear. The estimate $\hat{K}$ [Fig. 2(e)] does not vary much.

Second Experiment

During the four weeks before the experiment, the running conditions of the reactor, i.e. the volumetric loading rate, B_V, and the mean retention time, θ, were 1 g COD/L day and 18 days, respectively. These values correspond to a substrate concentration, S_{in}, of 18 g COD/L and a dilution rate, D, of 0.054 day^{-1}. Here, again, the process remains in steady state and is characterized by a gas production rate, $r_{V\ gas}$, of 0.55 L gas/L day (61% CH$_4$) and a conversion, $Y_{CODr/CODo}$, of 80%. The pH is 7.35. The total volatile fatty acid concentration, TVA, is less than 0.5 g COD/L.

The same procedures as above are used to choose the values of S^* and $\hat{K}_0$, 3.4 g COD/L and 0.94 (g COD h/L CH$_4$ L day), respectively, with S and Q equal to 3.4 g COD/L and to 0.84 L CH$_4$/h (0.33 L CH$_4$/L day). The following values are chosen for the upper bound on the dilution rate, D_{max}, and the design parameters, $C_1(t)$ and C_2 (Table I):

$$D_{max} = 0.4\ \text{day}^{-1};$$

$$C_1(t) = 5\ \text{day}^{-1}; \quad \text{and}$$

$$C_2 = 5\ \text{h}^2/\text{L}^2\ \text{CH}_4\ \text{day}^2.$$

In this adaptive control experiment, a step of the influent substrate concentration, S_{in}, from 18 to 30 g COD/L, is applied to the process. The results are shown in Figure 4. The step in charge was maintained for 44 h, the duration of the experiment, after which time the controller was switched

Figure 2. Adaptive control with a square influent substrate concentration charge (first experiment).
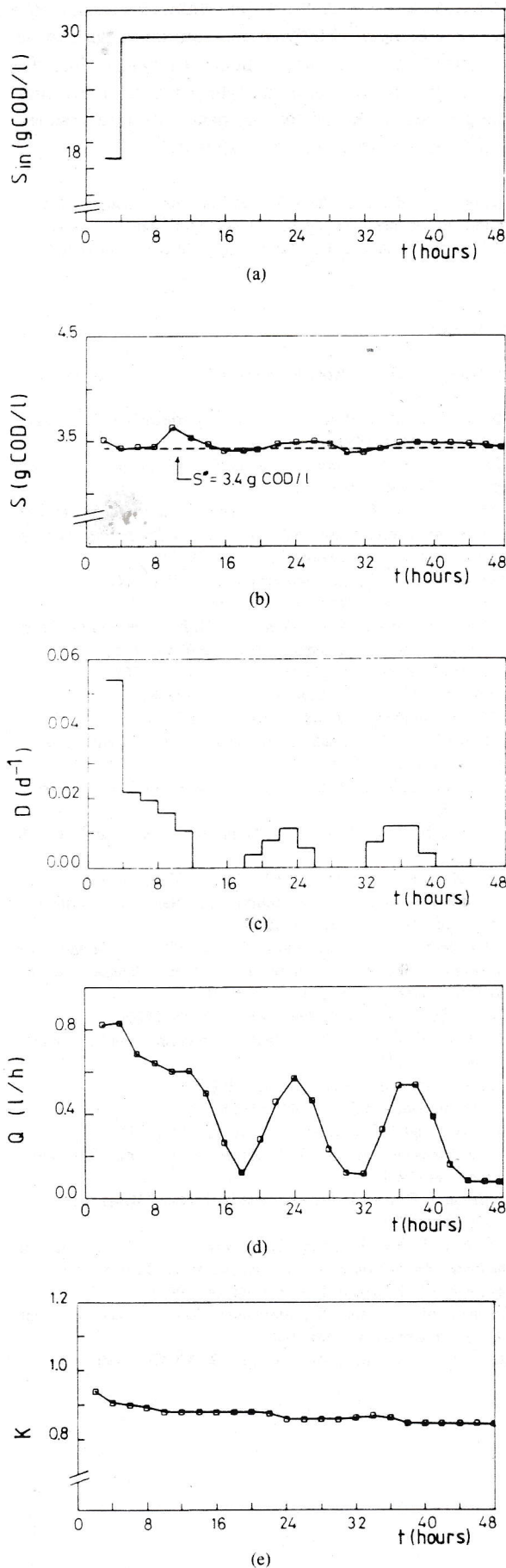
Figure 4. Adaptive control with a step of the influent concentration (second experiment).

off and the volumetric loading rate returned to its initial value. During the following days, the COD of the effluent was measured daily. No significative deviation from the value of 3.4 g COD/L was observed.

Remarks similar as for the first experiment can be drawn here. The variations of the calculated dilution rate, D, are modulated with respect to S, S_{in}, Q, and also $\hat{K}$. Parameter D is reduced when the step of S_{in} was detected. The variations of the estimate $\hat{K}$ remain unimportant.

DISCUSSION

As clearly shown in Figures 2 and 4, the real-life implementation and validation of the control algorithm, eqs. (12), (13), and (9), on a pilot-scale anaerobic digester has been successful. The control objective, as defined above, has been achieved, and the performances of the controller are remarkable: the regulation error $(S^\star - S)$ has been kept very low in both adaptive control experiments (below 2.7 and 6.7%, respectively) [Figs. 2(b) and 4(b)]. The advantages of the control are twofold here. The controller has prevented large variations of the substrate biomass concentration, S. Moreover, without the controller, variations of the influent substrate biomass concentration such as those applied in both experiments, could have led the process to a washout. In both cases, the reactor remained stable under the action of the controller. Figure 3 substantiates this in the case of the first experiment by showing the stability of the fatty acid concentrations in the days following the control experiment.

This was realized with a quite nervous control action, in order to compensate the important disturbance applied to the process through the large variations of the influent substrate concentration, S_{in}. Figures 2(c) and 4(c) show indeed large variations of the dilution rate D. Note also that the methane gas production rate, Q, closely followed the variations of D [Figs. 2(d) and 4(d)], and the estimate $\hat{K}$ did not change much [Figs. 2(e) and 4(e)].

Such a good control performance was achieved despite the delay imposed to the control decision by the duration of the COD analysis. These performances would undoubtedly be improved with more frequently available measurements. However, the control algorithm proved very robust with respect to such a constraint.

The process chosen, namely the biomethanation of spent liquor from citric acid fermentation, is at the same time a process stable in the steady-state conditions selected, yet a process very unstable as soon as running parameters are modified. Indeed, it shows a deficience in iron, which was not corrected. It had been previously shown[23] that in the CSTR mode, the volumetric loading rate, B_V, of 3 g COD/L day is the maximum value for reliable running of the digester. At this loading rate, because of the iron deficiency, the aceticlastic bacteria work at a low rate and acetate accumulates up to point. Besides, it has been long established as a rule of the thumb[24] that the volumetric loading rate should be doubled only with 20% in increases distributed over one mean hydraulic residence time.

The choice of the upper bound on the dilution rate, D_{max}, and of the design parameters, $C_1(t)$ and C_2, has been the object of careful preliminary simulation studies on a PDP 11/34 minicomputer (Digital, Maynard). Table I summarizes the control parameters. The chosen D_{max} ($D_{max} = 0.4$ day^{-1}) approximately corresponds to the maximum value of the specific growth rate usually reported to the anaerobic digestion process.

Two types of the design parameter $C_1(t)$ were tried in our control experiments: $C_1(t) = C_1 Q(t)$, with C_1 being a positive constant ($C_1 > 0$) in the first experiment and $C_1(t) = C_1$, i.e. C_1 is a positive constant ($C_1 > 0$) in the second experiment. In fact, our control scheme has proven convergent for these two choices. Furthermore, as already mentioned above, the design parameters $C_1(t)$ and C_2 have an important influence on the rate of convergence of the control algorithm. The time constant corresponding to the chosen design parameters are much shorter than the mean retention time.

It is usually considered that if the value of the yield coefficient K changes, its variation is a slow and a long-term variation. This explains the choice of a low value, in the first experiment, for C_2 ($C_2 = 0.1$). Nevertheless, a larger value of C_2 ($C_2 = 5$) was also successfully implemented in the second experiment.

The initial value of $\hat{K}_0$, in both experiments ($\hat{K}_0 = 1.09$ and $\hat{K}_0 = 0.94$) are in accordance with a rough estimate (equal to 1.15) which can be deduced from a simple line of reasoning on the conversion of the substrate into methane gas in an anaerobic digestion process.[15]

Our controller had to be implemented by using off-line COD measurements. The positive control results obtained highlight the need to develop reliable (and inexpensive) on-line sensors which would allow a completely automatic implementation of such control schemes.

CONCLUSIONS

In this article, we have shown the results of a successful computer control application to an anaerobic digestion pilot plant. An adaptive control algorithm of the effluent biomass substrate concentration has been implemented which takes advantage of the nonlinear structure of the system. Moreover, this algorithm does not require any analytical expression for the specific growth rate, μ, and has been previously proven stable and convergent.

It is worth noting that the presented control law is independent of the form of the equation(s) describing the process [in this case. eq. (4)] and hence of any assumption(s) made concerning methanogenesis. Other starting equations will lead through an analogous line of reasoning to similar adaptive control schemes.[25] Furthermore, similar ideas can be used to estimate on-line biochemical parameters (like the specific growth rate, μ[26]) and variables (like the active biomass concentration, X, or the biomass substrate concentration, S[25]) or for other control purposes.[6,26]

References

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