Improving the operational stability in continuous anaerobic digestion processes

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Abstract— **In this paper, the robust regulation of volatile fatty acids (VFA) and total alkalinity (TA) is addressed to improve the stability of continuous anaerobic digestion (AD) processes. The control scheme is derived from an AD model that includes TA as state variable and it is composed of a model-based multiple-input multiple-output feedback control and an extended Luenberger observer which not only allows the estimation of the process kinetics and variations in the influent composition but also the introduction of an anti-windup structure. The wastewater flow rate is used to regulate the VFA concentration, whereas an alkali solution is added directly to the digester to regulate TA. The controller performance is evaluated via numerical simulations showing excellent responses under the influence of control input saturations, noisy measurements, load disturbances and uncertain kinetics. Finally, it is shown that well-known practical stability criteria for AD processes can be also fulfilled by the proposed control scheme © UdG-AMCA.**

Key words: Robust Control, Anaerobic Digestion, Wastewater Treatment.

I. INTRODUCTION

Anaerobic digestion (AD) has regained the interest of the wastewater treatment scientific and industrial community to reduce the organic matter from industrial and municipal effluents because of its low initial and operational costs, high organic removal efficiency and low sludge production, combined with a net energy benefit through the production of biogas. However, its widespread application has been limited, because of the intrinsic difficulties involved in achieving the efficient operation of this processes such as: i) highly nonlinear behavior; ii) load disturbances; iii) system uncertainties; iv) constraints on manipulated and state variables; and v) limited on-line measurement information. Recently, Steyer et al. (Steyer *et al.*, 2006) have reported a summary of the advantages and drawbacks of several control schemes when applied to AD processes, and pointed out that most of the model-based control schemes that have been applied for AD stabilization, may appeared to be limited because they have been focused on the regulation of a single substrate without taking into account that certain operating conditions may impose limitations on the process physiological properties (Hill *et al.*, 1987).

From an operating safety viewpoint, anaerobic digestion is intrinsically a very unstable process, variations of the input variables (hydraulic flowrate, influent organic load) may easily lead the process to a washout (Bailey y Ollis, 1986). This phenomenon takes place under the form of volatile fatty acids (VFA) accumulation which induces an overflow of protons that decompose the liquid phase biocarbonates to produce CO_2 (increasing its composition in the gas phase) and decreasing the digester pH. However, pH can only be used as an indicator of the process stability in wastewaters with low buffering capacity, because high biocarbonate concentrations may compensate the pH changes due to VFA's accumulation (Rozzi, 1991). Thus, instead of pH, it is preferable to monitor the process alkalinity, which allows the timely detection of changes of the process buffer capacity and, as a consequence, a more accurate information about the risk of process failure by the VFA accumulation. In this regard, Pohland (1962) emphasized the need for a balance between total alkalinity (TA) and VFA concentrations for normal digestion and implied that appreciable variations occurred only after consistent variations of the ratio VFA/TA (in fact, this ratio gives an idea of the relative amount of the buffer capacity that is still available to neutralize VFA). Zickefoose and Hayes (Zickefoose y Hayes, 1976) suggested that such a ratio should be maintained within the range 0.1-0.35 $\frac{\text{mEq/L}}{\text{mEq/L}}$ to improve the digester operational stability. Moreover, it is well known that AD processes operate under stable conditions if the VFA concentration is kept under 25 mEq/L and the buffer capacity of the system (given by the bicarbonate concentration) is enough to maintain the system pH close to neutral (Hill *et al.*, 1987). These conditions are known as the normal operating conditions (NOC) (Hill *et al.*, 1987).

From the above, it is clear that control strategies in AD processes should not only be focused on using the pollution level as control variable but also on those variables related to the process operational stability such as alkalinity. In this paper, the operational stability problem in AD processes is addressed from a multiple-input multiple-output (MIMO) control point of view by the robust regulation of both, VFA and TA. The proposed control scheme is derived from a mathematical model of a typical AD process which includes TA as a state variable. The control objectives are achieved by manipulating the wastewater flow rate to regulate the VFA concentration and an alkali solution is added directly to the digester to maintain TA at a desired set-point (restoring the bicarbonate buffer capacity as it is destroyed by VFA). By controlling both these variables, it is possible to satisfy the before mentioned practical criteria while improving the operational stability of AD processes. The paper is outlined as follows. First, the model used in the synthesis of the MIMO robust scheme is briefly described. Then, TA, NOC and the control problem are stated in terms of this model. Thereafter, the robust scheme is proposed and evaluated via numerical simulations under different operating conditions including load disturbances, noisy measurements, uncertain kinetics and control input constraints. Finally, some conclusions are pointed out.

II. Model Description

II-A. *The physicochemical equilibrium*

TA is formally defined as the sum of equivalents of all the bases that can be titrated with a strong acid to the first equivalence point of the system (i.e., pH = 4.3) (Ripley *et al.*, 1985). In AD processes, alkalinity is due to many species but the most important are bicarbonates and VFA. In this paper, we propose the following expression to define TA:

$$\mathrm{TA} = f_{Tc}[\mathrm{HCO}_3^-] + f_{Ta}[S_2^-] \tag{1}$$

based on the following assumptions: (a) only VFA and bicarbonates are present as weak acids, (b) VFA are mainly composed by acetate and (c) the process is operated in the pH range 6<pH<8. Here, $[S_2^-]$ and $[\mathrm{HCO}_3^-]$ denote the concentrations of dissociated VFA and bicarbonates, while f_{Tc} and f_{Ta} are given by

$$f_{Tc} = \left(1 - \frac{10^{-pH} + K_c}{10^{-4,3} + K_c}\right), f_{Ta} = \left(1 - \frac{10^{-pH} + K_{ac}}{10^{-4,3} + K_{ac}}\right)$$

where K_c and K_{ac} (mmol/L) are the ionization constants for the equilibria $\mathrm{HCO}_3^-/\mathrm{CO}_2$ and S_2^-/S_2, respectively.

One of the most widely accepted expressions to represent the strong ions concentration in AD process is the following (Breusegem *et al.*, 1988):

$$\mathrm{Z} = [\mathrm{HCO}_3^-] + [S_2^-] \tag{2}$$

which, without loss of generality can be related to TA by

$$\mathrm{TA} = \mathrm{Z} - \beta \tag{3}$$

where β stands for the non titrated fraction of both bicarbonates and VFA beyond the first equivalence point which, for practical purposes, may be considered constant (Rozzi, 1991).

II-B. *The dynamical model*

A suitable mathematical model that describes the dynamics of continuous AD processes can be obtained by performing the mass balance of the species involved in the process (including TA) and by introducing the following additional assumption: (d) the process is operating under isothermal conditions and (e) the wastewater flow rate (Q_1) is much greater than the alkali flow (Q_2) (*i.e.*, $Q_1 \gg Q_2$), which means that the total dilution rate ($D = D_1 + D_2$) can be approximated by $D \approx D_1$. Thus, the resulting model is given by

$$\begin{aligned}
\dot{X}_1 &= (\mu_1(p_1, S_1, \mathrm{pH}, \mathrm{TA}) - \alpha D_1)X_1 \\
\dot{X}_2 &= (\mu_2(p_2, S_2, \mathrm{pH}, \mathrm{TA}) - \alpha D_1)X_2 \\
\dot{S}_1 &= (S_{1,in} - S_1)D_1 - k_1\mu_1(p_1, S_1, \mathrm{pH}, \mathrm{TA})X_1 \\
\dot{S}_2 &= (S_{2,in} - S_2)D_1 + k_2\mu_1(p_1, S_1, \mathrm{pH}, \mathrm{TA})X_1 \\
&\quad -k_3\mu_2(p_2, S_2, \mathrm{pH}, \mathrm{TA})X_2 \\
\dot{\mathrm{TA}} &= (\mathrm{TA}_{in} - \mathrm{TA})D_1 + (\mathrm{TA}_{in}^{'} - \mathrm{TA})D_2
\end{aligned} \tag{4}$$

where X_1, X_2, S_1, S_2 and TA denote, respectively, the concentrations of acidogenic bacteria (g/L), methanogenic bacteria (g/L), primary organic substrate (other than VFA) expressed as Chemical Oxygen Demand (COD) (g/L), VFA (mmol/L) and TA (mmol/L). The subscript in is used to identify the concentration of each component in the wastewater inlet flow rate. $\mathrm{TA}_{in}^{'}$ (mmol/L) represents the TA concentration of the alkali solution while D_1 and D_2 respectively represent the dilution rates associated to the wastewater and alkali solution flow rates (i.e., D_i(h^{-1})=Q_i(L/h)/ V(L) for $i = 1, 2$ and V is the digester volume). It has been shown that by introducing α in Model (4), it is possible to describe the dynamic behavior of various continuous bioreactor configurations (Bernard *et al.*, 2001). It is evident that by setting $\alpha = 1$, Model (4) describes the dynamics of the classical Continuous Stirred Tank Reactor (CSTR) where the biomass is completely suspended in the liquid phase. On the other hand, with $0 < \alpha < 1$, Model (4) has been successfully used to describe the dynamics of fluidized-bed and fixed-bed bioreactors operating under good mixing and recycling conditions together with a generous biogas production, making possible to neglect the effect of the axial dispersion (Escudie *et al.*, 2005). Finally, $\mu_1(p_1, S_1, \mathrm{pH}, \mathrm{TA})$ and $\mu_2(p_2, S_2, \mathrm{pH}, \mathrm{TA})$, respectively, represent the specific growth functions associated to the acidogenic and methanogenic populations, where p_1 and p_2 are kinetic parameters. In order to simplify the mathematic notation, from now on, the specific growth functions will be represented as $\mu_1(.)$ and $\mu_2(.)$.

II-C. *Normal Operating Conditions*

As previously noted, the operating conditions where the digester stability prevails are also known as the *Normal Operating Conditions* (NOC). In this section, these conditions are formally stated in terms of Model (4) by taking into account the stability criteria described in Section I. Thus, AD process (4) operates under NOC if the following conditions are fulfilled (Méndez-Acosta *et al.*, 2008):

(I) The biomass within the digester remains active, which in terms of the model implies that $X(t)_1, X(t)_2 > 0 \;\forall\; t \geq 0$. In addition,

$x^0 = x(t=0) > 0 \; \forall t \geq 0$ where $x = [X_1, X_2, S_1, S_2, \mathrm{TA}]'$.

(II) The VFA concentration (S_2) is less than 25 mEql/L, TA greater than 60 mEql/L and the ratio VFA's/TA is maintained within the range 0.1-0.3 $\frac{mEq/L}{mEq/L}$.

(III) $0 < \beta_1 \leq \int_t^{t+\delta_1} \Delta S_{1,in}(\tau) d\tau \quad \forall\; t \geq 0$ where $\Delta S_{1,in}(t)$ is defined as $\Delta S_{1,in}(t) \equiv S_{1,in}(t) - S_1(t)$ and, β_1 and δ_1 are positive constants.

(IV) $0 < \beta_2 \leq \left| \int_t^{t+\delta_2} \Delta S_{2,in}(\tau) d\tau \right| \quad \forall\; t \geq 0$ where $\Delta S_{2,in}(t)$ is defined as $\Delta S_{2,in}(t) \equiv S_{2,in}(t) - S_2(t)$ and, β_2 and δ_2 are positive constants.

There are other important process features that must be considered in the controller design when dealing with AD processes such as:

A1 VFA and TA are available on-line (Steyer *et al.*, 2002).

A2 $\mu_1(.)$ and $\mu_2(.)$ are unknown, but based on biological evidence, it is nonrestrictive to assume that, for controller design purposes, these functions are smooth, bounded and positive-definite (Grognard y Bernard, 2004).

A3 α is assumed to be uncertain but does vary in the open interval $0 < \alpha < 1$.

A4 The wastewater composition $S_{j,in}$ for $j = 1, 2$ and TA_{in} are assumed unknown but bounded, while the composition of the alkali solution (TA_{in}') is assumed constant and known.

A5 The wastewater (Q_1) and alkali solution (Q_2) flow rates used as the manipulated variables to regulate VFA and TA respectively are constrained in practice, because of the capacity of the pumps and to avoid undesired operating conditions such as the washout condition. Consequently, the dilution rates D_1 and D_2 are bounded by the following saturation function:

$$sat(D_i) = \begin{cases} D_i^{max}, & if \quad D_i \geq D_i^{max} \\ D_i, & if \quad D_i^{min} < D_i < D_i^{max} \\ D_i^{min}, & if \quad D_i \leq D_i^{min} \end{cases}$$

for $i = 1, 2$, where the upper and lower bounds are well-known and, from a practical point of view, $D_i \in \mathbb{R}^+$ is a piecewise constant function.

III. Controller Design

III-A. Control Problem Statement

As previously stated, the control of VFA and TA is of paramount importance in AD processes because these variables are directly related to the process operational stability. Hence, the control problem addressed in this paper can be stated as follows: the proposal of a MIMO control scheme capable to achieve the robust regulation of the VFA concentration and TA in order to improve the stability of continuous AD processes operating under NOC in the face of load disturbances, control input constraints and uncertain kinetics.

III-B. Linearizing Control Scheme

Here, the existence of a MIMO input-output linearizing control scheme capable to achieve the regulation of VFA and TA is demonstrated as a first step in the design of the robust control approach. For this purpose, let us rewrite Model (4) in the affine form for MIMO nonlinear systems (Isidori, 1995):

$$\dot{x} = f(x) + \sum_{i=1}^{m} g_i(x) u_i; \quad y_1 = h_1(x), \ldots, y_m = h_m(x)$$

where m is the number of state variables to be regulated, $f(x)$, $g_i(x)'s$ are smooth vector fields, $h_i(x)'s$ are smooth functions defined on $U \subset \mathbb{R}_+^5$ and U the set of NOC. y_i represents the output functions whereas the control inputs are denoted by u_i. Particularly, the input and output vectors are given by $y = [S_2, \mathrm{TA}]$; $u = [D_1, D_2]$. Then, it can be easily shown that Model (4) has a well-defined relative degree vector $r = [1, 1]$ under NOC. Therefore, it is straightforward to demonstrate the existence of a local coordinate transformation $z = \Phi(x)$ in a neighborhood U^{o} of x^{o}, such that Model (4) can be recast in the following normal form:

$$\begin{aligned} \dot{z}_1 &= (S_{2,in} - z_1) D_1 \\ &\quad + (k_2 \mu_1(.) z_3 - k_3 \mu_2(.) z_4)(S_{2,in} - z_1)^{\alpha} \\ \dot{z}_2 &= (\mathrm{TA}_{in} - z_2) D_1 + (\mathrm{TA}_{in}' - z_2) D_2 \end{aligned} \tag{5a}$$

$$\begin{aligned} \dot{z}_3 &= z_3 \left(\mu_1(.) - \frac{\alpha k_3 \mu_2(.) z_4 - k_2 \mu_1(.) z_3}{(S_{2,in} - z_1)^{1-\alpha}} \right) \\ \dot{z}_4 &= z_4 \left(\mu_2(.) - \frac{\alpha k_3 \mu_2(.) z_4 - k_2 \mu_1(.) z_3}{(S_{2,in} - z_1)^{1-\alpha}} \right) \\ \dot{z}_5 &= z_5 \left(\mu_1(.) - \frac{\alpha k_1 k_3 \mu_2(.) z_4}{k_2 (z_3/z_5)^{1/\alpha} (S_{2,in} - z_1)^{1-\alpha}} \right) \end{aligned} \tag{5b}$$

where (5b) denotes the internal dynamics of the system, which is stable under NOC (Méndez-Acosta *et al.*, 2009). Then, the following output feedback control guarantees the exponential convergence of the output vector $y = [z_1, z_2]$ towards its set-point $y^* = [S_2^*, \mathrm{TA}^*]$:

$$[D_1, D_2]' = \mathbb{A}^{-1}(z) \begin{pmatrix} -L_f h_1(z) - v_1(z) \\ -L_f h_2(z) - v_2(z) \end{pmatrix} \tag{6}$$

where $v_1(z) = K_1(z_1 - S_2^*)$, $v_2(z) = K_2(z_2 - \mathrm{TA}^*)$ are such that the polynomials $P_1(s) = s + K_1 = 0$ and $P_2(s) = s + K_2 = 0$ are Hurwitz and K_1, K_2 are the control gains. Unfortunately, the output feedback control (6) cannot be directly implemented in practice, because it needs the knowledge of both the inlet composition and the kinetic terms, which is a condition difficult to satisfy under actual operating conditions (see A2 and A4). Hence, in the next section, a robust control approach is proposed to overcome these limitations from an extension of the previously reported ideas in (Alvarez-Ramírez *et al.*, 1997).

III-C. The Robust Control Approach

Now, without any loss of generality, let us consider that the wastewater concentration of VFA and TA can be described, respectively, by $S_{2,in} = \tilde{S}_{2,in} + \Delta_{S_2}$ and $\mathrm{TA}_{in} = \tilde{\mathrm{TA}}_{in} + \Delta_{\mathrm{TA}}$, where Δ_{S_2} and Δ_{TA} are uncertain and bounded functions associated to the variation of the wastewater composition around the well-known nominal values $\tilde{S}_{2,in}$ and $\tilde{\mathrm{TA}}_{in}$, which can be determined by a single off-line measurement of the wastewater to be treated. Thus, by defining the uncertain functions $\eta_1 \equiv (k_2\mu_1(.)z_3 - k_3\mu_2(.)z_4)(S_{2,in} - z_1)^{\alpha} + \Delta_{S_2}D_1$ and $\eta_2 \equiv \Delta_{\mathrm{TA}}D_1$, Model (5) can be recast in the following extended state-space representation

$$
\begin{aligned}
\dot{z}_1 &= \eta_1 + (\tilde{S}_{2,in} - z_1)D_1 \\
\dot{z}_2 &= \eta_2 + (\tilde{\mathrm{TA}}_{in} - z_2)D_1 - (\mathrm{TA}_{in}^{'} - z_2)D_2 \qquad (7)\\
\dot{\eta} &= \Xi(z); \qquad \eta = [\eta_1, \eta_2]^{'} \\
\dot{z}_i &= \Upsilon(z); \qquad for \;\; i = 3,4,5
\end{aligned}
$$

where it can be shown that the extended state-space (7) is equivalent to (5). Notice that the augmented state vector η can be reconstructed from on-line measurements of the output and input variables by means of an extended Luenberger observer (ELO). Then, one can devise the following robust control scheme by coupling the ELO to the output feedback control (6):

$$
\begin{aligned}
\dot{\hat{z}}_1 &= \hat{\eta}_1 + (\tilde{S}_{2,in} - \hat{z}_1)D_1 + \Gamma_1 g_{11}(z_1 - \hat{z}_1) \\
\dot{\hat{z}}_2 &= \hat{\eta}_2 + (\tilde{\mathrm{TA}}_{in} - \hat{z}_2)D_1 + (\mathrm{TA}_{in}^{'} - \hat{z}_2)D_2 \\
&\quad + \Gamma_2 g_{21}(z_2 - \hat{z}_2) \qquad (8a)\\
\dot{\hat{\eta}}_1 &= \Gamma_1^2 g_{12}(z_1 - \hat{z}_1) \\
\dot{\hat{\eta}}_2 &= \Gamma_2^2 g_{22}(z_2 - \hat{z}_2)
\end{aligned}
$$

$$
\begin{aligned}
D_1 &= sat\left\{ -\frac{1}{(\tilde{S}_{2,in} - \hat{z}_1)}\left[\hat{\eta}_1 + K_1(\hat{z}_1 - S_2^*)\right]\right\} \qquad (8b)\\
D_2 &= sat\left\{ -\frac{1}{(\mathrm{TA}_{in}^{'} - \hat{z}_2)}\left[(\tilde{\mathrm{TA}}_{in} - \hat{z}_2)D_1 + \hat{\eta}_2 \right.\right. \\
&\quad \left.\left. + \; K_2(\hat{z}_2 - \mathrm{TA}^*)\right]\right\} \qquad (8c)
\end{aligned}
$$

Equation (8a) allows the estimation of the uncertain states η_1 and η_2, while (8b) and (8c) induce the desired behavior on VFA and TA, respectively. g_{11}, g_{12}, g_{21} and g_{22} are chosen such that the matrix associated to the linear part of the estimation error vector e ($e_1^i = z_i - \hat{z}_i$ and $e_2^i = \eta_i - \hat{\eta}_i$ for $i = 1,2$) be Hurwitz, whereas Γ_i and K_i for $i = 1,2$ are the estimation and control gains (tuning parameters), respectively. In this way, it is possible to guarantee that $e \rightarrow \varepsilon$ as $t \rightarrow \infty$, where ε is an arbitrarily small neighborhood around the origin. In order to avoid undesired effects in the controller performance due to the control input saturations such as the windup phenomena (Hanus *et al.*, 1987), the constrained values of the dilution rates are fed back to the observer as shown in Figure 1, which induce a structure into the control scheme that resembles an observer-based antiwindup scheme (Kothare *et al.*, 1994).

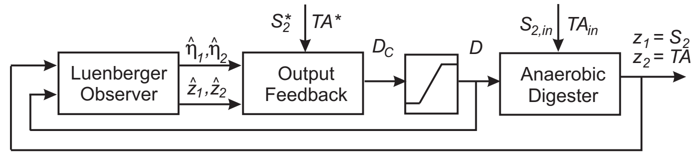

Figure 1. Block diagram of the robust regulator described by Eq. (8)

III-D. On the Closed-Loop Behavior

Here, we analyze the closed-loop behavior of the robust control scheme (8) under the following two possible situations:

S1. *The control inputs are not saturated* (*i.e.*, $D = D_C$); then, one obtains the following transfer functions

$$
\frac{\hat{z}_1}{z_1} = \frac{\Gamma_1 g_{11}}{s + \Gamma_1 g_{11} + K_1}; \quad \frac{\hat{z}_2}{z_2} = \frac{\Gamma_2 g_{21}}{s + \Gamma_2 g_{21} + K_2} \qquad (9)
$$

Clearly in this situation, the ELO acts as a first-order low-pass filter, where the cutoff frequency depends on the observer and control gains Γ_i, K_i, and, as a consequence, the effect of noisy measurements can be properly handled by selecting adequate values for both these parameters.

S2. *The control inputs are saturated* (*i.e.*, $D = D^{min,max}$); thus, the following expressions are obtained in the Laplace domain

$$
\begin{aligned}
\frac{\hat{z}_1}{z_1} &= \frac{\Gamma_1 g_{11}s + \Gamma_1^2 g_{12}}{s^2 + (\Gamma_1 g_{11} + D_1^{min,max})s + \Gamma_1^2 g_{12}} \qquad (10)\\
\frac{\hat{z}_2}{z_2} &= \frac{\Gamma_2 g_{21}s + \Gamma_2^2 g_{22}}{s^2 + (\Gamma_2 g_{21} + D_1^{min,max} + D_2^{min,max})s + \Gamma_2^2 g_{22}}
\end{aligned}
$$

In this case, the ELO has a structure of a second-order low-pass filter with a feed-forward action, which allows the continuous estimation of the uncertain state η, even when the control inputs saturate.

IV. Numerical Implementation

In this section, the performance of the robust approach (8) is evaluated via numerical simulations under different operating conditions and the most uncertain scenarios. The numerical implementation was carried out by using the Matlab-Simulink® software. The control parameters used were: $g_{11} = 2{,}0$, $g_{12} = 1{,}0$, $g_{21} = 2{,}0$, $g_{22} = 1{,}0$, $\Gamma_1, \Gamma_2(h^{-1}) = 0{,}3$ and $K_1, K_2(h^{-1}) = 0{,}3$. The model parameters used during the simulation runs were those used by (Bernard *et al.*, 2001), while the initial conditions were: $X_1(0) = 0{,}5$ g/L, $X_2(0) = 0{,}7$ g/L, TA(0) = 50 mEq/L, $S_1(0) = 2{,}0$ g/l, $S_2(0) = 30$ mEq/L, $\hat{z}_1(0) = 20$ mEq/L, $\hat{z}_2(0) = 60$ mEq/L, $\hat{\eta}_1(0) = 0$ mEq/L-h and $\hat{\eta}_2(0) = 0$ mEq/L-h. Note that there is a significant error between the initial conditions of the measured (z_i) and the estimated ($\hat{z}_i$) states, which was induced in order to test the ELO performance. The nominal values used for the wastewater influent composition were $\tilde{S}_{2,in} = 80$ mEq/L and $\tilde{\mathrm{TA}}_{in} =$

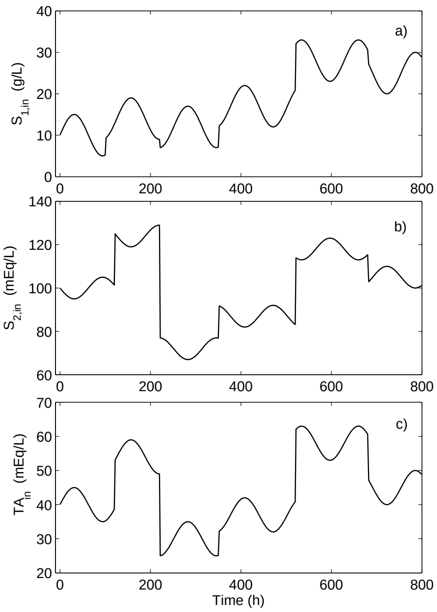

Figure 2. Wastewater influent composition

30 mEq/L, and the concentration of the alkali solution $\mathrm{TA}_{in}^{'}$ was fixed at 8,000 mEq/L based on the concentration that industrial soda normally has, which can vary between 8,000 and 10,000 mEq/L. The upper and lower bounds used in (II-C) to constrain the dilution rates were D_1 = [0.002,0.05] h^{-1} and D_2 = [0, 0.0005] h^{-1} (clearly, assumption (d) stated in Section II is satisfied since $D_1 >> D_2$).

In order to test the controller performance under the influence of noisy measurements, uniformly random white noises with an amplitude of ± 2.0 mEq/L and ± 10.0 mEq/L were added to the simulated measurements of VFA and TA, respectively. Furthermore, to ensure the digester stability, the set-point values were chosen by considering the NOC conditions stated in Section II-C as well as actual operating conditions for which the inlet composition was randomly varied around the values reported by (Bernard *et al.*, 2001) (see Figure 2). Several wastewater composition changes were also introduced during the simulation runs to test the performance of the control law (8) in the presence of load disturbances and set-point changes.

The response of the proposed robust scheme (8) on the regulation of the VFA concentration (S_2) is depicted in Figure 3a, where it can be noted that four set-point changes between 10-18 mEq/L were induced to evaluate the controller set-point tracking capability. As is seen, set-point changes were satisfactorily tracked under the influence of load disturbances while attenuating the noisy measurements. Notice also the dynamic response of the VFA concentration estimated by the ELO ($\hat{S}_2$), which was smoother than that measured (S_2), demonstrating the effect of the low-pass filter structure of the ELO on the controller response. The dynamic response of the wastewater dilution rate, which is the control input associated to the VFA concentration is

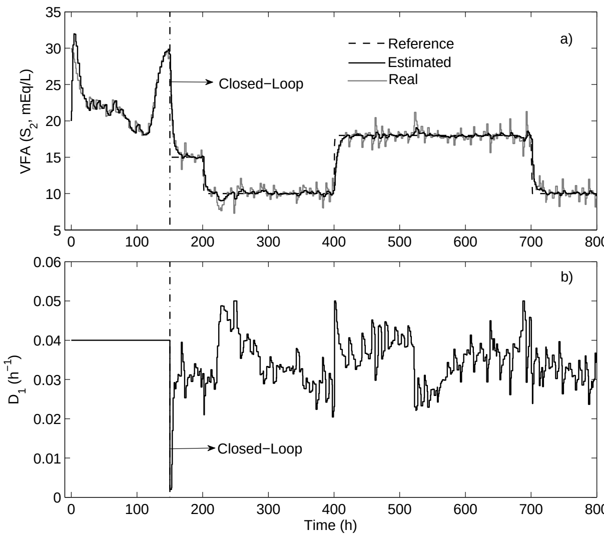

Figure 3. VFA concentration (S_2) and Wastewater dilution rate (D_1)

shown in Figure 3b. It is clear that D_1 saturated at various times; nevertheless, the controller performance did not deteriorate due to its antiwindup structure. Finally, Figure 3a shows that, during the open-loop, the process stability was on the edge of operational instability because of a drastic increase in the VFA concentration, which reached values higher than 20 mEq/L. Nevertheless, the robust controller was able to rapidly drive such concentration toward its set-point once the control loop was closed.

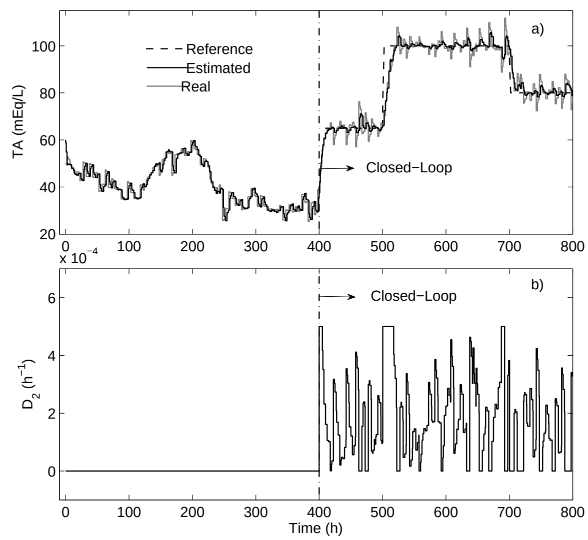

Figure 4. Total alkalinity (TA) and Alkali dilution rate (D_2)

Figure 4a shows the excellent response of TA in the face of both the set-point changes and load disturbances. Notice that the control scheme was capable to maintain TA above the stability limit (60 mEq/L) during the simulation run. The response of the alkali dilution rate D_2 is depicted in Figure 4b, where one can see the control effort to keep TA around its set-point and the effect of using the antiwindup structure in the robust control scheme. Although D_2 saturated at different times, it provided the appropriate control action to satisfy the controlled variable reference value.

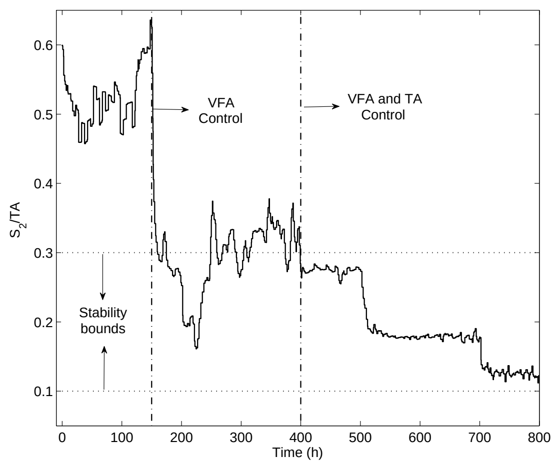

Figure 5. Behavior of the VFA/TA ratio during the simulation

The behavior of the S_2/TA ratio during the simulation run is depicted in Figure 5. At the start-up, the AD process was operated in an open-loop manner and clearly the S_2/TA ratio did not meet the required bounded response. Then, at 150 hours, we closed the VFA loop, which initially drove the S_2/TA response to the desired stability interval, but this response deteriorated and went off-bounds, despite the proper VFA control. The proposed MIMO robust control scheme was finally implemented at 400 hours providing the proper control action to maintain the S_2/TA ratio within bounds in the face of load disturbances, uncertain kinetics and set-point changes. Then, this figure allow us illustrate that, under certain circumstances, the proper VFA control does not guarantee the operational stability of continuous AD processes, which clearly state the benefits of using the proposed MIMO robust control scheme over a SISO controller.

V. Conclusions

A model-based MIMO robust control scheme was proposed to regulate both the VFA concentration and TA in order to improve the operational stability of continuous AD processes. This scheme is designed by using a previously reported AD model, which was modified to include TA as a state-variable. The control scheme combines an output feedback control with linearizing-like structure and an extended Luenberger observer, which not only allowed the estimation of the uncertain terms associated to the controlled states but induced a low pass filter and an antiwindup structure that improves the controller performance in the presence of noisy measurements and control input constraints. It was shown through numerical simulations that the proposed MIMO control scheme was capable to maintain the desired set-points in the face of the complete ignorance of either the processes kinetics or the wastewater influent composition while satisfying previously reported practical stability criteria. The experimental implementation of the proposed control scheme is currently under way and the results will be reported in the near future.

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