VFA robust control of an anaerobic digestion pilot plant: experimental implementation

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Abstract: In this work, an experimental implementation of a model based output feedback (OF) scheme applied to regulate the volatile fatty acids concentration of an anaerobic digestion pilot plant is presented. The OF scheme was developed within a global nonlinear dynamics framework, resulting in an interlaced control-observer design in the light of passivity, observability, and bifurcation properties. From a industrial control perspective, the resulting OF scheme is a saturated PI controller with: (i) systematic construction and tuning and (ii) nonlocal robust stability conditions in terms of control gains and limits. Experimental results demonstrate the applicability, robustness and reliability of the OF control scheme in a real environment under several load disturbances, severe uncertainties and typical operational failures.

Keywords: Waste treatment; Dynamic output feedback; Anaerobic Digestion; real-time control

1. INTRODUCTION

For decades, anaerobic digestion (AD) processes have been the center of interest to the scientific and industrial community focused on wastewater treatment to reduce and transform the organic matter from industrial and municipal effluents into a high-energy gas (Henze et al., 1997). Nevertheless, its widespread application has been limited, because of the difficulties to achieve its stable operation, that required the monitoring and tight control of certain operating conditions that may lead the system to the eventual crash even under tightly controlled pH and temperature conditions (Dochain, 2008; Méndez-Acosta et al., 2008). In addition, special attention must be paid to the effect of substrate and/or product inhibition on the behavior of AD processes such as the excessive accumulation of volatile fatty acids (VFA).

In the past, the regulation of the organic matter has been addressed by proposing many control techniques to keep, at a predetermined set-point, certain operating variables which are readily available, such as the chemical oxygen demand (COD) and the biogas production (Ahring and Angelidaki, 1997; Alvarez-Ramirez et al., 2002; Puñal et al., 2002; Schögerl, 2001). In particular, classical proportional-integral (PI) or proportional-integral-differential (PID) controllers have been recognized as a good alternative for the regulation of AD plants. However, it is well-known that its performance is strongly dependent on the tuning parameters, which, in the application to complex systems with nonlinear behavior, are only valid around a given operating point. Moreover, the presence of input constraints has been shown to seriously degrade the classical PI/PID performance limiting their practical applications (Olsson et al., 2005; Steyer et al., 2006).

It has been reported that the operational stability of AD processes is highly dependent on the accumulation of VFA. For instance, some authors recommend a VFA concentration below 1.5 g/l (25 mmol/l) (Angelidaki et al., 2005). Nevertheless, the problem of the operational instability due to the accumulation of VFAs remains open (Méndez-Acosta et al., 2008). Following these ideas, Méndez-Acosta et al. (2008, 2010) proposed a SISO robust regulator of VFA and a MIMO robust regulator of VFA and total alkalinity (TA). These regulators are composed of an output feedback (OF) control coupled with an extended Luenberger observer used to estimate the uncertain terms of the process and contains an antiwindup feedback scheme to suppress the influence of significant external disturbances that typically requires large control actions but limited by the presence of input constraints. However, in these contributions the limits for the inputs are selected empirically. On the other hand, Schaum et al. (2015) recently proved theoretically the asymptotic stability of an AD system regulated with an interlaced VFA OF control-observer designed in the light of passivity, observability, and bifurcation properties within a global nonlinear dynamics framework. The resulting scheme is a saturated linear PI controller with: (i) nonlocal robust stability conditions for control gains and limits, (ii) antiwindup protection, (iii) systematic construction, and (iv) conventional like gain tuning procedures. In this contribution, we present the results of the implementation of the OF scheme proposed...
by the authors (Schaum et al., 2015) in an AD pilot plant to regulate the VFA concentration. The performance and robustness of the OF scheme is tested in the face of uncertain parameters, load changes and measurement devices malfunctioning.

2. METHODOLOGY

2.1 AD model

For control design purposes it is considered a simplified macroscopic model of the AD process based on 2 main reactions (Bernard et al., 2001), where the complex organic matter ($S_1$) is degraded into VFA ($S_2$) by acidogetic bacteria ($B_1$), and then the VFA are degraded into methane $CH_4$ and $CO_2$ by methanogetic bacteria ($B_2$) according to the reaction scheme

$$\beta_1S_1K_{i1} + \beta_2S_2K_{i2} + B_1, \quad \beta_2S_2K_{i2} \rightarrow CH_4 + CO_2 + B_2,$$

where $\beta_1$, $\beta_2$, $K_{i1}$ and $K_{i2}$ are pseudo-stoichiometric coefficients associated to the bioreactions, while $K_i(S_i)$ and $B_i$ are the reaction rates. In particular, we consider a perfect mixed continuous anaerobic digester with volume $V$, where $S_1$ and $S_2$ are fed at volumetric rate $F$, with bounded from above and below time-varying mass concentrations ($S_{f1}$ and $S_{f2}$).

$$S_{f1}^L \leq S_{f1} \leq S_{f1}^U, \quad i = 1, 2.$$

The anaerobic digester's model is easily derived from the mass balances of the species involved in the AD (Bastin and Dochain, 1990; Bernard et al., 2001):

$$\dot{B}_1 = M_1(S_1)B_1 - DB_1$$

$$\dot{S}_1 = -\beta_1M_1(S_1)B_1 + D(S_{f1} - S_1)$$

$$\dot{B}_2 = M_2(S_2)B_2 - DB_2$$

$$\dot{S}_2 = \beta_2M_1(S_1)B_1 - \beta_2M_2(S_2)B_2 + D(S_{f2} - S_2)$$

where $D = F/V$ is the dilution rate. The specific bacterial growth rates, $M_1(S_1)$ and $M_2(S_2)$, are respectively Monod and Haldane kinetics (Dochain and Vanrolleghem, 2001; Van-Impe et al., 1998), i.e.

$$M_1(S_1) = \frac{K_{01}S_1}{S_1 + K_{s1}} \quad \text{and} \quad M_2(S_2) = \frac{K_{02}S_2}{S_2^2/K_{s2} + S_2 + K_{s2}}$$

where $K_{01}$, $K_{s1}$, $K_{02}$, $K_{s2}$ and $K_{s2}$ are the maximum bacterial growth rate and the half-saturation constant associated to the substrate $S_1$, the maximum bacterial growth rate in the absence of inhibition, and the saturation and inhibition constants associated to substrate $S_2$, respectively. Finally, it is considered that the VFA concentration is measured on-line, while the volumetric flow rate, $F$ (or the dilution rate, $D$) is the control variable. Schaum et al. (2015) have shown that system (1)-(4) with specific bacterial growth rates (1) has up to 6 equilibrium points depending on the parameters values, the dilution rate value and input substrates concentrations. In particular, this system has three bifurcations points: $D_*$, $D^*$ and $D^{**}$ (see Figure (1)). For dilutions rates below $D_*$ there exists only one stable equilibrium point in which both biomasses are alive (this equilibrium is the so called operational stable equilibrium). For $D_* < D < D^*$ there exists two stable equilibrium points, one is the operational stable equilibrium and the other one is the undesirable equilibrium where the methanogetic biomass is no longer active or dead. Then, for $D^* < D < D^{**}$ the only stable equilibrium is where the methanogetic biomass is death and, finally, above $D^{**}$ there is only one stable equilibrium, the washout of both biomasses.

2.2 Control design

Based on the bifurcation properties, the robust OF saturated control scheme proposed by Schaum et al. (2015) is described by equations

$$\dot{\chi} = -\omega \chi + \omega (aD - \omega y)$$

$$D = sat(\mu(\chi, y))$$

with

$$\mu(\chi, y) = \frac{k(y - y_{ref} + \chi + \omega y)}{a}$$

where $\chi$ is the dynamical variable which is used to estimate uncertainties, $y$ and $y_{ref}$ are the measured and reference variables, respectively, while $k > 0$ and $a = S_{f1}^i - y_{ref}$ are the feedback gains, and $\omega$ is the observer gain. The saturation function is defined as:

$$sat(\mu(\chi, y)) = \begin{cases} 
D^- & \text{if } \mu(\chi, y) < D^- \\
\mu(\chi, y) & \text{if } D^- \leq \mu(\chi, y) \leq D^+ \\
D^+ & \text{if } \mu(\chi, y) > D^+
\end{cases}$$

To guarantee nonlocal robust stability, the saturation limits must satisfy the following conditions

$$D^- < D_* \quad \text{and} \quad D^* < D^+ < D^{**}.$$  (10)

Without saturation, the controller (6)-(8) takes the form of a traditional PI

$$\tilde{D} = k_p \left( e + \tau_i e \int_0^\tau e(\tau) \, d\tau \right)$$

where $k_p = a^{-1}(k + \omega)$ and $\tau_i = k^{-1} + \omega^{-1}$ are the proportional gain and integral time, respectively, while $\tilde{D}$ and $e = y - y_{ref}$ are deviation variables.
**Tuning guidelines** From the closed-loop stability analysis presented in (Schaum et al., 2015), the next tuning guidelines follow (see González and Alvarez, 2005, and references therein):

1. Set measurements with typical (deterministic sinusoidal or random) measurement error.
2. Set the control limits $D_k^-$ and $D_k^+$ according to (10).
3. Set the control ($k$) and observer ($\omega$) gains conservatively at $k \approx 1$ to 2 and $\omega \approx 3$ to 5.
4. Increase $\omega$ until the response becomes oscillatory at $\omega^0$, and back-off to $\omega = (1/3$ to $1/2) \omega^0$.
5. Increase $k$ until the response becomes oscillatory at $k = k^0$, and back-off to $k = (1/3$ to $1/2) k^0$.
6. If necessary, adjust $(D_k^-, D_k^+)$ and $(k, \omega)$ repeating steps 4 and 5 to improve the behavior.

**2.3 Experimental set-up**

The influent characteristics Experiments were performed by using raw tequila vinasses as substrate. Tequila vinasses, whose compositions are similar to those described by Méndez-Acosta et al. (2010), were collected from a tequila factory located at Jalisco, Mexico. Such vinasses were stored in two tanks of 500 L storage capacity each one.

The AD process The pilot plant layout is shown in Figure 2. The AD process was carried out isothermally (around 37°C) in a 445 L stainless steel up-flow fixed-bed bioreactor packaged with PVC Cloisonyl pipes. This bioreactor has a recirculation stream with a flow rate approximately 10 times higher than the wastewater inflow rate to ensure homogeneous conditions, thus although the process is an up-flow fixed bed bioreactor, it can be modeled as a CSTR system. The first part of the process is conformed by a feeding tank of 200 L where known volumes of tap water and raw vinasses can be mixed in different proportions allowing the manipulation of the influent COD and VFA concentration. Within this tank the vinasses pH is regulated around 6.5 through an on-off control by adding a NaOH solution. A remotely controllable peristaltic pump that feeds the desired influent flow rate was connected from this tank to the bioreactor’s recirculation line, thus fresh substrate was mixed with the recirculated liquid just before entering the digester. The process effluent is collected by overflow in a receiving vessel.

On-line measurements The pilot plant is fully instrumented and automated. A National Instruments cRIO9004 device composed of analogical and digital cards was used in the acquisition, treatment and storage of the data. This device also included the appropriate ports and capabilities that allow the remote monitoring and control of the process from an internet connection. The programming of this device was carried out by using the LabVIEW 8.2 software. VFA and bicarbonate concentrations and partial and total alkalinity (PA and TA) were monitored with an automatic titrimetric Applitek analyzer Anasense. While a Siemens analyzer Ultramat 22P and a rotameter connected, after a condenser device which removes the humidity by cooling the biogas, to the top of the digester allowed to continuously measure the methane and carbon dioxide composition and biogas flow rate, respectively.

![Pilot plant diagram](image-url)

**Control implementation** A National Instruments cRIO-9004 device was used to close the control loop, using the VFA measurements provided by the automatic titrimetric Applitek analyzer Anasense while the vinasses inflow rate was used as the control variable. The control algorithm described in section 2.2 was programmed by using the LabVIEW® 8.2 software which communicates with the CRI09004. The proposed control scheme was tested during 20 days in the previously described AD process used for the treatment of Tequila vinasses. Such a scheme was implemented by using a sampling rate of 20 minutes, thus the measurements can be considered continuous.

The performance and robustness of the proposed control scheme were evaluated in the face of different set-point changes and load disturbances. Initially, the process was in an open loop steady state with low dilution rate and low VFA concentration (around 0.2 d⁻¹ and 300 mgVFA/L, respectively). Then, two set-point changes between 500 and 1000 mgVFA/L were induced at time $t = 0$ and 11 days, respectively.

On the other hand, since the composition of the tequila vinasses can change drastically from one tequila distillery to the next and even from batch to batch in the same distillery, an uncertain load disturbance was induced by choosing an influent COD concentration close to 10 gCOD/L during the first 15 days of the experimental run. Then, at day 15, a bigger load was induced by feeding an influent COD concentration close to 13 gCOD/L.

**3. RESULTS AND DISCUSSION**

After a procedure of parameter identification using previous open loop data of the pilot plant described in section 2.3 (Zárate, 2013), the control parameters were calculated as described in section 2.2 through closed loop simulations. Then, the resulting control parameters, which are presented in Table 1, were used in the LabView interface described in section 2.3 to carry out the experimental test.

The experimental results are presented in Figures 3 and 4. Figure 3a shows the VFA concentration measured by the Anasense sensor, as well as its set-point. Figure 3b shows the dilution rate computed by the control algorithm and...
Table 1. Control parameters.

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<th>Parameter</th>
<th>Value</th>
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<tbody>
<tr>
<td>$k$</td>
<td>$1 \text{ d}^{-1}$</td>
<td>$\omega$</td>
<td>$2.5 \text{ d}^{-1}$</td>
</tr>
<tr>
<td>$D^-$</td>
<td>$0.1 \text{ d}^{-1}$</td>
<td>$a$</td>
<td>$8 \text{ gVFA/L}$</td>
</tr>
<tr>
<td>$D^+$</td>
<td>$0.7 \text{ d}^{-1}$</td>
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the measured from a flow sensor, while Figure 4 depicts the dynamic behavior of the controller’s internal variable, $\chi$, which is associated to the uncertain VFA production and consumption rates.

Initially, the process was operated in an open loop manner around a steady state with low dilution rate and low VFA concentration (around $0.2 \text{ d}^{-1}$ and $300 \text{ mgVFA/L}$, respectively). Then, at $t = 0.3$ days the control loop was closed with an initial value of $-4000 \text{ mg/(Ld)}$ for $\chi$ (see Figure 4). Clearly, the OF control scheme failed to reach the set-point due mainly to an incorrect $\chi$ value and a series of process and measuring device malfunctions. For example, at $t = 0.7$ days, the Anasense sensor did not work for almost one day and as a consequence of this failure, the dilution rate saturated from the top which led to an expected VFA concentration increase (up to $1200 \text{ mg/L}$). As a result of this failure, the OF controller was shutdown for about 6 hours and reactivated once the sensor was repaired. At this instant the closed loop was reactivated with $\chi = -5000 \text{ mg/(L.d)}$. As seen, the manipulated variable saturated again from the top but, as expected, the VFA concentration approached it set-point. At $t = 2.8$ days of the AD operation, the controller was once again shutdown. This unfortunate incident lasted one day and caused an increase of the VFA concentration. Once the controller was again set for closed loop operation, the dilution rate saturated from below yielding an VFA concentration decrease. The series of unfortunate events continued at day 6 when a problem on the power lines caused a failure on the interface. Once the electricity was restored. Then, at $t = 5.8$ days when $\chi$ was approximately $-2389 \text{ mg/(L.d)}$ an electrical failure caused a shutdown in the interface, then when the interface was restored the initial value for $\chi$ was approximately $-3000 \text{ mg/(L.d)}$ and this difference produced a transitory increase in the dilution rate that caused a perturbation in the VFA. Then, for around 5 days there were no further failures and the controller was able to decrease the VFA concentration around the reference, while $\chi$ and $D$ attained a steady state around $-2050 \text{ mg/(L.d)}$ and $0.26 \text{ d}^{-1}$, respectively.

At $t \approx 10$ days an interface shutdown was detected, producing again a disturbance that was rejected in less than one day. Then, at $t = 10.76$ days a set-point change was induced from $500$ to $1000 \text{ mgVFA/L}$. The controller increased the dilution rate and the VFA concentration achieved the reference in approximately 4 days, in spite of a sensor failure around $t \approx 13.4$ days. Once the process was around the set-point and after a maintenance procedure was applied to the sensor, at $t \approx 15$ days a load disturbance was induced by increasing the influent COD concentration to around $13 \text{ gCOD/L}$. The controller was able to attenuate this load disturbance in less that 2 days, although an interface shutdown event befell at $t \approx 15.6$ days. The proposed OF control scheme was able to yield satisfactory response in the face of unexpected disturbances, power

4. CONCLUSIONS

In this work, an experimental implementation of a model based output feedback (OF) scheme applied to regulate the volatile fatty acids concentration of an anaerobic digestion pilot plant was presented. Experimental results demonstrate the excellent performance and robustness of the OF control scheme in a real scenario under several load disturbances, severe uncertainties and typical operational fail-
ures, such as operator’s mistakes, equipments and sensors’ failures, changes in the set-point and interface’s startups.

REFERENCES


