Journal of Process Control 34 (2015) 82–96

Contents lists available at ScienceDirect

Journal of Process Control

journal homepage: www.elsevier.com/locate/jprocont

On the dynamics and control of a class of continuous digesters☆

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ARTICLE INFO

Article history:
Received 7 January 2015
Received in revised form 15 June 2015
Accepted 16 July 2015
Available online 27 August 2015

Keywords:
Continuous anaerobic digestion
Stoichiometric invariants
Control input constraints
PI control
Robust nonlocal stability

ABSTRACT

The saturated OF control problem for a class of four-state anaerobic digesters with volatile fatty acids (VFAs) measurement is addressed. The reactor must operate about an optimal steady-state, with maximum VFA consumption that is locally open-loop stable but not necessarily structurally stable. The problem is addressed as an interlaced control-observer design in the light of passivity, observability, and bifurcation properties. The result is a saturated linear PI controller with: (i) systematic construction and tuning and (ii) nonlocal robust stability conditions in terms of control gains and limits. The proposed approach is illustrated with a representative case example through numerical simulations, including a comparison with the switching control approach.

1. Introduction

Anaerobic digestion (AD) processes are known to yield better overall performance when compared to aerobic processes, especially for wastewater treatment. This is due to its low initial and operating costs, short hydraulic retention times, low sludge production, and high organic load removal combined with energy benefit by biogas production. However, its widespread application has been limited because of the complexity of the process, and the difficulty of the associated control problem.

The AD is intrinsically a potentially unstable process. Variations on the input variables (hydraulic flowrate, influent organic load) may easily lead to washout operation [2–7], a phenomenon that takes place under volatile fatty acids (VFA) accumulation, decreasing the pH and leading to bioprocess breakdown. The control problem is further hindered by the lack of on-line measurements of key variables, constrained inputs, and model error of the growth rate function [5].

Basically the open-loop dynamics of AD processes has been studied by combining local stability analysis (in sufficiently small neighborhoods of the steady-states) with simulation-based multiplicity and transient assessments. According to these results [6]: (i) there are up to six steady states (SSs) with up to four possible washout SSs, (ii) there are no limit cycles, and (iii) the optimal (with respect to VFA consumption) SS is a local attractor accompanied by four washout SSs (three saddle repulsors and one attractor). Eventhough these results provide valuable insight on the AD reactor four-state open-loop dynamics, some fundamental questions for process and control design still remain open: (i) the characterization of the geometry of global open-loop dynamics, including robust stability in the sense of structural stability [8,9], and (ii) its consideration in the control design problem, with emphasis in nonlocal robustly stable closed-loop functioning over a delimited domain of attraction of the nominal SS. By structural instability it is meant qualitative changes in the geometry of the dynamics with changes in the system parameters, including appearance or disappearance of steady-states and/or limit cycles. From previous two [10] and three [11] state bioreactor studies with Haldane inhibited kinetics, it is known that: (i) open and closed-loop structural instability (with undesired washout SS attractors) and can be induced by changes in exogenous inputs, kinetics parameters and control gains, as well as control saturation, and (ii) consequently the nonlocality and structural stability properties play key roles in the solution of the robustly stabilizing saturated control problem.

The AD reactor Output-Feedback (OF) control problem has been addressed with several nonlocal-nonlinear approaches:

☆ A preliminary version of this paper has appeared in the Proceedings of the 8th IFAC Symposium on Advanced Control of Chemical Processes (AdChem), Singapore, 2012 [1].

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http://dx.doi.org/10.1016/j.jprocont.2015.07.004

detailed model-based time-optimal bang–bang [12,13], unconstrained robust [14], high-gain observer-based saturated [15], and saturated PI [16] control. These studies provide perspective on the complex saturated OF control problem of AD reactors, but the proposed controllers rise various applicability-oriented complexity and reliability concerns among practitioners: the strong nonlinearity and model dependency of the controllers, the lack of simple tuning guidelines, and the absence of nonlocal robust stability assurance. This in the understanding that the closed-loop AD bioreactor must operate in the presence of unmeasured time-varying feed substrate concentration disturbances, and that any advanced model-based OF controller inherits the modeling error of the bioreaction growth rate functions [2].

Industrial PI controllers have reduced model dependency, low development and maintenance costs, and yield good performance for nonlinear processes [17]. It is still an open question whether an anaerobic digester can be robustly and non-locally controlled through saturated PI control with anti-windup protection. In recent two and three-state bioreactor studies [10,11] it has been shown that: (i) the open-loop locally asymptotically stable (LAS) but structurally unstable equilibrium point of maximum biomass production rate can be robustly stabilized using a saturated PI controller with observer-based anti-windup protection, and (ii) the rather model-independent PI controller recovers (up-to observer convergence) the behavior of an exact model-based nonlinear saturated State-Feedback (SF) controller equivalent to an infinite-horizon nonlinear Model-Predictive Control (nMPC) without saturation. These results exploited: (i) the fundamental connection between robustness, passivity and optimality [18], and their relationship with the constrained control problem [19], and (ii) the idea of tailoring a model for control design in the light of passivity and observability [10,11]. These considerations motivate the present study on the development of a saturated PI control for the nonlocal robustly stable regulation of a four-state AD bioreactor at maximum VFA consumption operation.

In this study the saturated OF control problem for a class of four-state anaerobic digesters with volatile fatty acids (VFAs) measurement and steady-state operation at its (possibly structurally unstable) optimal steady-state (of maximum VFA consumption) is addressed. The objectives are the design of a controller as simple as possible in terms of nonlinearity and model dependency, and the attainment of closed-loop robust stability with preclusion of undesired washout SS attractors induced by control saturation. The technical difficulty resides in the extension to the four-state two-reaction AD bioreactor case (with Monod and Haldane kinetics) of the above mentioned control approach for two [10] and three [11]-state single-reaction bioreactors (with Haldane kinetics). The problem is addressed as an interlaced control-observer design within a global nonlinear dynamics framework in the light of passivity, observability, and bifurcation properties. First, the geometry of the global nonlinear open-loop dynamics is characterized. Then, the saturated nonlinear state-feedback robust control problem is addressed on the basis of the detailed reactor model. Finally, the behavior of this controller is recovered with an OF controller built with a simplified model tailored according to passivity and observability properties. The result 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. The proposed methodology is connected with standard PI as well as detailed model-based advanced nonlinear bang–bang and MP control approaches, and illustrated with a representative case example through numerical simulations, and compared with minimum time bang–bang control.

2. Control problem

2.1. Continuous anaerobic digester

Consider a continuous anaerobic digester with volume V, where organic substrate (S_1) and volatile fatty acid (VFA) (S_2) are fed at volumetric rate F, with bounded from above and below time-varying mass concentrations (S_{1f} and S_{2f}),

$$S_{f1}^- \le S_{f1}(t) \le S_{f1}^+, \quad S_{f2}^- \le S_{f2}(t) \le S_{f2}^+ \tag{1}$$

and converted into acidogenic (B_1) and methanogenic (B_2) biomass according to the reaction scheme

$$\beta_1 S_1 \xrightarrow{K_1} \beta_{12} S_2 + B_1, \quad \beta_2 S_2 \xrightarrow{K_2} \mathrm{CH}_4 + B_2, \tag{2}$$

the reaction rates

$$K_1(S_1, B_1) = M_1(S_1)B_1, \quad K_2(S_2, B_2) = M_2(S_2)B_2,$$
$$M_1(S_1) = \frac{K_{01}S_1}{S_1 + K_{s1}}, \qquad M_2(S_2) = \frac{K_{02}S_2}{S_2^2/K_I + S_2 + K_{s2}}$$

the flow dilution (D) and VFA consumption (P) rates, and the on-line VFA concentration measurement (Y)

$$D = \frac{F}{V}, \quad P = F(S_{f2} - S_2), \quad Y = S_2.$$

Here, β_1, β_{12}, and β_2 are the stoichiometric coefficients, M_1 (or M_2) is the acidogenic (or methanogenic) Monod (or Haldane) biomass growth rate function with mass action K_{01} (or K_{02}), substrate saturation K_{s1} (or K_{s2}), and substrate inhibition K_i constant. M_1 (or M_2) depends isotonically (or non-isotonically) on the organic substrate (or VFA) concentration S_1 (or S_2). The case of biogas production measurement $Y = K_2(S_2, B_2)$ can be addressed with the same approach.

Denoting by t_a the actual time, in dimensionless variables

$$s_1 = \frac{S_1}{\overline{S}_1}, \quad s_2 = \frac{S_2}{\overline{S}_2}, \quad b_1 = \frac{B_1}{\overline{B}_1}, \quad b_2 = \frac{B_2}{\overline{B}_2},$$
$$t = t_a\overline{D}, \quad d = \frac{D}{\overline{D}}, \quad y = \frac{Y}{\overline{S}_2}, \quad p = \frac{P}{\overline{D}\overline{S}_2},$$

with respect to nominal steady-state values ($\overline{S}_{i=1,2}$, $\overline{B}_{i=1,2}$, $\overline{D}$) the AD bioreactor dynamics are given by

$$\begin{aligned}
\dot{s}_1 &= d(s_{f1} - s_1) - \alpha_1\mu_1(s_1)b_1, && s_1(0) = s_{10},\\
\dot{b}_1 &= -db_1 + \mu_1(s_1)b_1, && b_1(0) = b_{10},\\
\dot{s}_2 &= d(s_{f2} - s_2) + \alpha_{12}\mu_1(s_1)b_1 - \alpha_2\mu_2(s_2)b_2), && s_2(0) = s_{20},\\
\dot{b}_2 &= -db_2 + \mu_2(s_2)b_2), && b_2(0) = b_{20},\\
y &= s_2, \quad p = d(s_{f2} - s_2), \quad s_{f1}^- \le s_{f1}(t) \le s_{f1}^+, \quad s_{f2}^- \le s_{f2}(t) \le s_{f2}^+,
\end{aligned} \tag{3}$$

where

$$\alpha_1 = \beta_1\frac{\overline{B}_1}{\overline{S}_1}, \quad \alpha_{12} = \beta_{12}\frac{\overline{B}_1\overline{S}_1}{\overline{S}_2}, \quad \alpha_2 = \beta_1\frac{\overline{B}_2}{\overline{S}_2}$$
$$\mu_1(s_1) = \frac{k_{01}s_1}{s_1 + k_{s1}}, \quad \mu_2(s_2) = \frac{k_{02}s_2}{s_2^2/k_I + s_2 + k_{s2}}, \tag{4}$$
$$k_{0i} = \frac{K_{0i}}{\overline{D}}, \quad k_{si} = \frac{K_{si}}{\overline{S}_i}, \quad k_I = \frac{K_I}{\overline{S}_2}.$$

In vector notation, the bioreactor control system (3) is written as follows:

$$\begin{aligned}
&\dot{\boldsymbol{x}} = \boldsymbol{f}[\boldsymbol{x}, d, \boldsymbol{s}_f], \quad \boldsymbol{x}(0) = \boldsymbol{x}_0,\\
&y = \boldsymbol{c}_y\boldsymbol{x} = s_2, \quad z = \boldsymbol{c}_z\boldsymbol{x} = s_2, \quad \boldsymbol{x} \in X_O \subset \mathbb{R}^4, \quad d \in \mathcal{U} = [d^-, d^+] \subset \mathbb{R},
\end{aligned} \tag{5}$$