Research Article

Hybrid Control Scheme for Anaerobic Digestion in a CSTR-UASB Reactor System

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Abstract
Anaerobic digestion is an important wastewater treatment technology for industrial wastewater. To achieve the target of global environmental regulation, process control plays an important role in the system operation. The control system for anaerobic digestion process is generally applied to each reactor separately without consideration of variables that mutually affect the operation of the other one. This work proposes a hybrid control scheme for a CSTR-UASB reactor system described by a PDE-ODE model. The CSTR system is employed to rapidly reduce the inlet COD concentration while the UASB reactor is used to accurately regulate the outlet COD concentration of the system. An input-output (I/O) linearization and proportional-integral (PI) control techniques are applied to formulate the control scheme for the process. The distributed variables are applied to the developed control system for handling the spatially distributed dynamics of the bacterial biomass. The COD concentration of both reactors are manipulated through the dilution rate and feed flow rate to achieve the desired targets. Simulation results of the closed-loop system illustrate that the developed control scheme regulates the controlled outputs to follow the desired trajectories and manipulate the control problems effectively.

Keywords: CSTR-UASB reactors, Input-output linearization control, Wastewater treatment, Coupled PDE-ODE system, Anaerobic digestion

1 Introduction

Due to the high capability to degrade various types of organic substrates, the anaerobic digestion system is a high potential technology for industrial wastewater treatment [1], [2]. The two-step biochemical reaction (acidogenesis-methanogenesis) is typically considered in the anaerobic digestion process. For the two-step reaction, the organic substrate (COD) is consumed by the group of acidogenic bacteria to formulate volatile fatty acid (VFA). The VFA is then converted to methane under anaerobic condition. The unit operations such the continuous stirred-tank reactor (CSTR) and the upflow anaerobic sludge blanket (UASB) reactor are widely applied to the wastewater treatment system for many industries. The advantage of the CSTR system is

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well mixing, which can promote the organic substrate degradation while the UASB reactor performs a rapid rate of methane production. For the UASB reactor, the substrates for both bacterial groups can be assumed to be a perfect mixing but the distribution of bacterial biomass presents a complex behavior with spatial dynamics. Due to the nonlinear behavior of the UASB reactor with recirculation, the control performance of the typically proportional-integral (PI) controller may be limited [3].

In order to regulate the nonlinear system, research works have proposed the advanced control approaches in the past decades [4], [5]. A fuzzy logic control that predicts the biogas and methane production are applied to a UASB reactor with multiple inputs multiple outputs (MIMO) fuzzy logic model in [4]. Waewsak and Nopharatana [6] used a neural network model to predict the state variables such total VFAs, alkalinity pH, which are considered as inputs for a fuzzy logic controller to manipulate the feeding rate of a hybrid reactor system. Results from the experiment showed a good response to regulate the system during start-up operation, overload condition, including the recovery phase [6]. A UASB reactor described by a PDE model is used to formulate a multivariable linear quadratic tracking (LQT) approach. The work employed its recycle/by-pass stream to track the effluent COD concentration [7]. In the UASB reactor, the dynamics of granular biomass perform spatial distribution while the soluble substrate can be lumped, hence the process model used to describe the dynamic behavior of the system can be considered as a coupled partial differential equation- ordinary differential equation (PDE-ODE) model. There are some works developed the control strategies for the chemical and biochemical processes described by coupled PDE-ODE models. Liu and Krstić proposed a back stepping boundary control for the system of Burgers' equation [8]. Recently, control schemes based on the input-output (I/O) linearization technique were applied to a thermal cracking furnace and a UASB reactor [9], [10]. However, the robustness of the control scheme may be limited for the wastewater treatment process with a wide range of the inlet concentration changes.

This work develops a control scheme for a CSTR-UASB reactor system. The PI controller and I/O linearization control technique are applied to the coupled PDE-ODE model. Since the process dynamics presents the bidirectional interconnection between the substrate and biomass, the control objective of the proposed controller aims to handle the organic substrate concentration for both reactors. An integrator is applied to provide the integral action that compensates the process-model mismatch during the operation. To investigate errors of the process and model, a nonlinear state observer is applied to estimate unmeasured variables of the system. The control performance tests are conducted under the conditions with changes of the inlet stream concentration to investigate the control robustness, and the results show good responses. The advantages of the proposed hybrid control scheme are the high performance to stabilize the process responses and avoid the inhibition effects.

2 Materials and Methods

2.1 *Problem formulation*

In order to describe the CSTR-UASB system for anaerobic digestion process, a model system described by the coupled partial differential equation and ordinary differential equation is proposed as follows:

$$\frac{\partial \xi_p^U(z,t)}{\partial t} = A\frac{\partial^2 \xi_p^U(z,t)}{\partial z^2} + B\frac{\partial \xi_p^U(z,t)}{\partial z} + f_p^U\left(\xi_p^U(z,t), \xi_o^U(t)\right) \tag{1}$$

$$\frac{\partial \xi_o^U(t)}{\partial t} = f_o^U\left(\xi_o^U(t), \tilde{\xi}_p^U(t), u_U(t)\right) \tag{2}$$

$$\frac{\partial \xi_o^C(t)}{\partial t} = f_o^C\left(\xi_o^C(t), u_C(t)\right) \tag{3}$$

$$y^U = h_U\left(\xi_p^U(z,t), \xi_o^U(t)\right)\Big|_{z=H} \tag{4}$$

$$y^C = h_C\left(\xi_o^C(t)\right) \tag{5}$$

with the initial and Danckwerts-boundary conditions, [Equations (6)–(10)]

$$A\frac{\partial \xi_p^U}{\partial z} = B\left(\xi^U{}_{(z=0)} - \xi_{in}^U\right), \qquad z = 0 \tag{6}$$

$$\frac{d\xi_p^U}{dz} = 0, \qquad z = H \tag{7}$$

$$\xi_p^U(z,0) = \xi_{p,}^U(z) \tag{8}$$

$$\xi_o^U(0) = \xi_{o,0}^U \tag{9}$$

$$\xi_o^C(0) = \xi_{o,0}^C \tag{10}$$

where $\xi_p^U(z,t)$ denotes the vector of state variables of the UASB reactor depending on the spatial coordinate (distance) and time, $\xi_o^U(t)$ and $\xi_o^C(t)$ denote the vector of time-dependent state variables of the UASB reactor and the CSTR, respectively. $\tilde{\xi}_p^U(t)$ denotes the state of ξ_p^U at the outlet position of the UASB reactor, y^U and y_C denote the vectors of the output variables of the UASB reactor and the CSTR, respectively. $z \in [0, H]$ is the spatial coordinate, $n_U(t)$ and $n_C(t)$ denote the vector of manipulated variables of the UASB reactor and the CSTR, respectively. $t \in [0, \infty]$ is the time, f_p^U, f_o^U and f_o^C are the vectors of nonlinear functions. A and B are matrices and h_U and h_C are nonlinear functions.

2.2 *Input-output linearization technique for the coupled PDE-ODE model*

For the proposed system, the control structure of the UASB reactor is based on an I/O linearization control technique. The dynamic model of state variables $\xi_p^U(z,t)$ and $\xi_o^U(t)$ are considered to formulate the I/O linearization controller. The system in Equations (1)–(5) can be written in the compact form as

$$\dot{\xi}^U = f^U\left(\xi^U, \xi_z^U, \xi_{zz}^U, u_U\right), \tag{11}$$

$$y_H^U = h_U\left(\xi^U\right)\Big|_{z=H} \tag{12}$$

where $\xi^U = \left[\xi_p^U, \xi_o^U\right]^T$ denotes the vector of state variables, $\xi_{zz}^U = \xi^U / \partial z^2$ is the second-order spatial derivatives of ξ^U, $\xi_z^U = \partial\xi^U / \partial z$ denotes the first-order spatial derivatives of ξ^U, u_U is the manipulated variable and y_H^U is the controlled output y_H^U at the outlet position of active volume y^U. f^U and h_U are the vectors of nonlinear functions.

To examine the relation between the controlled output and manipulated input, the relative order of the system in Equations (11) and (12) is denoted by r. The following equation can be used to explain the finite parameter.

$$y_H^U = h_U\left(\xi^U\right)\Big|_{z=H} \tag{13}$$

$$\frac{dy_H^U}{dt} = \left[\frac{\partial h_U}{\partial \xi^U}\frac{\partial \xi^U}{\partial t}\right]\Bigg|_{z=H} = h_U^1\left(\xi^U, \xi_z^U, \xi_{zz}^U\right)\Big|_{z=H} \tag{14}$$

$$\frac{d^2 y_H^U}{dt^2} = \left[\frac{\partial h_U^1}{\partial \xi^U}\frac{\partial \xi^U}{\partial t}\xi^U + \frac{\partial h_U^1}{\partial \xi_z^U}\left(\xi_z^U\right)^{(1)} + \frac{\partial h_U^1}{\partial \xi_{zz}^U}\left(\xi_{zz}^U\right)^{(1)}\right]_{z=H} = h_U^2\left(\xi^U, \xi_z^U, \xi_{zz}^U, \left(\xi_z^U\right)^{(1)}, \left(\xi_{zz}^U\right)^{(1)}\right)\Big|_{z=H} \tag{15}$$

$$\vdots$$

$$\frac{d^{r-1} y_H^U}{dt^{r-1}} = \left[\frac{\partial h_U^{r-2}}{\partial \xi^U}\frac{\partial \xi^U}{\partial t} + \frac{\partial h_U^{r-2}}{\partial \xi_z^U}\left(\xi_z^U\right)^{(1)} + \frac{\partial h_U^{r-2}}{\partial \xi_{zz}^U}\left(\xi_z^U\right)^{(1)} + \ldots + \left(\xi_{zz}^U\right)^{(r-2)}\right]_{z=H} = h_U^{r-1}\left(\xi^U, \xi_z^U, \xi_{zz}^U, \left(\xi_z^U\right)^{(1)}, \ldots, \left(\xi_z^U\right)^{(r-2)}, \left(\xi_{zz}^U\right)^{(1)}, \ldots, \left(\xi_{zz}^U\right)^{(r-2)}\right)\Big|_{z=H} \tag{16}$$

$$\frac{d^{r} y_H^U}{dt^{r}} = \left[\frac{\partial h_U^{r-1}}{\partial \xi^U}\frac{\partial \xi^U}{\partial t} + \frac{\partial h_U^{r-1}}{\partial \xi_z^U}\left(\xi_z^U\right)^{(1)} + \frac{\partial h_U^{r-1}}{\partial \xi_{zz}^U}\left(\xi_{zz}^U\right)^{(1)} + \ldots + \left(\xi_{zz}^U\right)^{(r-1)}\right]_{z=H} = h_U^{r}\left(\xi^U, \xi_z^U, \xi_{zz}^U, \left(\xi_z^U\right)^{(1)}, \ldots, \left(\xi_z^U\right)^{(r-1)}, \left(\xi_{zz}^U\right)^{(1)}, \ldots, \left(\xi_{zz}^U\right)^{(r-1)}, u_U\right)\Big|_{z=H} \tag{17}$$

where $\left(\xi_z^U\right)^{(1)} = d^l \xi_z^U / dt^l$ and $\left(\xi_{zz}^U\right)^{(1)} = d^l \xi_{zz}^U / dt^l$.

An application of I/O linearization technique is conducted to obtain a linear response of the closed-loop actual output at the outlet position of the UASB reactor $\left(y_H^U\right)$. It can be written as the following form:

$$\left(\beta D + 1\right)^r y_H^U = y_{sp}^U \tag{18}$$

where D is defined as the differential operator ($D=d/dt$), $\boldsymbol{y}_{sp}^{U}$ is the output set point, and β denotes the tuning parameter.

The Equations (13)–(18) are applied to formulate the feedback controller. Then the manipulated input of the UASB reactor $\boldsymbol{n}_U$ can be solved and expressed in the compact form by [Equation (19)]

$$\boldsymbol{u}_U = \Psi\left(\bar{\xi}^{\boldsymbol{U}}, \bar{\xi}_z^{\boldsymbol{U}}, \bar{\xi}_{zz}^{\boldsymbol{U}}, \boldsymbol{y}_{sp}^{U}\right)\Big|_{z=H} \quad (19)$$

where $\bar{\xi}^{\boldsymbol{U}}$ is the vector of state variables which can be measured. $\bar{\xi}_z^{\boldsymbol{U}}$ and $\bar{\xi}_{zz}^{\boldsymbol{U}}$ are the first- and second-order spatial derivatives of $\bar{\xi}^{\boldsymbol{U}}$, respectively.

2.3 *Process model of the CSTR-UASB reactor system for anaerobic digestion*

A simplified schematic of a CSTR-UASB reactor system is shown in Figure 1.

The wastewater treatment for an industrial wine distillery is applied in this work. Anaerobic digestion processes for the wine distillery wastewater have been studied in various works [11]–[13]. In general, the organic content (COD) of the process can be varied in the range 10–40 kg/m^3 (10,000–40,000 mg/L) [13]. In order to explain the dynamic behaviors for the UASB reactor, some works developed the process models with the distributed parameters (PDE model) [14]–[16]. For the purpose of developing process control technique, the PDE model is significantly useful to formulate the model-based control system [7]. Bernard's kinetic model and distributed parameter model presented in [7] and [17] are applied in this work. The soluble substrates for both bacterial groups can be assumed to be a perfect mixing while the granular biomass can be assumed to be solid phase and presents a behavior with spatial dynamics in the liquid phase [16], [18].

Although the UASB reactor with a high performance controller may regulate the COD concentration of the outlet stream to follow the desired setpoint, common uncertainties in the inlet stream concentration could limit the process stability. Thus the CSTR system is employed to manipulate the inlet COD concentration of the UASB reactor and avoid the inhibition effects. The UASB reactor is a circular column with a recycle stream R_r. For both CSTR and UASB reactors, the state variables of the output stream

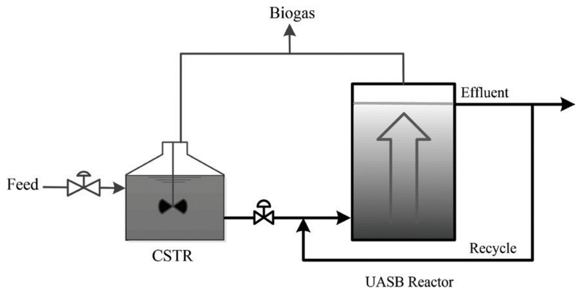

Figure 1: Schematic of a CSTR-UASB reactor system.

can be measured at the medium outlet of the active volume. In order to develop a nonlinear model for the CSTR-UASB reactor system, additional assumptions are defined to simplify the process model:

1) There are two main groups of bacterial population in the system, acidogenic and methanogenic bacteria.

2) The flow of medium and biomass in the UASB reactor is one-dimensional consideration.

3) The pH of the medium is controlled effectively for both reactors, in a range pH 7–8.

The developed mathematical model is based on a two-step mass balance equation [17]. For the CSTR system, it is assumed that the substrates (COD, VFAs) and biomass concentrations are perfect mixing and can be modeled by the ODE system. For the UASB reactor, the substrate concentrations can be explained by the ODE system while the biomass concentrations are spatially distributed and modeled by the PDE system. The concept of the distributed parameter models [7] is applied, and the coupled PDE-ODE model is expressed as the following equation.

$$\frac{\partial X_1^U}{\partial t} = D_x \frac{\partial^2 X_1^U}{\partial z^2} - u_{up,x}\frac{\partial X_1^U}{\partial z} + y_1\mu_1\left(S_1^U\right)X_1^U \quad (20)$$

$$\frac{\partial X_2^U}{\partial t} = D_x \frac{\partial^2 X_2^U}{\partial z^2} - u_{up,x}\frac{\partial X_2^U}{\partial z} + y_2\mu_2\left(S_2^U\right)X_2^U \quad (21)$$

$$\frac{\partial S_1^U}{\partial t} = D_U\left(S_1^C - S_1^U\right) - y_3\mu_1\left(S_1^U\right)X_1^U \quad (22)$$

$$\frac{\partial S_2^U}{\partial t} = D_U\left(S_2^C - S_2^U\right) - y_4\mu_2\left(S_2^U\right)X_2^U + \left(1 - y_1\right)\mu_1\left(S_1^U\right)X_1^U \quad (23)$$

$$\frac{\partial X_1^C}{\partial t} = D_C \cdot \left(X_{1,in} - X_1^C\right) + y_1 \cdot \mu_1\left(S_1^C\right)\cdot X_1^C \quad (24)$$

$$\frac{\partial X_2^C}{\partial t} = D_C \cdot \left(X_{2,in} - X_2^C\right) + y_2 \cdot \mu_2\left(S_2^C\right) \cdot X_2^C \quad (25)$$

$$\frac{\partial S_1^C}{\partial t} = D_C \cdot \left(S_{1,in} - S_1^C\right) - y_3 \cdot \mu_1\left(S_1^C\right) \cdot X_1^C \quad (26)$$

$$\frac{\partial S_2^C}{\partial t} = D_C \cdot \left(S_{2,in} - S_2^C\right) - y_4 \mu_2\left(S_2^C\right) X_2^C + \left(1 - y_1\right)\mu_1\left(S_1^C\right) X_1^C \quad (27)$$

with the Danckwerts boundary conditions [19],

$$D_x \frac{\partial X_1^U}{\partial z} = u_{up,x}\left(X_{1(z=0)}^U - X_1^C\right), \qquad z = 0 \quad (28)$$

$$\frac{dX_1^U}{dz} = 0, \qquad z = H \quad (29)$$

$$D_x \frac{\partial X_2^U}{\partial z} = u_{up,x}\left(X_{2(z=0)}^U - X_2^C\right), \qquad z = 0 \quad (30)$$

$$\frac{dX_2^U}{dz} = 0, \qquad z = H \quad (31)$$

and initial conditions

$$X_1^U\left(z,0\right) = X_{1,0}^U, \quad X_2^U\left(z,0\right) = X_{2,0}^U \quad (32)$$

$$S_1^U\left(z,0\right) = S_{1,0}^U, \quad S_2^U\left(z,0\right) = S_{2,0}^U \quad (33)$$

$$X_1^C\left(z,0\right) = X_{1,0}^C, \quad X_2^C\left(z,0\right) = X_{2,0}^C \quad (34)$$

$$S_1^C\left(z,0\right) = S_{1,0}^C, \quad S_2^C\left(z,0\right) = S_{2,0}^C \quad (35)$$

$$S_1^C\left(z,0\right) = S_{1,0}^C, \quad S_2^C\left(z,0\right) = S_{2,0}^C \quad (36)$$

where X_1^U and X_2^U are the biomass concentration of acidogenic and methanogenic bacteria of the UASB reactor. X_1^C and X_2^C are the biomass concentration of acidogenic and methanogenic bacteria of the CSTR. S_1^U and S_2^U are the COD and total VFA concentrations of the UASB reactor. S_1^C and S_2^C are the COD and total VFA concentrations of the CSTR. D_x is the coefficient of axial dispersion for the solid phase of biomass. D_U and D_C are dilution rates of the UASB reactor and the CSTR, respectively. $u_{up,x} = \alpha \times v$, where $u_{up,x}$ is the velocity of bacterial biomass in the upflow direction, v is the media velocity and α is constant. y_1, y_2, y_3, y_4 are the yield coefficients of acidogenic biomass, methanogenic biomass, COD concentration and VFA concentration, respectively. For inlet stream variables of the CSTR system, $S_{1,in}$, $S_{2,in}$ are the inlet concentration while $X_{1,in}$ and $X_{2,in}$ are the influent biomass concentration, respectively. The growths of acidogenic bacteria are described by the Monod kinetics for growths as [Equations (37) and (38)]:

$$\mu_1\left(S_1^U\right) = \mu_{1,\max} \frac{S_1^U}{K_{S1} + S_1^U} \quad (37)$$

$$\mu_1\left(S_1^C\right) = \mu_{1,\max} \frac{S_1^C}{K_{S1} + S_1^C} \quad (38)$$

For the methanogenic bacteria, the Haldane kinetics are used as [Equations (39) and (40)]:

$$\mu_2\left(S_2^U\right) = \mu_{2,\max} \frac{S_2^U}{K_{S2} + S_2^U + \frac{\left(S_2^U\right)^2}{K_{I2}}} \quad (39)$$

$$\mu_2\left(S_2^C\right) = \mu_{2,\max} \frac{S_2^C}{K_{S2} + S_2^C + \frac{\left(S_2^C\right)^2}{K_{I2}}} \quad (40)$$

where μ_1 and μ_2 are the maximum bacterial growth rate for acidogens and methanogens, respectively. K_{S1} denotes the half-saturation constant, this parameter is related to organic substrate property. K_{S2} is the half-saturation constant while K_{I2} is the inhibition constant, these parameters are related to the VFA property. The model parameters for the CSTR-UASB reactor system are listed in Table 1.

2.4 *Control system formulation*

2.4.1 PI controller for the CSTR system

In this work, the PI controller is proposed to regulate the CSTR system [20], [21]. The main function of the reactor is providing the COD concentration within a suitable range before feeding to the UASB reactor. The manipulated input for the CSTR system can be formulated as [Equation (41)]

$$u_C(t) = K_p \cdot e(t) + K_i \cdot \int_0^t e(t') \quad (41)$$

where u_C is the manipulated input of the UASB reactor. $e(t) = y_{sp}^C - y^C(t)$ is the error. K_p and K_i are tuning parameters.

Table 1: Parameters of the CSTR-UASB reactor system.

Symbol	Value	Meaning
K_{S1}	0.5 kg m^{-3}	Half-saturation constant for COD concentration
K_{S2}	0.15 kg m^{-3}	Half-saturation constant for VFA concentration
K_{I2}	0.5 kg m^{-3}	Inhibition constant
H_i	3.35 m	Medium height
h_i	3.5 m	Reactor height
D_x	0.1 m^{2}d^{-1}	Axial dispersion coefficient
D_i	0.6 m	Reactor diameter
V	0.948 m^3	Media effective volume
$\mu_{1,\max}$	0.4 d^{-1}	Maximum growth rate of acidogens
$\mu_{2,\max}$	0.7 d^{-1}	Maximum growth rate of methanogens
y_1, y_2	0.05,0.05	Yield coefficients
y_3, y_4	1,1	Yield coefficients
α	0.001	Constant

2.4.2 Application of I/O linearization control technique for the PDE-ODE model

For the UASB reactor, the I/O linearization control technique is applied to support the complex behavior of the coupled PDE-ODE model [9], [10]. The application of the linear response explained in Equation (18) is conducted to formulate the controller of the UASB reactor. Although the biomass concentrations are considered as distributed parameters, both substrate concentration of the reactor is considered as a lumped variable. The controlled output of the reactor is then linearized as [Equation (42)]

$$\beta_U D S_1^U + S_1^U = y_{sp}^U \quad (42)$$

where y_{sp}^U denotes the set point of the UASB reactor and β_U is a tuning parameter. An application of the process model of Equations (20)–(36) is conducted, and then solved to obtain the control action. The compact form of the developed controller with feedback variables can be written as

$$u_U(t) = \Psi\left(S_1^U, S_{1,sp}^U, \beta, X_{1,H}^U, R_r\right) \quad (43)$$

The developed controller is then applied in each time instant to compute the control action for the system. The recirculation-to-feed ratio R_r is an important parameter that affects the inlet stream concentration of the UASB reactor, and the recycle stream must be included in the manipulated input calculation.

2.4.3 Nonlinear state observer and compensator

A nonlinear state observer is combined to the control system to estimate the unmeasured variables. The state observer can be applied to investigate the process-model mismatch and support the condition without the measured variables [22], [23]. The open-loop state observer is applied to the UASB reactor as the following equation.

$$\dot{\hat{\boldsymbol{\xi}}}_{\boldsymbol{p}}^{\boldsymbol{U}} = \boldsymbol{f}_1\left(\hat{\boldsymbol{\xi}}_{\boldsymbol{p}}^{\boldsymbol{U}}(\boldsymbol{z},\boldsymbol{t}), \hat{\boldsymbol{\xi}}_{\boldsymbol{p,z}}^{\boldsymbol{U}}, \hat{\boldsymbol{\xi}}_{\boldsymbol{p,zz}}^{\boldsymbol{U}}, \hat{\boldsymbol{\xi}}_{\boldsymbol{o}}^{\boldsymbol{U}}(\boldsymbol{t})\right) \quad (44)$$

$$\dot{\hat{\boldsymbol{\xi}}}_{\boldsymbol{o}}^{\boldsymbol{U}} = \boldsymbol{f}_2\left(\hat{\boldsymbol{\xi}}_{\boldsymbol{p,H}}^{\boldsymbol{U}}(\boldsymbol{t}), \hat{\boldsymbol{\xi}}_{\boldsymbol{o}}^{\boldsymbol{U}}(\boldsymbol{t}), \boldsymbol{u}_U(\boldsymbol{t})\right) \quad (45)$$

$$\hat{\boldsymbol{y}}^{\boldsymbol{U}} = \boldsymbol{h}_U\left(\hat{\boldsymbol{\xi}}_{\boldsymbol{p}}^{\boldsymbol{U}}(\boldsymbol{z},\boldsymbol{t}), \hat{\boldsymbol{\xi}}_{\boldsymbol{o}}^{\boldsymbol{U}}(\boldsymbol{t})\right) \quad (46)$$

$\hat{S}_1^U$

where the state observer vectors are $\hat{\boldsymbol{\xi}}_{\boldsymbol{p}}^{\boldsymbol{U}} = \left[\hat{\boldsymbol{X}}_1^U, \hat{\boldsymbol{X}}_2^U\right]^T$ and $\hat{\boldsymbol{\xi}}_{\boldsymbol{o}}^{\boldsymbol{U}} = \left[\hat{\boldsymbol{S}}_1^U, \hat{\boldsymbol{S}}_2^U\right]^T$. $\hat{\boldsymbol{y}}^{\boldsymbol{U}} = \left[\hat{\boldsymbol{S}}_1^{\boldsymbol{U}}\right]$ is the controlled output observer of the UASB reactor. $\hat{S}_1^U$ denotes the estimated variable of S_1^U, $\hat{S}_2^U$ denotes the estimated variable of S_2^U, $\hat{X}_1^U$ denotes the estimated variable of X_1^U, and $\hat{X}_2^U$ denotes the estimated variable of X_2^U. $\hat{\boldsymbol{\xi}}_{\boldsymbol{p,z}}^{\boldsymbol{U}}$ and $\hat{\boldsymbol{\xi}}_{\boldsymbol{p,zz}}^{\boldsymbol{U}}$ denote the vectors of the first- and second-order state gradients, respectively.

For the proposed control scheme, the process model is important to formulate the controller and compute the control actions. In the case that the offset from the process-model mismatch is observed, a compensator such the integrator could be applied to eliminate the errors. The equation of an integrator used in this work can be written as

$$\dot{\eta}_U = \lambda_U\left(\boldsymbol{y}^U - \boldsymbol{y}_{sp}^U\right) \quad (47)$$

$$\boldsymbol{v}_U = \boldsymbol{y}_{sp}^U - \left(\boldsymbol{y}_{sp}^U - \eta_U\right) \quad (48)$$

where η_U denotes the process integrator. $\boldsymbol{v}_U$ is a compensated set point of the UASB reactor. The control approach is developed by an application of the I/O

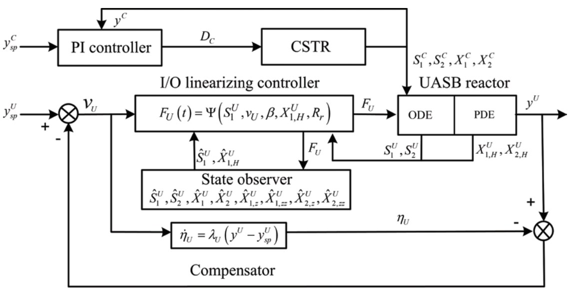

Figure 2: Schematic diagram of the proposed control scheme for the CSTR-UASB system.

linearizing controller of Equation (43), the open-loop state observer of Equations (43)–(46) and the integrator of Equations (47) and (48). The developed control scheme can be expressed as [Equation (49)]

$$\boldsymbol{u_U}(\boldsymbol{t}) = \Psi\left(\boldsymbol{y^U}, \boldsymbol{v_U}, \boldsymbol{\beta_U}, \boldsymbol{X^U_{1,H}}, \boldsymbol{R_r}\right) \tag{49}$$

where β_U is the tuning parameter of the UASB system controller. The schematic of the developed control structure can be illustrated as Figure 2.

3 Results and Discussion

3.1 *Closed-loop responses*

The coupled PDE-ODE model is applied to simulate the dynamic behaviors of the CSTR-UASB reactor system. The computing software that can solve the finite-element models such as COMSOL Multiphysics or MATLAB is applied to perform the simulations [10], [24], [25]. An application of the developed control scheme is conducted to investigate the process responses. The effluent COD concentrations of the CSTR (S_1^C) and UASB reactor (S_1^U) are controlled by manipulation of the feed flow rate (F_U) and dilution rate (D_C), respectively. In the process simulation, the initial and Danckwerts-boundary conditions are applied; $S^C_{1,0}$ = 20 kg/m^3, $S^C_{2,0}$ = 6 kg/m^3, $X^C_{1,0}$ = 7 kg/m^3, and $X^C_{2,0}$ = 6 kg/m^3.

For the wastewater treatment process, the inlet stream contains 20 kg/m^3 (20,000 mg/L) of COD concentration which needs to be reduced, y^U_{sp} = 0.12 kg/m^3 (120 mg/L), according to the environmental regulation. The CSTR system applied the PI controller with tuning parameters, K_p = –0.05 and K_I = –0.03. The developed I/O linearization control scheme is

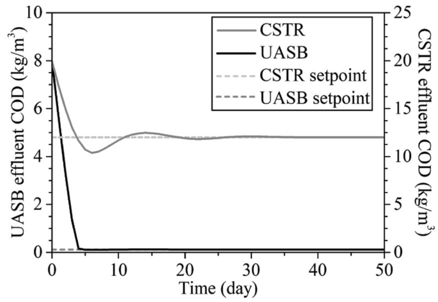

Figure 3: Closed-loop responses of the COD concentration of the CSTR and UASB reactors.

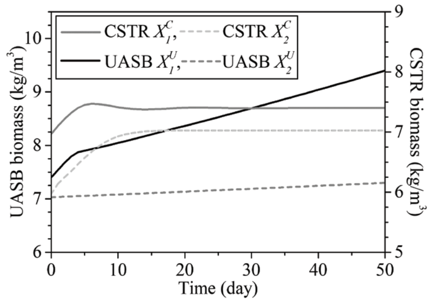

Figure 4: Biomass concentration of the CSTR and UASB reactors corresponding to the operation of Figure 3.

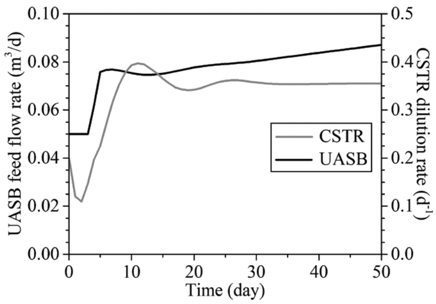

Figure 5: Dilution rate of the CSTR and feed flow rate of the UASB reactor corresponding to the operation of Figure 3.

applied to the UASB reactor with a set of tuning parameters; β_U = 0.2, γ_U = 0.01 and R_r = 0.5. The responses of the closed-loop system are shown in Figures 3–5 for the controlled outputs (effluent COD concentrations S_1^C and S_1^U), the biomass concentration (X_1^C, X_1^U) and the computational profiles of the manipulated inputs (F_U, D_C). As shown in the figures, the PI controller manipulates the CSTR system to follow the desired setpoint with oscillations. However, the I/O linearization-based control scheme applied to the UASB reactor has the ability to handle the fluctuations of the inlet stream (outlet stream of the CSTR system) and force the process response to the requested trajectories, effectively. Additionally, it is found that the proposed control scheme presents less oscillation of the output COD concentration response when compared to a previous work that applied a linear quadratic tracking (LQT) approach [7].

3.2 *Control performances*

The change in the inlet stream concentration is an important problem that commonly occurs in the wastewater treatment process. The variation of the inlet concentration could strongly affect the bacterial growth of the anaerobic digestion, especially the effect of substrate inhibition. The effect could limit the methane production and the outlet stream still contains a high organic content. Thus the control problem is determined as the process disturbances in this work.

In order to investigate the control performances of the developed control strategy, control problems are introduced to the closed-loop system in this work. An increase of the inlet COD concentration of the CSTR system is applied after the controlled outputs reached the desired set point: the inlet COD concentration is normally set at 20 kg/m^3, then it is changed to be 28 kg/m^3 (increase by 40%). The COD concentration and the manipulated inputs are shown in Figures 6 and 7, respectively. After the disturbance is added, the COD concentration of the CSTR outlet stream contains the oscillations around the desired setpoint before feeding to the UASB reactor.

The results demonstrate that the proposed control scheme has performance to handle the introduced control problem and successfully reject the disturbance. For the adjustment of manipulated inputs, it can be seen that the controller of the CSTR system needs to reduce the dilution rate to maintain the setpoint. And the controller of the UASB reactor adjusts the feed flow rate to handle the fluctuation of the inlet stream, effectively.

Since the inlet COD concentration is the main carbon source to formulate the VFA and methane, a decrease of the parameter could reduce the methane production of the wastewater treatment process. Thus a decrease of the COD concentration of the CSTR inlet stream is added as a control problem of the proposed control system. The inlet COD concentration is initially set at 20 kg/m^3, then it is increased by 20%. The COD concentration and the manipulated inputs are shown in Figures 8 and 9, respectively.

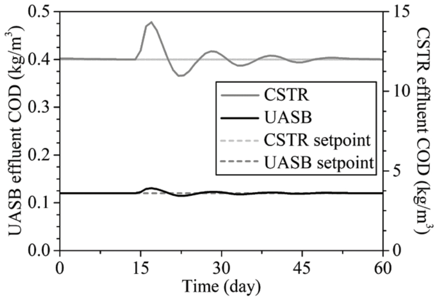

Figure 6: Effluent COD concentration of the CSTR and UASB reactors under the condition with an increase in the inlet stream concentration.

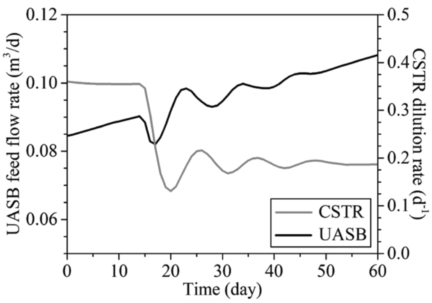

Figure 7: Dilution rate of the CSTR and feed flow rate of the UASB reactor corresponding to the operation of Figure 6.

It is clear from the results that the controllers of both reactors have the ability to manipulate the problem and force the responses to follow the desired trajectories. Since the disturbance affects the controlled output of the CSTR system, the manipulated input is increased to sustain the target. To regulate the problem, the I/O linearizing controller of the UASB reactor generates paths for setpoint tracking and stabilizes the process responses while the integrator is applied to compensate the estimated errors and quickly reject process disturbances.

Due to a high concentration of organic materials in wastewater, lagoons, activated sludge plants or

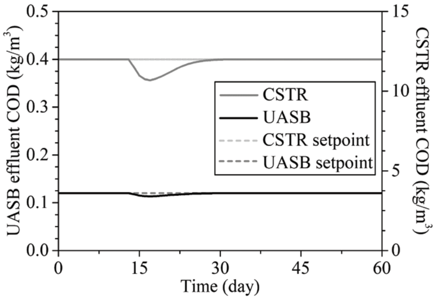

Figure 8: Effluent COD concentration of the CSTR and UASB reactors under the condition with a decrease in the inlet stream concentration.

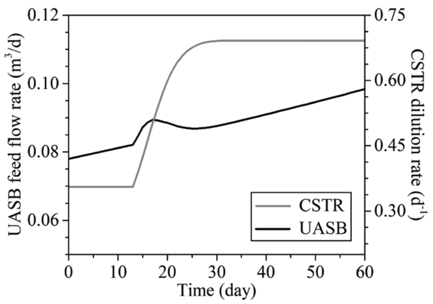

Figure 9: Dilution rate of the CSTR and feed flow rate of the UASB reactor corresponding to the operation of Figure 8.

evaporation ponds are generally used to reduce the COD concentration. In cases that the system fails to control the outlet concentration, the effluents are usually disposed or managed by public courses, and cause a serious environmental problem [11]. The proposed CSTR-UASB system and control scheme can be applied as an alternative way to rapidly reduce and accurately regulate the COD concentration of the effluents effectively. Note that this work proposed a control scheme to improve control performance of the CSTR-UASB reactor system, further investigation of the developed control strategy can be conducted by implementing the control algorithm to the industrial-scale process. To support the control structure, the application of the variables from the feedback monitoring system and the state observer are required. Since the proposed model-based controller was developed under the assumptions, some model parameters may need to be adjusted to fit the experimental data, especially the parameters associated with the bacterial growth [26], [27]. For the process model applied in this work, we realize that the mismatch between the process and model can be presented, hence the integral action has been added to compensate the errors and handle the problem.

4 Conclusions

This paper has developed a hybrid control scheme for a CSTR-UASB reactor system. The control strategy is designed to handle the variables that mutually affect the reactors and reject the unmeasured disturbances. The feedback control system, based on the concept of I/O linearization and PI control technique, regulates the outlet COD concentrations by manipulating the dilution rate (CSTR) and feed flow rate (UASB). The proposed controller was combined with an integrator and nonlinear state observer to investigate the unmeasured disturbances and compensate the process-model mismatch. The developed control scheme was applied to the process system to investigate the closed-loop responses by the simulations. Results showed that the proposed control scheme successfully regulates the COD concentration of the outlet stream. The process operation with changes in inlet COD concentration demonstrated that the control system has the ability to reduce a high organic substrate and force the control output to follow the desired target accurately.

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