

Linear Parameter-Varying Control of a Copolymerization Reactor

Hossam S. Abbas* Sandy Rahme** Nader Meskin**
Christian Hoffmann*** Roland Tóth****
Javad Mohammadpour†

* *Electrical Engineering Dept., Faculty of Engineering, Assiut University, 71515 Assiut, Egypt, (e-mail: hossam.abbas@aun.edu.eg).*

** *Dept. of Electrical Engineering, College of Engineering, Qatar University, PO Box 2713 Doha, Qatar, (e-mail: [sandy.rahme,nader.meskin](mailto:{sandy.rahme,nader.meskin}@qu.edu.qa))@qu.edu.qa*

*** *Institute of Electrical Engineering in Medicine, Universität zu Lübeck, 23558 Lübeck, Germany (christian.hoffmann@uni-luebeck.de)*

**** *Control Systems Group, Eindhoven University of Technology, P.O. Box 513, 5600 MB Eindhoven, The Netherlands, (r.toth@tue.nl)*

† *Complex Systems Control Lab, College of Engineering, The University of Georgia, Athens, GA 30602, USA, (javadm@uga.edu)*

Abstract: This paper demonstrates the application of the *linear parameter-varying* (LPV) framework to control a copolymerization reactor. An LPV model representation is first developed for a nonlinear model of the process. The LPV model complexity in terms of the model order and the number of scheduling variables is then reduced by truncating those system states that have insignificant direct influence on the input-output behavior of the system and do not directly appear in the output equations of the model. It is important to note that these truncated states are still preserved in the reduced model by affecting the scheduling parameters and hence enabling the representation of the input-output map. Using the derived model, a *linear fractional transformation* (LFT) based LPV controller synthesis approach is used to synthesize a controller for the process. Simulation based studies of the closed-loop behavior of the system regulated by the designed LPV controller demonstrate that the LPV controller solution outperforms a model predictive control designed previously for this system in terms of the achieved control performance and the online computational effort.

© 2015, IFAC (International Federation of Automatic Control) Hosting by Elsevier Ltd. All rights reserved.

1. INTRODUCTION

From an industrial perspective, there are interesting incentives for efficient control of polymer reactors. These include maximization of production quality and rates and minimization of transition losses. Control of polymerization reactors is a difficult task due to several challenges specific to them. Basically, they are highly *nonlinear* (NL) systems whose states exhibit strongly coupled behavior (Soroush and Kravari, 1993), (Özkana et al., 2003), and hence they exhibit multiple steady states, as well as oscillatory or unstable modes at some operating points (Congalidis and Richards, 1998). They can be highly exothermic and result in a reactor thermal runaway in the absence of an appropriate control strategy. Several control approaches have been investigated for these processes, see (Richards and Congalidis, 2006). Classical control methods such as standard PID and multi-loop PID control, e.g., (Congalidis et al., 1989), are commonly used as they require minimal process knowledge. However, they are not adequate to cope with such a multivariate control problem with strong interactions between the controlled variables. A further relevant control approach relies on model predictive control techniques based on simple process models, e.g., (Maner and Doyle, 1997), (Özkana et al., 2003), that allow a rapid

transition between different operating points. Nonlinear control (Soroush and Kravari, 1993) has been considered as well, which depends on the availability of a highly accurate nonlinear model and online measurements of *time-varying* (TV) model parameters.

Generally, optimal control techniques are preferred if a good process model is available because of their presumed superior performance (Embirucu et al., 1996). Moreover, adaptive control strategies can be applied in order to take the time-varying nature of the process into account, provided that online measurements/estimations are available. *Linear parameter-varying* (LPV) control (Mohammadpour and Scherer, 2012) is a promising candidate for control design that can be classified as an adaptive control technique based on the extension of powerful linear time-invariant (LTI) approaches such as $\mathcal{H}_2/\mathcal{H}_\infty$ optimal control, see e.g., (Mohammadpour and Scherer, 2012), to address the control design problem for NL and TV systems. LPV systems are represented by dynamical models capable of describing TV behaviors in terms of a linear structure depending on the so-called scheduling variables, which are assumed to be online measurable. The variation of the scheduling variables represents time-variance, changing operating conditions, etc. The capability of LPV framework to model and control NL/TV systems has been

demonstrated in terms of several successful applications, e.g., (Abbas et al., 2014) and (Bachnas et al., 2014).

In this paper, we investigate an LPV control design strategy to control a free radical solution polymerization in a *continuously stirred tank reactor* (CSTR). To implement the LPV controller, an LPV representation in state-space form of the nonlinear model developed in (Congalidis et al., 1989) is first introduced. The complexity of the model is reduced by a specific model reduction approach which also preserves the input-output behavior of the full model. Then, the operating region is specified and the control design approach introduced in (Scherer, 2001) is utilized to design an LPV controller based on the mixed sensitivity design. The objectives of this study are twofold: (1) Show a practically applicable LPV modeling technique and complexity reduction approach for copolymerization processes. (2) Assess the capabilities of LPV control to providing high performance control solution for polymerization reactors.

The paper is organized as follows: In Section 2, the NL dynamical model of the copolymerization reactor is reviewed and an LPV state-space model representation of the process is developed. In Section 3, a low-complexity LPV representation for the process is developed. The synthesis of an LPV controller and the closed-loop implementation are demonstrated in Section 4. Finally, conclusions are drawn in Section 5. Notations considered in this paper are standard. The subscripts a, b, i, s, t, z are used to indicate monomer A, monomer B, initiator, solvent, chain transfer, and inhibitor, respectively. The symmetric completion of a matrix is denoted by $*$ and $\ker[X]$ denotes the nullspace of a matrix X .

2. PRELIMINARIES

2.1 Nonlinear Dynamic Model of The Process

In this paper, we consider the model of the solution copolymerization reactor developed in (Congalidis et al., 1989); a flow diagram of the process is shown in Fig. 1. Monomers A and B, which are methyl methacrylate and vinyl acetate, respectively, are continuously added with initiator, solvent, and chain transfer agent, which are benzene, azobisisobutyronitrile and acetaldehyde, respectively. Moreover, an inhibitor, which is m-dinitrobenzene, may enter with the fresh feed as an impurity. The reactor is assumed to be a jacketed, well-mixed tank. A coolant flows through the jacket to remove the heat of polymerization. The NL mathematical model is based on a free radical mechanism with 27 separate reactions, and is given by

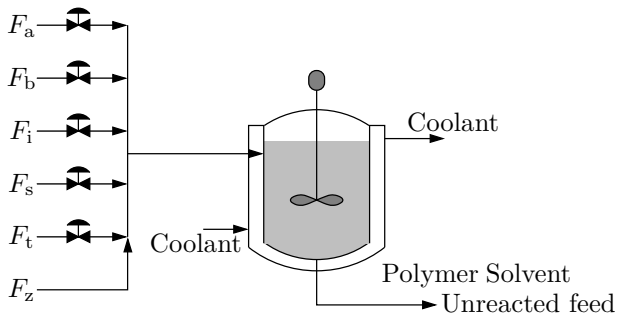


Fig. 1. Copolymerization reactor

$$\frac{d}{dt}C_k = -\phi_l C_k + \frac{F_k}{M_k V} - \frac{\sum_k F_k}{V\rho} C_k, \quad (1a)$$

$$(k, l) = \{(a, 1), (b, 2), (i, 3), (s, 4), (t, 5), (z, 6)\},$$

$$\frac{d}{dt}T_r = \frac{\phi_7}{c\rho} C_a + \frac{\phi_8}{c\rho} C_b - \frac{SU}{Vc\rho} (T_r - T_j) + \frac{\sum_k F_k}{V\rho} (T_{rf} - T_r), \quad (1b)$$

$$\frac{d}{dt}\lambda_a = \phi_1 C_a - \frac{\sum_k F_k}{V\rho} \lambda_a, \quad (1c)$$

$$\frac{d}{dt}\lambda_b = \phi_2 C_b - \frac{\sum_k F_k}{V\rho} \lambda_b, \quad (1d)$$

$$\frac{d}{dt}\psi_0 = \phi_9 C_a + \phi_{10} C_b - \frac{\sum_k F_k}{V\rho} \psi_0, \quad (1e)$$

$$\frac{d}{dt}\psi_1 = \phi_{11} C_a + \phi_{12} C_b - \frac{\sum_k F_k}{V\rho} \psi_1, \quad (1f)$$

$$\frac{d}{dt}\psi_2 = \phi_{13} C_a + \phi_{14} C_b - \frac{\sum_k F_k}{V\rho} \psi_2, \quad (1g)$$

where the inputs are the flow rates $F_a, F_b, F_i, F_s, F_t, F_z$ in kg/h, and the reactor jacket temperature T_j is in K. The states are the concentrations $C_a, C_b, C_i, C_s, C_t, C_z$ in kmol/m³, the reactor temperature T_r in K, the molar concentrations of monomer in polymer λ_a, λ_b and the moments of molecular weight distribution are ψ_0, ψ_1, ψ_2 . The variables $\phi_i, i \in \{1, \dots, 14\}$ are NL functions of the states (see Appendix A for their definition). The constant parameters are the molecular weights $M_a, M_b, M_i, M_s, M_t, M_z$, the reactor volume V , the heat surface area S , the overall heat transfer coefficient U , the heat capacity c , the density ρ and the reactor feed temperature T_{rf} ; see Table 1 for the value of these constant parameters. The important reactor output variables for control are T_r and

$$G_{pi} = M_a V \phi_1 C_a + M_b V \phi_2 C_b, \quad (2a)$$

$$Y_{ap} = \phi_{15} \lambda_a, \quad (2b)$$

$$M_{pw} = \phi_{16} \psi_2, \quad (2c)$$

where G_{pi}, Y_{ap}, M_{pw} are the polymer production rate in kg/h, mole fraction of monomer A in the polymer and the weight average molecular weight in kg/kmol, respectively; for the definition of ϕ_{15} and ϕ_{16} , see Appendix A.

Table 1. Constant parameters for the model

M_a	100.10	kg/kmol	V	1	m ³
M_b	86.09	kg/kmol	S	4.6	m ²
M_i	164.00	kg/kmol	U	216.00	J/m ² ·s·K
M_s	78.11	kg/kmol	c	2.01	kJ/kg·K
M_t	44.05	kg/kmol	ρ	879.00	kg/m ³
M_z	168.11	kg/kmol	T_{rf}	353.00	K

The main control objective in (Özkana et al., 2003) was to provide a fast transition between two operating points as shown in Table 2 with a controller that is robust against unmeasured disturbances. The first operating point, OP1, given in (Congalidis et al., 1989) was obtained for a monomer feed ratio $F_A/F_B = 0.2$ while the second one, OP2, was obtained by increasing the ratio by 0.25 keeping F_B constant. In order to achieve this objective, the manipulated variables to control the four previously specified output variables are chosen –based on the investigation in (Congalidis et al., 1989)– to be F_a, F_b, F_t and T_j . For comparison purposes, we consider in this work the same

Download English Version:

<https://daneshyari.com/en/article/711589>

Download Persian Version:

<https://daneshyari.com/article/711589>

[Daneshyari.com](https://daneshyari.com)