Journal of Process Control 24 (2014) 1358–1370

Contents lists available at ScienceDirect

Journal of Process Control

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

A model structure-driven hierarchical decentralized stabilizing control structure for process networks

Katalin M. Hangos [a,b,*], Gábor Szederkényi [a,c]

[a] Process Control Research Group, Computer and Automation Research Institute HAS, Budapest, Hungary
[b] Department of Electrical Engineering and Information Systems, University of Pannonia, 8200 Veszprém, Hungary
[c] Faculty of Information Technology and Bionics, Pázmány Péter Catholic University, 1083 Budapest, Hungary

ARTICLE INFO

Article history:
Received 22 October 2013
Received in revised form 6 June 2014
Accepted 6 June 2014
Available online 15 July 2014

Keywords:
Process control
Process modeling
Stability
Plant wide control

ABSTRACT

Based on the structure of process models a hierarchically structured state-space model has been proposed for process networks with controlled mass convection and constant physico-chemical properties. Using the theory of cascade-connected nonlinear systems and the properties of Metzler and Hurwitz matrices it is shown that process systems with controlled mass convection and without sources or with stabilizing linear source terms are globally asymptotically stable. The hierarchically structured model gives rise to a distributed controller structure that is in agreement with the traditional hierarchical process control system structure where local controllers are used for mass inventory control and coordinating controllers are used for optimizing the system dynamics. The proposed distributed controller is illustrated on a simple non-isotherm jacketed chemical reactor.

1. Introduction

It is widely known that process systems are highly nonlinear and form complex networks. The complex dynamics of a process network is caused partially by the complex dynamic behavior of the component subsystems, but also by the effect of complex interactions. In order to cope with this complex nonlinear dynamics two principally different controller design approaches exist: the centralized and decentralized ones. The latter is most widely used in complex process plants because it offers the possibility to handle nonlinearities locally, i.e., controlling the operating units, for example, and then handle the interactions plant-wide. One of the key critical steps in this approach is to decompose the process plant and/or its control system into hierarchical and decentralized structures.

The plantwide control problem that deals with designing a complex distributed controller for a given complex process plant is widely investigated in process control and forms a traditional area of it. The first approaches were based on linear or linearized dynamic models and their properties (such as steady-state gains), see e.g. [1] or [2], where graph-theoretic approaches could be applied for efficient solution [3]. Applying the theory of linear systems, systematic approaches have been developed for complex large scale systems in general (see e.g. [4]), and for process plants in particular [5], [6]. The performance limitations in decentralized control were also investigated in the linear(ized) model case [7].

The above-mentioned systematic methods include heuristic elements in determining the controller structure, i.e., the matching of controlled and manipulated variables in the plant ([5], [6]), and they have identified controller layers that should form a hierarchy. One of such heuristics is to regulate the inventories, most notably the masses in each operating unit using the lowest controller layer of the hierarchy [8].

Although the general modern theory of possibly nonlinear, hierarchical, multilevel and distributed systems and control is well developed (see e.g. [9] for an early, and [10] for a recent reference), there are powerful and specially developed techniques for nonlinear process systems [11], too.

More recently, modern robust control techniques have been proposed for distributed control of plantwide chemical processes [12], with an attempt to extend it to the nonlinear case [13] using the notion of dissipativity [14]. This approach was further extended to the decentralized case using a Hamilton–Jacobi equation approach [15]. Powerful distributed and hierarchical variants of the popular model predictive control (MPC) have also been developed and applied to complex process plants, see [10] for a recent review.

* Corresponding author at: Process Control Research Group, Computer and Automation Research Institute HAS, Budapest, Hungary. Tel.: +36 1 279 6101; fax: +36 1 466 7503.

E-mail addresses: hangos@scl.sztaki.hu, hangos.katalin@virt.uni-pannon.hu (K.M. Hangos), szeder@scl.sztaki.hu (G. Szederkényi).

http://dx.doi.org/10.1016/j.jprocont.2014.06.003
0959-1524/

Utilizing the engineering insight into the physics and chemistry of the system, the thermodynamic passivity approach [16] as a special control approach has been proposed for nonlinear process systems that is based on controlling its inventories [17]. The controller design method has been combined with passivity [18], too. In addition to the material and energy connections in the process network, information and communication interconnections of the system and its control system are also considered in a framework integrating physics communication and computation in [19]. Furthermore, the inventory control scheme was also extended with local nonlinear controllers (see [20] and [21]) to construct stabilizing controllers to arbitrary steady-state points. Further improvements of the physically motivated nonlinear controller design have been achieved by using passivity [22], control Lyapunov [23] and Hamiltonian approaches [24–26] to nonlinear process systems.

Although the above physically motivated nonlinear control approaches exploit successfully the properties of nonlinear process systems, but the hierarchical structure of process models has not been fully utilized in process control structure design. This inherent hierarchy of dynamic process models seems useful to apply, because powerful approaches in systems and control theory (for example the contraction theory approach [27] or the theory of interconnected nonlinear systems [28]) have been proposed for stability analysis for hierarchically decomposed systems.

A critical and still generally open problem in applying the powerful hierarchical decentralized control techniques is how to decompose a complex nonlinear plant into subsystems [10]. This is traditionally performed using the process layout, where the operating units form the subsystems, but then the complex interactions, the material and energy recycles, for example, are difficult to handle.

1.1. Aim and problem statement

The main aim of this paper is to propose a decomposition that is based on the hierarchical structure of dynamic process models based on first engineering principles and use this for designing decentralized control systems to stabilize the subsystems and maintain stable performance as these subsystems are integrated into the complex process system.

The first paper in this direction [29] investigated the simplest case of process systems with constant holdup in each balance volume to show that such systems with constant pressure and no source are structurally asymptotically stable. The restriction of constant mass holdup, however, is a severe limitation which does not hold in almost any practical situation.

The above results are extended in this paper for the case of process systems with time-varying regulated mass holdup and stabilizing sources. For analyzing the stability of process systems in this extended case, the dynamic model of the system based on first engineering principles will be brought to a nonlinear cascade-connected state-space model form [28].

The usual hierarchical structure of process models and that of process control systems is considered in this paper, where low level controllers provide regulated mass holdup in each balance volume and the high level controller(s) are used for controlling the other process variables, such as temperatures and concentrations. This structure enables to partition the system model into a controlled mass subsystem that acts as a "driving subsystem" to the other part that is the "driven subsystem" [28].

With the above model structure, the stabilizing controller design will be performed by using local distributed controllers acting on the driven subsystem.

1.2. Basic assumptions

The starting point of the analysis is the general form of the state equation of a lumped process system originated from the differential conservation balances of the conserved extensive quantities over perfectly stirred regions or balance volumes. A perfectly stirred region is the smallest elementary part of the process system over which conservation balances are constructed. The following *assumptions* are made about the regions:

$\mathcal{A}1$ *Physico-chemical properties, like heat capacity, density, and heat transfer coefficients are constant.*

$\mathcal{A}2$ *The pressure is assumed to be constant.*

The above two assumptions imply that only incompressible liquid phases can be present in the regions.

The paper is organized as follows. We start with the lumped dynamic model of the mass subsystem of a process system in the next section. Thereafter we develop the model of the energy and component mass subsystems in Section 3. The stability analysis of the hierarchically decomposed state-space model is given in Section 4. The distributed controller structure driven by the hierarchically decomposed model structure is described in Section 5 illustrated by a simple case study of a jacketed CSTR. Finally conclusions are drawn.

2. The mass subsystem

Following the philosophy and the incremental approach of building a process model [31], we distinguish two subsystems in a process system. The basic subsystem is the mass subsystem upon which the energy and component mass subsystem is built.

2.1. The mass conservation balance equations

The *overall mass balance* of the perfectly stirred region j is given by the equation

$$\frac{dm^{(j)}}{dt} = v_{in}^{(j)} - v_{out}^{(j)}, \quad j = 1, \ldots, \mathcal{C} \tag{1}$$

where $v_{in}^{(j)}$ and $v_{out}^{(j)}$ are the *mass in- and out-flow rates* respectively and $\mathcal{C}$ is the number of the regions.

Note that a definite flow direction (i.e. in or out to/from a balance volume) is associated to any mass flow, i.e. $v_{in}^{(j)} \geq 0$ and $v_{out}^{(j)} \geq 0$. This implies, that two separate flows are defined to pipes where flow in either direction is allowed.

As the overall mass is conserved, the balance (1) has no source term. As we only consider process systems with incompressible fluid phases under assumptions $\mathcal{A}1$ and $\mathcal{A}2$, only the convection of the overall mass is present in the conservation balance equations.

2.1.1. Convective flows

Like in any process network, the regions are connected by flows, that can be convective flows or transfer flows. In order to describe the general case let us assume that the outlet flow of region j is divided into parts described by ratios $\alpha_\ell^{(j)}$ satisfying the equation

$$\sum_{\ell=0}^{\mathcal{C}} \alpha_\ell^{(j)} = 1, \quad j = 0, \ldots, \mathcal{C} \tag{2}$$

where $\alpha_j^{(\ell)}$ is the ratio of the outlet flow $v^{(\ell)}$ of region ℓ flowing into region j. Fig. 1 illustrates the notation. By using flows to and from