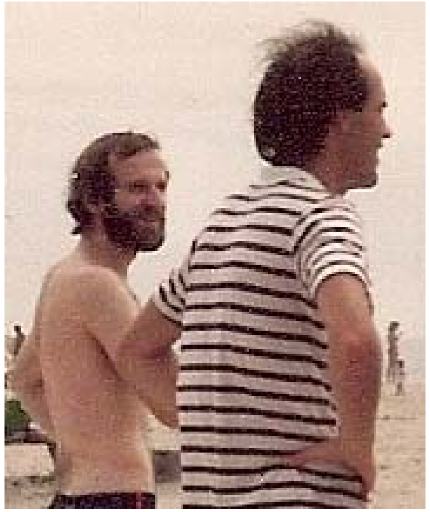
MIMO Controllability and Decentralized and Plantwide Control



Sigurd Skogestad on Manfred Morari's contributions

The start (1977): A PhD thesis on plantwide control

STUDIES IN THE SYNTHESIS OF CONTROL STRUCTURES

FOR CHEMICAL PROCESSES

A THESIS

SUBMITTED TO THE FACULTY OF THE GRADUATE SCHOOL

OF THE UNIVERSITY OF MINNESOTA

BY

MANFRED MORARI

IN PARTIAL FULFILLMENT OF THE REQUIREMENTS

FOR THE DEGREE OF

DOCTOR OF PHILOSOPHY

DECEMBER 1977

To: Terje Hertzberg & Kristian Lien
From: Sigurd (April 1984):
"Morari himself feels that his thesis is somewhat outdated and contains material of mainly academic interest"

Til Terje H. / Kristian Lien
Morari mener selv at avhandlingen er noe foreldet og inneholder mye stoff av mer akademisk interesse.
SiS 5/4-84

Acknowledgements

It is virtually impossible for me to express sufficient gratitude to Professor Stephanopoulos who did not only direct my research toward those interesting topics (sometimes through night long discussions), but who also offered me the psychological support which was so important in moments of disappointing failures. I would like to thank Professor Aris for his initial guidance and especially for being this unreachable example which increased my efforts substantially.

I would also like to thank Yaman Arkun, a friend and office mate, for those hours of discussions from which I hope he learned as much as I did.

Finally my thanks go to the National Science Foundation for the financial support during the past two years and to Ms. Chris Olsen for bringing my notes into this readable form.

AIChE Journal (Vol. 26, No. 2) March, 1980 Page 227

Studies in the Synthesis of Control Structures for Chemical Processes

Part I: Formulation of the Problem. Process Decomposition and the Classification of the Control Tasks. Analysis of the Optimizing Control Structures.

MANFRED MORARI
YAMAN ARKUN
GEORGE STEPHANOPOULOS

Department of Chemical Engineering
and Materials Science
University of Minnesota
Minneapolis, Minn. 55455

Part I of this series presents a unified formulation of the problem of synthesizing control structures for chemical processes. The formulation is rigorous and free of engineering heuristics, providing the framework for generalizations and further analytical developments on this important problem.

Decomposition is the underlying, guiding principle, leading to the classification of the control objectives (regulation, optimization) and the partitioning of the process for the practical implementation of the control structures. Within the framework of hierarchical control and multi-level optimization theory, mathematical measures have been developed to guide the decomposition of control tasks and the partitioning of the process. Consequently, the extent and purpose of the regulatory and optimizing control objectives for a given plant well defined, and alternative control structures can be generated for the signer's analysis and screening.

In addition, in this first part we examine the features of various optimi control strategies (feedforward, feedback; centralized, decentralized) and velop methods for their generation and selective screening. Application o these principles is illustrated on an integrated chemical plant that offers eno variety and complexity to allow conclusions about a real life situation.

AIChE Journal (Vol. 26, No. 2) March, 1980 Page 227

The origin of “Self-optimizing control”

In attempting to synthesize a feedback optimizing control structure, our main objective is to translate the economic objective into process control objectives. In other words we want to find a function c of the process variables (in the above distillation example, c is the pressure drop) which when held constant, leads automatically to the optimal adjustment of the manipulated variables, and with it, the optimal operating conditions.

In more concrete terms, we look for control structures such that

if

$$c(m, d) = c_d \Rightarrow \Phi = \Phi^* \qquad (21)$$

In many cases we will not be able to find c so that relationship (22) is satisfied, or that c is independent of the disturbances d. In general, we look for feedback optimizing structures which satisfy the following approximation

$$c(m, d) = c_d = \Rightarrow \Phi \simeq \Phi^* \qquad (23)$$

i.e., holding certain process variables at their set points, keeps the objective function approximately at the optimum. The ad-

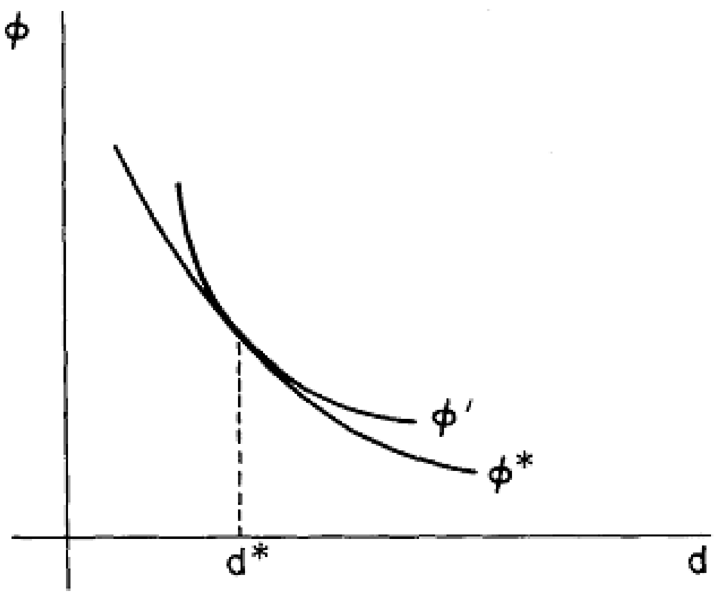


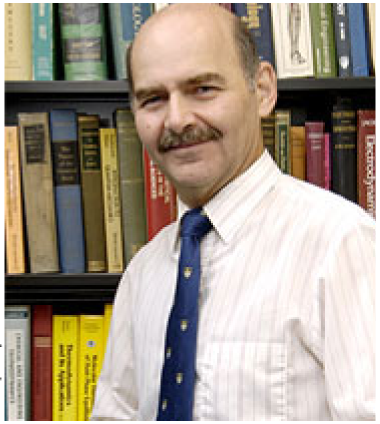
Figure 4. Optimal (Φ^*) and suboptimal (Φ') control policies with disturbance changes.

But:
- Largely forgotten until late 90’s
- No unconstrained examples
- Measurement noise not included

Controllability: *Chem. Eng. Sci.* Series on "Design of resilient processing plants"

I. Process Design Under Consideration of Dynamic Aspects (w/ Lenhoff), 1982. (40 citations)
II. Design and Control of Energy Management Systems (w/ Marselle & Rudd), 1982 (46 citations)
III. A General framework for the assessment of dynamic resilience (alone), 1983 **(112 citations)**
IV. Some New Results on Heat Exchanger Network Synthesis (w/ Saboo), 1984 (42 citations)
V. The Effect of Deadtime on Dynamic Resilience (w/ Holt). 1985 (69 citations)
VI. The Effect of Right-Half-Plane Zeros on Dynamic Resilience (w/ Holt), 1985 (85 citations)
VII. Design of Energy Management System for Unstable Reactors: New Insights (w/ Grimm, Ogelsby, Prosser), 1985 (20 citations)
VIII. A Resilience Index for Heat Exchanger Networks (w/ Saboo, Woodcock), 1985, (49 citations)
IX. Effect of Model Uncertainty on Dynamic Resilience (w/ Skogestad), 1987 (30 citations)
NaN. New characterization of the effect of RHP zeros (w/ Zafiriou and Holt), 1987 (9 citations)

Chemical Engineering Science Vol. 37, No. 2, pp. 245–258, 1982
Printed in Great Britain.



DESIGN OF RESILIENT PROCESSING PLANTS—I

PROCESS DESIGN UNDER CONSIDERATION OF DYNAMIC ASPECTS

ABRAHAM M. LENHOFF and MANFRED MORARI*
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(*Received* 15 *February* 1980; *accepted* 22 *June* 1981)

Abstract—Rising costs of natural resources have forced designers to seek more efficient processing schemes, but practical applications have been limited because of operational difficulties. A design approach is proposed which considers economic and dynamic aspects simultaneously. Optimisation of dynamic behaviour is achieved by a bounding technique based on Lagrangian theory taking advantage of the typical modular structure of processing installations. The theoretical derivation of the new approach is followed by a case study of heat-integrated distillation columns, whose results demonstrate the utility of the method.

1. INTRODUCTION

In recent years, rising raw material and energy costs have brought about a change in the philosophy of plant design. Economic necessity has forced designers to strive for more efficient utilisation of natural resources. Case studies, e.g. [2, 4, 21, 25] demonstrate the im- exchanger [28]. On the other hand, many investigations of the dynamic behavior of particular *unit operations* are available. The influence of design parameters on the dynamic behavior of CSTR's [33] and heat exchangers [8] can be mentioned as examples. The only systematic general approach to process design including dynamic

40 citations

pressive improvements obtainable by the new approach. The weakness in the improvements in design efficiency is that they usually lead to stronger dynamic interactions between the different units in a plant, which can result in operational difficulties. By making surge tanks and storage tanks smaller, or eliminating them completely, changeover times, inventory, and capital investment are reduced; there is, however, an accompanying reduction in the decoupling action of these pieces of equipment.

Assessment of an alternative should consider

(1) Steady-state aspects, i.e. an economic performance index (EPI), which is to be kept as small as possible, for example, and

(2) Operational aspects, i.e. a dynamic performance index (DPI), which should also be kept small.

(Those indices will be defined in more detail later.)

DPI = Integral squared error J

$$J = \sum_{i=1}^{p} J_i$$

$$= \sum_{i=1}^{p} 1/2 \left\{ y_{i_1}^T P_{i_1} y_{i_1} + \int_{t_0}^{t_1} [y_i^T Q_i y_i + u_i^T R_i u_i + z_i^T S_i z_i]\, dt \right\}. \tag{31}$$

Chemical Engineering Science Vol. 38, No. 11, pp. 1881–1891, 1983
Printed in Great Britain.

0009–2509/83 $3.00 + .00
Pergamon Press Ltd.

DESIGN OF RESILIENT PROCESSING PLANTS—III

A GENERAL FRAMEWORK FOR THE ASSESSMENT OF DYNAMIC RESILIENCE

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(*Received* 9 *June* 1982; *accepted* 7 *March* 1983)

Abstract—With increased process integration it has become very important to evaluate and compare the dynamic operability characteristics (dynamic resilience) of alternate designs. Based on recent results in multivariable frequency response theory a new framework is developed for this purpose. It is shown that dynamic resilience is determined by characteristics inherent in the system and that it is independent of the imposed controller structure and type. This gives the new method considerable intuitive appeal and allows it to avoid the lengthy optimization procedures which are typical for the previously published techniques.

1. INTRODUCTION

Process design can be viewed as a sequence of decision and evaluation steps. Decisions are made on the structure of the plant (e.g. how many reactors are to be used and of what type they should be, where heat exchangers are to be placed, etc.), the parameters (equipment sizing) and the control structure (what variables are to be measured, estimated, controlled or manipulated). Each decision is followed by an evaluation which can lead to the modification of earlier decisions. A dual objective is pursued in this iterative process: Some measure of profitability and the "resilience" of the plant are to be maximized. We use the term "resilience" to describe the ability of the plant to move fast and smoothly from one

which is the subject of our study. When we refer to "(dynamic) resilience" in this paper we mean the quality of the regulatory and the servo behavior which can be obtained for the plant by feedback.

Dynamic resilience clearly depends on the specific type of controller used and the selected control parameters. These choices are made by the control engineer at the end of the design phase or during the initial operation. However, it has long been recognized by both industry and academia that modifications of the physical system itself can sometimes affect the resilience significantly more than changes in the controller. Usually these process modifications for improved resilience must be balanced carefully against increases in capital and

2. DYNAMIC RESILIENCE AND SYSTEM INVERTIBILITY

The traditional method for evaluating the controllability properties of a system is to select a control structure, to tune the controller "optimally", and then to judge the performance of the system under closed loop control. *Our approach is based on the fundamental insight that the ultimate closed loop behavior is determined by constraints in the system.*

Chemical Engineering Science, Vol. 42, No. 10, pp. 2425–2428, 1987.
Printed in Great Britain.

0009–2509/87 $3.00 + 0.00

DESIGN OF RESILIENT PROCESSING PLANTS. NEW CHARACTERIZATION OF THE EFFECT OF RHP ZEROS

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and

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(*Received 11 August 1986; accepted in revised form 19 February 1987*)

Abstract—Right Half Plane (RHP) zeros restrict the achievable closed loop performance independent of controller design. A new characterization of all achievable closed loop setpoint/output transfer matrices is provided in terms of "zero-directions". The zero directions also give some insight into what forms of partial decoupling are preferable.

INTRODUCTION

Through extensive studies by a number of researchers during the last few years it has been established rigorously that for both Single-Input–Single-Output (SISO) and Multi-Input–Multi-Output (MIMO) systems RHP zeros limit the achievable closed loop performance *independent* of the control system design. RHP zeros are characteristics of the plant which can be affected only by changes of the plant itself, for example the selection of a different set of manipulated variables. A thorough understanding of the effect of RHP zeros on the achievable closed loop behavior can help the design engineer avoid process options which have inherently bad dynamic performance regardless of how well the control system is designed.

Holt and Morari (1985) (referenced as H&M in the

Definition 1

Let z_i be a zero of $G(s)$. The vector λ_i ($\lambda_i \neq 0$) satisfying $\lambda_i^T G(z_i) = 0$ is called the *direction* of the zero z_i.

Note that $G(z_i)$ is of rank $n - 1$ because the zero was assumed to be of degree one. λ_i is the eigenvector of $G(z_i)^T$ associated with the eigenvalue zero. λ_i is called zero direction because for any system input with frequency z_i the output in the direction of λ_i is identically equal to zero.

Let H_{oi} denote the transfer matrix between output o and input i. In particular, with respect to Fig. 1A we can define the following four relations

$$H_{ur} = C(I + GC)^{-1} \tag{1}$$

$$H_{ud} = -H_{ur} \tag{2}$$

Ind. Eng. Chem. Res. **1987**, *26*, 2323–2330

Implications of Large RGA Elements on Control Performance

Sigurd Skogestad[†] and Manfred Morari*

Chemical Engineering, 206-41, California Institute of Technology, Pasadena, California 91125

Large elements in the RGA imply a plant which is fundamentally difficult to control. (1) The plant is very sensitive to *uncorrelated* uncertainty in the transfer matrix elements. (2) The closed-loop system with an inverse-based controller is very sensitive to diagonal input uncertainty. With a diagonal controller, the system is *not* sensitive to diagonal input uncertainty, but the controller does not correct for the strong directionality of the plant and may therefore give poor performance even without uncertainty.

1. Introduction

Each element in the Relative Gain Array (RGA) is defined as the open-loop gain divided by the gain between the same two variables when all *other* loops are under "perfect" control (Bristol, 1966)

$$\lambda_{ij} = \frac{(\partial y_i/\partial u_j)_{u_{k\neq j}}}{(\partial y_i/\partial u_j)_{y_{k\neq i}}} = \frac{\text{gain all other loops open}}{\text{gain all other loops closed}} \qquad (1)$$

The elements λ_{ij} form the RGA Λ. Definition 1 is compactly written in terms of the transfer matrix $\mathbf{G}(s) = \{g_{ij}\} = \{(\partial y_i/\partial u_j)_{uk}\}$ and its inverse $\mathbf{G}^{-1}(s) = \{\hat{g}_{ij}\} = \{(\partial u_i/\partial y_j)_{yk}\}$ as

$$\Lambda(\mathbf{G}) = \{\lambda_{ij}\} = \{g_{ij}\hat{g}_{ij}\} = \mathbf{G}(s) \times \mathbf{G}^{-1}(s)^T \qquad (2a)$$

basis. The RGA was originally defined as a measure of interactions (eq 1) when using single-loop controllers on a multivariable plant (decentralized control), and Grosdidier et al. (1985) have proved a number of results which demonstrate the usefulness of the RGA in this respect. In particular, pairings corresponding to negative RGA elements should be avoided whenever possible. Shinskey (1967, 1984) uses the RGA extensively as a tool for selecting control configurations for distillation columns. Also he interprets the RGA primarily as an interaction measure. However, if this were the case, then an RGA analysis would be of no interest if multivariable rather than decentralized controllers were chosen. Experience indicates that this is not the case, and that the RGA is a measure of achievable control quality in a much wider sense than just as a tool

97 citations

Ind. Eng. Chem. Res. 1987, 26, 2029–2035

Effect of Disturbance Directions on Closed-Loop Performance

Sigurd Skogestad† and Manfred Morari*
Chemical Engineering, 206-41, California Institute of Technology, Pasadena, California 91125

The effectiveness of disturbance suppression in a multivariable control system can depend strongly on the direction of the disturbance. The "disturbance condition number" is introduced to quantify the effect of disturbance direction on closed-loop performance. As an example a two-point composition control system for a distillation column is analyzed for various disturbance and set-point changes.

I. Introduction

Disturbance rejection is often the main objective of process control. For multivariable systems, usually each disturbance affects all the outputs. As an example, consider a distillation column. A feed composition disturbance corresponding to an increased amount of light component in the feed leads to an *increase* of *both* product compositions y_D and x_B. (Here y_D and x_B correspond to the mole fraction of light component in the top and bottom products.) In this paper we define as "disturbance direction" the direction of the system output vector resulting from a specific disturbance. As we will show, some disturbance directions may be easily counteracted by the control system, while others may not. This has also been pointed out by Shimizu and Matsubara (1985) and Stanley et al. (1985). These papers are discussed in some detail later. The aim of this paper is to develop simple measures which may be used to indicate how the disturbances are "aligned" with the plant and thus how well they can be rejected.

Consider the linear control system in Figure 1. The process model is

$$y(s) = \mathbf{G}(s)m(s) + \mathbf{G}_d(s)z(s)$$

We will consider two different effects of disturbance directions. One is in terms of the magnitude of the manipulated variables m needed to cancel the influence of the disturbance on the process output completely at steady state. It is independent of the controller C. This may be used to identify problems with constraints at steady state. However, the issue of constraints at steady state is not really a *control* problem, but rather a plant *design* problem. Any well-designed plant should be able to reject disturbances at steady state. The second and most important effect of disturbance directions is on closed-loop performance. Here we mean by performance the behavior of the controlled outputs y in the presence of disturbances.

II. Singular Value Decomposition

Throughout this paper we will make use of the singular value decomposition (SVD) of a matrix (Klema and Laub, 1980). Any complex $n \times n$ matrix A can be written in the form

$$\mathbf{A} = \mathbf{U}\boldsymbol{\Sigma}\mathbf{V}^H \quad (4)$$

where U and V are unitary matrices ($\mathbf{U}^H = \mathbf{U}^{-1}$) and $\boldsymbol{\Sigma}$ is a diagonal matrix with real nonnegative entries

67 citations

Robust Control of Ill-Conditioned Plants: High-Purity Distillation

SIGURD SKOGESTAD, MANFRED MORARI, MEMBER, IEEE, AND JOHN C. DOYLE

Abstract—**Ill-conditioned plants are generally believed to be difficult to control. Using a high-purity distillation column as an example, the physical reason for the poor conditioning and its implications on control system design and performance are explained. It is shown that an acceptable performance/robustness trade-off cannot be obtained by simple loop-shaping techniques (via singular values) and that a good understanding of the model uncertainty is essential for robust control system design. Physically motivated uncertainty descriptions (actuator uncertainty) are translated into the H_∞/structured singular value framework, which is demonstrated to be a powerful tool to analyze and understand the complex phenomena.**

I. Introduction

It is well known that ill-conditioned plants may cause control problems [1]–[5]. By ill-conditioned we mean that the plant gain is strongly dependent on the input direction, or equivalently that the plant has a high condition number

$$\gamma(G(j\omega)) = \frac{\bar{\sigma}(G(j\omega))}{\underline{\sigma}(G(j\omega))}. \tag{1}$$

Here $\bar{\sigma}(G)$ and $\underline{\sigma}(G)$ denote the maximum and minimum singular values of the plant

$$\bar{\sigma}(G) = \max_{u \neq 0} \frac{\|Gu\|_2}{\|u\|_2}, \quad \underline{\sigma}(G) = \min_{u \neq 0} \frac{\|Gu\|_2}{\|u\|_2} \tag{2}$$

For tight control of ill-conditioned plants the controller should compensate for the strong directionality by applying large input signals in the directions where the plant gain is low; that is, a controller similar to G^{-1} in directionality is desirable. However, because of uncertainty, the direction of the large input may not correspond exactly to the low plant-gain direction, and the amplification of these large input signals may be much larger than expected from the model. This will result in large values of the controlled variables y (Fig. 1), leading to poor performance or even instability.

The concept of directionality is clearly unique to multivariable systems, and extensions of design methods developed for SISO systems are likely to fail for multivariable plants with a high degree of directionality. Furthermore, since the problems with ill-conditioned plants are closely related to how uncertainty affects the particular plant, it is very important to model the uncertainty as precisely as possible. Most multivariable design methods (LQG, LQG/LTR, DNA/INA, IMC, etc.) do not explicitly take uncertainty into account, and these methods will in general not yield acceptable designs for ill-conditioned plants.

A distillation column will be used as an example of an ill-conditioned plant. Here the product compositions are very sensitive to changes in the external flows (high gain in this direction), but quite insensitive to changes in the internal flows (low gain in this direction). In this paper the main emphasis is on the general properties of ill-conditioned plants, rather than the control system design for real distillation columns. We therefore choose to use a very simple model of the column where the condition number as a function of f...

152 citations

Aside (and small comfort for young researchers):

In fact, Early M^2 work is not highly cited…

- **1st journal paper**
  Title: **FINITE STABILITY REGIONS FOR LARGE-SCALE SYSTEMS WITH STABLE AND UNSTABLE SUBSYSTEMS**
  Author(s): MORARI M; STEPHANOPOULOS G; ARIS R
  Source: **INTERNATIONAL JOURNAL OF CONTROL** Volume: 26 Issue: 5 Pages: 805-815 Published: **1977**
  Times Cited: 2 (from Web of Science).

- **2nd journal paper**
  Title: **FINDING GENERIC RANK OF A STRUCTURAL MATRIX - COMMENTS**
  Author(s): MORARI M; STEPHANOPOULOS G;**;**
  Source: **IEEE TRANSACTIONS ON AUTOMATIC CONTROL** (Technical Note), 509-510**, 1978**
  Tines Cited: 5

- **3rd journal paper**
  Title: **SYNTHESIS OF DISTILLATION SCHEMES WITH ENERGY INTEGRATION**
  Author(s): FAITH DC; MORARI M
  Source: **COMPUTERS & CHEMICAL ENGINEERING** Volume: 3 Issue: 1-4 Pages: 269-272 Published: **1979**
  Times Cited: 3

- **4th journal paper**
  Title: **STABILITY ANALYSIS OF STRUCTURED CHEMICAL-ENGINEERING SYSTEMS VIA DECOMPOSITION**
  Author(s): MORARI M; STEPHANOPOULOS G; ARIS R
  Source: **CHEMICAL ENGINEERING SCIENCE** Volume: 34 Issue: 1 Pages: 11-15 Published: **1979**
  Times Cited: 4

- Title: **STABILITY OF MODEL-REFERENCE ADAPTIVE-CONTROL SYSTEMS**
  Author(s): MORARI M
  Source: **INDUSTRIAL & ENGINEERING CHEMISTRY PROC. DES. & DEV**. Volume: 19 Issue: 2 Pages: 279-281 Published: **1980**
  Times Cited: 0

- Title: **USING AN ADAPTIVE-OBSERVER TO DESIGN POLE-PLACEMENT CONTROLLERS FOR DISCRETE-TIME-SYSTEMS**
  Author(s): SHAHROKHI M; MORARI M
  Source: **INTERNATIONAL JOURNAL OF CONTROL** Volume: 36 Issue: 4 Pages: 695-710 Published: **1982**
  Times Cited: 0

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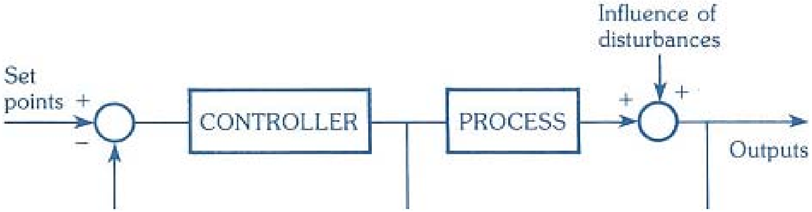

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OVERVIEW

DAY 1	INTRODUCTION TO PROCESS DYNAMICS AND CONTROL
DAY 2	SISO INTERNAL MODEL CONTROL
DAY 3	MULTIVARIABLE SYSTEM DYNAMICS AND CONTROL
DAY 4	THE RELATIVE GAIN ARRAY MIMO INTERNAL MODEL CONTROL
DAY 5	CONSTRAINED CONTROL INTERACTIONS BETWEEN PROCESS DESIGN AND CONTROL

DAY 1 INTRODUCTION TO PROCESS DYNAMICS AND CONTROL

I. OVERVIEW

II. PROCESS DYNAMICS

1. MODELLING
2. LAPLACE TRANSFORM AND TRANSFER FUNCTIONS
3. FREQUENCY RESPONSE

III. FEEDBACK CONTROL

1. BLOCK DIAGRAM
2. CLOSED LOOP POLES AND ZEROS
3. STABILITY CRITERIA

IV. DESIGN OF FEEDBACK CONTROLLERS

1. FUNDAMENTALS
2. BASIC CONTROLLER TYPES
3. ZIEGLER-NICKOLS TUNING RULES
4. PERFORMANCE CRITERIA
5. ROBUSTNESS CRITERIA

V. (MORE) ADVANCED CONTROL SCHEMES

1. FEEDFORWARD CONTROL
2. CASCADE CONTROL
3. MULTIVARIABLE CONTROL

DAY 2 SINGLE INPUT SINGLE OUTPUT INTERNAL MODEL CONTROL (SISO-IMC)

I. IMC STRUCTURE AND DESIGN

1. STRUCTURE
2. BASIC DESIGN PROCEDURE
3. OPTIMAL FACTORIZATION OF NMP ELECTS
4. TUNING FOR SETPOINTS VS. DISTURBANCES
5. FEEDFORWARD IMC
6. ANTI-RESET WINDUP
7. BUMPLESS TRANSFER
8. CASCADE IMC

II. IMC VS. CLASSICAL CONTROL

1. IMC VS. PID
2. IMC VS. SMITH PREDICTOR

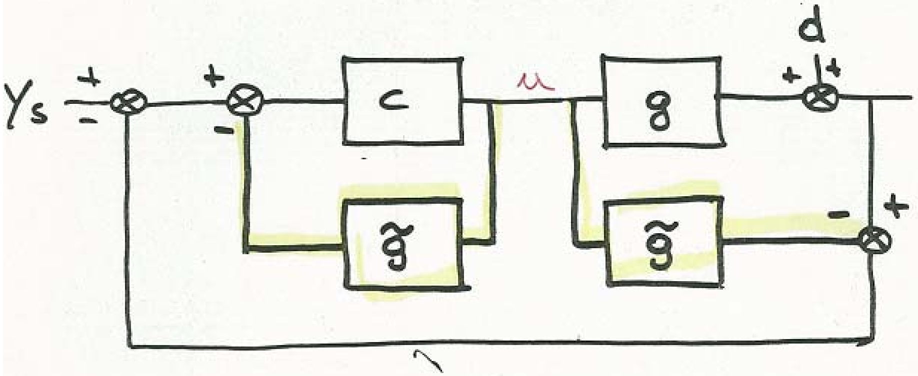
III. DESIGN UNDER MODEL UNCERTAINTY

1. TYPES OF UNCERTAINTY DESCRIPTIONS
2. CONTROLLER DESIGN UNDER UNCERTAINTY
3A. APPLICATION: TUNING OF SMITH PREDICTOR IN PRESENCE OF TIME DELAY UNCERTAINTIES
3B. APPLICATION: TUNING GUIDELINES FOR PID CONTROLLERS
4. EXISTENCE OF STABILIZING FILTER ("INT GRA CONTROLLABILITY")
5. ON-LINE TUNING OF IMC
6. IMC VS. PID

I. IMC STRUCTURE & DESIGN

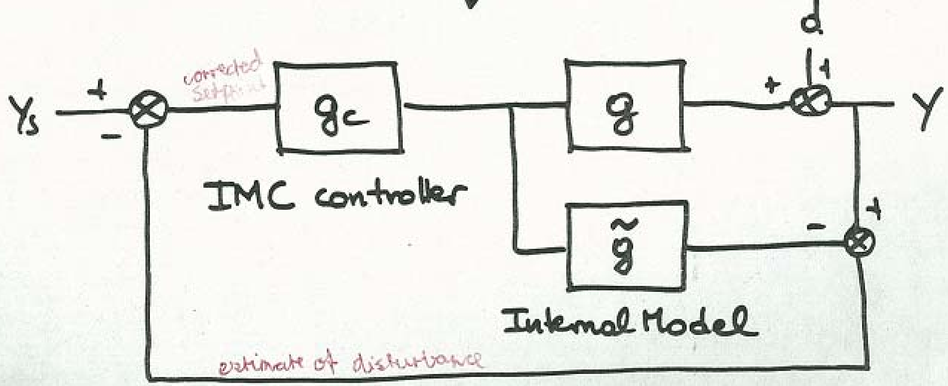
1. Structure

Linear, no constraints



NOTE: The two blocks $\tilde{g}$ cancel each other

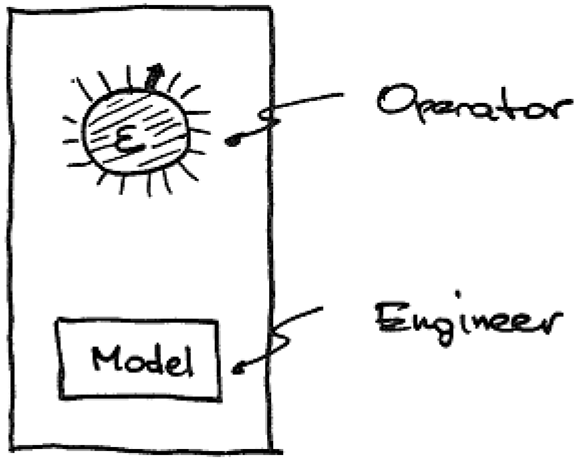
⇓



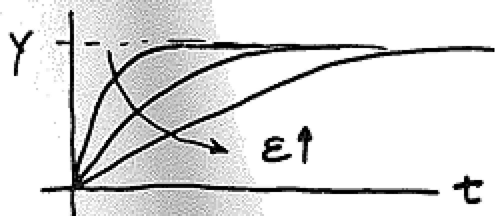
IMC Structure

IMC is the basis of a

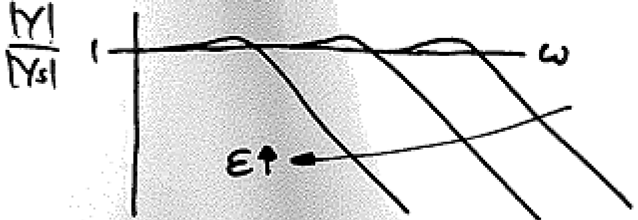
"ONE-KNOB CONTROLLER"



Knob is <u>directly</u> related to performance



Speed of closed loop response



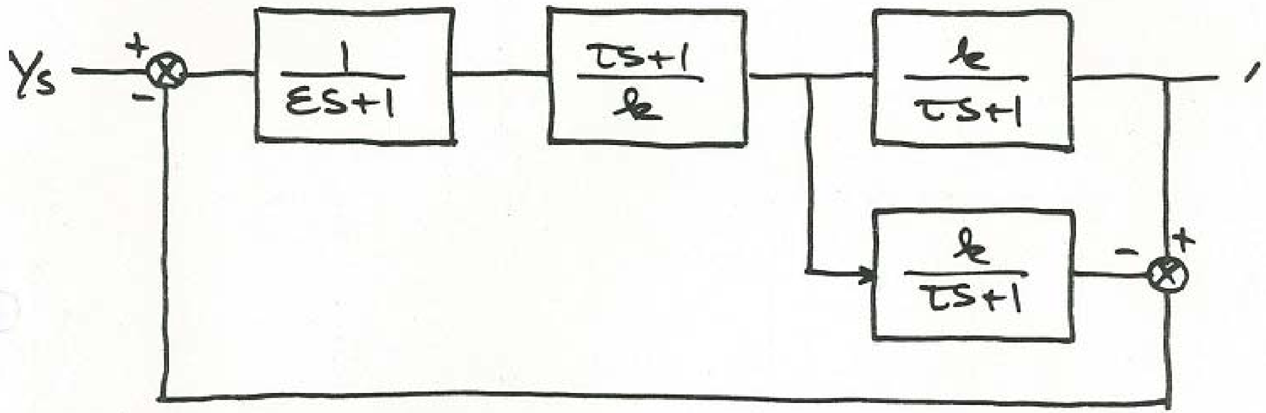
Closed loop bandwidth

1. IMC vs. PID

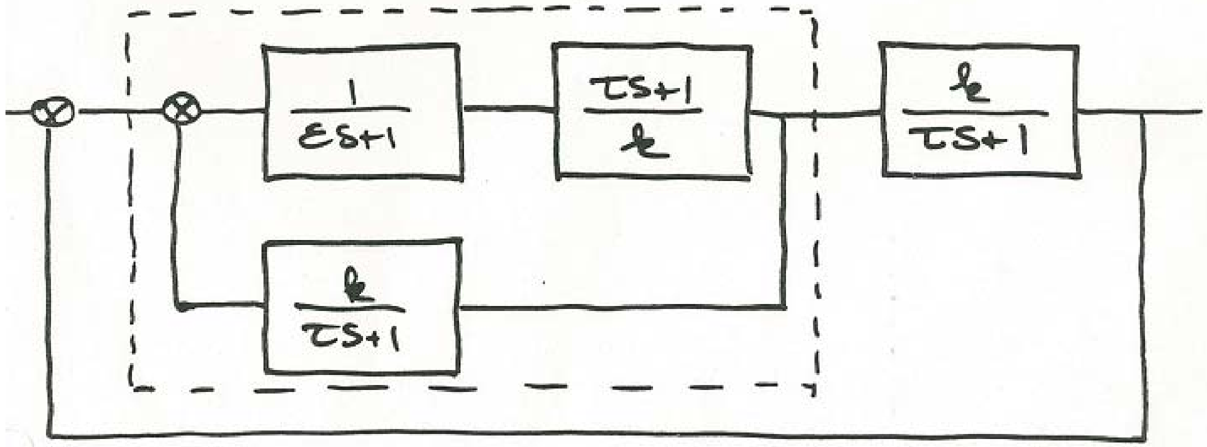
Important!
Most people would have
this question at this point.

NMP Systems

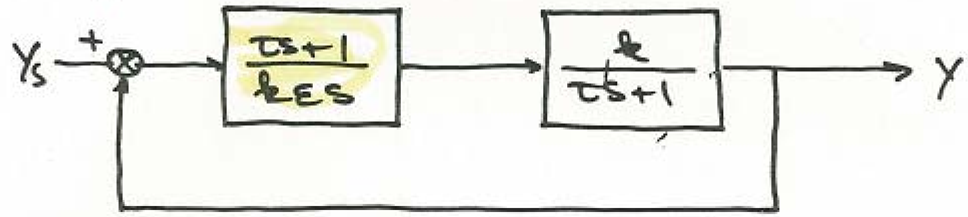
1st order system



$\Downarrow$



Equivalent classical feedback structure.



1st order system, MP

$$PI \iff IMC$$

$$k_c\left(1+\frac{1}{\tau_I s}\right) \qquad \frac{\tau s+1}{k \varepsilon s}$$

$$\boxed{\begin{aligned} k_c &= \frac{\tau}{k \varepsilon} \\ \tau_I &= \tau \end{aligned}}$$

$\leftarrow$ tuning

$\leftarrow$ model

$k_c > \frac{1}{k} \Rightarrow \tau > \varepsilon$ speed up by contro[l]

$k_c < \frac{1}{k} \Rightarrow \tau < \varepsilon$ slow down by cont[rol]

	Model	$\frac{y}{y_s} = \tilde{g}_+ f$	Controller	$K_c k$	τ_I	τ_D	τ_F
A	$\frac{k}{\tau s+1}$	$\frac{1}{\epsilon s+1}$	$\frac{1}{k}\frac{\tau s+1}{\epsilon s}$	$\frac{\tau}{\epsilon}$	τ	—	—
B	$\frac{k}{(\tau_1 s+1)(\tau_2 s+1)}$	$\frac{1}{\epsilon s+1}$	$\frac{(\tau_1 s+1)(\tau_2 s+1)}{k\epsilon s}$	$\frac{\tau_1+\tau_2}{\epsilon}$	$\tau_1+\tau_2$	$\frac{\tau_1\tau_2}{\tau_1+\tau_2}$	—
C	$\frac{k}{\tau^2 s^2+2\zeta\tau s+1}$	$\frac{1}{\epsilon s+1}$	$\frac{\tau^2 s^2+2\zeta\tau s+1}{k\epsilon s}$	$\frac{2\zeta\tau}{\epsilon}$	$2\zeta\tau$	$\frac{\tau}{2\zeta}$	—
D	$k\frac{-\beta s+1}{\tau s+1}$	(1) $\frac{-\beta s+1}{\epsilon s+1}$	$\frac{\tau s+1}{k(\beta+\epsilon)s}$	$\frac{\tau}{\beta+\epsilon}$	τ	—	—
E	$k\frac{-\beta s+1}{\tau s+1}$	(2) $\frac{-\beta s+1}{(\beta s+1)(\epsilon s+1)}$	$\frac{\tau s+1}{ks(\beta\epsilon s+2\beta+\epsilon)}$	$\frac{\tau}{2\beta+\epsilon}$	τ	—	$\frac{\beta\epsilon}{2\beta+\epsilon}$
F	$k\frac{-\beta s+1}{\tau^2 s^2+2\zeta\tau s+1}$	(1) $\frac{-\beta s+1}{\epsilon s+1}$	$\frac{\tau^2 s^2+2\zeta\tau s+1}{k(\beta+\epsilon)s}$	$\frac{2\zeta\tau}{\beta+\epsilon}$	$2\zeta\tau$	$\frac{\tau}{2\zeta}$	—
G	$k\frac{-\beta s+1}{\tau^2 s^2+2\zeta\tau s+1}$	(2) $\frac{-\beta s+1}{(\beta s+1)(\epsilon s+1)}$	$\frac{\tau^2 s^2+2\zeta\tau s+1}{ks(\beta\epsilon s+2\beta+\epsilon)}$	$\frac{2\zeta\tau}{2\beta+\epsilon}$	$2\zeta\tau$	$\frac{\tau}{2\zeta}$	$\frac{\beta\epsilon}{2\beta+\epsilon}$
H	$\frac{k}{s}$	(3) $\frac{1}{\epsilon s+1}$	$\frac{1}{k\epsilon}$	$\frac{1}{\epsilon}$	—	—	—
I	$\frac{k}{s}$	(4) $\frac{2\epsilon s+1}{(\epsilon s+1)^2}$	$\frac{2\epsilon s+1}{k\epsilon^2 s}$	$\frac{2}{\epsilon}$	2ϵ	—	—
J	$\frac{k}{s(\tau s+1)}$	(3) $\frac{1}{\epsilon s+1}$	$\frac{\tau s+1}{k\epsilon}$	$\frac{1}{\epsilon}$	—	τ	—
K	$\frac{k}{s(\ s+1)}$	(4) $\frac{2\epsilon s+1}{(\epsilon s+1)^2}$	$(\tau s+1)(2\epsilon s+1)$	$\frac{2\epsilon+\tau}{\epsilon^2}$	$2\epsilon+\tau$	$\frac{2\epsilon\tau}{2\epsilon+\tau}$	—

Ind. Eng. Chem. Process Des. Dev. **1986**, *25*, 252–265

Internal Model Control. 4. PID Controller Design

Daniel E. Rivera, Manfred Morari,* and Sigurd Skogestad

Chemical Engineering, 206-41, California Institute of Technology, Pasadena, California 91125

For a large number of single input–single output (SISO) models typically used in the process industries, the Internal Model Control (IMC) design procedure is shown to lead to PID controllers, occasionally augmented with a first-order lag. These PID controllers have as their only tuning parameter the closed-loop time constant or, equivalently, the closed-loop bandwidth. On-line adjustments are therefore much simpler than for general PID controllers. As a special case, PI- and PID-tuning rules for systems modeled by a first-order lag with dead time are derived analytically. The superiority of these rules in terms of both closed-loop performance and robustness is demonstrated.

I. Introduction

Synthesis and tuning of control structures for SISO systems comprises the bulk of process control problems. In the past, hardware considerations dictated the use of the PID controller, but through the use of computers, controllers have now advanced to the stage where virtually any conceivable control policy can be implemented. Despite these advances, the most widely used controller is still of the PID type. Finding design methods which lead to the optimal operation of PID controllers is therefore of significant interest.

For controller tuning, simplicity, as well as optimality, is important. The three modes of the ordinary PID controller, k_c, τ_I, and τ_D, do not readily translate into the

occasionally augmented by a first-order lag. Furthermore, the proposed procedure provides valuable insight regarding controller tuning effects on both performance and robustness.

II. Performance and Robustness Measures

Probably the best indicator of performance is the *sensitivity function*

$$S = \frac{1}{1 + gc} = \frac{e}{y_s - d} \tag{1}$$

(The nomenclature should be apparent from Figure 1.) It is desirable to keep the sensitivity function small over as wide a frequency range as possible. For any proper system,

DAY 3 MULTIVARIABLE SYSTEM DYNAMICS AND CONTROL

I. PROCESS DYNAMICS

1. MODELLING
2. LAPLACE TRANSFORMS AND TRANSFER MATRICES
3. POLES AND ZEROS OF G(s)
4. "GAIN" AND "AMPLITUDE RATIO" FOR MIMO SYSTEMS
5. MATRIX NORMS

II. MIMO FEEDBACK CONTROL FUNDAMENTALS

1. CONTROL STRUCTURES
2. BLOCK DIAGRAMS
3. CLOSED LOOP POLES AND ZEROS
4. STABILITY CRITERIA

III. DESIGN OF MIMO FEEDBACK CONTROLLERS

1. $\boxed{\text{THE PAIRING PROBLEM – PROCESS INTERACTIONS}}$
2. THE MODEL ERROR SENSITIVITY PROBLEM

DAY 4 MULTIVARIABLE CONTROL

I. THE RELATIVE GAIN ARRAY

1. DEFINITION
2. ALGEBRAIC PROPERTIES
3. THEORETICAL IMPLICATIONS OF THE RGA
4. ROBUSTNESS AND THE RGA

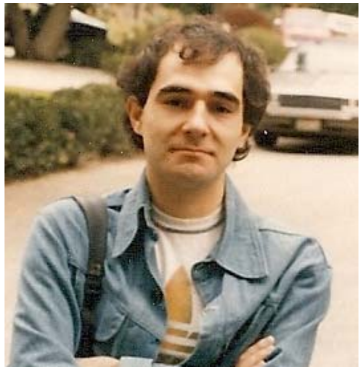
II. MIMO INTERNAL MODEL CONTROL (IMC)

1. STRUCTURE
2. BASIC DESIGN PROCEDURE

Decentralized control

Ind. Eng. Chem. Fundam. **1985**, *24*, 221–235

Closed-Loop Properties from Steady-State Gain Information



Pierre Grosdidier and Manfred Morari*

Department of Chemical Engineering, California Institute of Technology, Pasadena, California 91125

Bradley R. Holt†

Department of Chemical Engineering, University of Wisconsin, Madison, Wisconsin 53706

Closed-loop properties of open-loop stable multivariable systems are explored when the controllers include integral action. The studied properties comprise closed-loop stability, sensor and actuator failure tolerance, feasibility of decentralized control structures, and robustness with respect to modeling errors. All the results are based on steady-state gain information *only*. The relationship between the new theorems and the Relative Gain Array (RGA) is elucidated.

Introduction

Modeling uncertainties and constantly changing operating conditions make it very difficult to develop reliable dynamic models for chemical processes. Often, only steady-state gain information is available. In multi-input multi-output (MIMO) systems, these data may be represented as a matrix of steady-state gains $\mathbf{G}(0)$. Since this matrix $\mathbf{G}(0)$ is often the only information available on the system, any method that will allow the extraction of *useful*

than the RGA. Next, the relationship between Right Half Plane (RHP) zeros and the RGA will be briefly examined. Finally, a new theorem will show how the RGA can be used to predict the sensitivity of a multivariable system to modeling errors.

Many of the results presented in this paper were conjectured previously by other researchers. However, as we shall show, some of these conjectures are incorrect, some are partially correct, and some are correct but the argu-

Algebraic Properties of the Relative Gain Array. **Property 1.** The sum of the elements of each row and each column of the relative gain array is always unity.

Proof. From $\mathbf{G}^{-1}\mathbf{G} = \mathbf{I}$

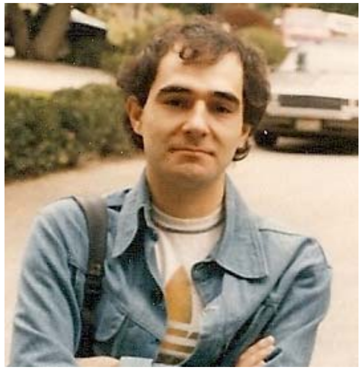
$$\sum_{j=1}^{n} \hat{g}_{ij} g_{jk} = \delta_{ik}$$

Thus

$$\sum_{j=1}^{n} \hat{g}_{ij} g_{ji} = \sum_{j=1}^{n} \lambda_{ij} = 1$$

Similarly from $\mathbf{G} \times \mathbf{G}^{-1} = \mathbf{I}$

$$\sum_{i=1}^{n} \lambda_{ij} = 1$$



Conjecture 1. For the $n \times n$ transfer matrix **G** the minimum condition number γ^* is bounded by

$$\gamma^* \leq 2 \max\left[\|\Lambda\|_1, \|\Lambda\|_\infty\right] \qquad (34)$$

Theorem 1. Assume $\mathbf{H}(s)$ is a proper rational transfer matrix. $\mathbf{H}(s)$ is integral stabilizable only if $\det(\mathbf{H}(0)) > 0$.

Proof. See Appendix A. For SISO systems theorem 1 becomes necessary and sufficient.

Relationship to the Relative Gain Array. Theorem 6. If $\lambda_{jj}(\mathbf{G}) < 0$ then for any compensator $\mathbf{C}(s)$ with the properties (a) $\mathbf{G}(s)\mathbf{C}(s)$ is proper, (b) $c_{jl} = c_{lj} = 0 \ \forall \ l \neq j$ (y_j affects u_j only, u_j is affected by y_j only) and any $k > 0$ the closed loop system shown in Figure 4 has at least one of the following properties: (a) The closed loop system is unstable. (b) Loop j is unstable by itself, i.e., with all the other loops opened. (c) The closed loop system is unstable as loop j is removed.

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Pergamon Journals Ltd.

Interaction Measures for Systems Under Decentralized Control*

PIERRE GROSDIDIER and MANFRED MORARI†

A new interaction measure predicts the stability of decentralized control systems and measures the performance loss caused by these control structures.

Key Words—Multivariable control systems; decentralized control; (interaction measures); (Structured Singular Value); large-scale systems; Nyquist criterion; stability criteria.

Abstract—A shortcoming of many of the currently available measures of interactions is their limited theoretical basis. Using the notion of Structured Singular Value (SSV), a new dynamic interaction measure is defined for multivariable systems under feedback with diagonal or block diagonal controllers. This measure can be used not only to predict the stability of decentralized control systems but also to measure the performance loss caused by these control structures. In particular, its steady state value provides a sufficient condition for achieving offset-free performance with the closed loop system. The relationship of this new interaction measure with Rijnsdorp's interaction measure (1965) and Rosenbrock's Direct Nyquist Array (1974) is clarified.

INTRODUCTION

LET $\mathbf{G}(s)$ be an $n \times n$ rational transfer function matrix relating the vector of system inputs $\mathbf{u}$ to the vector of system outputs $\mathbf{y}$. Let $\mathbf{r}$ be the vector of reference signals or set points for the closed loop

the controller hardware costs could be high if an implementation through analogue circuitry is required. The purpose of an interaction measure would be to measure the performance degradation caused by the block diagonal controller. These conditions are relevant, for example, for large networks of power stations where the distances between the stations can be significant. Hardware issues are generally irrelevant in the context of process control; in all modern plants all measurement signals are sent into a central control room from where all the actuator signals originate.

2. *Design simplicity*. If the block $\mathbf{G}_{ij}(s) = 0$ $(i \neq j)$ then each controller $\mathbf{C}_i(s)$ can be designed for the

Theorem 3. Assume that $\mathbf{G}(s)$ and $\tilde{\mathbf{G}}(s)$ have the same RHP poles and that $\tilde{\mathbf{H}}(s)$ is stable. Then the closed loop system $\mathbf{H}(s)$ is stable if

$$\sigma_{\max}(\tilde{\mathbf{H}}(j\omega)) < \mu^{-1}(\mathbf{E}(j\omega)) \quad \forall\,\omega. \tag{23}$$

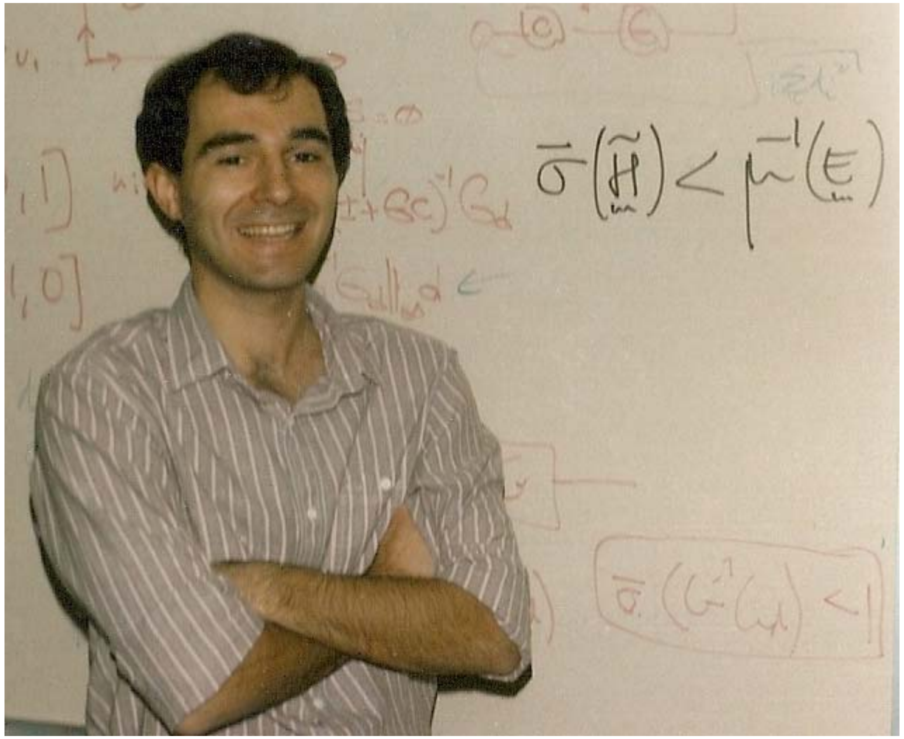
This is the tightest norm bound, in the sense that if there is a system $\tilde{\mathbf{H}}_1(s)$ which violates (23)

$$\sigma_{\max}(\tilde{\mathbf{H}}_1(j\omega)) > \mu^{-1}(\mathbf{E}(j\omega)) \tag{24}$$

then there exists another system $\tilde{\mathbf{H}}_2(s)$ such that

$$\sigma_{\max}(\tilde{\mathbf{H}}_1(j\omega)) = \sigma_{\max}(\tilde{\mathbf{H}}_2(j\omega)) \tag{25}$$

for which (8) is violated and $\mathbf{H}(s)$ is unstable.



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Printed in Great Britain.

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Pergamon Press plc

Brief Paper

Robust Performance of Decentralized Control Systems by Independent Designs*

SIGURD SKOGESTAD† and MANFRED MORARI†

Key Words—Decentralized control; robustness; large-scale systems; (structured singular value).

Abstract—Decentralized control systems have fewer tuning parameters, are easier to understand and retune, and are more easily made failure tolerant than general multivariable control systems. In this paper the decentralized control problem is formulated as a series of independent designs. Simple bounds on these individual designs are derived, which when satisfied, guarantee robust performance of the overall system. The results provide a generalization of the μ-interaction measure introduced by Grosdidier and Morari (*Automatica*, **22**, 309–319 (1986)).

1. *Introduction*

Robust performance. The goal of any controller design is that the overall system is stable and satisfies some minimum performance requirements. These requirements should be

(the set Π of possible plants G_p) as norm bounded perturbations (Δ_i) on the nominal system. Through weights each perturbation is normalized to be of size one:

$$\bar{\sigma}(\Delta_i) \leq 1, \quad \forall \omega. \tag{3}$$

The perturbations, which may occur at different locations in the system, are collected in the diagonal matrix Δ_U (the subscript U denotes uncertainty)

$$\Delta_U = \text{diag}\,\{\Delta_1, \ldots, \Delta_n\} \tag{4}$$

and the system is rearranged to match the structure in Fig. 1. The interconnection matrix M in Fig. 1 is determined by the nominal model (G), the size and nature of the uncertainty, the performance specifications and the controller. For Fig. 1 the robust performance condition (1b) becomes (Doyle *et*

AICHE Annual Meeting, Washington DC, Nov. 1988

VARIABLE SELECTION FOR DECENTRALIZED CONTROL

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Manfred Morari
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Pasadena, CA 91125

Summary

Decentralized controllers (single-loop controllers applied to multivariable plants) are often preferred in practice because they are robust and relatively simple to understand and to change. The design of such a control system starts with pairing inputs (manipulated variables) and outputs (controlled variables). For a $n \times n$ plant there are $n!$ possible pairings, and there is a great need for screening techniques to quickly eliminate undesirable pairings. In this paper we present several tests for eliminating pairings which are not decentralized integral controllable (DIC). A system is DIC if there exists a stabilizing decentralized controller with integral action such that the gains of the individual loops may be reduced *independtly* without introducing instability. Note that DIC is a property of the plant and the chosen pairings. The tests presented are in terms of different measures of the sign of steady state gain matrix; including the RGA, the determinant and eigenvalues. The relationship to previously presented results is discussed in detail.

Achievable Closed-Loop Properties of Systems Under Decentralized Control: Conditions Involving the Steady-State Gain

Peter J. Campo and Manfred Morari, *Member, IEEE*

***Abstract*—The question of the existence of decentralized controllers for open-loop stable multivariable systems which provide particular closed-loop properties is investigated. In particular, we study the existence of decentralized controllers which provide integral action (Type 1 closed-loop performance) and also demonstrate one or more of: unconditional stability, integrity with respect to actuator and sensor failure, and decentralized unconditional stability. Necessary, sufficient, and, in some cases, necessary and sufficient conditions on the open-loop steady-state gain are derived such that there exists a controller which provides these desired closed-loop characteristics. These results provide the basis for a systematic approach to control structure selection for decentralized controller design.**

I. Introduction

DESPITE the closed-loop performance advantages of multivariable controllers, the use of single loop controllers for multivariable plants is the rule in industrial process control applications. In addition to its inherent simplicity, a decentralized control system consisting of independent controller sub-

which must be specified in each SISO design is typically much smaller than a full multivariable design.

4) *Simplified Tuning*. Individual subsystems can be (manually) tuned and retuned on-line to accommodate the effects of (slowly) changing process conditions.

The requirement that the control system be decentralized introduces the pairing problem. The pairing problem is concerned with defining the control system structure, i.e., which of the available plant inputs is to be used to control each of the plant outputs. For a fully noninteracting plant, the choice is obvious, and the benefits of decentralized control discussed above accrue trivially. In any practical problem, there are (to a greater or lesser extent) interactions in the plant. This implies that even if the control system is decentralized, subsystems of the closed-loop system are not independent of each other. To the extent that the control system can be designed to make the closed-loop subsystems independent, the idealized characteristics outlined above can be realized.

DAY 5

I. CONSTRAINED CONTROL

II. INTERACTIONS BETWEEN PROCESS DESIGN AND CONTROL.

A little confusion on "constrained control":

- **Is it active constraints control (steady-state) ?**
- **Is it dealing with constraints dynamically (today: MPC) ?**

CONSTRAINED CONTROL

Motivation:

Operating objective of chemical process industry:

Make product of specified quality at minimum cost

Major financial benefits:

Steady state optimization

→ Optimum occurs usually at the intersection of operating constraints

Good regulation is required to maintain the optimal operating conditions in the presence of constraints

⇒ constrained control

Rule of plantwide control:
Control active constraints
(steady-state optimal operation)

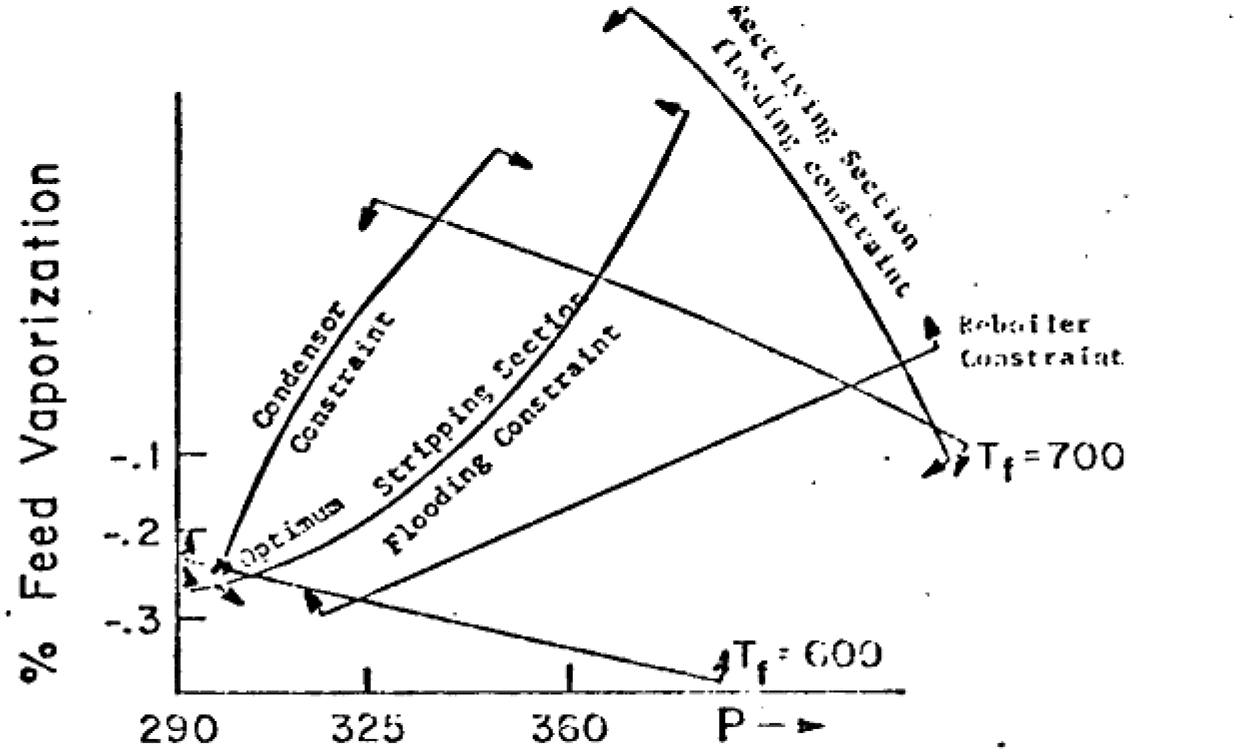

Figure 8. The Feasible Operating Region for the Depropanizer at the Optimum Design Conditions.

Approaches for dealing with constraints

1) Select operating point safely away from constraint. Design linear controller such that constraint is not violated for typical disturbance.

DRAWBACKS:

- financial loss due to nonoptimal operating point
- linear controller is conservative <u>all</u> the time not to violate constraints for <u>extreme</u> disturbances
- does not guarantee that constraints will be maintained
- does not help in moving from one operating condition to another

2) Design control system such that
   - it is linear in the feasible region
   - it becomes nonlinear as the constraints are hit

??

2) Design control system such that
- it is linear in the feasible region
- it becomes nonlinear as the constraints are hit

??

Manfred, seems you are mixing

1. optimally active constraints (steady-state)
2. dynamic saturation

Example car:

$J = T$ (minimize driving time from A to B)

u = engine power

Constraints: $u < u_{max}$

$v < v_{max}$ (speed limit)

1. Steady-state. What should we control?

Flat road: CV = v (active constraint with setpoint = vmax – backoff)

Steep hill: CV = u (active constraint with setpoint = umax)

Switching between these two cases is trivial.

2. Dynamic. Flat road with increase in speed limit (vmax):

May temporary reach umax-constraint (use MPC?)

PROCESS CONTROL AND DESIGN

AT CALTECH

Caltech group 1987

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SUMMARY OF RESEARCH ACTIVITIES

The overall research program encompasses the following seven areas:

1. Robust control system design
2. Control system design for nonlinear systems
3. Decentralized control
4. The effects of process design on process control
5. Process design
6. Concurrent processing and other new computer architectures
7. Artificial Intelligence and Expert Systems

Most of the research projects described in the following part of this report belong to more than one area. Often a researcher is involved in more than one project. Some of the main concepts and ideas are introduced next.

Caltech ChE brochure 1988

16

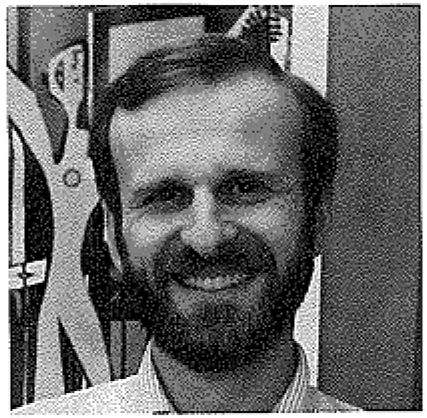
Manfred Morari, Professor

Manfred Morari received his diploma in Chemical Engineering from the Swiss Federal Institute of Technology in 1974 and the Ph.D. degree from the University of Minnesota in 1977. He was on the faculty of the University of Wisconsin from 1977 to 1983, when he assumed his current position. He also held short-term positions with Exxon Research & Engineering and ICI.

His general interests are in the areas of process control and design, in particular robust and decentralized control and the effect of design on operability. He and his students have developed several control analysis and synthesis techniques that have found widespread application in industry. Many of these results are summarized in his book *Robust Process Control*, published by Prentice Hall.

Honors and Awards

- Donald P. Eckman Award, American Automatic Control Council (1980)
- Robert W. Vaughan Lectureship, California Institute of Technology (1983)
- Allan P. Colburn Award, American Institute of Chemical Engineers (1984)
- NSF Presidential Young Investigator Award (1984)
- Ernest W. Thiele Lectureship, University of Notre Dame (1987)
- Gulf Visiting Professor of Chemical Engineering, Carnegie-Mellon University (1987)



John H. Seinfeld, Louis E. Nohl Professor and Executive Officer for Chemical Engineering

John Seinfeld joined the Caltech faculty in 1967 after receiving a B.S. from the University of Rochester and a Ph.D. in Chemical Engineering from Princeton University. Since 1973 he has served as Executive Officer for the department.

His research interests are in the atmospheric chemistry and physics of air pollution and estimation of the properties of dynamic systems, the latter being a long-term continuation of work begun in his Ph.D. thesis. Professor Seinfeld directs one of the most comprehensive research programs on air pollution chemistry and physics in the world. His research encompasses both theoretical and experimental aspects and is particularly well known for its influence in mathematical modeling of air pollution and in elucidating the fundamental processes by which aerosols form and grow. He has served as an advisor to the U.S. Environmental Protection Agency, the State of California, NASA, and industry on problems relating to air pollution and atmospheric chemistry.

Professor Seinfeld is a member of the National Academy of Engineering.

Honors and Awards

MODELING REQUIREMENTS FOR PROCESS CONTROL

The concept of what constitutes an adequate model for control system design has varied through the decades. In recent years, the robust control paradigm (i.e., controller design taking into account model uncertainty) has redefined conventional notions of dynamic modeling, allowing for plant descriptions that consist of sets of linear plants and thus take into consideration the effects of nonlinearities, parameter variations, and unmodelled dynamics which are not adequately captured by the conventional single linear plant model. The following modeling issues are examined within this context:

1. *Control-relevant model reduction* [36]. The capability to incorporate control considerations in the model simplification problem is desirable, as this can lead to simpler yet more effective plant descriptions for controller design purposes. Towards this end a variety of frequency- weighted regression problems have been developed, in which the weight incorporates explicitly the uncertainty description, the desired closed-loop response, and the setpoint/disturbance characteristics of the problem.

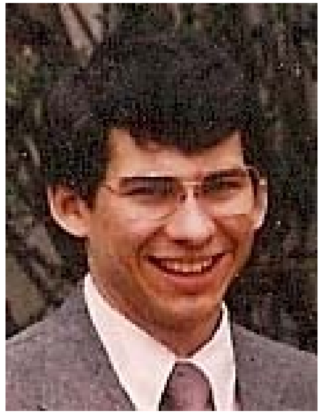
2. *Control-relevant identification.* The techniques of spectral analysis have been related to identification for control system design purposes. Through spectral smoothing of appropriately generated data, it is possible to obtain the norm-bound uncertainty description and a frequency response representing the nominal plant. The model reduction procedure described in (1) then serves as a control-relevant parameter estimation routine. As a final validation step, the robust performance theorems are used to determine rigorously the effectiveness of the identified model in a control system design application.

Because the Internal Model Control structure relates models directly to feedback controllers, our investigations on model reduction and identification have had direct implications for low-order controller tuning. The modeling issues investigated have thus led to efforts in the following areas: 1) General low-order controller tuning, 2) PID controller tuning [17] and 3) Smith Predictor tuning in the presence of uncertainty [35].

Software to perform the methods described above has been developed for CONSYD. Some of the analysis has also been incorporated into the ROBEX expert system.

D. E. Rivera

MODELING UNCERTAINTY FOR PROCESS CONTROL

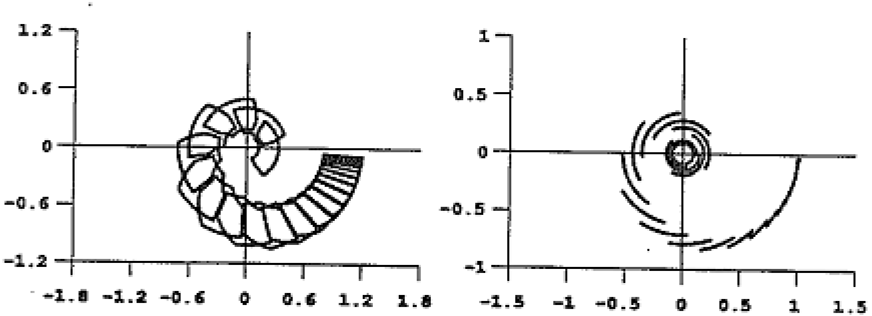


The modern robust controller design methods require that model uncertainty be expressed in a specific mathematical form, usually as norm bounded perturbations on a nominal process model. For these design methods to lead to practically useful controllers the mathematical uncertainty expressions must be precise representations of the physical reality. This motivates the development of techniques for accurately translating physically motivated uncertainty descriptions, e.g., parameter bounds, into mathematically convenient forms, like norm bounds on frequency responses.

In the course of this research a complete SISO controller design technique has been developed that begins with physically meaningful uncertainty descriptions. First the set of all possible frequency responses generated by a transfer function with real parameter uncertainty is located [32]. The possible responses are pictured as model uncertainty regions on the complex plane. Typical model uncertainty regions corresponding to a process with uncertain gain, time-constant and time delay are shown in Fig. 1. Those corresponding to a process with uncertain time-delay are shown in Fig. 2. Next a controller is designed using a method based on the Internal Model Control structure [35]. The controller is used to map the process uncertainty regions to the Nyquist plane for stability and performance analysis. It is then easy to discern a controller design with adequate robustness margins (Fig. 3) from a potentially unstable design (Fig. 4). This region-mapping technique was used to compare robust performance of PID and Smith Predictor controllers. The technique was also used to facilitate Structured Singular Value (μ) analysis and "μ-optimal" controller synthesis for models with gain, time-constant, and time-delay uncertainty.

D. Laughlin

Figure 1



ROBUST DIGITAL CONTROLLER DESIGN

The goal of this project is the development of a rigorous procedure for the synthesis of digital multi-input multi-output (MIMO) control systems, that directly address "typical" engineering objectives, like good disturbance rejection and setpoint tracking, trade-offs between decoupled and non-decoupled responses, sampling time selection, etc. The major objective is to provide sound theoretical answers to the real life problem where the actual process does not behave in exactly the same manner as its model (model uncertainty). In addition to that, special attention is paid to the fact that the plant operators have to be provided with on-line tuning parameters which have clear physical meaning and effect. Finally the sampled-data nature of the implemented control system is to be taken into account in every stage of the synthesis procedure. Our efforts resulted in a two-step design method based on the following theoretical developments:

- PERFORMANCE

i) *SISO (single-input single-output) systems*: The theoretical study of a number of well known control algorithms led to a simple algorithm that combines the advantages of each, (no overshoot, no intersample rippling, speed of response, etc.) but is free of their disadvantages [14,27].

ii) *MIMO systems*: The effect of inherent performance limiting process characteristics, like time delays and zeros in certain parts of the complex plane, were quantified and yielded measures for the various trade-offs between decoupled and non-decoupled responses. The results from the SISO case were then extended to MIMO systems to design control systems with similar properties. The capability to treat general multivariable time-delay structures [45] is of major importance for chemical engineering systems.

- ROBUSTNESS

Robustness is handled within the Structured Singular Value framework by adjusting the IMC robustness filter [55]. The filter parameters are obtained from an optimization problem which involves a relatively small number of variables, for which good initial guesses are available [19].

- SAMPLING TIME SELECTION

An iterative method for sampling time selection is an integral part of the synthesis procedure [27].

E. Zafiriou

ROBUST CONTROLLER DESIGN FOR DISTILLATION COLUMNS

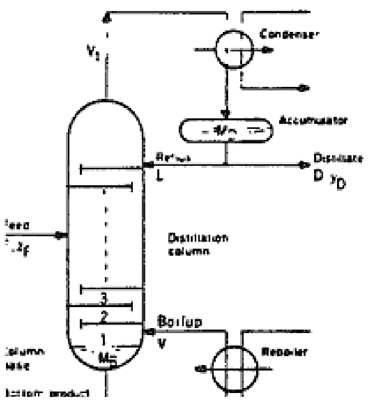
One of the main goals of this project is to develop insight into the fundamental static and dynamic behavior of distillation columns (Fig. 1). Some of the problems arising in the design of a control system for controlling the compositions y_D and x_B are

1. Strongly nonlinear system and dynamic behavior which is often difficult to understand [59].
2. Many possible choices of manipulated variables for composition control (2 independent combinations of L, V, V_T, B, D) [54].
3. System often ill-conditioned (large RGA-elements) [48,50]. It is well known that ill-conditioned systems are sensitive to model uncertainty.

So far the issue of uncertainty has been largely overlooked in the chemical engineering literature [38] but it turns out to be one of the keys to understand why some distillation columns are difficult to control. As an example let us consider a column with L and V as manipulated variables for composition control (often the "default" choice) with product purity specifications $y_D = 1 - x_B = 0.01$ [60]. In this case it is found that the compositions are very sensitive to changes in the external flows (B and D), but less sensitive to changes in the internal flows (i.e., when $\Delta L \approx \Delta V$). Thus the gains depend strongly on the input "direction" (high gain when $\Delta L \approx -\Delta V$ and low gain when $\Delta L \approx \Delta V$), i.e., the column is ill-conditioned. For good control it is necessary to make large changes in the internal flows L and V for which the plant gain is small and to keep the external flows for which the plant gain is large, almost constant. Because of uncertainty (e.g., valve resolution) it is essentially impossible to make large changes in L and V while keeping their difference almost constant. This leads to very poor response characteristics with large overshoot (Fig. 2).

The project involves an appropriate blend of modelling [56,57,59], theoretical developments [47,48] and case studies [60].

S. Skogestad



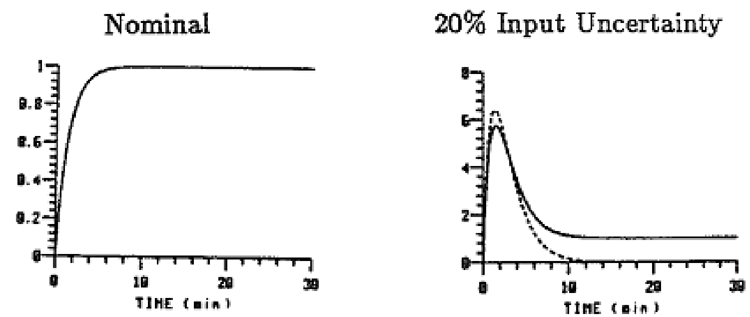
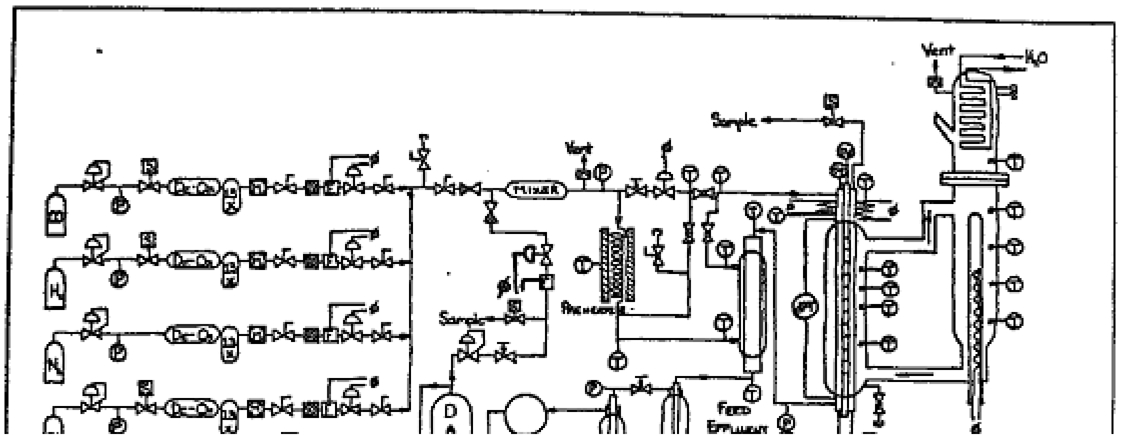


Fig. 2. Closed-loop response to setpoint change in y_D using

CONTROL OF A FIXED BED CHEMICAL REACTOR WITH MASS AND ENERGY RECYCLE

The design of advanced control systems for a fixed-bed catalytic reactor is addressed. The difficulties associated with the control of fixed- bed reactors are well known and originate from a number of factors. The thermal coupling between the fluid and the catalytic bed slows the propagation of temperature changes considerably. This gives rise to large time delays and therefore to fundamental limitations on the achievable closed-loop performance with certain control configurations. In many cases the controller implementation is complicated by the unavailability of on-line measurements for some of the variables which are to be controlled, such as product concentrations. The most important obstacles in the synthesis of advanced control schemes for these reactor systems reside in the difficulties associated with obtaining accurate mathematical descriptions of such systems and in the overall complexity of their dynamic behavior.

In the present work the issue of model uncertainty is incorporated explicitly into the controller design procedure. The work benefits from recent, significant advances in control system design theory: The IMC design procedure is combined with analysis and synthesis techniques based on the Structured Singular Value (μ), into a systematic procedure for robust control system design. Robust controllers are known as those capable of maintaining stability and achieving certain performance requirements even in the face of plant-model mismatch. The present work includes the first application of several of the advances in this area to a fixed-bed reactor process, in some cases even the first application to a complex chemical process ever reported [28,33].



The work is carried out in the context of a laboratory, fixed-bed non- adiabatic reactor. In the experimental set-up energy is recycled through the use of a feed-effluent heat exchanger while mass is recycled through a pump and recycle stream. Both recycle rates are controlled with flow control valves.

The reaction studied is the methanation of carbon monoxide on a nickel catalyst. The reactor is a 27" long single stainless steel tube filled with a mixture of catalyst and inert alumina pellets. A thermowell runs down the center of the reactor allowing for measurement of axial temperatures. The reactor tube is placed in a Dowtherm heat exchanger bath keeping the reactor wall temperature constant.

Various temperature measurements are available throughout the system including the twelve thermocouples which measure the axial temperature gradient. Pressure measurements are made with two pressure transducers. One transducer measures the pressure drop across the reactor bed while the other can be switched to measure the gauge pressure at various points in the system. Finally, gas composition is measured with a gas chromatograph. A series of solenoid valves is used in selecting which stream to sample.

Other operating parameters which are easily set include the inlet gas flowrates using mass flow controllers, the preheater operating temperature, the system pressure, and the reactor wall temperature.

J. Mandler, C. Webb

NONLINEAR CONTROLLER DESIGN

Strong similarities between control theory and the theory on the solution of operator equations have been observed and basic results in control theory have been derived from operator theory arguments. The purpose of this work is to investigate the theory of controller design as an application of basic operator theory principles and to establish a unified framework in which control theory can benefit from a "rich" operator theory. The major impact is anticipated in nonlinear feedback control theory: controller design can be formulated as selection of an iterative algorithm to solve a nonlinear operator equation corresponding to the control objective. Stability and analysis is based on algorithm convergence properties. As an example, controllers induced by the method of successive substitution and the Newton method were introduced and the corresponding analysis and synthesis issues were studied [16]. Applied to linear systems, the proposed concepts have a straightforward interpretation in terms of familiar notions in linear controller design theory.

An application to an exothermic stirred tank reactor shows that the newly developed Newton controllers can stabilize systems under conditions where all linear controllers fail.

C. Economou

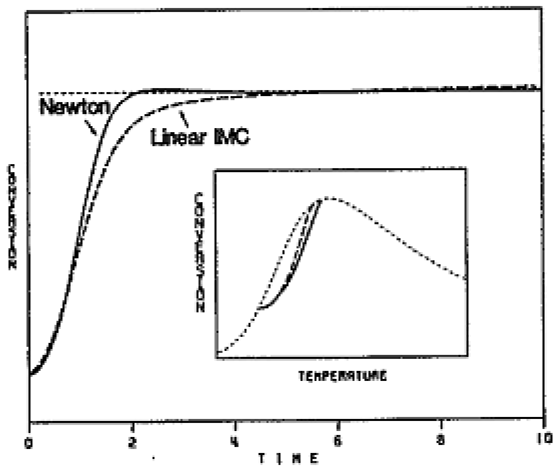


Fig. 1 Starting of exothermic reactor with conversion controlled by temperature.

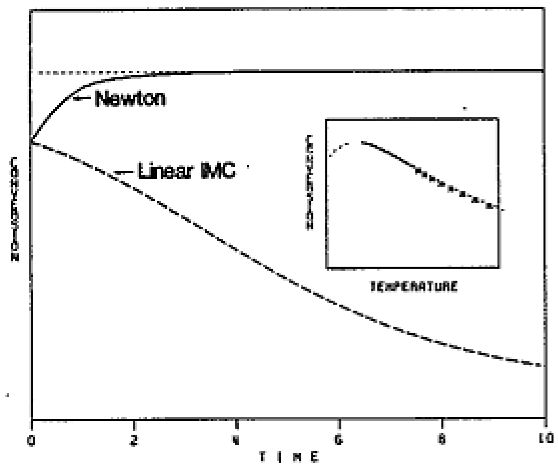
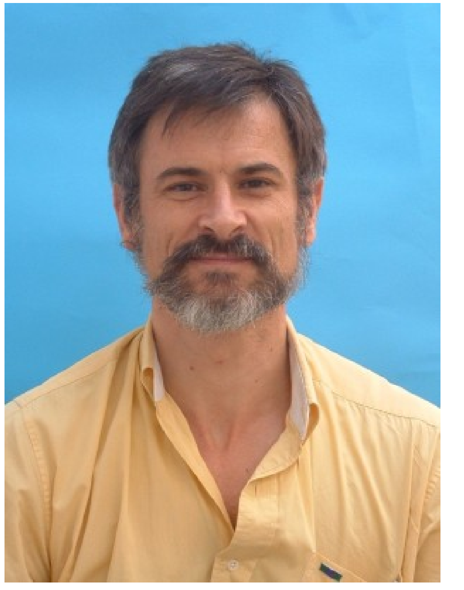


Fig. 2. Unstable disturbance response with linear controller, stabilization with Newton controller.

ROBEX: ROBUST CONTROL SYNTHESIS VIA EXPERT SYSTEM



This research addresses two related issues within the process control field. The first of these is the application of expert system technology to the problem of process control system design. The second involves the interpretation of qualitative user-heuristics ("rules of thumb") regarding observed (simulated) responses for designed controllers, and its use in guiding the search for an acceptable compensator.

Engineers practicing in the control system design field often experience difficulty in staying abreast of recent developments. Since most existing control design packages require some level of expertise with the inherent techniques, they cannot effectively disseminate new design technology. A system structure has been developed which will allow all users access at skill levels commensurate with their backgrounds.

The general structure embodies both the theory and heuristics surrounding control design, and has allowed the generation of an interactive control design environment, the ROBEX system, in which both novices and experts can benefit from the latest developments in theory and methodology.

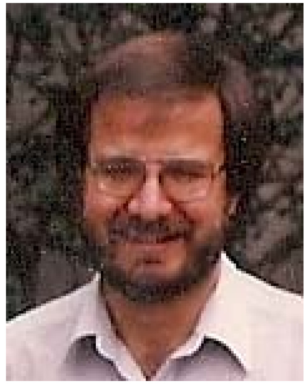
Performance specifications in process control are usually quite subjective and qualitative in nature. Here, process uncertainty adds another dimension to the dilemma faced by novice users, and "expert system" type decision-making is needed to aid them. The "bottom-line" on any design is its time domain response, and this could be specified a priori by the novice as typical step response attributes (e.g., settling time, overshoot, etc.). The expert system should have the capacity to both check if these specifications are reasonable and also to translate them into a form required by the design methods, which is usually in the frequency domain.

A successful algorithm for process control system design should incorporate a means for processing qualitative judgements from novice users on the simulated time domain response. When collated into evaluation records, these user-heuristics could be used in an optimization scheme whose objective is a user-acceptable design.

The current implementation allows the user to generate a SISO design for a plant described by a rational transfer function with uncertain coefficients. Extensions are forseen to include more general process descriptions. The long term goal is the development of a package addressing the MIMO problem.

D. R. Lewin, R. Heersink, A. Skjellum, M. Creed

SYNTHESIS AND ANALYSIS OF RESILIENT HEAT EXCHANGER NETWORKS



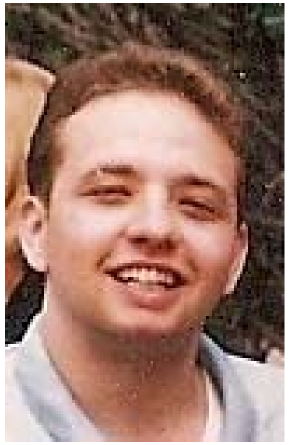
Research in heat exchanger network (HEN) synthesis has been performed by various investigators for over 15 years. Many methods have been developed to synthesize HENs which are economically optimal for assumed nominal design conditions. However, *"black box"* computer methods which automatically synthesize a HEN are not flexible enough to deal with practical constraints like the number and location of stream splits; *"pencil and paper"* techniques, although flexible, are difficult to apply to typical industrial size problems. We are developing a prototype software package RESHEX [29,30] which can handle large problems, yet which is flexible enough to allow the designer to impose practical constraints. Planned research includes making RESHEX flexible enough to design revamps of existing HENs.

In addition to being economically optimal, a practical HEN must also be able to tolerate changes in operating conditions and uncertainties in design parameters; i.e., the HEN must be *resilient.* In the past, designers have tested the resilience of a HEN simply by determining if it could operate at the extremes of an expected range of uncertain variables (such as supply temperatures and flowrates). Unfortunately, HEN resilience is a nonconvex problem; in general, it is not sufficient to check only the extremes of an uncertainty range. For example, consider the HEN in Fig. 1, with its feasible operating region shown in Fig. 2. (The boundaries for this feasible region are the ΔT_m and load constraints of exchanger 2, assuming that the HEN always uses minimum utilities. Other constraints lie outside of this region). Even though the HEN is feasible at the vertices of uncertainty range Θ, it is not feasible for all points in Θ.

We have established conditions when checking operation at the extremes of a specified uncertainty range is sufficient to guarantee resilience of a given HEN, and have developed tests for some of the cases when the HEN resilience problem is not convex [40,41] (e.g., HENs with temperature dependent heat capacities, stream splits, flowrate variations, or large temperature variations). We have also derived a *resilience index* [11] useful for comparing the resilience of two or more HENs, and a target for the maximum resilience which can be achieved for specified nominal stream conditions [62]. Planned research areas include the development of a minimum area target for resilient HENs, and of techniques to predict the *dynamic resilience* of a HEN (its ability to respond quickly and smoothly to temperature and flowrate disturbances).

R. Colberg, P. Engel

DEVELOPMENT OF ALGORITHMS, ANALYSIS AND MODELING METHODS FOR CONCURRENT SOLUTION OF NONLINEAR MATHEMATICAL PROGRAMS



The development of powerful multiple-instructure multiple-data (MIMD) concurrent computer systems offers both new opportunities and significant challenges to the engineer. If such computers can be harnessed effectively, problems of genuine interest will be solvable at higher speeds than currently possible. Furthermore, more sophisticated and computationally-demanding engineering problems will come under scrutiny because "super-computer" power will be widely available and less expensive than it is today.

However, because of their architectural characteristics, such computers require new solution methods which often differ radically from traditional (sequential) solution algorithms and which take full advantage of the ability to compute concurrently. Otherwise, minimal advantage will be realized from this emerging technology. Specifically, a well-know theoretical result (Amdahl's law) bounds concurrent performance according to the inverse of the "inherently sequential temporal fraction" of an algorithm. This bound severely limits potential benefits of a concurrent computer in the case where the "sequential fraction" is significant. Two other important issues to be addressed are load balancing of computation among nodes and the cost of communication between nodes.

The above considerations have led us to investigate asynchronous iterative approaches for solving basic optimization building blocks. Theory already exists concerning the convergence of certain classes of linear and nonlinear iterative processes which run asynchronously under relatively weak assumptions. These methods have been termed "chaotic relaxation" and "asynchronous iteration" in the literature. Our goal is to ascertain the efficiency of such procedures as compared to direct (non-iterative) matrix algorithms. These algorithms implicitly ease the problems of load balancing and communication and also can take advantage of problem sparsity.

Once low-level building blocks are understood, we hope to tackle higher level optimizations by placing "outer loops" on the building blocks. These steps will require careful attention to Amdahl's law to avoid deleterious effects on performance. Finally, we will be able to solve an interesting example problem or problems in the area of optimal control and offer performance results and intuition for other engineering optimization problems in a concurrent setting. (Specific systems to be considered are large natural gas transmission networks and distillation columns). We also hope to refine the sufficient conditions for algorithmic convergence offered in the existing literature.

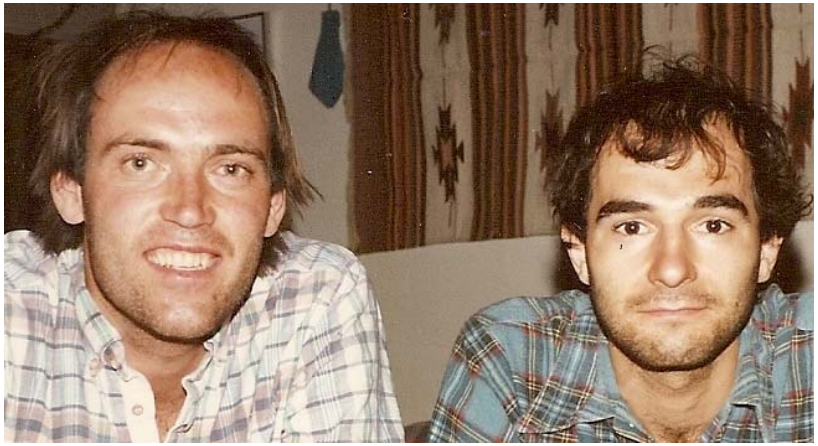
A. Skjellum

DECENTRALIZED CONTROL

Decentralized control systems have fewer tuning parameters, are easier to understand and tune, and are more easily made failure tolerant than general multivariable control systems. The restriction to a (block-) diagonal controller structure leads to performance deterioration, however, when compared to the system with a full controller matrix.

The design of a decentralized control system involves two steps: 1) the choice of pairings between measured outputs and manipulated variables (control structure) and 2) the design of each controller block. Because the Relative Gain Array provides necessary conditions for the existence of a fault-tolerant decentralized controller [5,42], it is presently the most effective tool for addressing the first step, particularly when many alternatives are available.

In the second step our approach is to design each controller block *independently*. We have derived a bound (μ Interaction Measure [26]) on the magnitude of the closed loop transfer function of the *individual* loops which guarantees closed loop stability with all loops operating simultaneously. We also know how to translate (H_∞) performance specifications and uncertainty descriptions into bounds which when satisfied by the individual loops lead to robust performance of the overall system with the decentralized controller [58]. Fig. 1 shows the μ-Interaction Measure (1) and the closed loop transfer function magnitude for three different loop designs (2-4) for a distillation column with three inputs and three outputs [26]. The proximity of design 2 to the stability boundary manifests itself in a highly oscillatory response (Fig. 2) when compared with design 4 (Fig. 3).



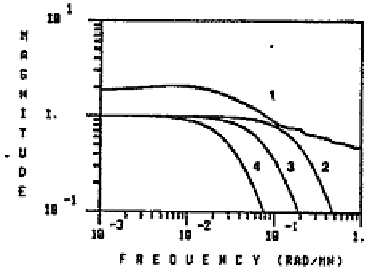


Fig. 1. 1: μ interaction measure, 2-4: magnitudes for different designs.

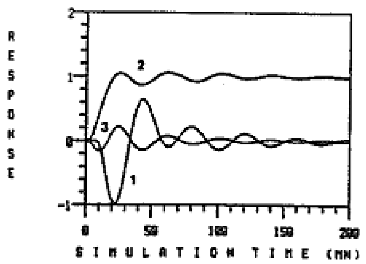


Fig. 2. Response for design 2.

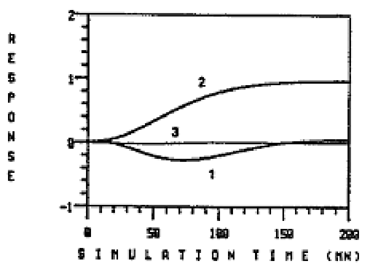
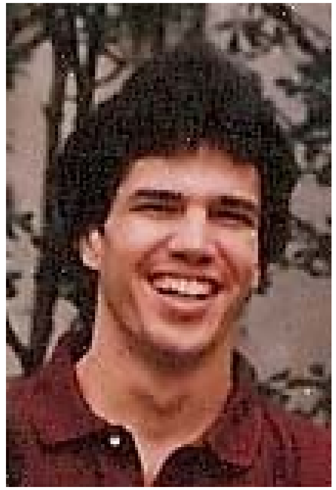


Fig. 3. Response for design 4.

Recently the μ-optimal controller synthesis method has been extended to enable synthesis of a μ-optimal two-block *decentralized* controller. The two blocks can be either SISO or MIMO controllers. The new method represents a significant advance beyond previous methods that require sequential loop-closing or independent controller designs. Repeated applications of this decentralized controller synthesis method can be used to design a decentralized controller with more than two blocks.

P. Grosdidier, S. Skogestad, D. Laughlin

CONTROLLER DESIGN FOR SYSTEMS WITH CONSTRAINTS



Economic optimization of chemical processing plants generally dictates operating points which lie on one or more process constraints. These may be constraints on process inputs, outputs, or states. This situation has prompted the development of multivariable control schemes which handle constraints. The most promising of these schemes is *model predictive control*. Model predictive controllers are implemented as on-line optimization problems whose objective is to minimize future tracking errors predicted by a process model.

We have developed a number of model predictive formulations, differing principally in the nature of the objective function, each corresponding to particular control objectives [21]. One such formulation uses the infinity- norm, resulting in the minimization of the maximum predicted tracking error. This formulation provides a simple physical interpretation of controller parameters and demonstrates regulatory behavior particularly suited for process control.

In practice the model can never exactly represent the behavior of the physical system. This model error can significantly effect the performance of the closed loop system. Current research focuses on the generalization of model predictive control to uncertain systems. The objective now is to minimize the worst tracking error for an entire family of possible plants. Recent results indicate that the required min-max problem is tractable for an uncertainty set defined by a nominal model and linear perturbations from that model. Furthermore it is evident that such a set of plants is sufficiently rich to include most meaningful uncertainty descriptions. While the definition of uncertainty sets is straightforward in the single-input single-output case, the most advantageous parametrization for the multivariable case remains a topic of research.

P. Campo

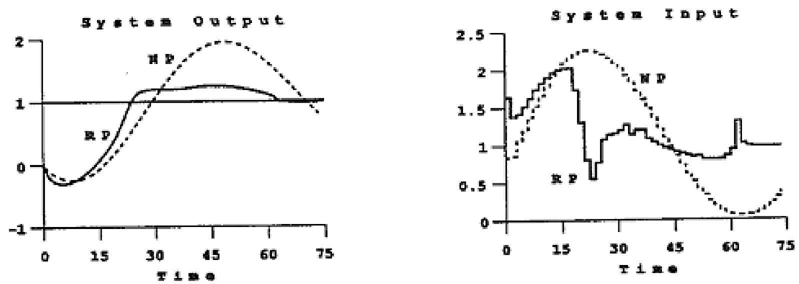


Figure 1. This example indicates the impact of model uncertainty on system performance. The curve labeled NP is the result of optimizing nominal performance (i.e., tracking error predicted by the nominal model). The curve labeled RP is the result of optimizing robust performance (worst tracking error predicted for a family of plants).

March 1990

Process Control & Design Group

California Institute of Technology

Professor Manfred Morari

CURRENT GRADUATE STUDENTS AND THESIS PROJECTS

3/8/90

Name of Student and Undergraduate School	*Research Project*	*Degree*	*Start Date*	*Expected Graduation*
Bekiaris, Nikolaos (NTU, Greece)	Azeotropic Distillation	Ph.D.	9/89	12/93
Braatz, Richard (Oregon State University)	Robust Process Control With Minimum Modeling Effort	Ph.D.	9/88	12/92
Doyle, Frank (Princeton University)	Nonlinear Process Control	Ph.D.	9/86	12/90
Gelormino, Marc (University of Rochester)	State-Space Model Predictive Control	M.S.	9/87	(leaving 3/16)
Holcomb, Tyler (University of Texas)	Neural Networks for Process Control	Ph.D.	9/87	12/92
Laroche, Lionel (Ecole Polytechnique, France)	Homogeneous Azeotropic Distillation	Ph.D.	9/86	12/90
Lee, Jay (University of Washington)	Robust Inferential Control	Ph.D.	9/86	12/90
Raven, Douglas (University of Texas)	Robust Control of Constrained Systems	Ph.D.	9/89	12/93
Skjellum, Anthony (Caltech)	Concurrent Dynamic Simulation	Ph.D.	6/84	5/90

No plantwide, decentralized or controllability/resilience.

2 on distillation design!

Conclusion: Contributions M^2

- Plantwide: Some major, a little confusion?
- Controllability (“dynamic resilience”): Major
- Decentralized: Major

Top 3 robustness papers

1. Title: Robust constrained model predictive control using linear matrix inequalities
   Author(s): Kothare MV; Balakrishnan V; Morari M
   Source: AUTOMATICA Volume: 32 Issue: 10 Pages: 1361-1379 DOI: 10.1016/0005-1098(96)00063-5 Published: OCT 1996
   Times Cited: 567 (from Web of Science)

2. Title: ROBUST-CONTROL OF ILL-CONDITIONED PLANTS - HIGH-PURITY DISTILLATION
   Author(s): SKOGESTAD S; MORARI M; DOYLE JC
   Source: IEEE TRANSACTIONS ON AUTOMATIC CONTROL Volume: 33 Issue: 12 Pages: 1092-1105 DOI: 10.1109/9.14431 Published: DEC 1988
   Times Cited: 151 (from Web of Science)

3. Title: ROBUST-CONTROL OF PROCESSES SUBJECT TO SATURATION NONLINEARITIES
   Author(s): CAMPO PJ; MORARI M
   Source: COMPUTERS & CHEMICAL ENGINEERING Volume: 14 Issue: 4-5 Pages: 343-358 DOI: 10.1016/0098-1354(90)87011-D Published: MAY 1990
   Times Cited: 124 (from Web of Science)

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