Putting Optimization into process control

Sigurd Skogestad

Department of Chemical Engineering
Norwegian University of Science and Technology (NTNU)
Trondheim

Bratilslava
03 May 2024

Putting Optimization into process control. Abstract

- How can you control a complex plant effectively using simple elements with a minimal amount of modelling?
- How can you put optimization into the control layer?

Industry has been using simple and effective as "advanced regulatory control" (ARC) schemes based on PID controllers for almost 100 years. The objective of my work is to provide a systematic approach for designing such control systems.

- The main competitor to ARC is MPC (model predictive control), but the costs of implementing and maintaining MPC solutions are high. Moreover, in most cases ARC solutions (including cascade, ratio, split range and selector control) are more flexible and easier to tune. The main problem right now is that the knowledge and competence about ARC strategies is very low, especially in academia, but also in industry the knowledge is dying out. The result is that people turn off good ARC applications, simply because they don't understand what they are doing.

- The reason for the lack of training and knowledge is that there has a been belief in academia since the 1980s, that ARC solutions (and PID control) are old-fashioned and will soon be replaced by MPC. However, MPC has now been around for 50 years, and yet the use of MPC is far from increasing as expected. The latest hype is that, if MPC is too complex, then machine learning is the solution. No, it is not, because or the lack of rich data (with sufficiently large input excitations) in most control applications, in particular in process control.

In summary, there is a need to change the mindset of people, both in academia and industry, People need to realize that ARC solutions should be a central part of the future. MPC of course has its place, but mainly as an improvement for large-scale applications that can afford the effort.

The talk will emphasize the above points and in addition present a systematic approach to ARC methods based on my recent paper (which is open access).

Reference: Sigurd Skogestad, "Advanced control using decomposition and simple elements".
Published in: Annual Reviews in Control, vol. 56 (2023), Article 100903 (44 pages).

Sigurd Skogestad is a Professor in chemical engineering at the Norwegian University of Science and Technology (NTNU) in Trondheim. He received his PhD from Caltrech in 1987 and he is the principal author together with Ian Postlethwaite of the book "Multivariable feedback control" published by Wiley in 1996 (first edition) and 2005 (second edition). The goal of his research is to develop simple yet rigorous methods to solve problems of engineering significance. Research interests include the use of feedback as a tool to (1) reduce uncertainty (including robust control), (2) change the system dynamics (including stabilization), and (3) generally make systema more well-behaved (including self-optimizing control). Other interests include limitations on performance in linear systems, control structure design and plantwide control, interactions between process design and control, and distillation column design, control and dynamics. His other main interests are mountain skiing (cross country), orienteering (running around with a map) and grouse hunting.


Midnight or midday?


Geiranger fjord


Trondheim

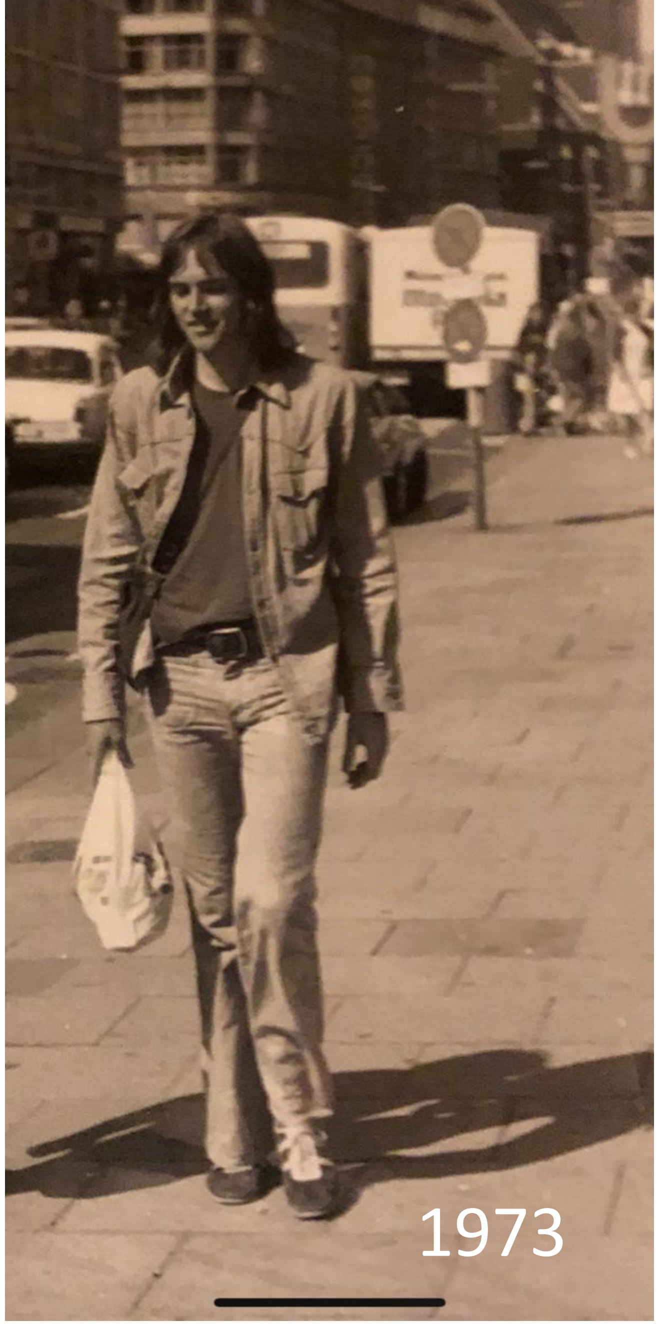

Norwegian Sea
Trondheim
Arctic circle
Sweden
Faroe Islands
Norway
Oslo
Stockholm
North Sea
Edinburgh
Copenhagen
Denmark
United Kingdom
Isle of Man
Manchester
Liverpool
Dublin
Ireland
Hamburg
Amsterdam
Netherlands
Berlin
Poland
London
Brussels
Cologne
Belgium
Germany
Frankfurt
Prague
Luxembourg
Czechia

About Sigurd Skogestad

- 1955: Born in Flekkefjord, Norway
- 1956-1961: Lived in South Africa
- 1974-1978: MS (Siv.ing.) studies in chemical engineering at NTNU
- 1979-1983: Worked at Norsk Hydro co. (process simulation)
- 1983-1987: PhD student at Caltech (supervisor: Manfred Morari)
- 1987-present: Professor of chemical engineering at NTNU
- 1994-95: Visiting Professor UC Berkeley
- 2001-02: Visiting Professor UC Santa Barbara
- 1999-2009: Head of ChE Department, NTNU
- 2015-..: Director SUBPRO (Subsea research center at NTNU)

Non-professional interests:

- **mountain skiing (cross country)**
- **orienteering (running around with a map)**
- **grouse hunting**



"The goal of my research is to develop simple yet rigorous methods to solve problems of engineering significance"


https://folk.ntnu.no/skoge/

did v2 rockets use der... adapti skogestad KLM Royal Dutch Airli... All Bookmarks

Chemical Engineering

Sigurd Skogestad Professor

Department of Chemical Engineering, Norwegian University of Science and Technology (NTNU), N7491 Trondheim, Norway

Start here...

- About me - CV - Powerpoint presentations - How to reach me - *Email: skoge@ntnu.no*
- Teaching: Courses - Master students - Project students
- Research: My Group - Research - Ph.D. students - Academic tree

"The overall goal of my research is to develop simple yet rigorous methods to solve problems of engineering significance"

"We want to find a self-optimizing control structure where close-to-optimalo operation under varying conditions is achieved with constant (or slowly varying) setpoints for the controlled variables (CVs). The aim is to move more of the burden of economic optimization from the slower time scale of the real-time optimization (RTO) layer to the faster setpoint control layer. More generally, the idea is to use the model (or sometimes data) off-line to find properties of the optimal solution suited for (simple) on-line feedback implementation"

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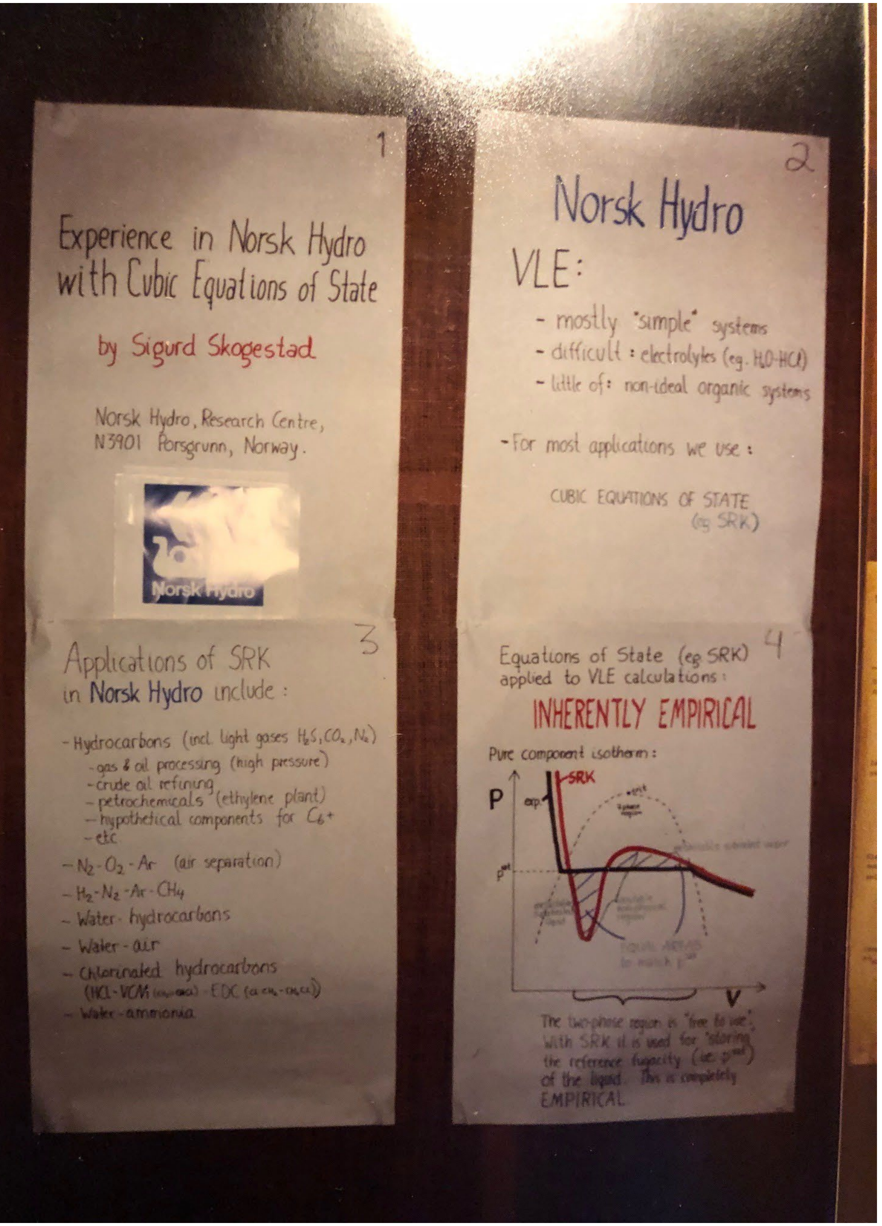
- 27 Oct. 2021: Plenary talk on "Advanced process control - A newe look at the old" at the Brazilian Chemical Engineering Conference, COBEQ 2021, Gramado, Brazil (virtual) [slides]
- 13 Oct. 2021: Plenary talk on "Advanced process control" at the Mexican Control Conference, CNCA 2021 (virtual) [video and slides]
- Nov. 2019: Sigurd receives the "Computing in chemical engineering award from the American Institute of Chemical Engineering (Orlando, 12 Nov. 2019)
- June 2019: Best paper award at ESCAPE 2019 conference in Eindhoven, The Netherlands
- July 2018: PID-paper in JPC that verifies SIMC PI-rules and gives "Improved" SIMC PID-rules for processes with time delay (taud=theta/3)
- June 2018: Video of Sigurd giving lecture at ESCAPE-2018 in Graz on how to use classical advanced control for switching between active constraints
- Feb. 2017: Youtube videos of Sigurd giving lectures on PID control and Plantwide control (at University of Salamanca, Spain)
- 06-08 June 2016: IFAC Symposium on Dynamics and Control of Process Systems, including Biosystems (DYCOPS-2016), Trondheim, Norway.
- Videos and proceedings from DYCOPS-2016
- Aug 2014: Sigurd recieves IFAC Fellow Award in Cape Town
- 2014: Overview papers on "control structure design and "economic plantwide control"
- OLD NEWS

Books...

- Book: S. Skogestad and I. Postlethwaite: MULTIVARIABLE FEEDBACK CONTROL-Analysis and design. Wiley (1996; 2005)
- Book: S. Skogestad: CHEMICAL AND ENERGY PROCESS ENGINEERING CRC Press (Taylor&Francis Group) (Aug. 2008)
- Bok: S. Skogestad: PROSESSTEKNIKK- Masse- og energibalanser Tapir (2000; 2003; 2009).

More information ...

- **Publications** from my Google scholar site
- **Download publications** from my official publication list .... or look HERE if you want to download our most recent and upublished work
- Proceedings from conferences - some of these may be difficult to obtain elsewhere
- Process control library - We have an extensive library for which Ivar has made a nice on-line search
- Photographs that I have collected from various events (maybe you are included...)
- International conferences - updated with irregular intervals
- SUBPRO (NTNU center on subsea production and processing) [Annual reports] [Internal]
- Nordic Process Control working group - in which we participate
- 5-year Master program in Chemical and Biochemical Engineering at NTNU (MTKJ) - Sigurd Skogestad is Program Leader 2019-2025.



Fluid Phase Equilibria, 13 (1983) 179—188
Elsevier Science Publishers B.V., Amsterdam — Printed in The Netherlands

EXPERIENCE IN NORSK HYDRO WITH CUBIC EQUATIONS OF STATE

SIGURD SKOGESTAD

Norsk Hydro, Research Centre, N3901 Porsgrunn, Norway

ABSTRACT

The paper presents some specific applications of cubic equations of state (EOS) in Norsk Hydro and points out some aspects of such equations that one should be aware of when using them or when developing new equations. Is is emphasized that the use of EOS to calculate vapor-liquid equilibrium is inherently empirical. Activity coefficients predicted for some systems by the Soave-Redlich-Kwong (SRK) equation of state are presented. The limitations of the van Laar equation for activity coefficients which may be derived from SRK at infinite pressures does not necessarily apply at finite pressures. The shortcomings of the SRK equations of state are pointed out and suggestions are given on how to develop an extended SRK-equation.

1983-87: Caltech

1. Robust control
2. Distillation
3. PID (IMC)

STUDIES ON ROBUST CONTROL

OF

DISTILLATION COLUMNS

Thesis by

Sigurd Skogestad

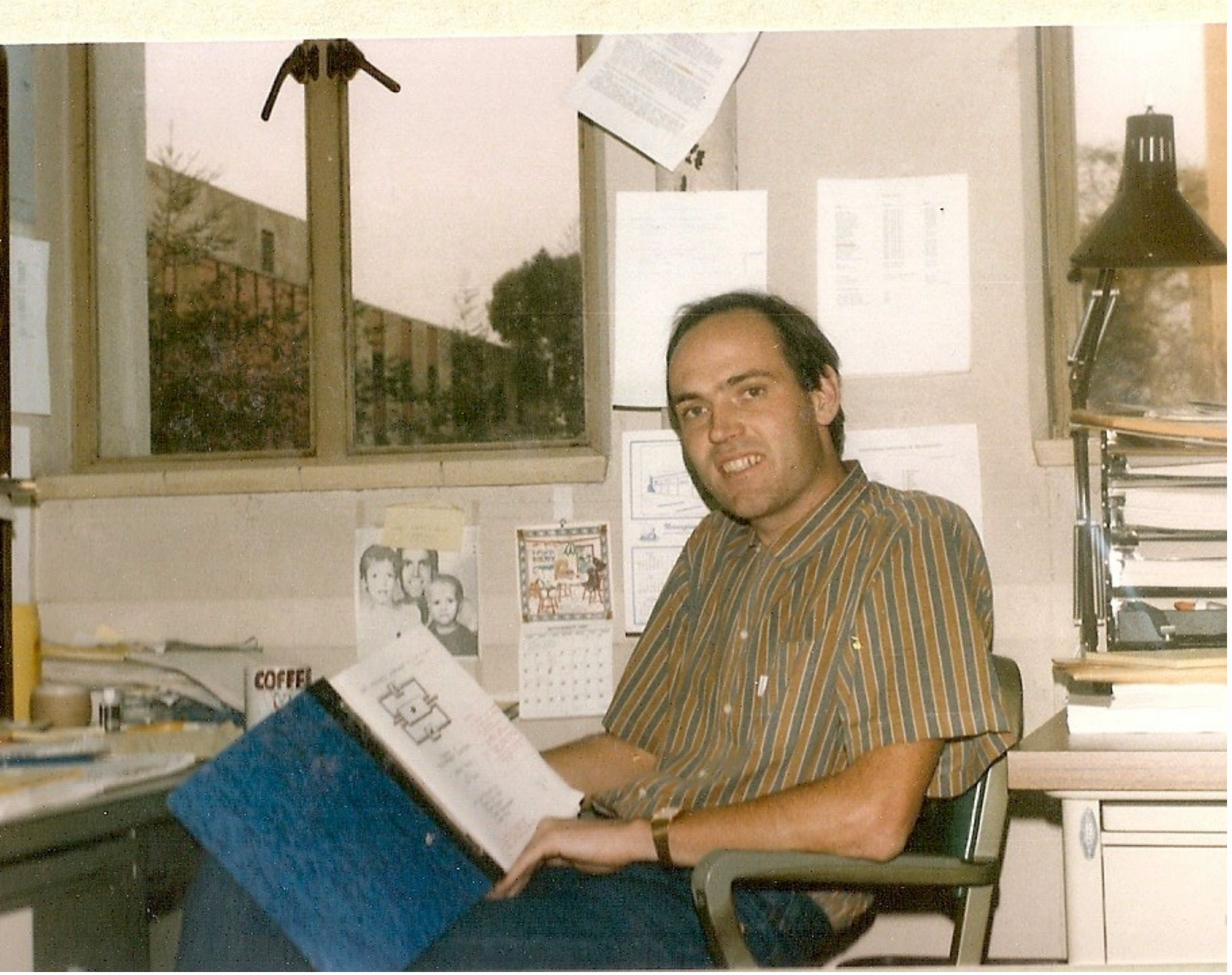
California Institute of Technology

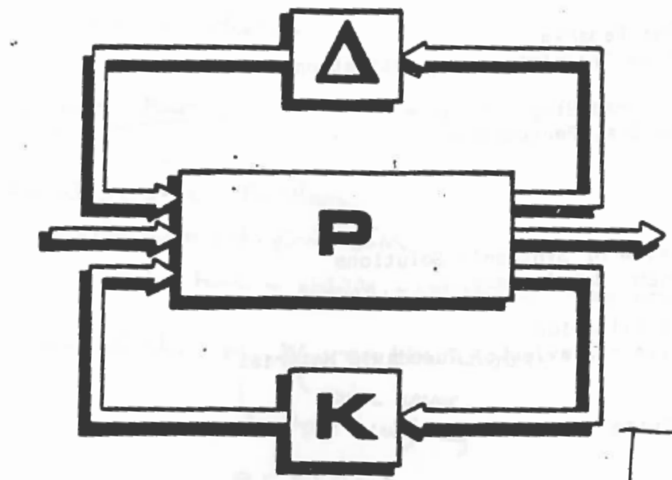
Pasadena, California

1987

(Submitted January 26, 1987)

October 1984
Robust control takes off!





ONR / HONEYWELL WORKSHOP Oct. 8-10, 1984, Minneapolis Minnesota

Advances in
MULTIVARIABLE CONTROL

Lecture Notes
by
John Doyle

with contributions by
Cheng-Chih Chu
Bruce Francis
Pramod Khargonekar
Gunter Stein

Contents:

1. Own notes from Workshop (
2. Gunter Steins overview (Oc
3. John Doyles notes (Jan-May 85-version)

H^∞-min.
3. Some own notes from Hankel-stuff (April 8
4. Chester's (Cheng-Chih Ch paper on General Dist. P (April 85)

5. Bruce Francis' overhead Stabilization & H^∞ o (Oct. 84)

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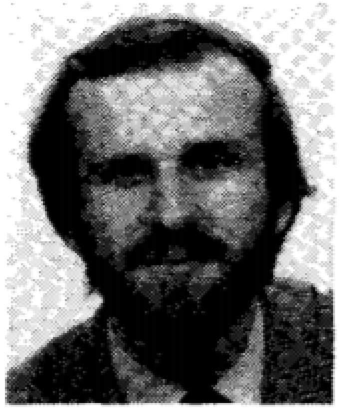
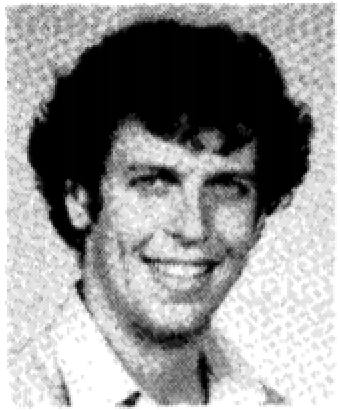
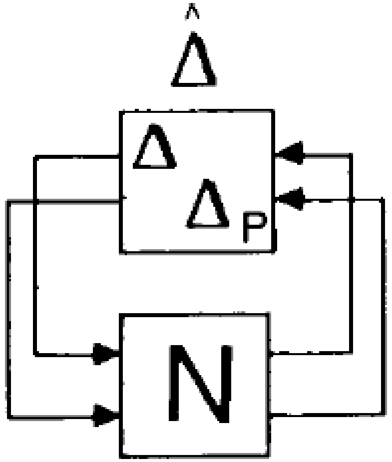
NTNU

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

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



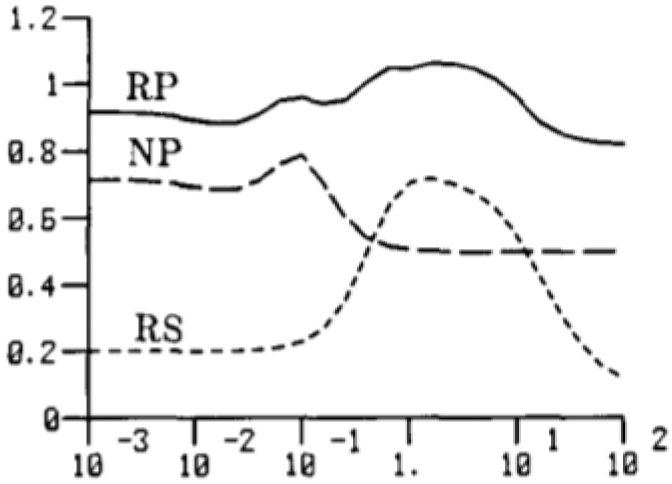
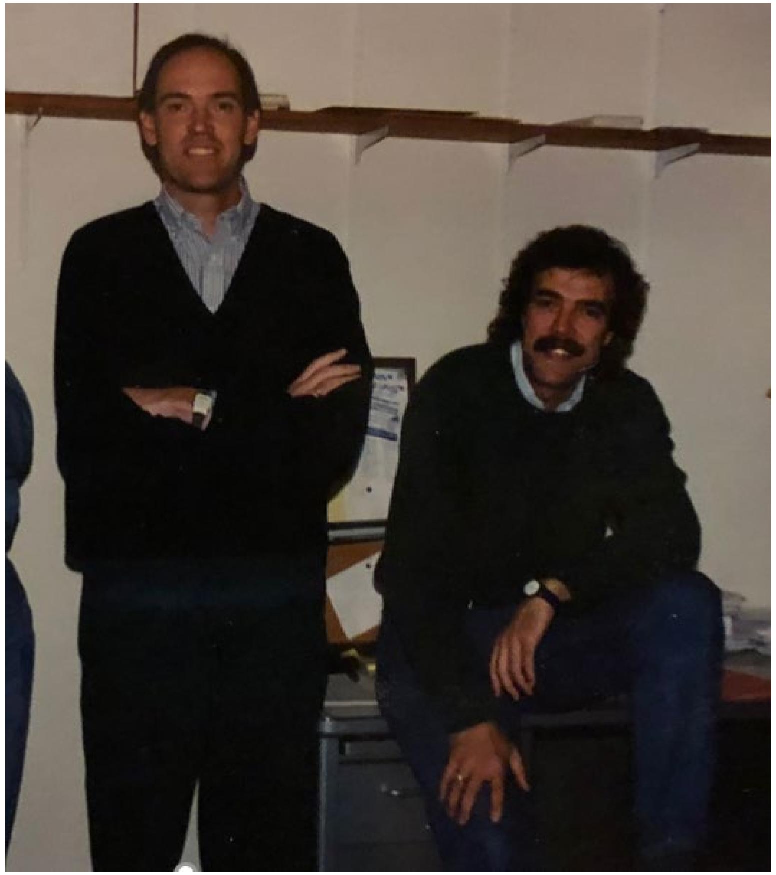
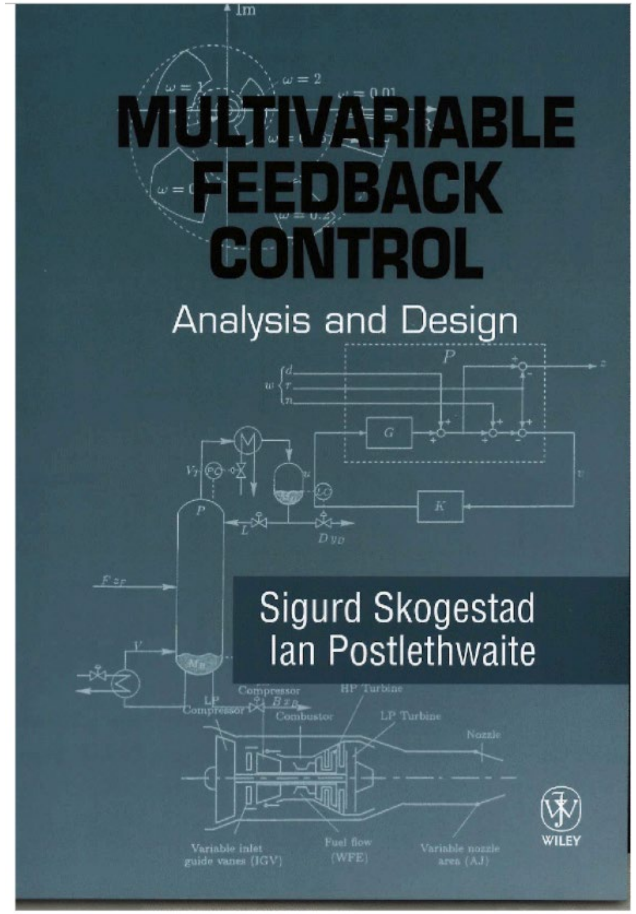


Fig. 13. LV-configuration. μ-plots for μ-optimal controller $C_\mu(s)$.

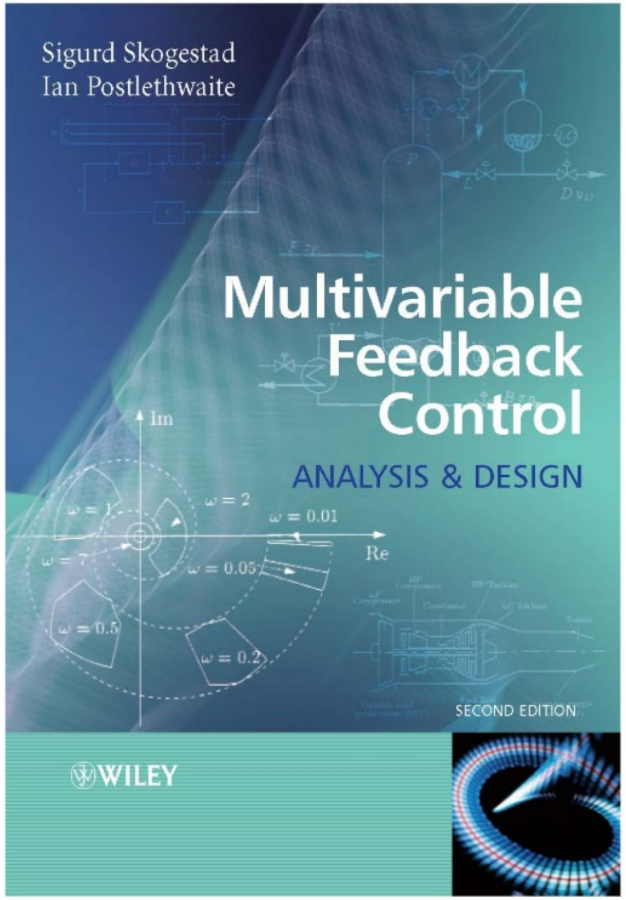
Robust control



Berkeley, Dec. 1994



1996



2005

Distillation

At home doing moonshine distillation (1979)

Chemical Engineering Research and Design

Trans IChemE, Part A, January 2007

THE DOS AND DON´TS OF DISTILLATION COLUMN CONTROL

S. Skogestad*

Department of Chemical Engineering, Norwegian University of Science and Technology, Trondheim, Norway.

Abstract: The paper discusses distillation column control within the general framework of plant-wide control. In addition, it aims at providing simple recommendations to assist the engineer in designing control systems for distillation columns. The standard LV-configuration for level control combined with a fast temperature loop is recommended for most columns.

AMERICAN CONTROL CONFERENCE
San Diego, California
June 6-8, 1984

IMPLICATIONS OF INTERNAL MODEL CONTROL FOR PID CONTROLLERS

Manfred Morari
Sigurd Skogestad

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

Daniel F. Rivera

University of Wisconsin
Department of Chemical Engineering
Madison, Wisconsin 53706

252 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.

NTNU

SIMC* PID tuning rule (2001,2003)

[19] S. Skogestad, Probably the best simple PID tuning rules in the world. AIChE Annual Meeting, Reno, Nevada, November 2001

$$g(s) = \frac{k}{(\tau_1 s + 1)(\tau_2 s + 1)} e^{-\theta s}$$

$$K_c = \frac{1}{k} \frac{\tau_1}{\tau_c + \theta}$$

$$\tau_I = \min\{\tau_1, 4(\tau_c + \theta)\}$$

$$\tau_D = \tau_2$$

Tuning paramete

$$\tau_c \geq \theta$$

ELSEVIER

Journal of Process Control 13 (2003) 291–309

JOURNAL OF PROCESS CONTROL

www.elsevier.com/locate/jprocont

Simple analytic rules for model reduction and PID controller tuning☆

Sigurd Skogestad*

Department of Chemical Engineering, Norwegian University of Science and Technology, N-7491 Trondheim, Norway

Received 18 December 2001; received in revised form 25 June 2002; accepted 11 July 2002

Abstract

The aim of this paper is to present analytic rules for PID controller tuning that are simple and still result in good closed-loop behavior. The starting point has been the IMC-PID tuning rules that have achieved widespread industrial acceptance. The rule for the integral term has been modified to improve disturbance rejection for integrating processes. Furthermore, rather than deriving separate rules for each transfer function model, there is a just a single tuning rule for a first-order or second-order time delay model. Simple analytic rules for model reduction are presented to obtain a model in this form, including the "half rule" for obtaining the effective time delay.

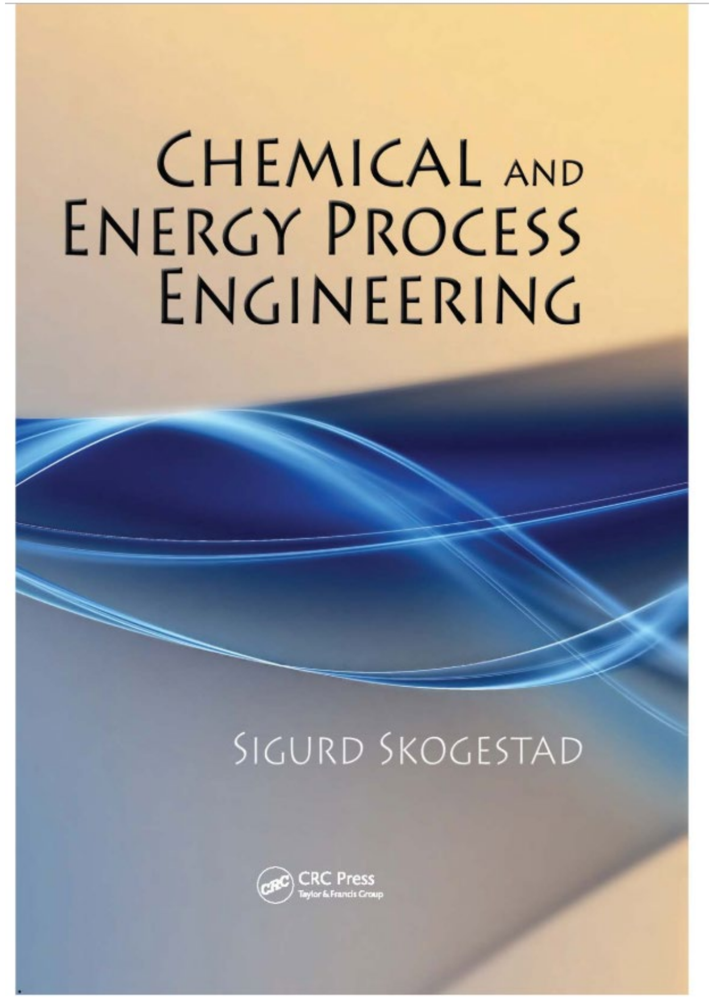
*SIMC = Simple/Skogestad IMC

NTNU

Chemical Engineering

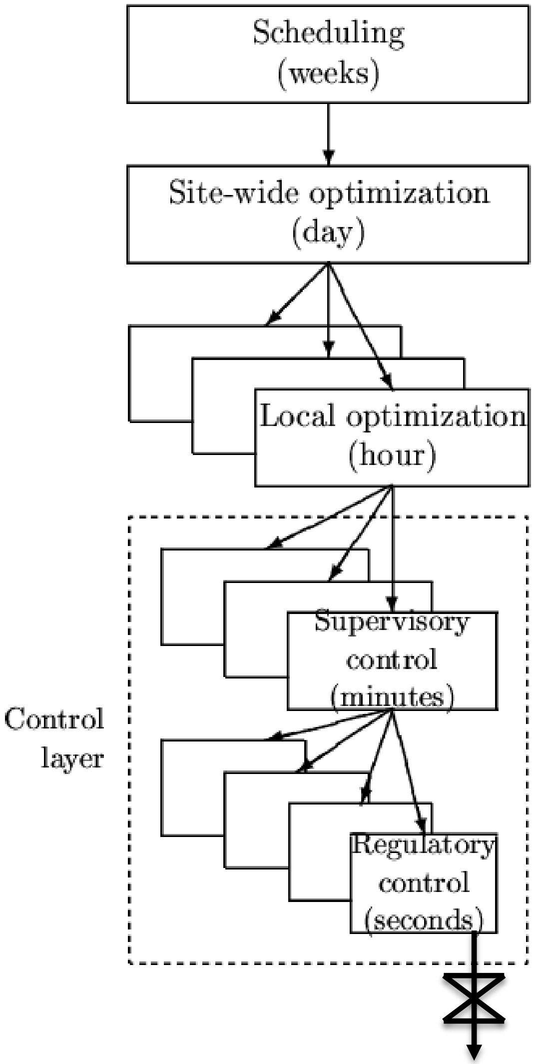
2000, 2003, 2009

2000



2009

Process control: Hierarchical decision system based on time scale separation



Control structure design = Plantwide control

Alan Foss ("Critique of chemical process control theory", AIChE Journal,1973):

*The central issue to be resolved ... is the determination of control system structure. **Which variables should be measured, which inputs should be manipulated and which links should be made between the two sets?** There is more than a suspicion that the work of a genius is needed here, for without it the control configuration problem will likely remain in a primitive, hazily stated and wholly unmanageable form. The gap is present indeed, but contrary to the views of many, it is the theoretician who must close it.*

Control structure design = Plantwide control

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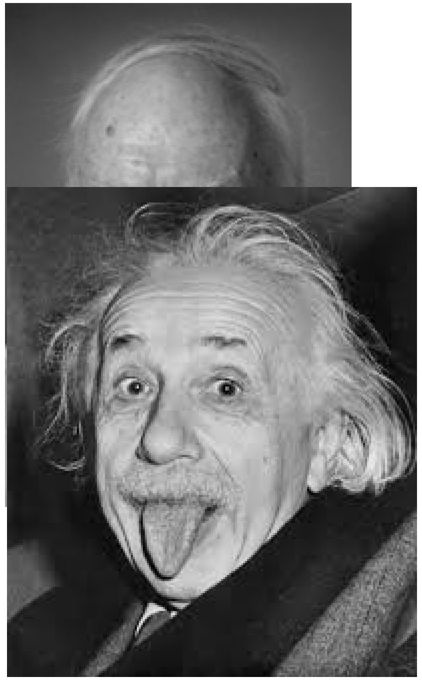

The central issue to be resolved ... is the determination of control system structure. ***Which variables should be measured, which inputs should be manipulated and which links should be made between the two sets?*** *There is more than a suspicion that the work of a genius is needed here, for without it the control configuration problem will likely remain in a primitive, hazily stated and wholly unmanageable form. The gap is present indeed, but contrary to the views of many, it is the theoretician who must close it.*

Well, I'm not a genius, but I didn't give up.
I started on this in 1983. 40 years later:

Chemical Engineering

Sigurd Skogestad Professor

Department of Chemical Engineering, Norwegian University of Science and Technology (NTNU), N7491 Trondheim, Norway

Start here...

- About me - CV - Powerpoint presentations - How to reach me - *Email: skoge@ntnu.no*
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Annual Reviews in Control 56 (2023) 100903

Contents lists available at ScienceDirect

Annual Reviews in Control

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

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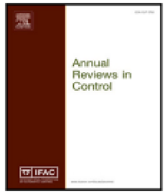


Review article

Check for updates

Advanced control using decomposition and simple elements

Sigurd Skogestad

Department of Chemical Engineering, Norwegian University of Science and Technology (NTNU), Trondheim, Norway

ARTICLE INFO

Keywords:
Control structure design
Feedforward control
Cascade control
PID control
Selective control
Override control
Time scale separation
Decentralized control
Distributed control
Horizontal decomposition
Hierarchical decomposition
Layered decomposition
Vertical decomposition
Network architectures

ABSTRACT

The paper explores the standard advanced control elements commonly used in industry for designing advanced control systems. These elements include cascade, ratio, feedforward, decoupling, selectors, split range, and more, collectively referred to as "advanced regulatory control" (ARC). Numerous examples are provided, with a particular focus on process control. The paper emphasizes the shortcomings of model-based optimization methods, such as model predictive control (MPC), and challenges the view that MPC can solve all control problems, while ARC solutions are outdated, ad-hoc and difficult to understand. On the contrary, decomposing the control systems into simple ARC elements is very powerful and allows for designing control systems for complex processes with only limited information. With the knowledge of the control elements presented in the paper, readers should be able to understand most industrial ARC solutions and propose alternatives and improvements. Furthermore, the paper calls for the academic community to enhance the teaching of ARC methods and prioritize research efforts in developing theory and improving design method.

Contents

How we design a control system for a complete chemical plant?

- Where do we start?
- What should we control? and why?
- etc.
- etc.

Optimal steady-state operation (economics)

- Typical cost function*:

J [$/s] = cost feed + cost energy – value products

*No need to include fixed costs (capital costs, operators, maintainance) at ”our” time scale (hours)

Note: J=-P where P= Operational profit

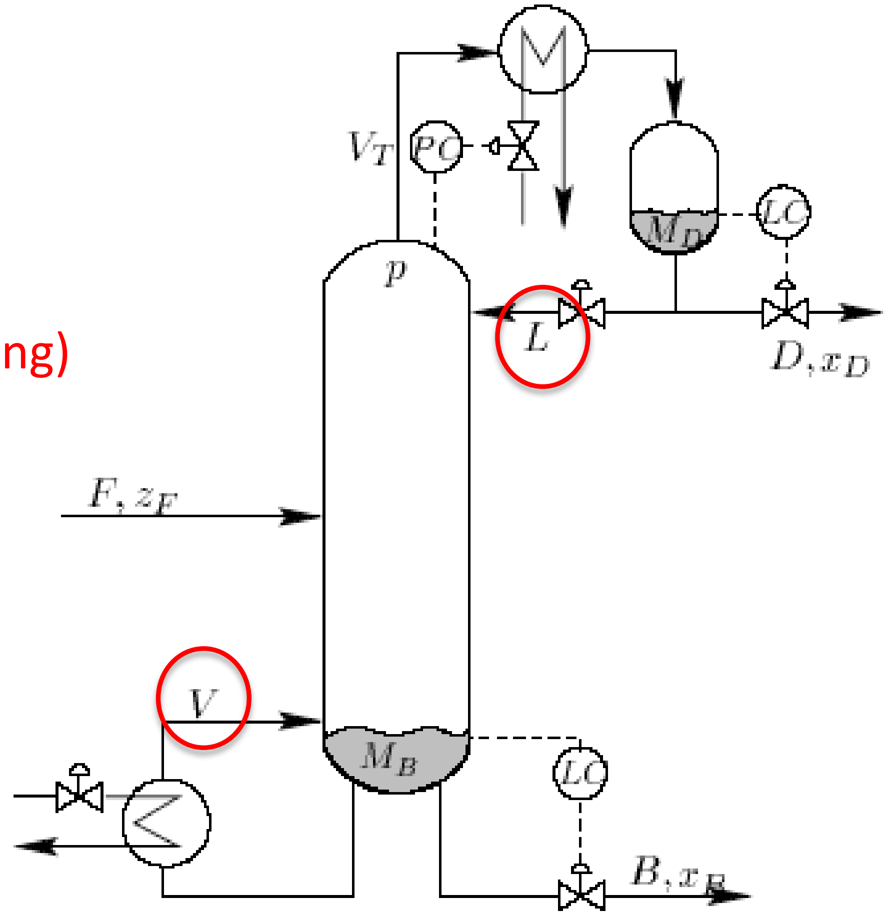
Example: distillation column

Cost J [$s] to be minimized (economics):

cost energy (heating + cooling)

$$J = -P \quad \text{where} \quad P = \underbrace{p_D D + p_B B}_{\text{value products}} - p_F F - p_V V$$

cost feed



Subject to Constraints:

Purity D: For example, $x_{D,\ impurity} \leq \max$

Purity B: For example, $x_{B,\ impurity} \leq \max$

Flow constraints: $\min \leq D, B, L$ etc. $\leq \max$

Column capacity (flooding): $V \leq V_{max}$, etc.

- Optimal operation: Minimize J with respect to steady-state degrees of freedoms (inputs u)
  - u = [reflux L; heat input V]

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Process control: Hierarchical decision system

Objective

min J (economics)

hour

Setpoint control

(+ look after other variables)

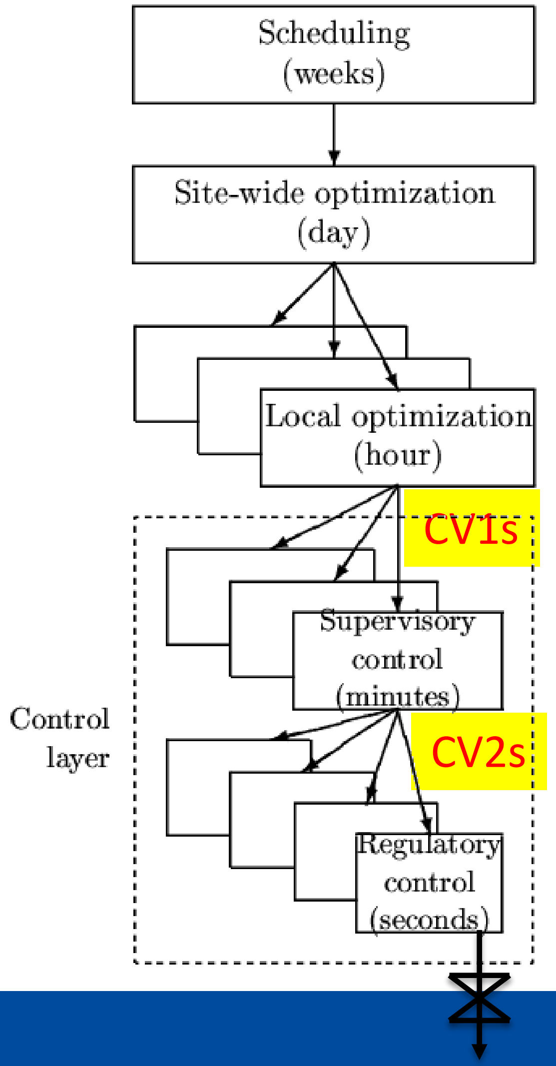
$J_c = (y-y_s)^2 + \Delta u^2$

minute

Stabilize + avoid drift

Gain margin…

second



Responsible

Manager

Process engineer

Operator/RTO (usually steady-state)

Supervisory control layer

"Advanced control"/MPC

Regulatory (basic) control layer

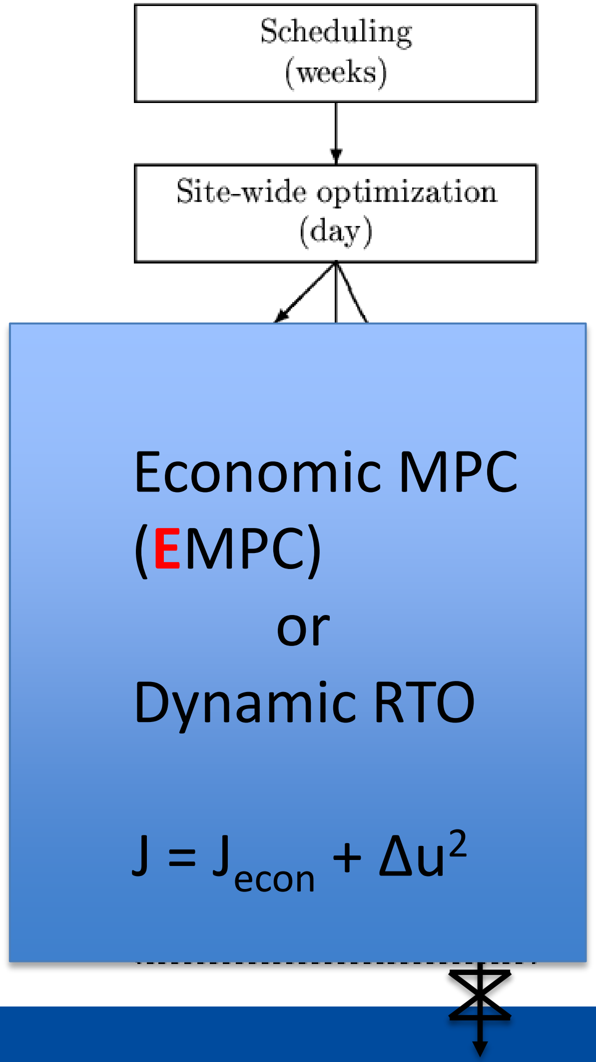
PID-control

u (MV) = valves

CV = controlled variable (with setpoint)

Process control: Hierarchical decision system

Objective |



Manager

Process engineer

min J (economics)

hour

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(+ look after other variables)

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minute

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second

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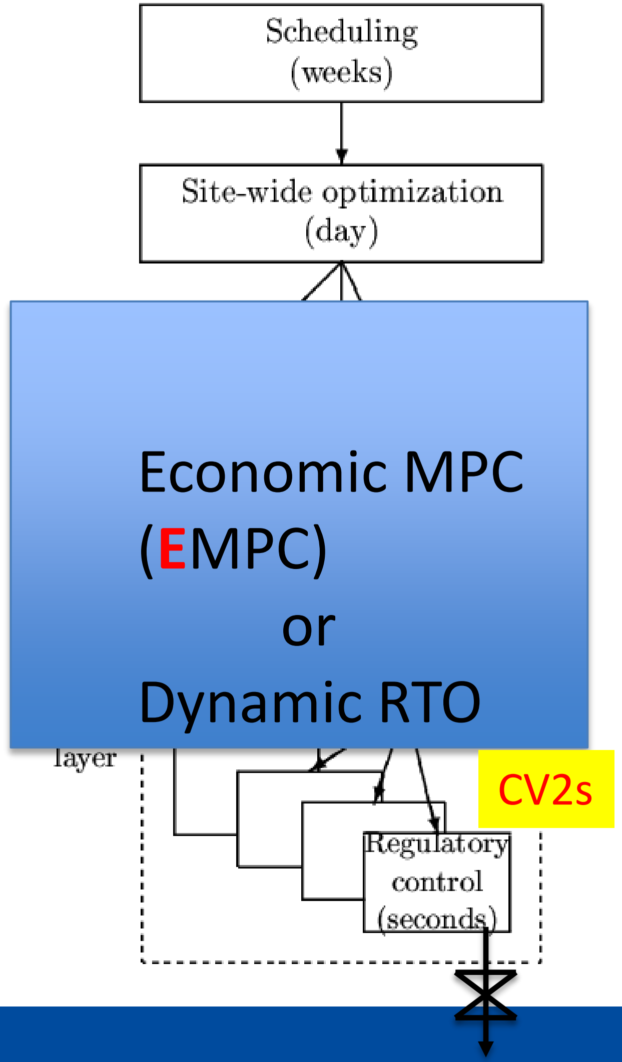
Process control: Hierarchical decision system

Objective |



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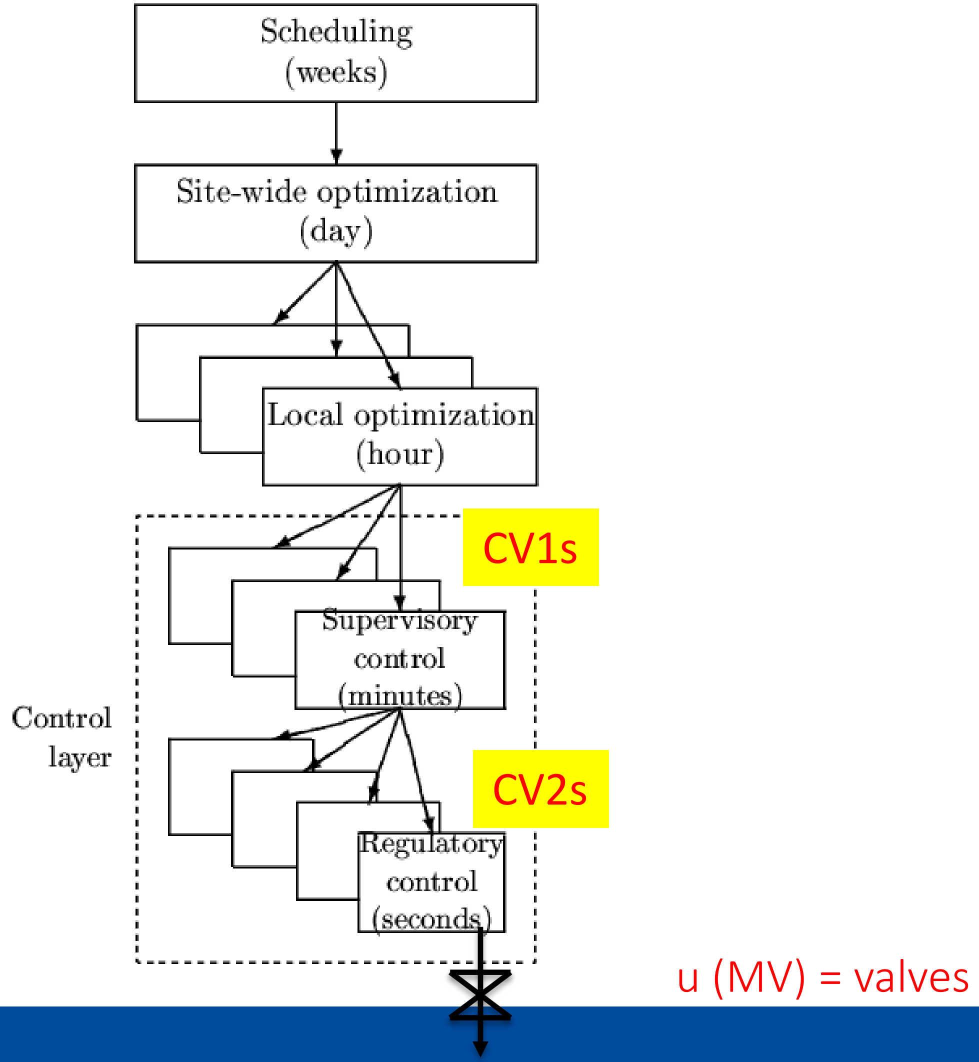
u (MV) = valves

Process control: Hierarchical decision system



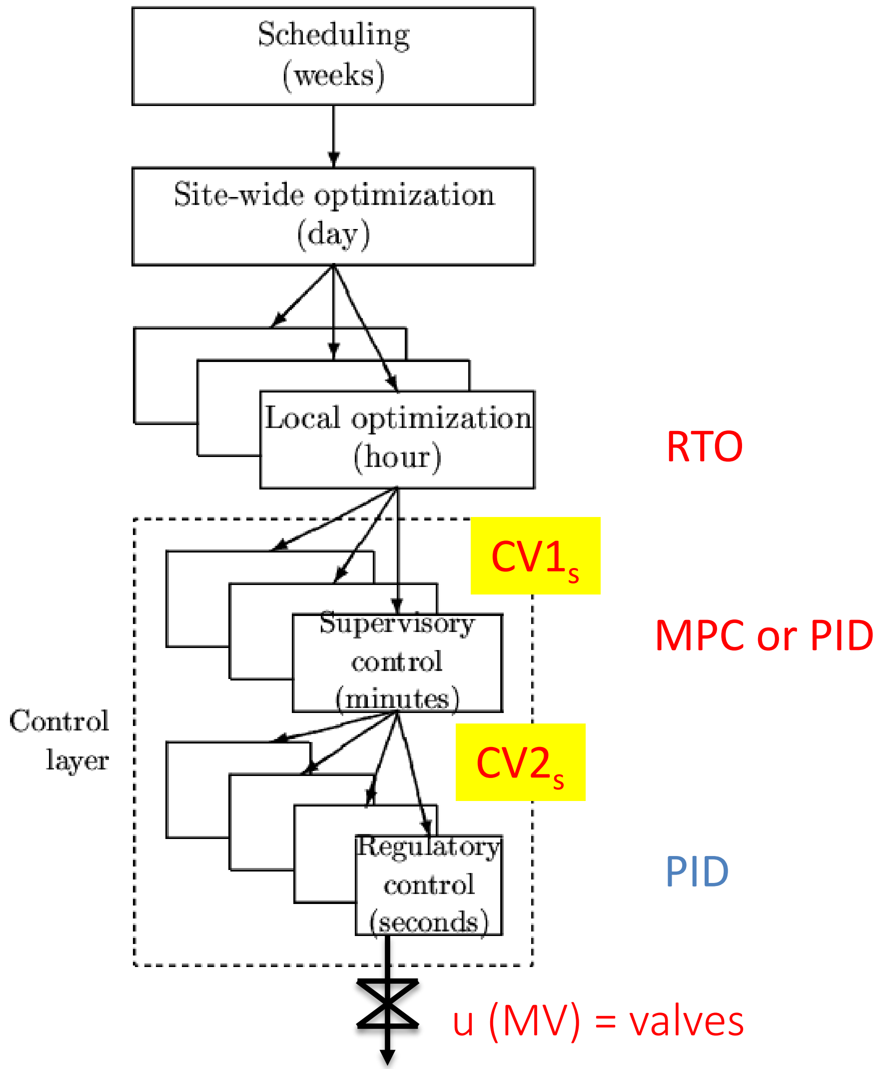
Process control layers

- **Real-time optimization layer (RTO):**
  - Optimize setpoints $CV1_s$ based on detailed nonlinear model (usually steady state)

- **Supervisory/"Advanced" control:**
  - Follow set points for CV1
  - Switch between active constraints (change CV1)
  - Look after regulatory layer (avoid that MVs saturate, etc.)

  Implementation:
  Alternative 1: "Advanced PID" (ARC) based on "simple elements"
  Alternative 2: MPC (model predictive control)

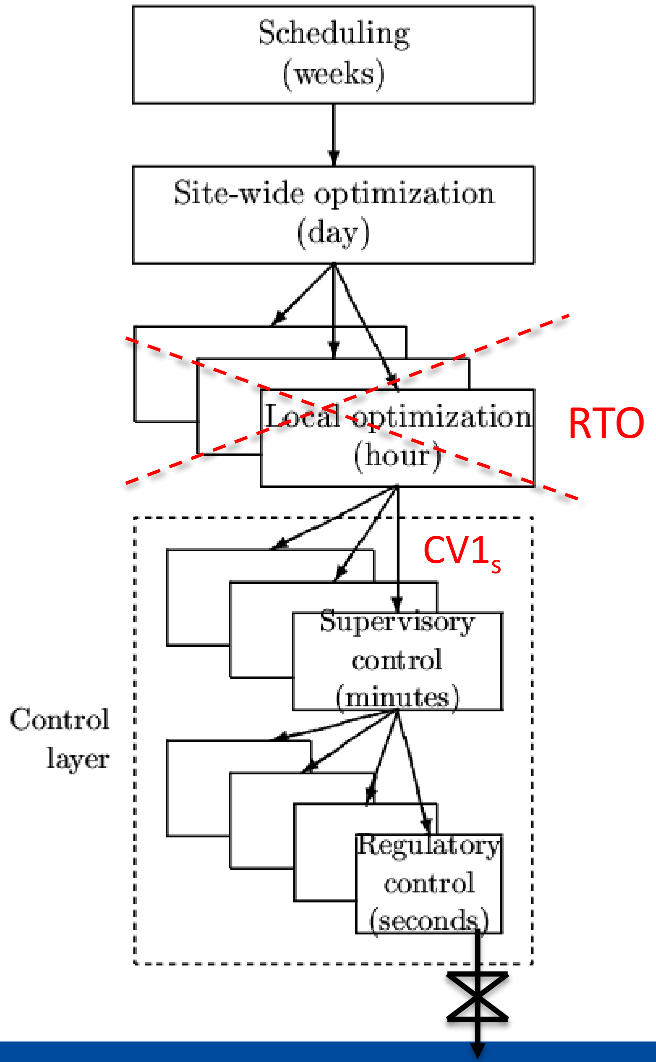
- **Regulatory control (PID):**
  - Stable operation (CV2)



CV = controlled variable. MV = manipulated variable. ARC = Advanced regulatory (PID) control

Move optimization into the control layer

- Try to eliminate RTO-layer
- = «Feedback-optimizing control»
- Unconstrained case: Select CV1 using «self-optimizing» control
  - Ideal: Control cost gradient to zero, $CV1=J_u=0$
- Changing active constraints: More complicated *(but I think we now have solved it ☺)*
  - I. Primal-dual optimizing control (with control of constraints on slow timescale)
    - Can use PID control
    - May add fast override control for constraints
  - Region-based control (fast control of constraints)
    - II. More inputs (MVs) than constraints: Can use PID with selectors
    - III. General case: MPC with changing cost function (swicth CV1)

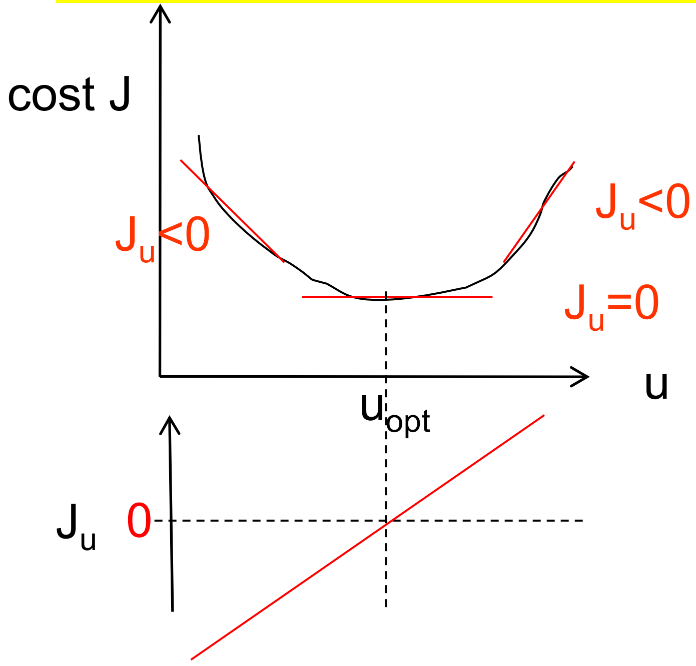


Optimal steady-state operation

$\min_u J(u,d)$

s.t. $g(u,d) \geq 0$ (constraints)

- J = economic cost [$/s]
- Unconstrained case: Optimal to keep gradient J_u □ $\partial J/\partial u = 0$



- Constrained case: KKT-conditions: $g_A = 0$, $L_u = J_u + \lambda^T g_u = 0$

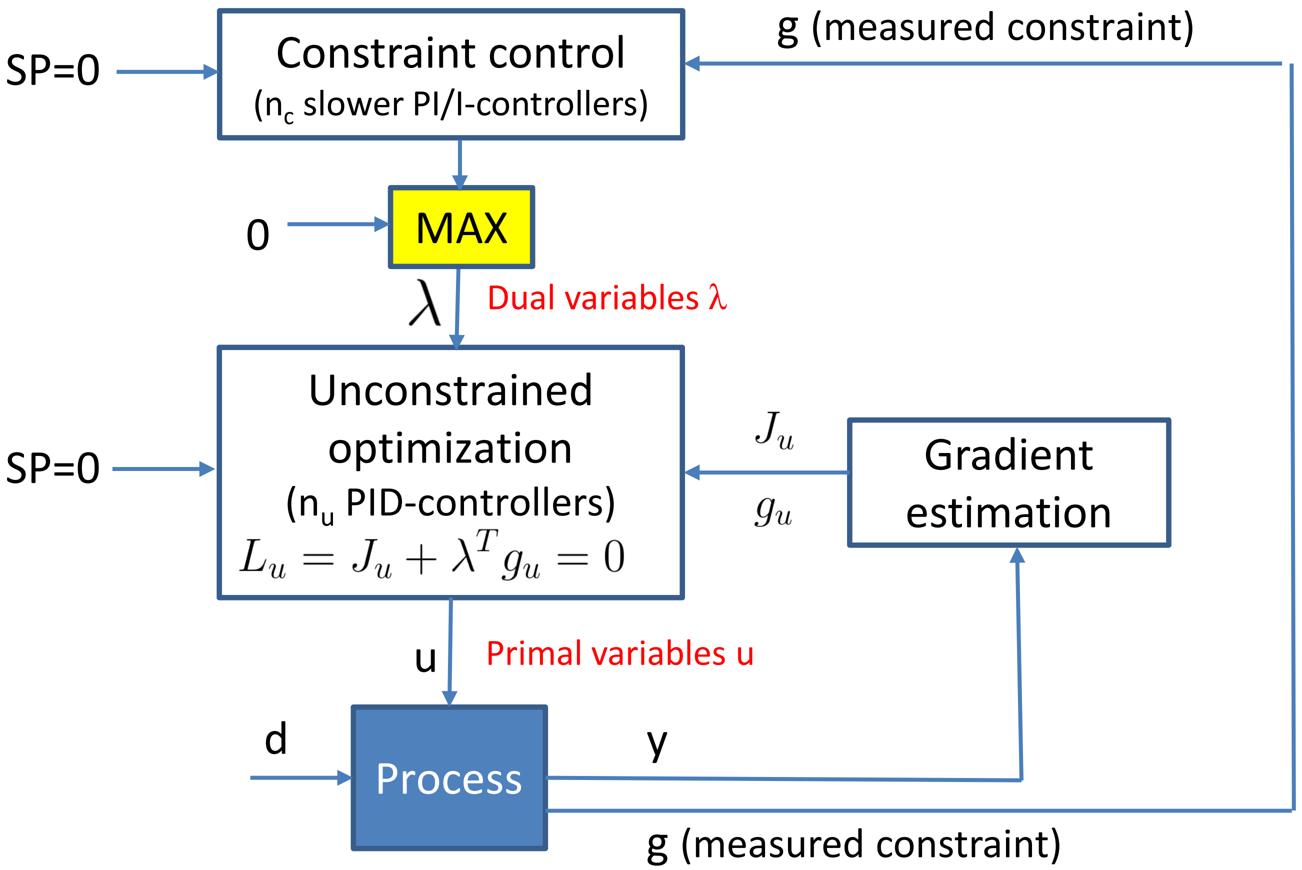
Optimal steady-state operation

Want tight control of active constraints for economic reasons

- Active constraint: $g_A=0$
- Tight control of g_A minimizes «back-off»

- How can we identify and control active constraints?
- How can we switch constraints?

I. Primal-dual control based on KKT conditions: Feedback solution that automatically tracks active constraints by adjusting Lagrange multipliers (= shadow prices = dual variables) λ



$$L_u = J_u + \lambda^T g_u = 0$$

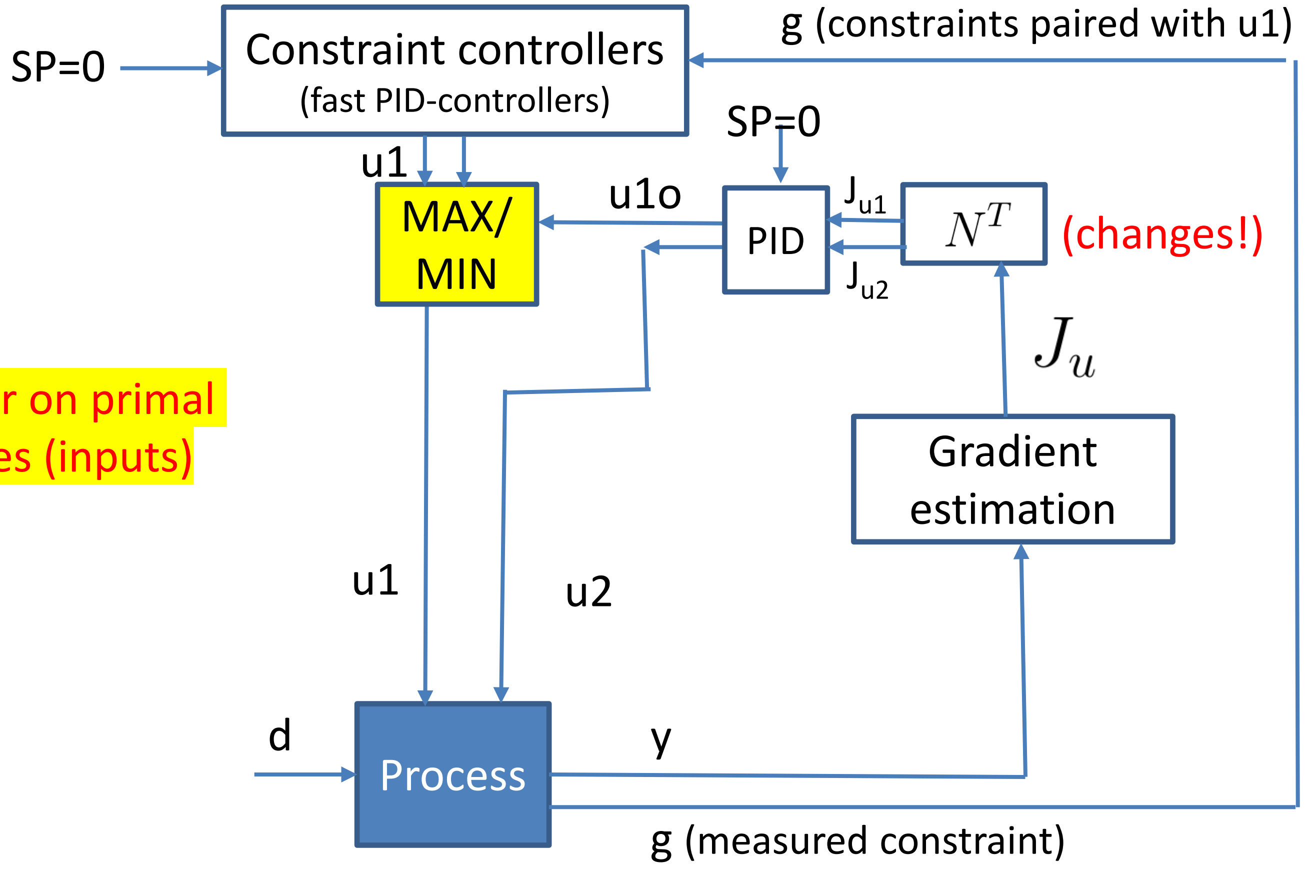
Inequality constraints: $\lambda \geq 0$

Primal-dual feedback control.

- Makes use of «dual decomposition» of KKT conditions
- Selector on dual variables λ
- Problem: Constraint control using dual variables is on slow time scale

- D. Krishnamoorthy, A distributed feedback-based online process optimization framework for optimal resource sharing, J. Process Control 97 (2021) 72–83,
- R. Dirza and S. Skogestad . Primal–dual feedback-optimizing control with override for real-time optimization. J. Process Control, Vol. 138 (2024), 103208.

II. Region-based feedback solution with «direct» constraint control (for case with more inputs than constraints)



- Selector on primal variables (inputs)

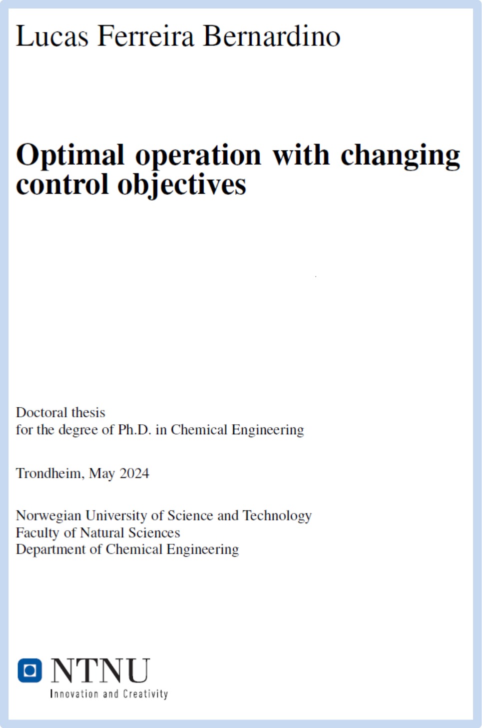
KKT: $L_u = J_u + \lambda^T g_u = 0$

Introduce N: $N^T g_u = 0$

Control
1. Reduced gradient $N^T J_u = 0$
   - «self-optimizing variables»)
2. Active constrints g_A = 0.

- Jaschke and Skogestad, «Optimal controlled variables for polynomial systems». S., J. Process Control, 2012
- D. Krishnamoorthy and S. Skogestad, «Online Process Optimization with Active Constraint Set Changes using Simple Control Structure», I&EC Res., 2019
- Bernardino and Skogestad, Decentralized control using selectors for optimal steady-state operation with changing active constraints, J. Process Control, Vol. 137, 2024

Static gradient estimation:
Very simple and works well!


Lucas Ferreira Bernardino

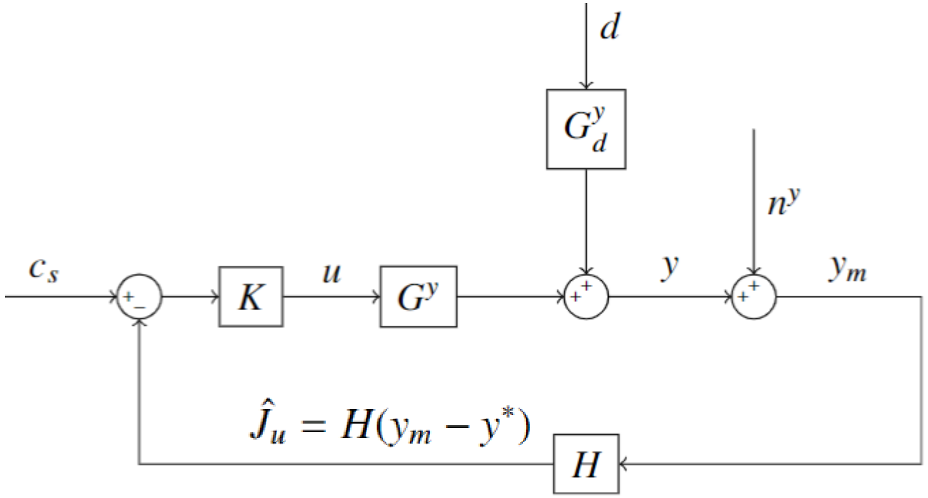
Optimal operation with changing control objectives

Doctoral thesis
for the degree of Ph.D. in Chemical Engineering

Trondheim, May 2024

Norwegian University of Science and Technology
Faculty of Natural Sciences
Department of Chemical Engineering

NTNU
Innovation and Creativity



From «exact local method» of self-optimizing control:

$$H^J = J_{uu}\left[G^{yT}\left(\tilde{F}\tilde{F}^T\right)^{-1}G^y\right]^{-1}G^{yT}\left(\tilde{F}\tilde{F}^T\right)^{-1}$$

where $\tilde{F} = [FW_d \quad W_{n^y}]$ and $F = \frac{dy^{opt}}{dd} = G_d^y - G^y J_{uu}^{-1} J_{ud}$.

III. Region-based MPC with switching of cost function (for general case)

Standard MPC with fixed CVs: Not optimal

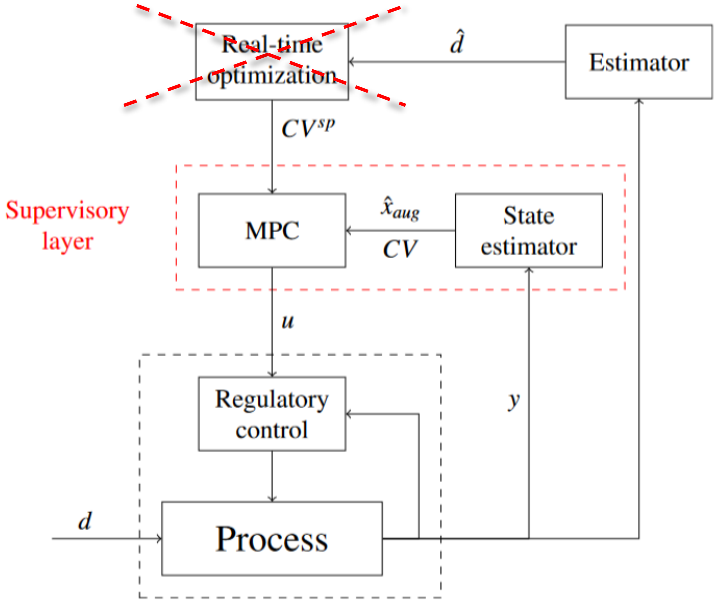


Figure 1: Typical hierarchical control structure with standard setpoint-tracking MPC in the supervisory layer. The cost function for the RTO layer is J^{ec} and the cost function for the MPC layer is J^{MPC}. With no RTO layer (and thus constant setpoints CV^{sp}), this structure is not economically optimal when there are changes in the active constraints. For smaller applications, the state estimator may be used also as the RTO estimator.

$$J^{MPC} = \sum_{k=1}^{N} \|CV_k - CV^{sp}\|_Q^2 + \|\Delta u_k\|_R^2$$

Proposed: With changing cost (switched CVs)

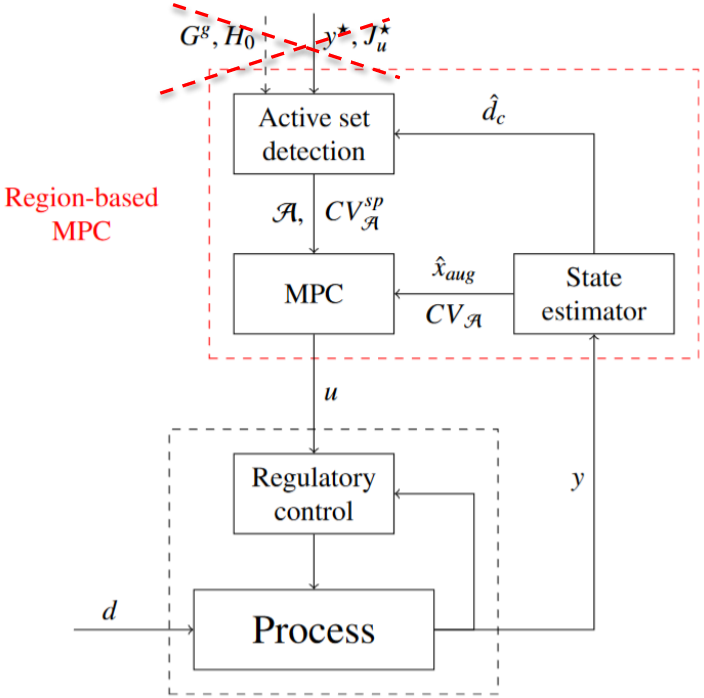


Figure 2: Proposed region-based MPC structure with active set detection and change in controlled variables. The possible updates from an upper RTO layer (y^*, J_u^* etc.) are not considered in the present work. Even with no RTO layer (and thus with constant setpoints $CV_{\mathcal{A}}^{sp}$, see (14) and (15), in each active constraint region), this structure is potentially economically optimal when there are changes in the active constraints.

$$J_{\mathcal{A}}^{MPC} = \sum_{k=1}^{N} \|CV_{\mathcal{A}} - CV_{\mathcal{A}}^{sp}\|_{Q_{\mathcal{A}}}^2 + \|\Delta u_k\|_{R_{\mathcal{A}}}^2$$

$$CV_{\mathcal{A}} = \begin{bmatrix} g_{\mathcal{A}} \\ c_{\mathcal{A}} \end{bmatrix} = \begin{bmatrix} g_{\mathcal{A}} \\ N_{\mathcal{A}}^T H_0 y \end{bmatrix} \quad (14)$$

$$H_0 = \begin{bmatrix} J_{uu} & J_{ud} \end{bmatrix} \begin{bmatrix} G^y & G_d^y \end{bmatrix}^\dagger$$

- Bernardino and Skogestad, Optimal switching of MPC cost function for changing active constraints. J. Proc. Control, 2024 (submitted)

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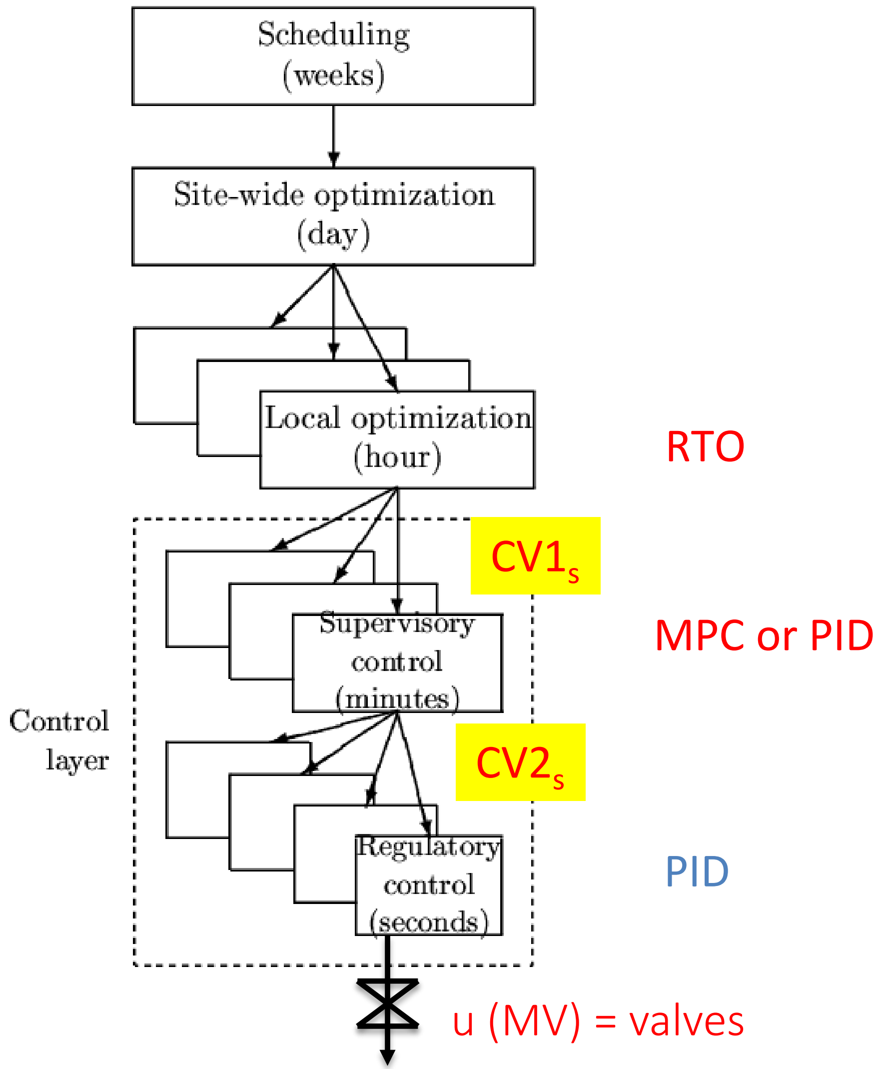
Process control layers

- **Real-time optimization layer (RTO):**
  - Optimize setpoints $CV1_s$ based on detailed nonlinear model (usually steady state)

- **Supervisory/"Advanced" control:**
  - Follow set points for CV1
  - Switch between active constraints (change CV1)
  - Look after regulatory layer (avoid that MVs saturate, etc.)

Implementation:
Alternative 1: "Advanced PID" (ARC) based on "simple elements"
Alternative 2: MPC (model predictive control)

- **Regulatory control (PID):**
  - Stable operation (CV2)



CV = controlled variable. MV = manipulated variable. ARC = Advanced regulatory (PID) control

«Advanced» control

- This is a relative term
- Usually used for anything than comes in addition to (or in top of) basic PID loops
- Main options
  - ARC using advanced control elements
    - PID + Cascade, feedforward, selectors, etc.
    - This option is preferred if it gives acceptable performance and it's not too complicated
  - Model predictive control (MPC)
    - Requires more effort to implement and maintain

ARC = Advanced regulatory (PID) control

Academia: MPC

- MPC
  - General approach, but we need a dynamic model
  - Also: MPC is usually implemented only after some time of operation
  - Furthermore: Not all problems are easily formulated using MPC

- So we need something in addition to MPC

Research question: Alternative simpler solutions to MPC

- Would like: Feedback solutions that can be implemented without a detailed models

- Machine learning?
  - Requires a lot of data
  - Can only be implemented after the process has been in operation

- Solution: "Classical advanced control" (ARC) based on single-loop PIDs
  - Extensively used by industry
  - Problem for engineers: Lack of design methods
    - Has been around since 1930's
    - But almost completely neglected by academic researchers
  - Main fundamental limitation: Based on single-loop (need to choose pairing)

ARC = Advanced regulatory (PID) control

QUIZ
What are the three most important inventions of process control?

- Hint 1: According to Sigurd Skogestad
- Hint 2: All three are from the 1930's

SOLUTION

1. PID controller, in particular, I-action
2. Cascade control
3. Ratio control

Standard Advanced control elements

Sigurd Skogestad, "Advanced control using decomposition and simple elements". Annual Reviews in Control, vol. 56 (2023), Article 100903 (44 pages).

- Each element links a subset of inputs with a subset of outputs
- Results in simple local tuning

First, there are some elements that are used to improve control for cases where simple feedback control is not sufficient:

E1*. Cascade control[2]
E2*. Ratio control
E3*. Valve (input)[3] position control (VPC) on extra MV to improve dynamic response.

Next, there are some control elements used for cases when we reach constraints:

E4*. Selective (limit, override) control (for output switching)
E5*. Split range control (for input switching)
E6*. Separate controllers (with different setpoints) as an alternative to split range control (E5)
E7*. VPC as an alternative to split range control (E5)

All the above seven elements have feedback control as a main feature and are usually based on PID controllers. Ratio control seems to be an exception, but the desired ratio setpoint is usually set by an outer feedback controller. There are also several features that may be added to the standard PID controller, including

E8*. Anti-windup scheme for the integral mode
E9*. Two-degrees of freedom features (e.g., no derivative action on setpoint, setpoint filter)
E10. Gain scheduling (Controller tunings change as a given function of the scheduling variable, e.g., a disturbance, process input, process output, setpoint or control error)

In addition, the following more general model-based elements are in common use:

E11*. Feedforward control
E12*. Decoupling elements (usually designed using feedforward thinking)
E13. Linearization elements
E14*. Calculation blocks (including nonlinear feedforward and decoupling)
E15. Simple static estimators (also known as inferential elements or soft sensors)

Finally, there are a number of simpler standard elements that may be used independently or as part of other elements, such as

E16. Simple nonlinear static elements (like multiplication, division, square root, dead zone, dead band, limiter (saturation element), on/off)
E17*. Simple linear dynamic elements (like lead–lag filter, time delay, etc.)
E18. Standard logic elements

[2] The control elements with an asterisk * are discussed in more detail in this paper.
[3] In this paper, Valve Position Control (VPC) refers to cases where the input (independent variable) is controlled to a given setpoint ("ideal resting value") on a slow time scale. Thus, the term VPC is used for other inputs (actuator signals) than valve position, including pump power, compressor speed and flowrate, so a better term might have been Input Position Control.

“Classical Advanced control” (ARC) using simple control elements

E1. Cascade control

- Have Extra output (state) measurements

E2. Ratio and feedforward control

- Have measured disturbance

E12. Decoupling elements

- Have interactive process

E13. Linearization elements / Adaptive gain

- Have Nonlinear process

Often static nonlinear «function block»
One unifying approach is «Transformed inputs» (similar to feedback linearization)

E5-E7. Split-range control (or multiple controllers or VPC)

- Need extra inputs (MV) to handle all conditions (steady state) (MV-MV switch)

E3. Valve position control (VPC) (Input resetting/Midranging control)

- Have extra inputs dynamically

E4. Selectors

- Have changes in active constraints (CV-CV switch)

ARC = Advanced regulatory (PID) control

How design classical APC elements?

- Industrial literature (e.g., Shinskey).

  Many nice ideas. But not systematic. Difficult to understand reasoning

- Academia: Very little work
  - I feel alone

Annual Reviews in Control 56 (2023) 100903

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Review article

Advanced control using decomposition and simple elements

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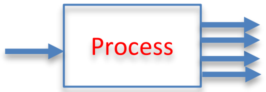
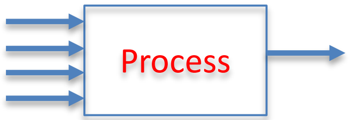
Constraint switching
(because it is optimal at steady state)

- CV-CV switching
  - Control one CV at a time

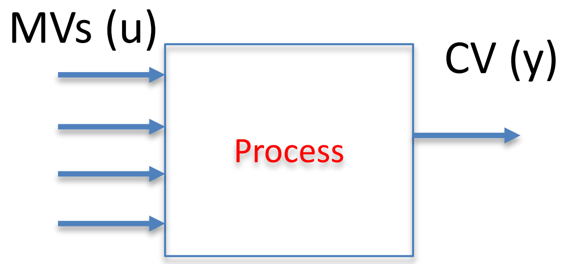


- MV-MV switching
  - Use one MV at a time



- MV-CV switching
  - MV saturates so must give up CV
  1. Simple («do nothing»)
  2. Complex (repairing of loops)

MV-MV switching



For cases with one CV (y) and many inputs (MVs)

- Need several MVs to cover whole steady state range
- Example 1: Not both heating (u_1) and cooling (u_2) to control temperature (y=T)
- Example 3: Need both gas (u_1) and brake (u_2) to control car speed (y)

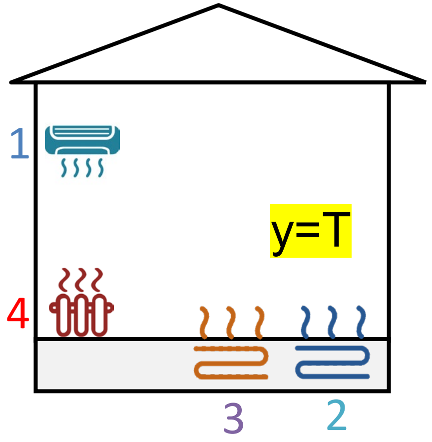
Three alternatives

E5. Split range control

E6. Multiple controllers with different setpoints

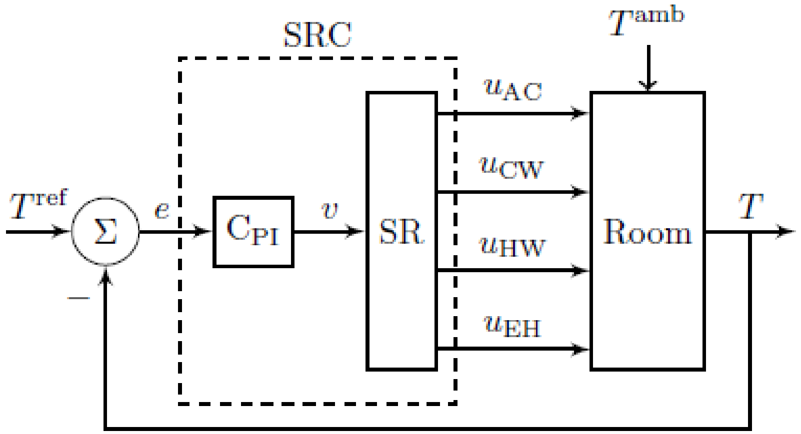
E7. Valve position control

Example split range control (E5) : Room temperature with 4 MVs

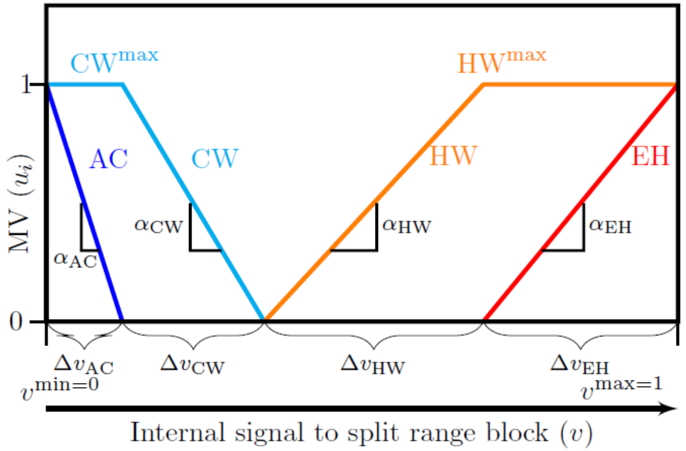


MVs (two for summer and two for winter):

1. AC (expensive cooling)
2. CW (cooling water, cheap)
3. HW (hot water, quite cheap)
4. Electric heat, EH (expensive)

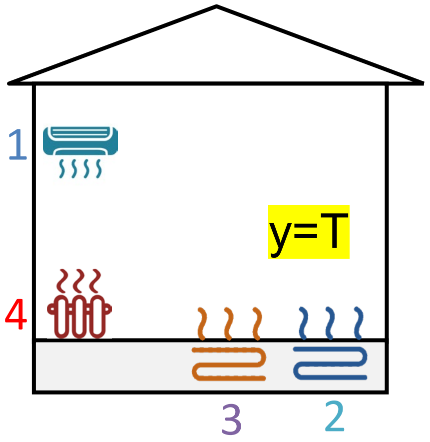


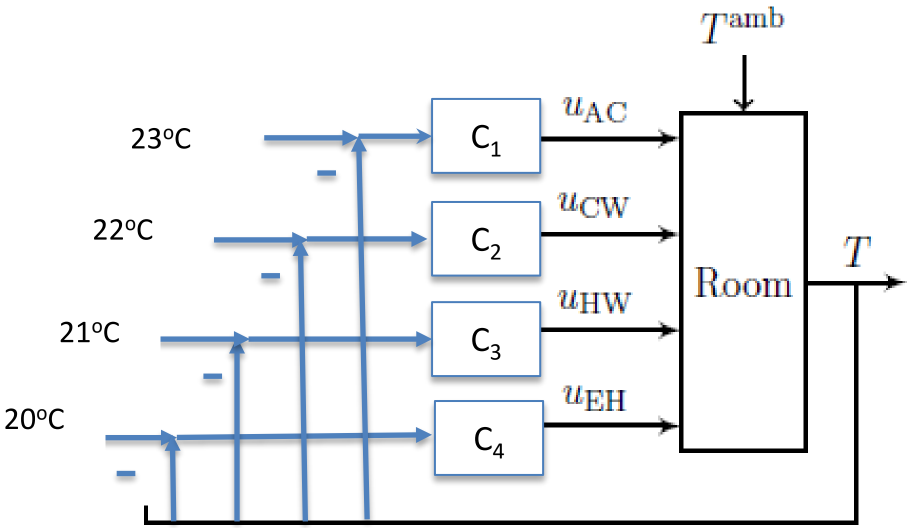
SR-block:



C_{PI} – same controller for all inputs (one integral time)
But get different gains by adjusting slopes α in SR-block

Alternative: Multipliple Controllers with different setpoints (E6)





Disadvantage (comfort):

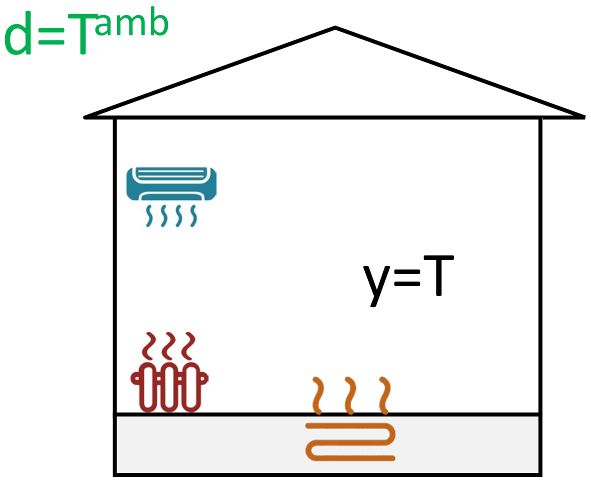
- Different setpoints

Advantage (economics) :

- Different setpoints (energy savings)

Simulation Room temperature

- Dashed lines: SRC (E5)
- Solid lines: Multiple controllers (E6)



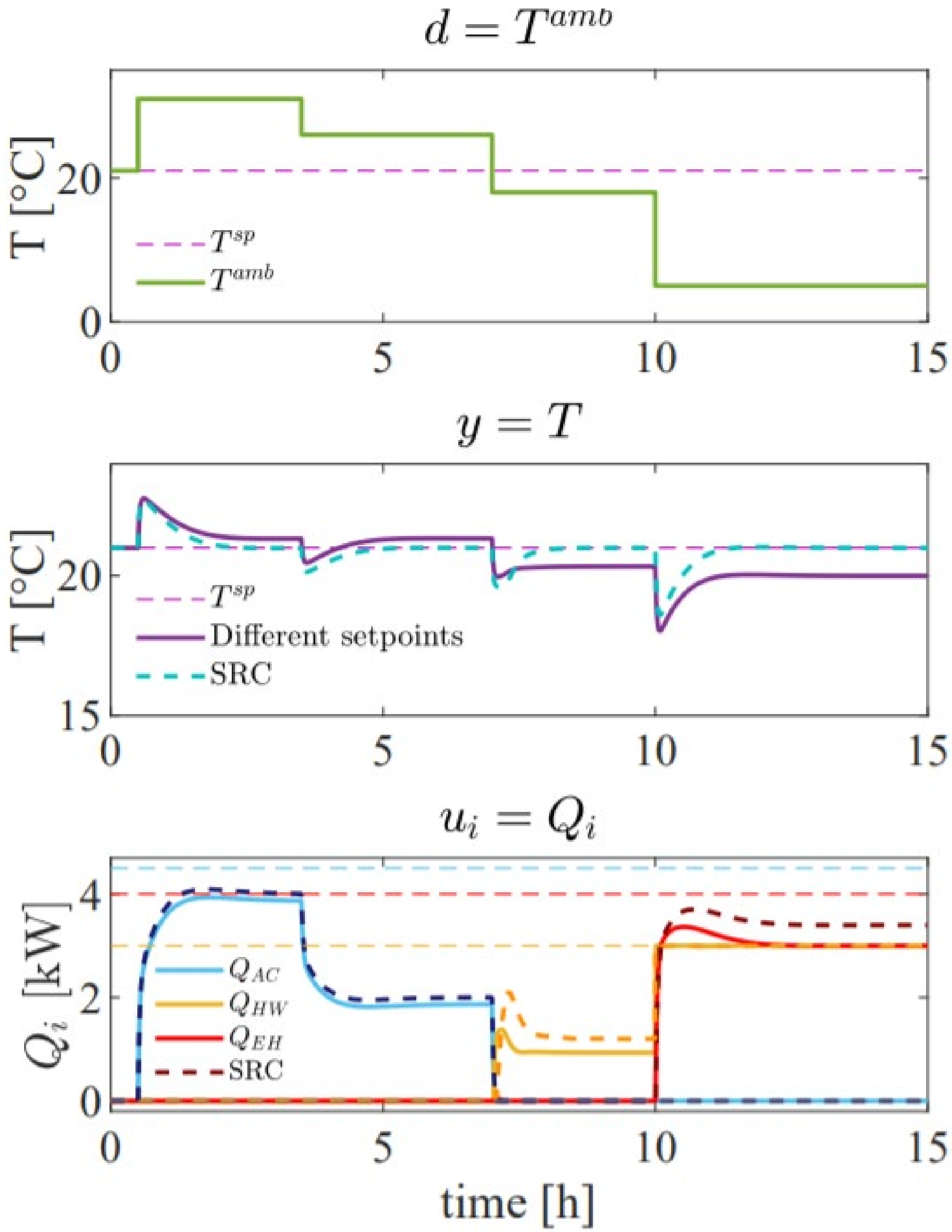


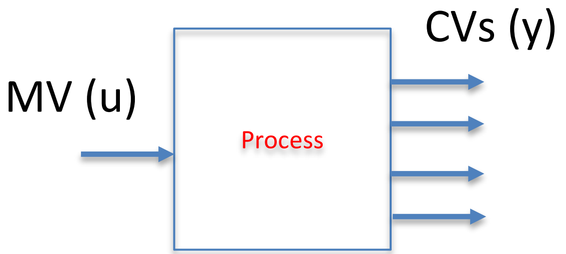
Table 1. Ranges for available inputs (u_k).

Input (u_k)	Description	Nominal	Min	Max	Units
$u_1 = Q_{AC}$	air conditioning	0	0	4.5	kW
$u_2 = Q_{HW}$	heating water	0	0	3.0	kW
$u_3 = Q_{EH}$	electrical heating	0	0	4.0	kW

SRC = split range control

A Reyes-Lua, S Skogestad. Multiple-Input Single-Output Control for Extending the Steady-State Operating Range - Use of Controllers with Different Setpoints. Processes 7 (12), 941 (2019)

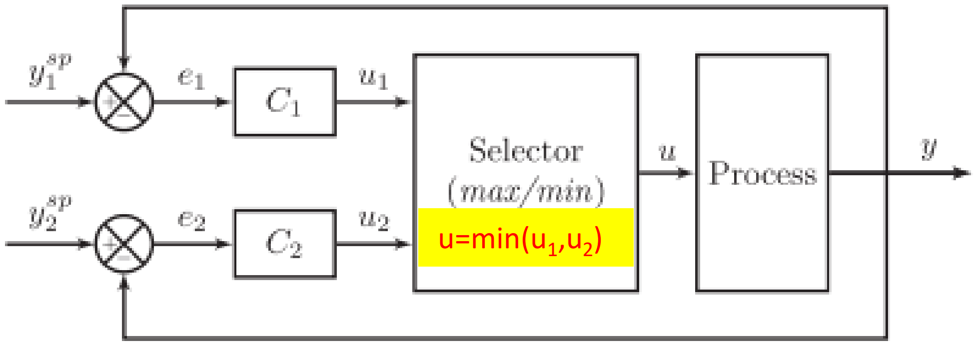
CV-CV switching

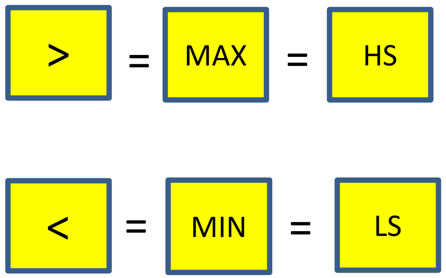


- Only one input (MV) controls many outputs (CVs)
  - Typically caused by change in active constraint
  - Example 1: Control car speed (y_1) - but give up if too small distance (y_2) to car in front.
  - Example 2: Control power (y_1) - but give up if too high engine temperature (y_2).

- Use max- or min-selectors (E4)

E4. Selector: One input (u), several outputs (y_1,y_2)





- Note: The selector is on the input u, even though the setpoint/constraint is on the output y
- Sometimes called “override”
  - OK name for temporary dynamic fix, but otherwise a bit misleading
- Selectors are used for output-output (CV-CV) switching
- Selectors work well, but require pairing each constraint with a given input (not always possible)

Furnace control with safety constraint

Input (MV)

u = Fuel gas flowrate

Output (CV)

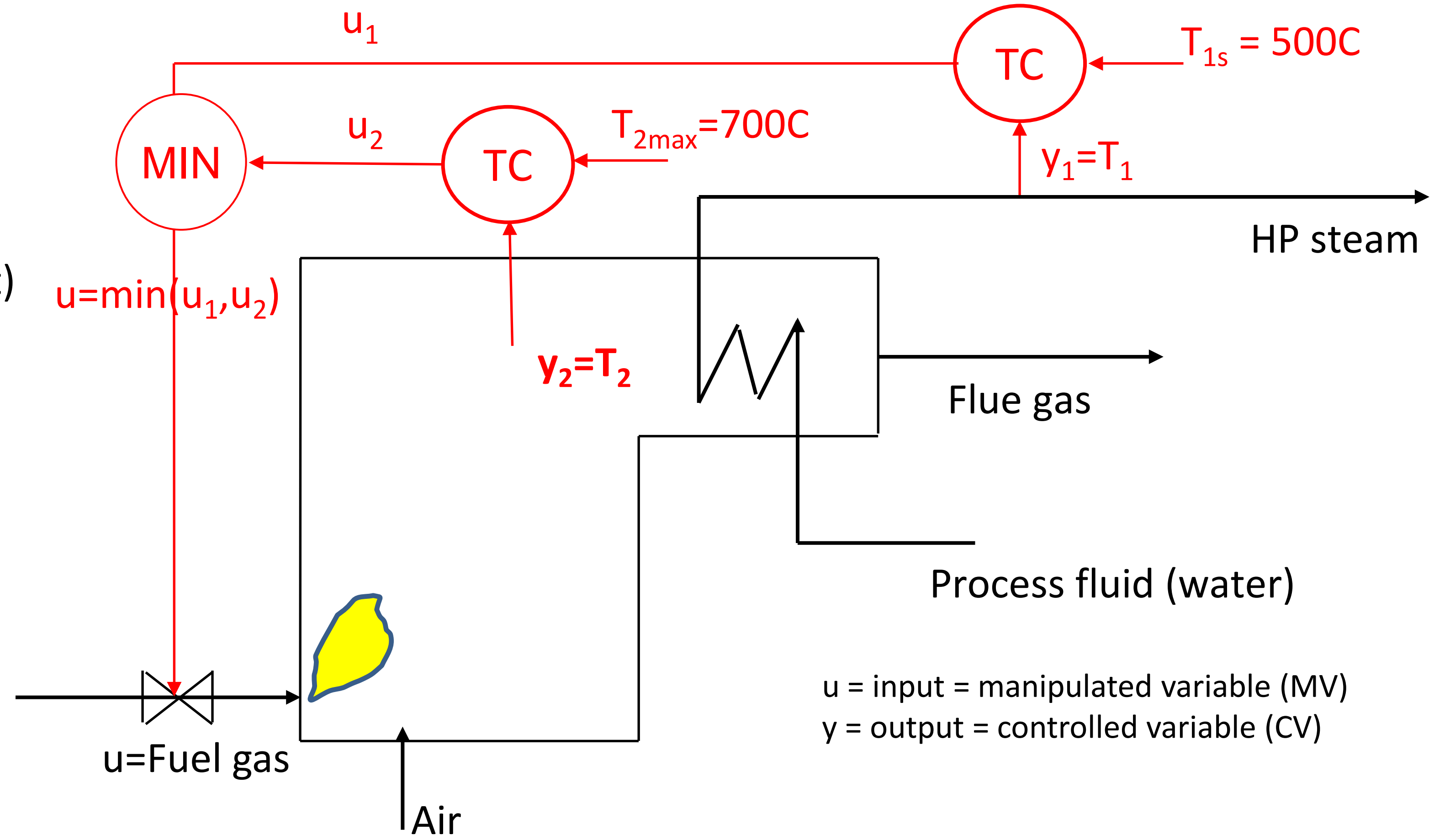
y_1 = process temperature T_1

(desired setpoint or max constraint)

y_2 = furnace temperature T_2

(T2max= 700C)

Rule: Use min-selector for constraints that are satisfied with a small input



u = input = manipulated variable (MV)

y = output = controlled variable (CV)

Design of selector structure

Rule 1 (max or min selector)

- Use max-selector for constraints that are satisfied with a large input
- Use min-selector for constraints that are satisfied with a small input

Rule 2 (order of max and min selectors):

- If need both max and min selector: Potential infeasibility
- Order does not matter if problem is feasible
- If infeasible: Put highest priority constraint at the end

"Systematic design of active constraint switching using selectors."
Dinesh Krishnamoorthy , Sigurd Skogestad. Computers & Chemical Engineering, Volume 143, (2020)

Example. Maximize flow with pressure constraints

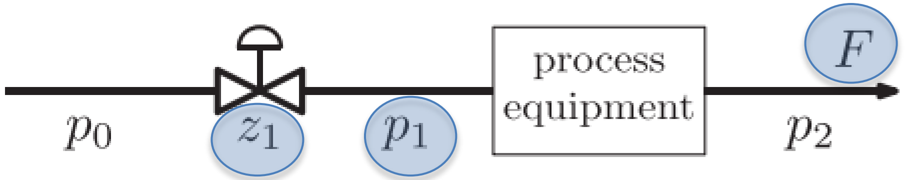


Fig. 6. Example 2: Flow through a pipe with one MV ($u = z_1$).

Input u = z_1
Want to maximize flow, J=-F:

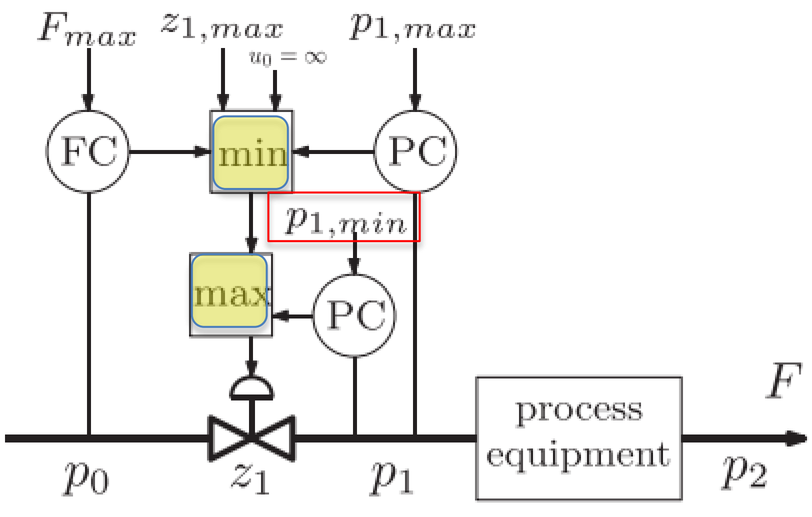
Optimization problem is:

$$\max_{z_1} F$$

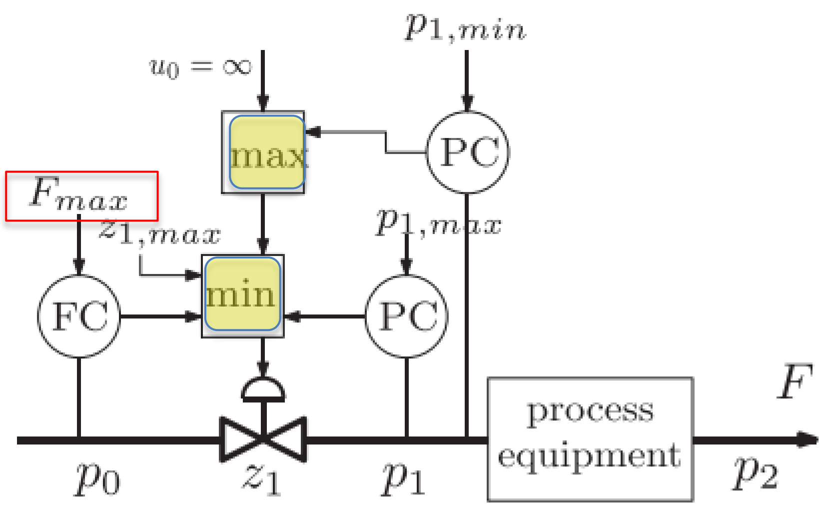
s.t.

$$\boxed{\begin{aligned} &F \leqslant F_{max} \\ &p_1 \leqslant p_{1,max} \\ &p_1 \geqslant p_{1,min} \\ &z_1 \leqslant z_{1,max} \end{aligned}} \qquad (15)$$

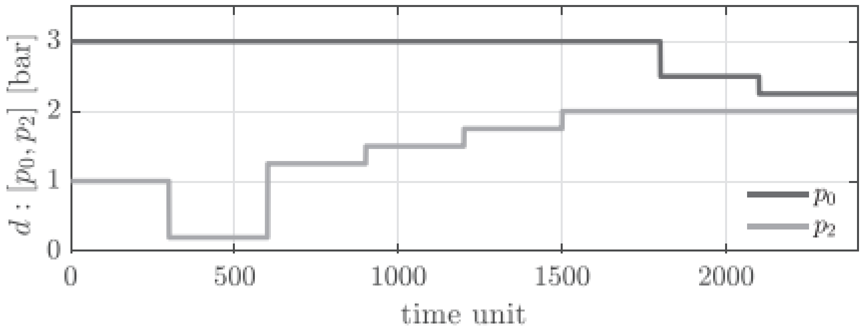
where $F_{max} = 10$ kg/s, $z_{1,max} = 1$, $p_{1,max} = 2.5$ bar, and $p_{1,min} = 1.5$ bar. Note that there are both max and min- constraints on p_1. De-



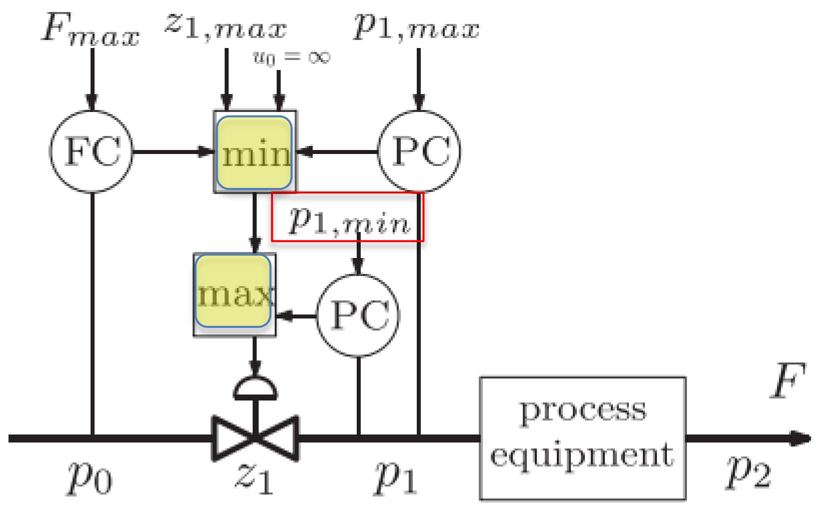
(a)



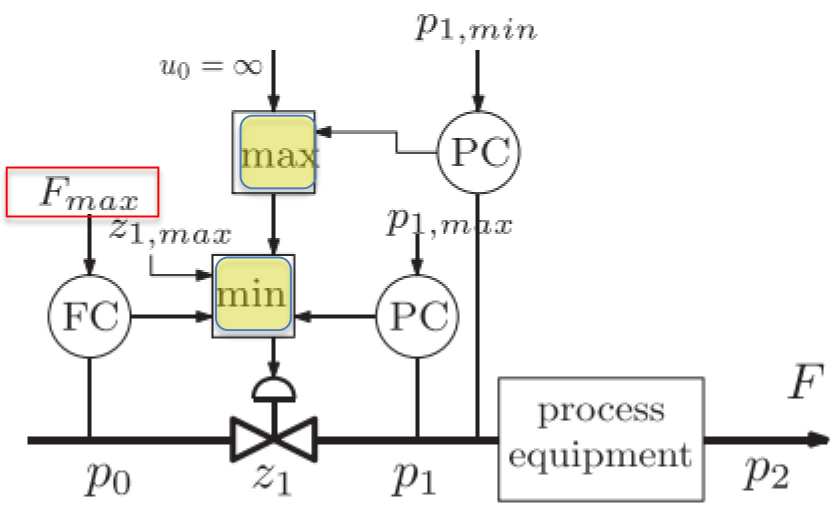
(b)



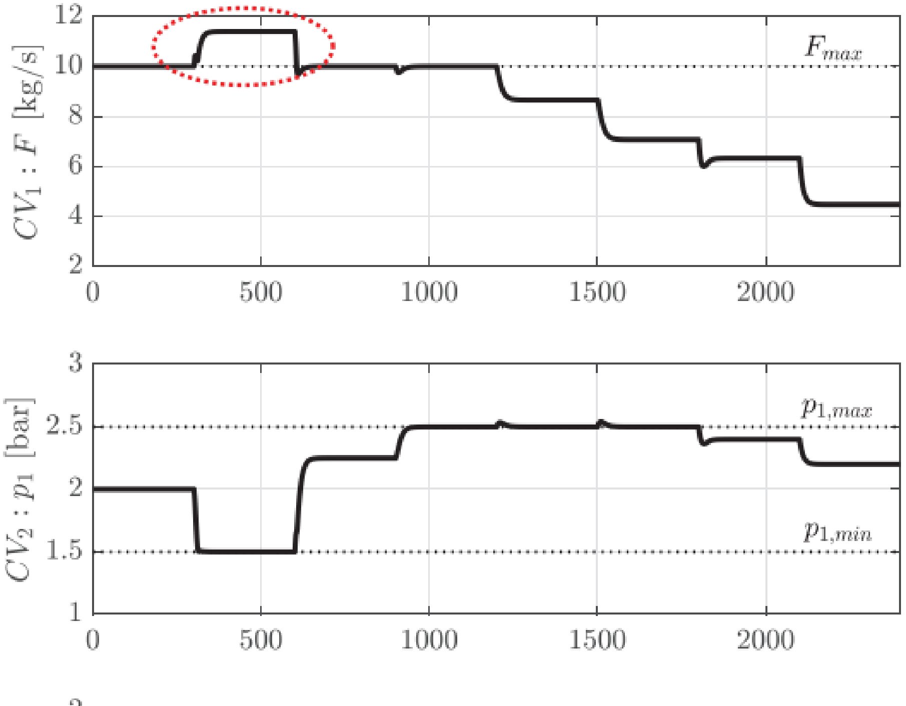
Disturbances in p_0 and p_2 (unmeasured)

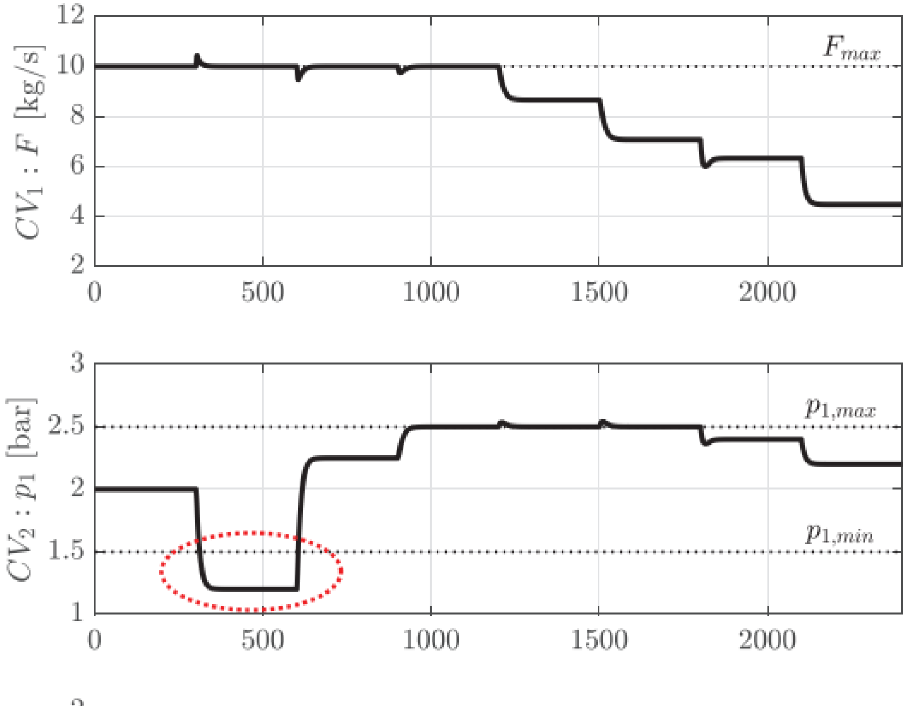


(a)



(b)





t>1800: u=zmax=1

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Example «simple» MV-CV switching (no selector)

Anti-surge control (= min-constraint on F)

Minimize recycle (MV=z) subject to

CV= $F \geq F_{min}$

MV ≥ 0

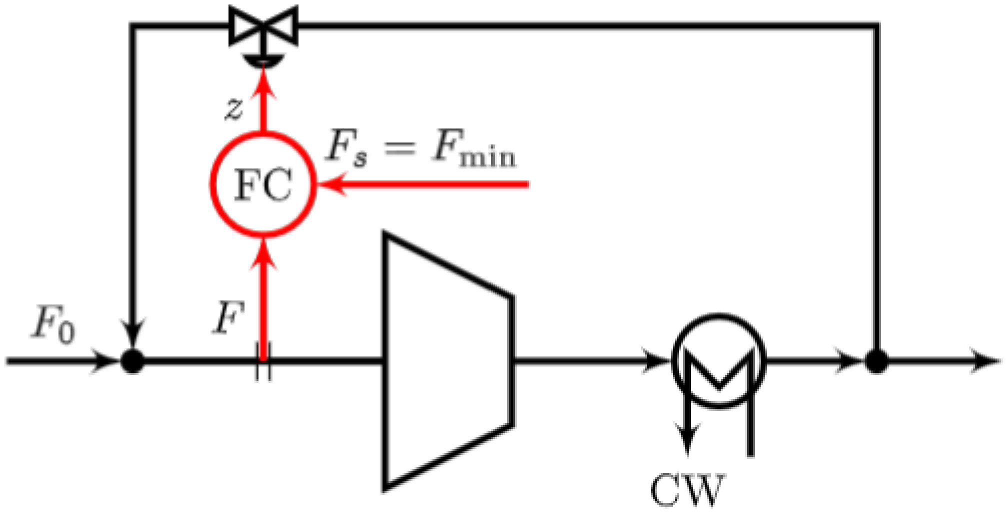
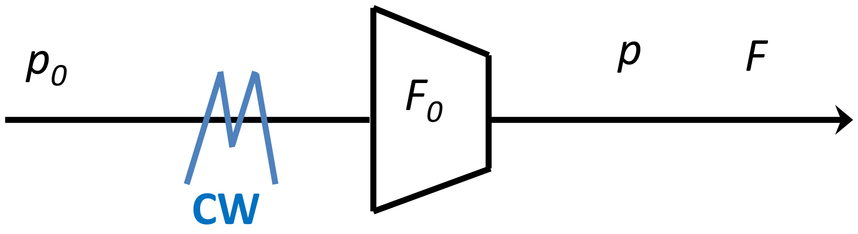
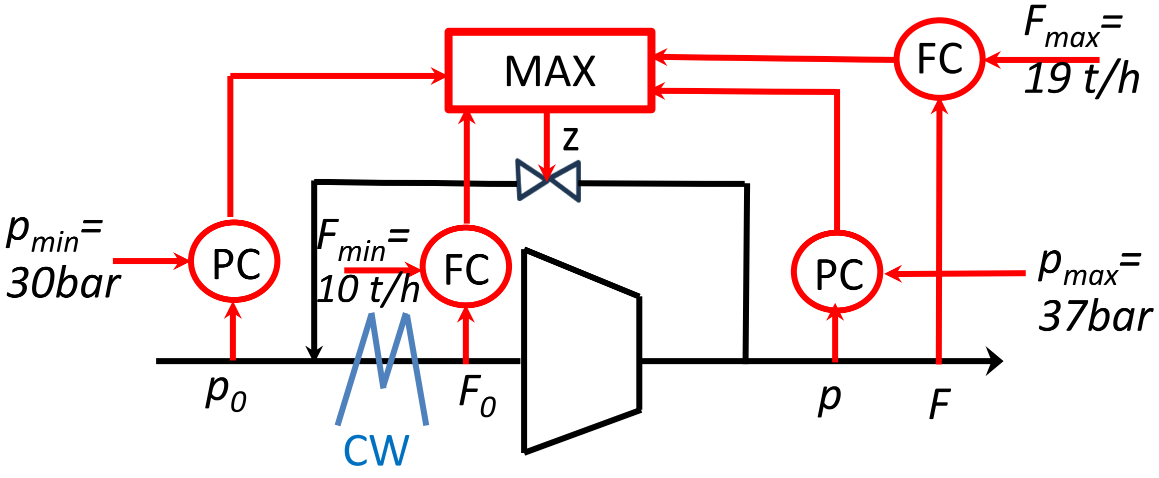


Fig. 32. Flowsheet of anti-surge control of compressor or pump (CW = cooling water). This is an example of simple MV-CV switching: When MV=z (valve position) reaches its minimum constraint ($z = 0$) we can stop controlling CV=F at $F_s = F_{min}$, that is, we do not need to do anything except for adding anti-windup to the controller. Note that the valve has a "built in" max selector.

- No selector required, because MV=z has a «built-in» max-selector at z=0.
- Generally: «Simple» MV-CV switching (with no selector) can be used if we satisfy the input saturation rule: «*Pair a MV that may saturate with a CV that can be given up (when the MV saturates at z=0)"*

QUIZ
Compressor control





Suggest a solution which achieves

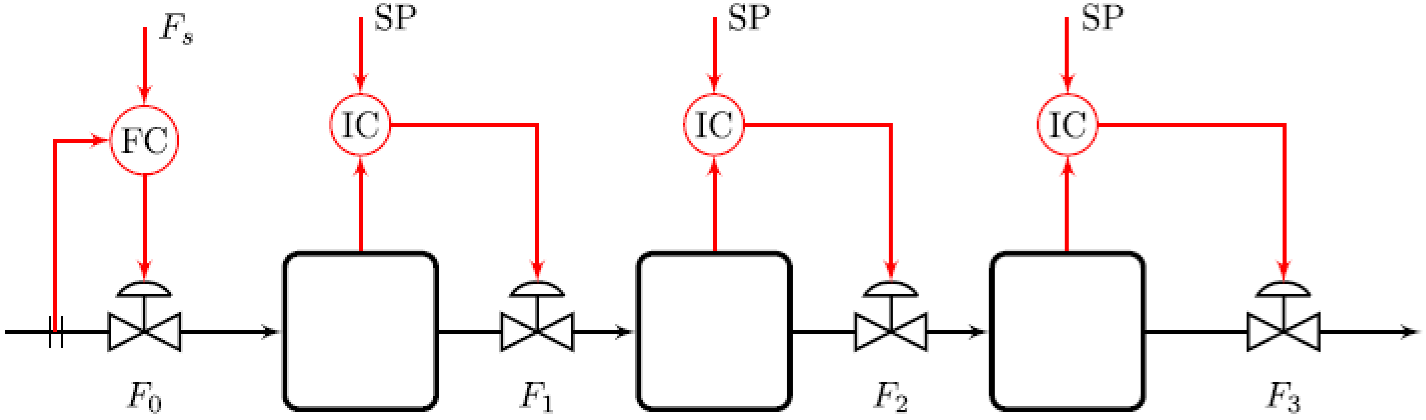
- $p < p_{max} = 37$ *bar* *(max delivery pressure)*
- $P_0 > p_{min} = 30$ *bar* *(min. suction pressure)*
- $F < F_{max} = 19$ *t/h* *(max. production rate)*
- $F_0 > F_{min} = 10$ *t/h* *(min. through compressor to avoid surge)*

Rule CV-CV switching: Use max-selector for constraints that are satisfied by a large input (MV) (here: valve opening z)

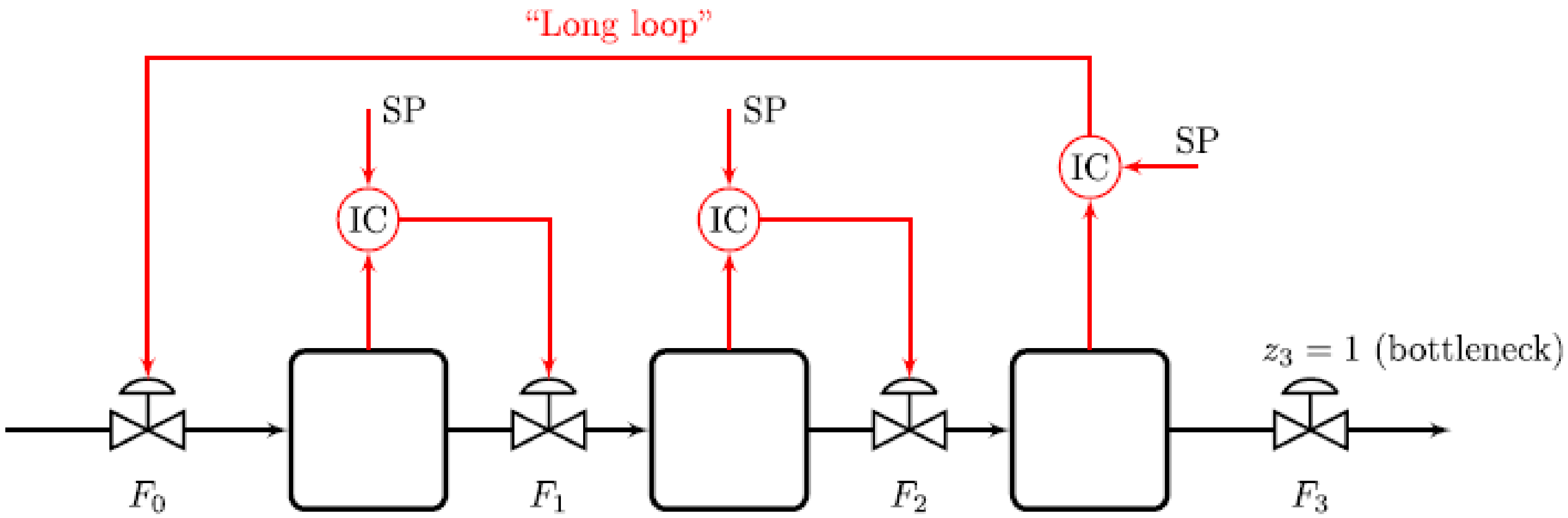
Complex MV-CV switching

- = CV-CV switch followed by MV-MV switch
- Example inventory control: Avoid «long loop» (dynamic issue)

Example: Inventory control



(a) Inventory control in direction of flow (for given feed flow, TPM = F_0)



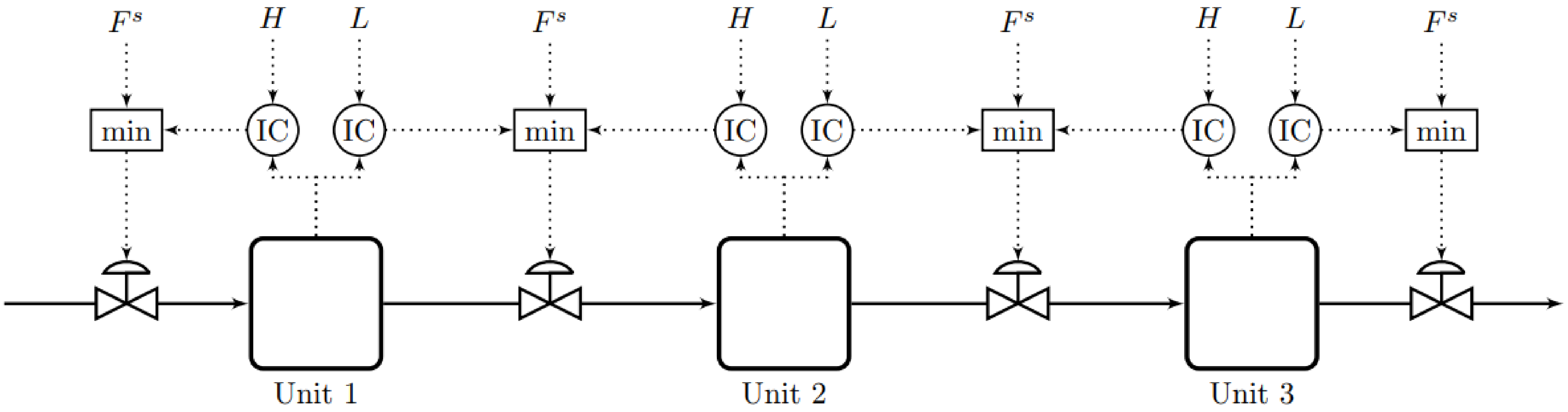
(d) Inventory control with undesired "long loop", not in accordance with the "radiation rule" (for given product flow, TPM= F_3)

TPM = throughput manipulator

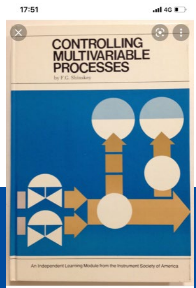
Example . Very smart selector strategy: **Bidirectional inventory control**

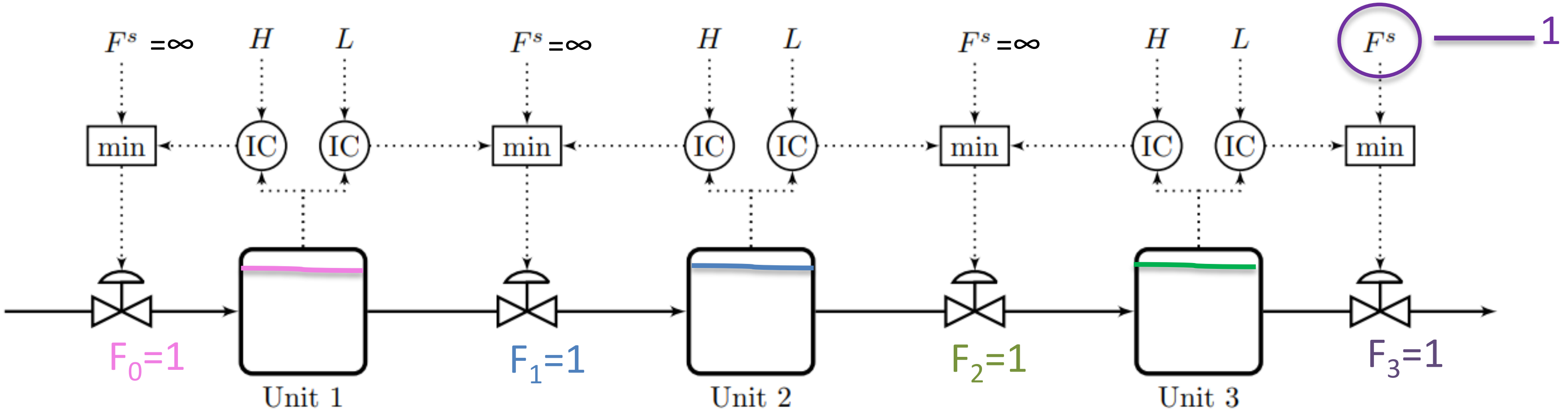
Reconfigures automatically with optimal buffer management!!

Max flow:
$F^s = \infty$



F.G. Shinskey, «Controlling multivariable processes», ISA, 1981
C. Zotica, S. Skogestad and K. Forsman, Comp. Chem. Eng, 2021




$F^s = \infty$
H
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$F^s = \infty$
H
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$F^s = \infty$
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IC
min
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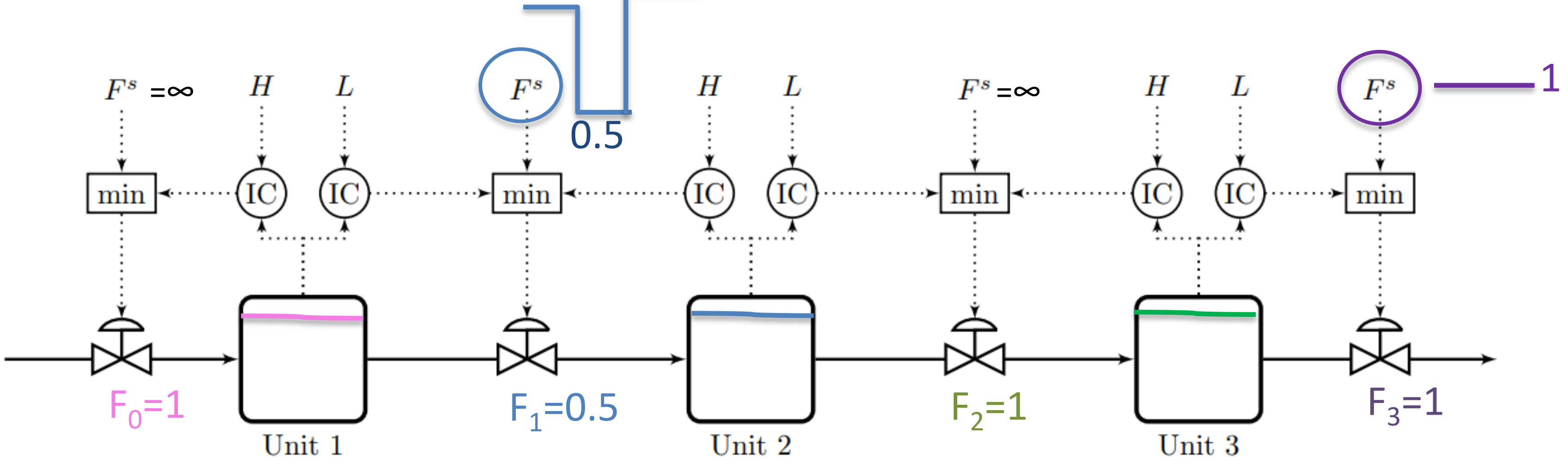
$F_0=1$
Unit 1
$F_1=1$
Unit 2
$F_2=1$
Unit 3
$F_3=1$


$F^s = \infty$
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$F^s = \infty$
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IC
min
IC
IC
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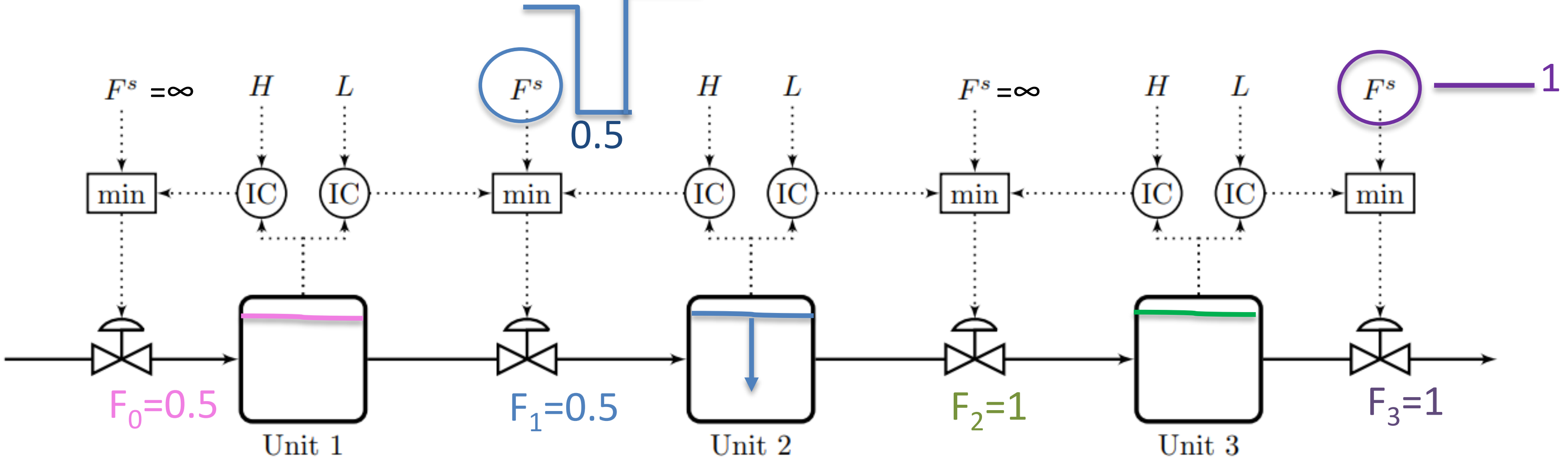
$F_0=1$
Unit 1
$F_1=0.5$
Unit 2
$F_2=1$
Unit 3
$F_3=1$

NTNU



$F^s = \infty$
H
L
F^s
0.5
H
L
$F^s = \infty$
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F^s
1
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min
IC
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IC
IC
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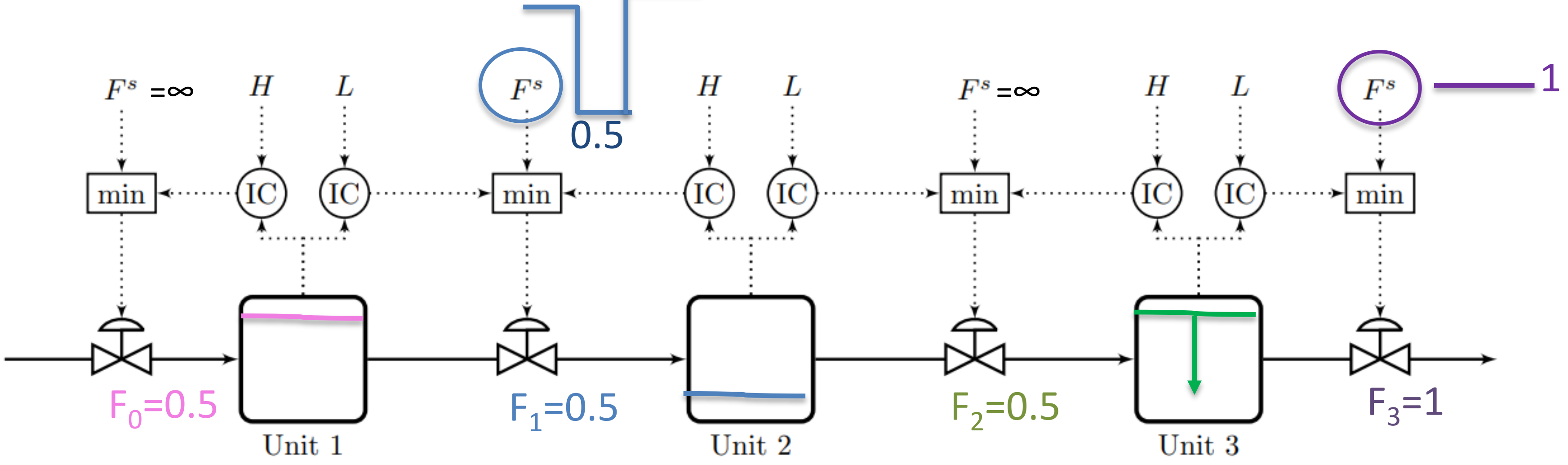
$F_0=0.5$
Unit 1
$F_1=0.5$
Unit 2
$F_2=1$
Unit 3
$F_3=1$

NTNU



$F^s = \infty$
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0.5
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$F^s = \infty$
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F^s
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min
$F_0=0.5$
Unit 1
$F_1=0.5$
Unit 2
$F_2=0.5$
Unit 3
$F_3=1$

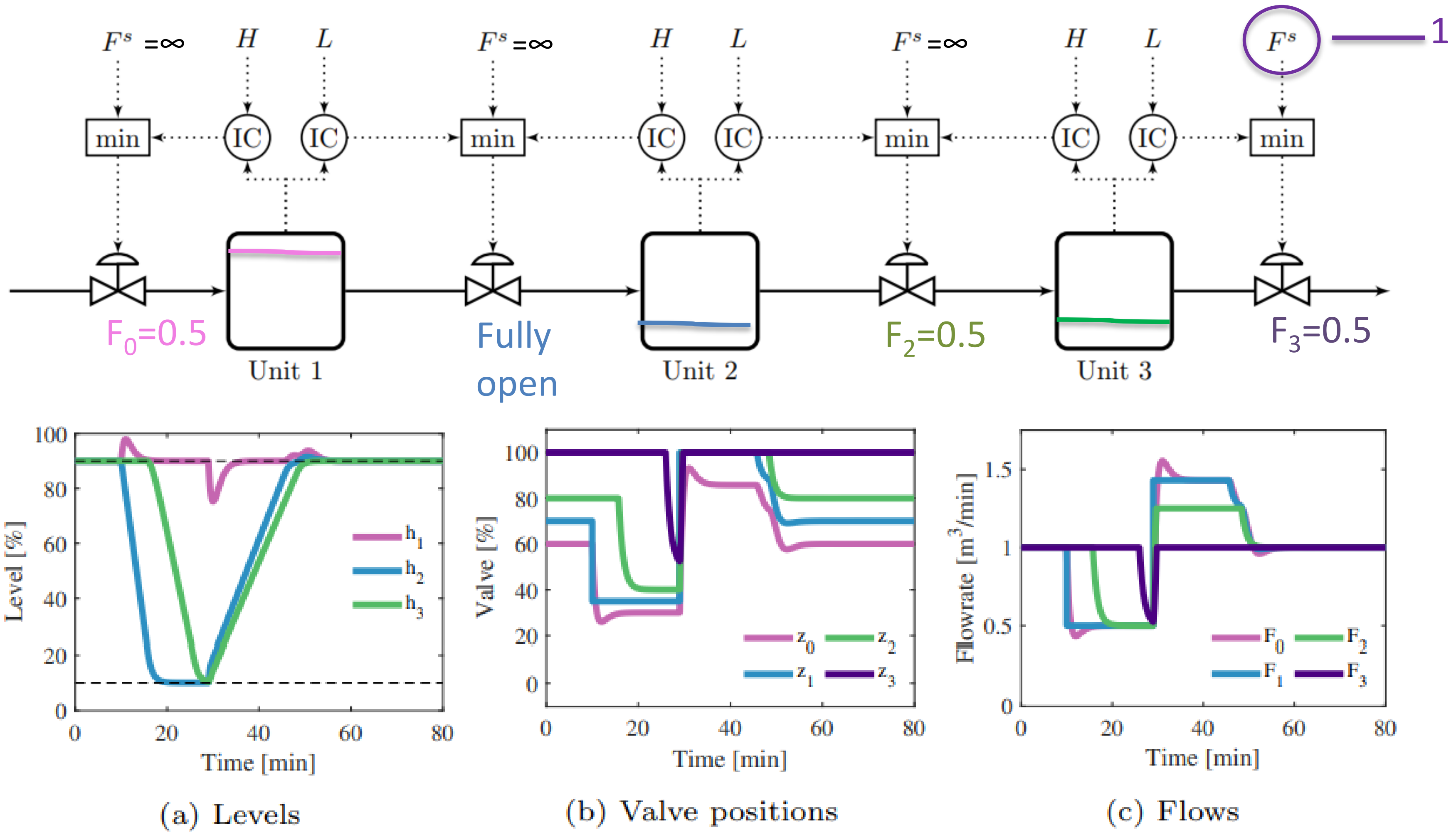
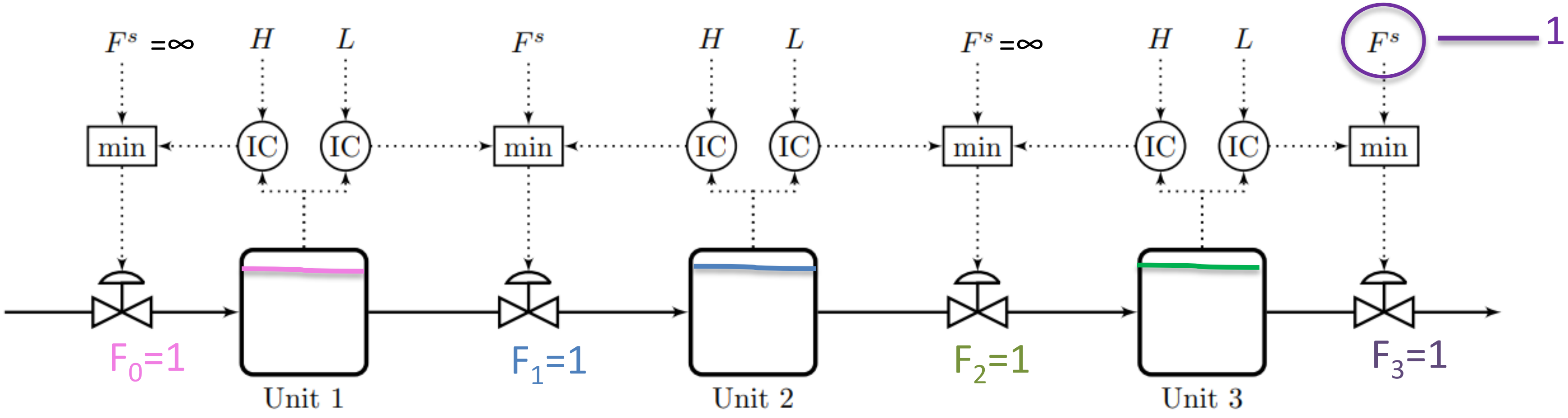


Figure 12: Simulation of a 19 min temporary bottleneck in flow F_1 for the control structures in Fig. 3d with the TPM downstream of the bottleneck.



Challenge: Can MPC be made to do his? Optimally reconfigure loops and find optimal buffer?

- Yes, possible with standard setpoint-based MPC if we use
  - Trick: All flow setpoints = infinity (unachoevable setpoint)
- What about Economic MPC? Cannot do it easily; may try scenario-MPC

Example adaptive cruise control: CV-CV switch followed by MV-MV switch

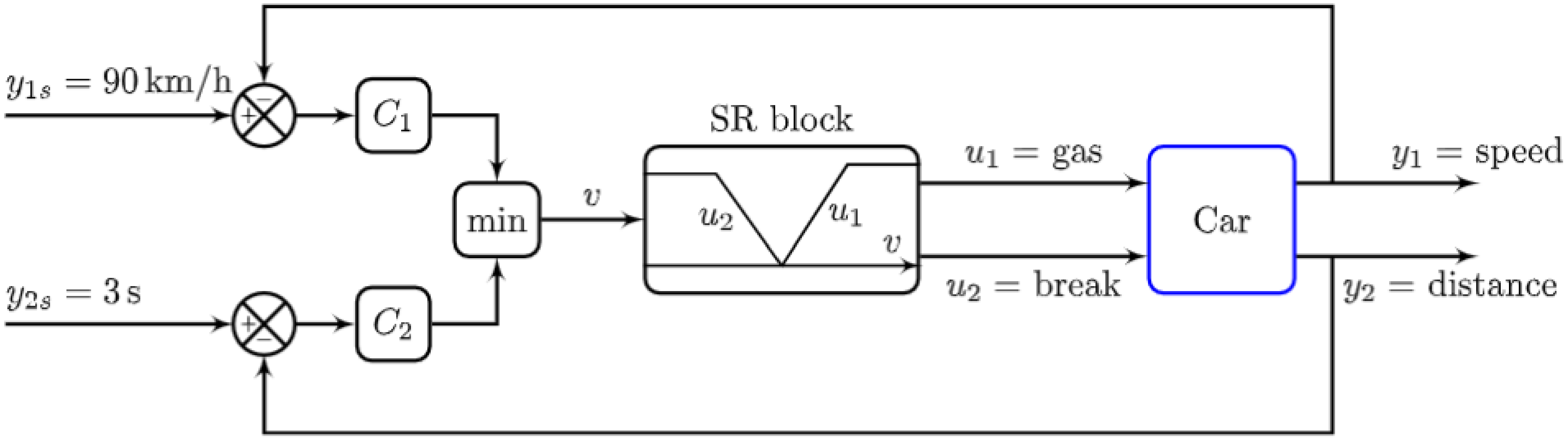


Fig. 31. Adaptive cruise control with selector and split range control.

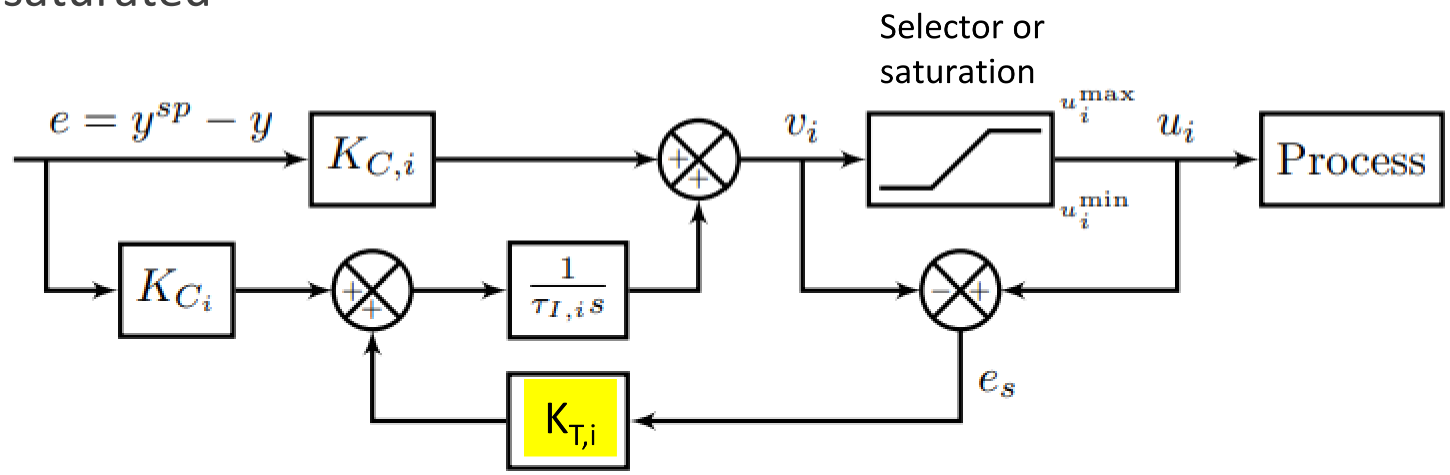
Note: This is not Complex MV-CV switching, because then the order would be opposite.

Important insight

- Many problems: Optimal steady-state solution always at constraints
- In this case optimization layer may not be needed
  - if we can identify the active constraints and control them using selectors

E8. Anti-windup

- All the controllers shown need anti-windup to «stop integration» during periods when the control action (v_i) is not affecting the process:
  - Controller is disconnected (because of selector)
  - Physical MV u_i is saturated



Anti-windup using back-calculation. Typical choice for tracking constant, K_T=1

Challenges selector design

- Standard approach requires pairing of each active constraint with a single input
  - May not be possible in complex cases
- Stability analysis of switched systems is still an open problem
  - Undesired switching may be avoided in many ways:
    - Filtering of measurement
    - Tuning of anti-windup scheme
    - Minimum time between switching
    - Minimum input change

When use MPC?

When conventional APC performs poorly or becomes complex

- Cases with many changing constraints (where we cannot assign one input to each constraint)
- Interactive process
- Know future disturbances and setpoint changes (predictive capability)

Conclusion Advanced process control (APC)

- Classical APC, aka «Advanced regulatory control» (ARC) or «Advanced PID»:
  - Works very well in many cases
  - Optimization by feedback (active constraint switching)
  - Need to pair input and output.
    - Advantage: The engineer can specify directly the solution
    - Problem: Unique pairing may not be possible for complex cases
  - Need model only for parts of the process (for tuning)
  - Challenge: Need better teaching and design methods

- MPC may be better (and simpler) for more complex multivariable cases
  - But MPC may not work on all problems (Bidirectional inventorycontrol)
  - Main challenge: Need dynamic model for whole process
  - Other challenge: Tuning may be difficult

Academic process control community fish pond

Optimal centralized Solution (**EMPC**)

Simple solutions that work (**ARC** = PID++)