PID is the future of Advanced Control

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Norwegian University of Science and Technology (NTNU)
Trondheim

IFAC World Congress Workshop
Busan, Korea
PID Workshop 23 Aug 2026

Abstract

PID is the future of advanced control

This talk explains why “advanced regulatory control” (ARC), where the PID controller is a central component, will often be the preferred solution for future advanced control. Keywords are simplicity and flexibility. The main competitor to ARC is MPC (model predictive control), but the costs of implementing and maintaining MPC solutions are high, especially for process control. Moreover, in most cases ARC solutions (including cascade, ratio, split range and selector control) are more flexible and easier to tune. A main problem right now is that the knowledge and competence about ARC strategies is low, especially in academia, but also in industry the knowledge is dying out. The result is that people may 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 in the process industry for 50 years, and yet the use of MPC is far from increasing as expected. The latest belief is that, if MPC is too complex, then machine learning is the solution. No, it is not, because of 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 in process control mainly as an improvement for large-scale applications that can afford the effort. The talk will emphasize the above points and in addition present systematic approaches to ARC methods.

Intro….

- 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

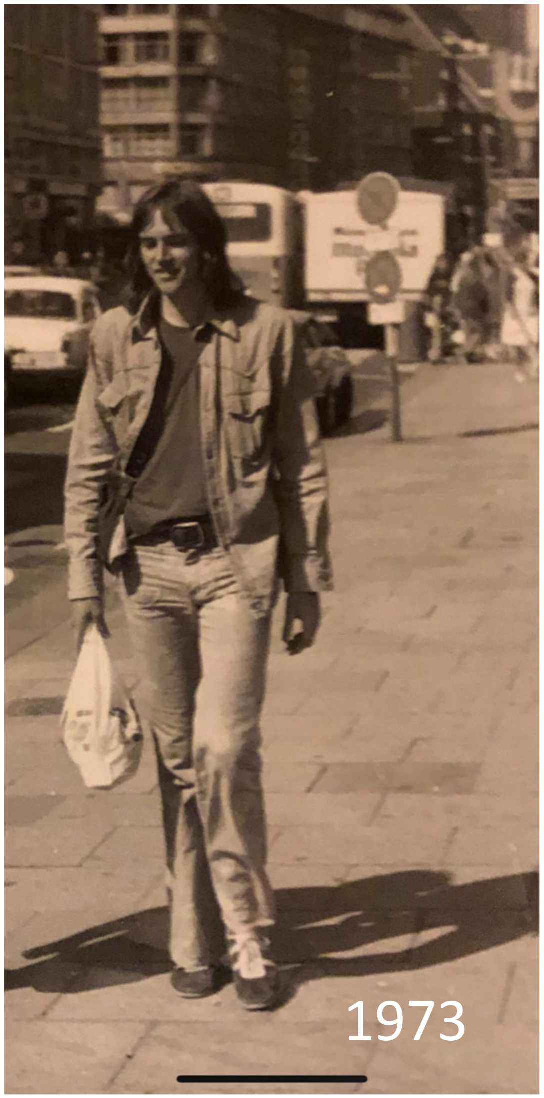

Norwegian Sea
Trondheim
Arctic circle
Faroe Islands
Norway
Sweden
Oslo
Stockholm
North Sea
Edinburgh
United Kingdom
Isle of Man
Denmark
Copenhagen
Dublin
Ireland
Manchester
Liverpool
Hamburg
London
Amsterdam
Netherlands
Berlin
Poland
Brussels
Cologne
Germany
Belgium
Frankfurt
Luxembourg
Prague
Czechia

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**



NTNU

“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"

"News"...

- 27 Nov. 2023: Welcome to the SUBPRO Symposium at the Britannia Hotel in Trondheim
- Aug. 2023: Tutorial review paper on "Advanced control using decomposition and simple elements". Published in Annual reviews in Control (2023). [paper] [tutorial workshop] [slides from Advanced process control course at NTNU]
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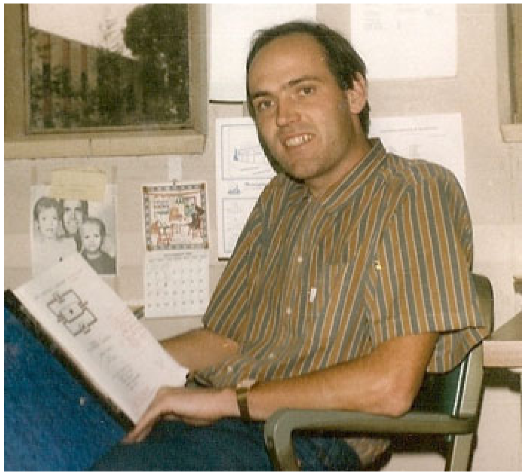
- 08 Dec. 2021: Plenary talk on "Nonlinear input transformations for disturbance rejection, decoupling and linearization" at Control Conference of Africa (CCA 2021), Magaliesburg, South Africa (virtual) [video and slides]
- 27 Oct. 2021: Plenary talk on "Advanced process control - A newe look at the 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 vidoes 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.



Sigurd at Caltech (1984)

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

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

Control system structure*

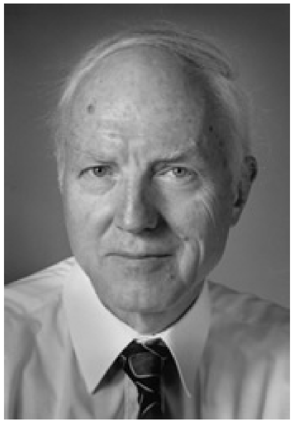
Alan Foss ("Critique of chemical process control theory", AIChE Journal,1973):

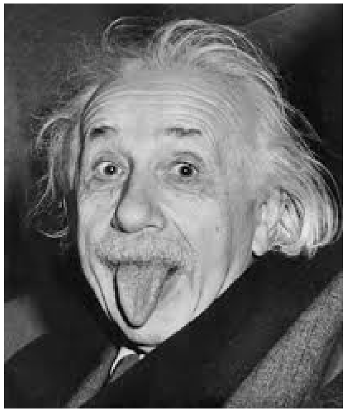
The central issue to be resolved ... is the determination of control system structure. **Which variables should be measured, which inputs should be manipulated and which links should be made between the two sets?***

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 [between theory and practice] is present indeed, but contrary to the views of many, it is the theoretician who must close it.

*Current terminology: Control system architecture





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

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Start here...

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Review article

Advanced control using decomposition and simple elements

Sigurd Skogestad

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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 enhancing the teaching of ARC methods and prioritize research efforts in developing theory and improving design method.

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Main objectives of a control system

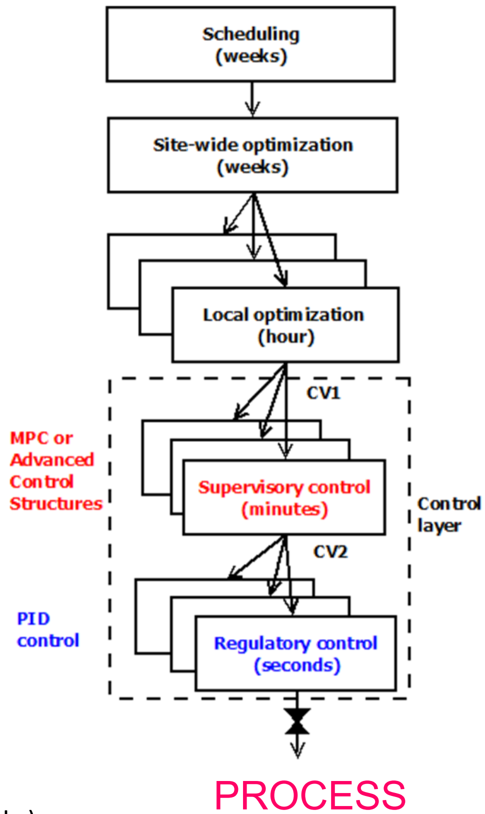
1. Economics: Implementation of acceptable (near-optimal) operation
2. Regulation: Stable operation

ARE THESE OBJECTIVES CONFLICTING?

- Usually NOT
  - Different time scales
    - Stabilization → fast time scale
  - Stabilization doesn't "use up" any degrees of freedom
    - Reference value (setpoint) available for layer above
    - But it "uses up" part of the time window (frequency range)
      - Time scale separation

Two fundamental ways of decomposing the controller

- Vertical (hierarchical; cascade)
- Based on time scale separation
- Decision: Selection of CVs that connect layers



- Horizontal (decentralized)
- Usually based on distance
- Decision: Pairing of MVs and CVs within layers

In addition: Decomposition of controller into smaller elements (blocks):
Feedforward element, nonlinear element, estimators (soft sensors), switching elements

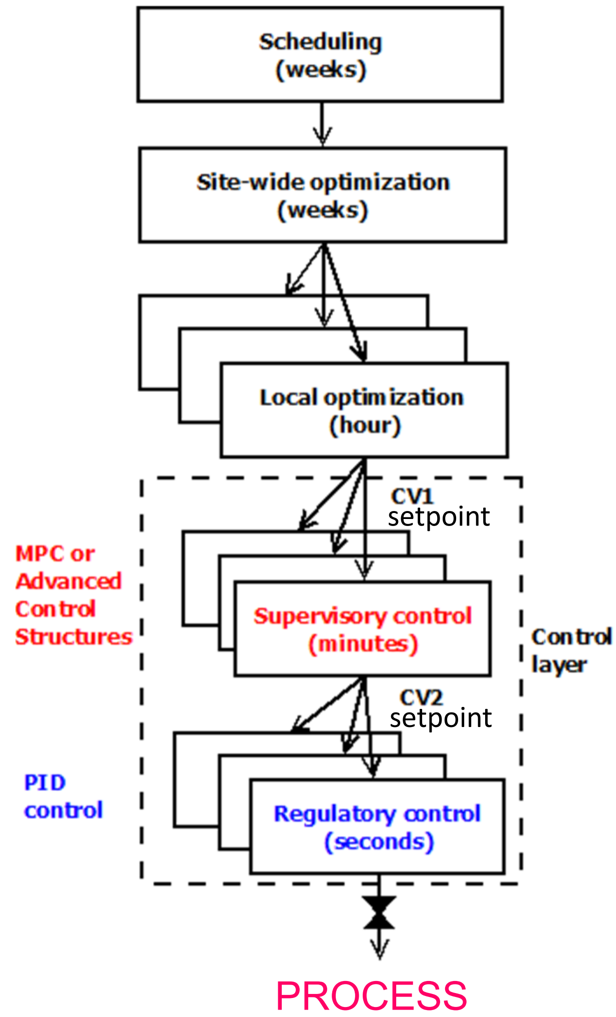
Control* layers

- **Supervisory ("advanced") control layer**

  Tasks:
  - Follow set points for CV1 from economic optimization layer
  - Switch between active constraints (change CV1)
  - Look after regulatory layer (avoid that MVs saturate, etc.)

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

*My definition of «control» is that the objective is to track setpoints



CV = controlled variable

Combine control and optimization into one layer?

EMPC: Economic model predictive “control”’

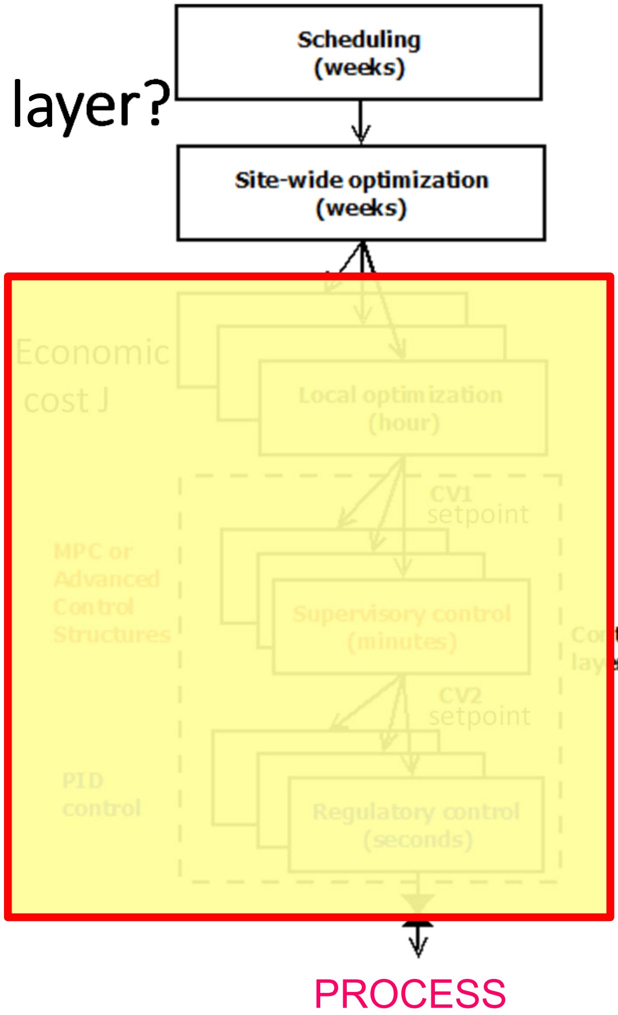
$J_{EMPC} = J + J_{control}$

- Economic cost: J [$/s] = cost feed + cost energy – value products
- Penalize input usage, $J_{control} = \Sigma \Delta u_i^2$

NO, combining layers is generally not a good idea!
(the good idea is to separate them!)

One layer (EMPC) is optimal theoreretically, but

- Need detailed dynamic model of everything
- Tuning difficult and indirect
- Slow! (or at least difficult to speed up parts of the control)
- Robustness poor
- Implementation and maintainance costly and time consuming



EMPC

(no setpoints, CV1, CV2)

CV = controlled variable
RTO = real-time optimization

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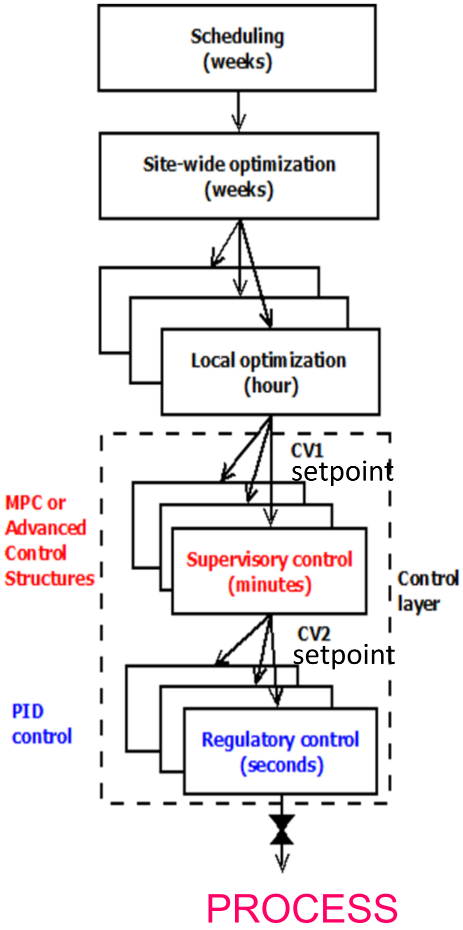
Alternative solutions for advanced process control

- **Model Predictive Control (MPC)**
  - Ye, it's OK
  - BUT: Has been around for more than 50 years (since 1970s) - the expected growth never came.
  - Process industry: MPC is still used mostly in large-scale plants (petrochemical and refineries).
  - Process industry: MPC is far from replacing PID as some expected in the 1990s.

- **Machine learning?**
  - Requires a lot of data, not realistic for process control
  - And: Can only be implemented after the process has been in operation

- **"Classical advanced regulatory control" (ARC) based on single-loop PIDs?**
  - YES!
  - Extensively used by industry
  - Feedback solutions that can be implemented with minimum need for models
  - 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 control

PID control

Excellent work – especially considering that it was published only 3 years after the PID controller came on the market (Taylor Model 100 Fullscope, 1939)

Trans. ASME, 64, 759-768 (Nov. 1942).

Optimum Settings for Automatic Controllers

By J.G. ZIEGLER[1] and N. B. NICHOLS[2] • ROCHESTER, N. Y.

In this paper, the three principle control effects found in present controllers are examined and practical names and units of measurement are proposed for each effect.

varying its output air pressure, repositions a diaphragm-operated valve. The controller may be measuring temperature, pressure, level, or any other variable, but we will completely divorce the

Ziegler-Nichols «open-loop» method (give similar tunings as «closed-loop» method)

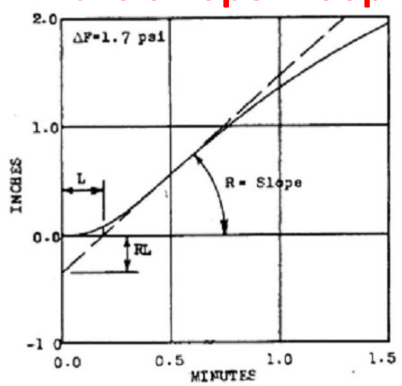


FIG. 8 REACTION CURVE

Reset-Rate Determination From Reaction Curve. Since the period of oscillation at the ultimate sensitivity proves to be 4 times the lag. A substitution of 4 L for P_u in previous equations for optimum reset rate gives an equation expressing this reset rate in terms of lag. For a controller with proportional and auto-matic-reset responses, the optimum settings become

$$Sensitivity = \frac{0.9}{R_1 L} \text{ psi per in.}$$

$$\text{Reset Rate} = \frac{0.3}{L} \text{ per min}$$

At these settings the period will be about 5.7L, having been increased, by both the lowering of sensitivity and the addition of automatic reset.

Very similar to SIMC for integrating process with τ_c=0 (aggressive!):

$0.9 \rightarrow 1$

$0.3 \rightarrow \frac{1}{4} = 0.25$

Disadvantages Ziegler-Nichols:

1. Aggressive settings
2. No tuning parameter
3. Poor for processes with large time delay (θ)

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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.

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

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dead time are derived
ustness is demonstrated.

258 Ind. Eng. Chem.

Table I. IMC-Based PID Controller Parameters[a]

	model	$y/y_s = \tilde{g}_+ f$	controller	$k_c k$	τ_I	τ_D	τ_F	comments
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	$\frac{k}{\tau^2 s^2+2\zeta\tau s+1}$	$\frac{1}{\epsilon s+1}$	$\frac{\tau^2 s^2+2\zeta\tau s+1}{k\epsilon s}$	$\frac{2\zeta\tau}{\epsilon}$				
D	$k\frac{-\beta s+1}{\tau s+1}$	$\frac{-\beta s+1}{\epsilon s+1}$	$\frac{\tau s+1}{k(\beta+\epsilon)s}$	$\frac{\tau}{\beta+\epsilon}$				
E	$k\frac{-\beta s+1}{\tau s+1}$	$\frac{-\beta s+1}{(\beta s+1)(\epsilon s+1)}$	$\frac{\tau s+1}{ks(\beta\epsilon s+2\beta+\epsilon)}$	$\frac{\tau}{2\beta+\epsilon}$				
F	$k\frac{-\beta s+1}{\tau^2 s^2+2\zeta\tau+1}$	$\frac{-\beta s+1}{\epsilon s+1}$	$\frac{\tau^2 s^2+2\zeta\tau s+1}{k(\beta+\epsilon)s}$	$\frac{2\zeta\tau}{\beta+\epsilon}$				
G	$k\frac{-\beta s+1}{\tau^2 s^2+2\zeta\tau s+1}$	$\frac{-\beta s+1}{(\beta s+1)(\epsilon s+1)}$	$\frac{\tau^2 s^2+2\zeta\tau s+1}{k(\beta\epsilon s+2\beta+\epsilon)s}$	$\frac{2\zeta\tau}{2\beta+\epsilon}$				
H	$\frac{k}{s}$	$\frac{1}{\epsilon s+1}$	$\frac{1}{k\epsilon}$	$\frac{1}{\epsilon}$				
I	$\frac{k}{s}$	$\frac{2\epsilon+1}{(\epsilon s+1)^2}$	$\frac{2\epsilon s+1}{k\epsilon^2 s}$	$\frac{2}{\epsilon}$				
J	$\frac{k}{s(\tau s+1)}$	$\frac{1}{\epsilon s+1}$	$\frac{\tau s+1}{k\epsilon}$	$\frac{1}{\epsilon}$	–	τ	–	–
K	k	$2\epsilon s+1$	$(\tau s+1)(2\epsilon s+1)$	$2\epsilon+\tau$	$2\epsilon+\tau$	$2\epsilon\tau$	–	(6)

Table II. IMC-Based PID Parameters for $g(s) = ke^{-\theta s}/(\tau s + 1)$ and Practical Recommendations for ϵ/θ

controller	kk_c	τ_I	τ_D	recommended ϵ/θ (> $0.1\tau/\theta$ *always*)
PID	$(2\tau + \theta)/(2\epsilon + \theta)$	$\tau + (\theta/2)$	$\tau\theta/(2\tau + \theta)$	>0.8
PI	$\theta/\tau = 0.1$	1.54		>1.7
improved PI	$(2\tau + \theta)/2\epsilon$	$\tau + (\theta/2)$		>1.7

Disadvantage IMC-PID (=Lambda tuning):
1. Many rules
2. Poor disturbance response for «slow» processes (with large τ_1/θ)

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Probably the best **simple** PID tuning rules in the world

Sigurd Skogestad*
Department of Chemical Engineering
Norwegian University of Science and Technology
N–7491 Trondheim Norway

Presented at AIChE Annual meeting, Reno, NV, USA, 04-09 Nov. 2001
Paper no. 276h ©Author
This version: November 19, 2001†

Abstract

The aim of this paper is to present analytic tuning rules which are as simple as possible and still result in a good closed-loop behavior. The starting point has been the IMC PID tuning rules of Rivera, Morari and Skogestad (1986) which have achieved widespread industrial acceptance. The integral term has been modified to improve disturbance rejection for integrating processes. Furthermore, rather than deriving separate rules for each transfer function model, we start by approximating the process by a first-order plus delay processes (using the "half method"), and then use a single tuning rule. This is much simpler and appears to give controller tunings with comparable performance. All the tunings are derived analytically and are thus very suitable for teaching.

1 Introduction

Hundreds, if not thousands, of papers have been written on tuning of PID controllers, and one must question the need for another one. The first justification is that PID controller is by far the most widely used control algorithm in the process industry, and that improvements in tuning of PID controllers will have a significant practical impact. The second justification is that the simple rules and insights presented in this paper may contribute to a significantly improved understanding into how the controller should be tuned.

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Journal of Process Control 13 (2003) 291–309

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Simple analytic rules for model reduction and PID controller tuning☆

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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.

Keywords: Process control; Feedback control; IMC; PI-control; Integrating process; Time delay

1. Introduction

Although the proportional-integral-derivative (PID) controller has only three parameters, it is not easy, without a systematic procedure, to find good values (settings) for them. In fact, a visit to a process plant will usually show that a large number of the PID controllers are poorly tuned. The tuning rules presented in this paper have developed mainly as a result of teaching this material, where there are several objectives:

1. The tuning rules should be well motivated, and preferably model-based and analytically derived.
2. They should be simple and easy to memorize.
3. They should work well on a wide range of processes.

Step 2. Derive model-based controller settings. PI-settings result if we start from a first-order model, whereas PID-settings result from a second-order model.

There has been previous work along these lines, including the classical paper by Ziegler amd Nichols [1], the IMC PID-tuning paper by Rivera et al. [2], and the closely related direct synthesis tuning rules in the book by Smith and Corripio [3]. The Ziegler–Nichols settings result in a very good disturbance response for integrating processes, but are otherwise known to result in rather aggressive settings [4,5], and also give poor performance for processes with a dominant delay. On the other hand, the analytically derived IMC-settings in [2] are known to result in a poor disturbance response for integrating processes (e.g., [6,7]), but are robust and

SIMC-PID Tuning Rules

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

For cascade form PID controller:

$$K_c = \frac{1}{k}\frac{\tau_1}{\tau_c + \theta} = \frac{1}{k'} \cdot \frac{1}{\tau_c + \theta} \qquad (1)$$

$$\tau_I = \min\{\tau_1, \frac{4}{k'\,K_c}\} = \min\{\tau_1, 4(\tau_c + \theta)\} \qquad (2)$$

$$\tau_D = \tau_2 \qquad (3)$$

Derivation:

1. First-order setpoint response with response time τ_c (IMC-tuning = "Direct synthesis")
2. Reduce integral time to get better disturbance rejection for slow or integrating process (but avoid slow cycling $\Rightarrow \tau_I \geq \frac{4}{k'\,K_c}$)

One tuning parameter: Desired closed-loop time constant τ_c

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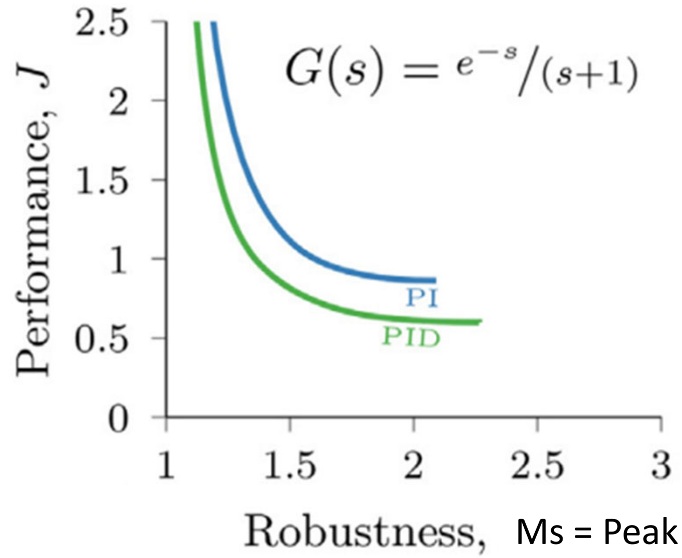
Can the SIMC PID-rules be improved?

- No, hardly. They are almost Pareto-optimal (vary τ_c)
- Optimal trade-off between robustness (Ms) and performance (J=IAE for setpoint and disturbance):

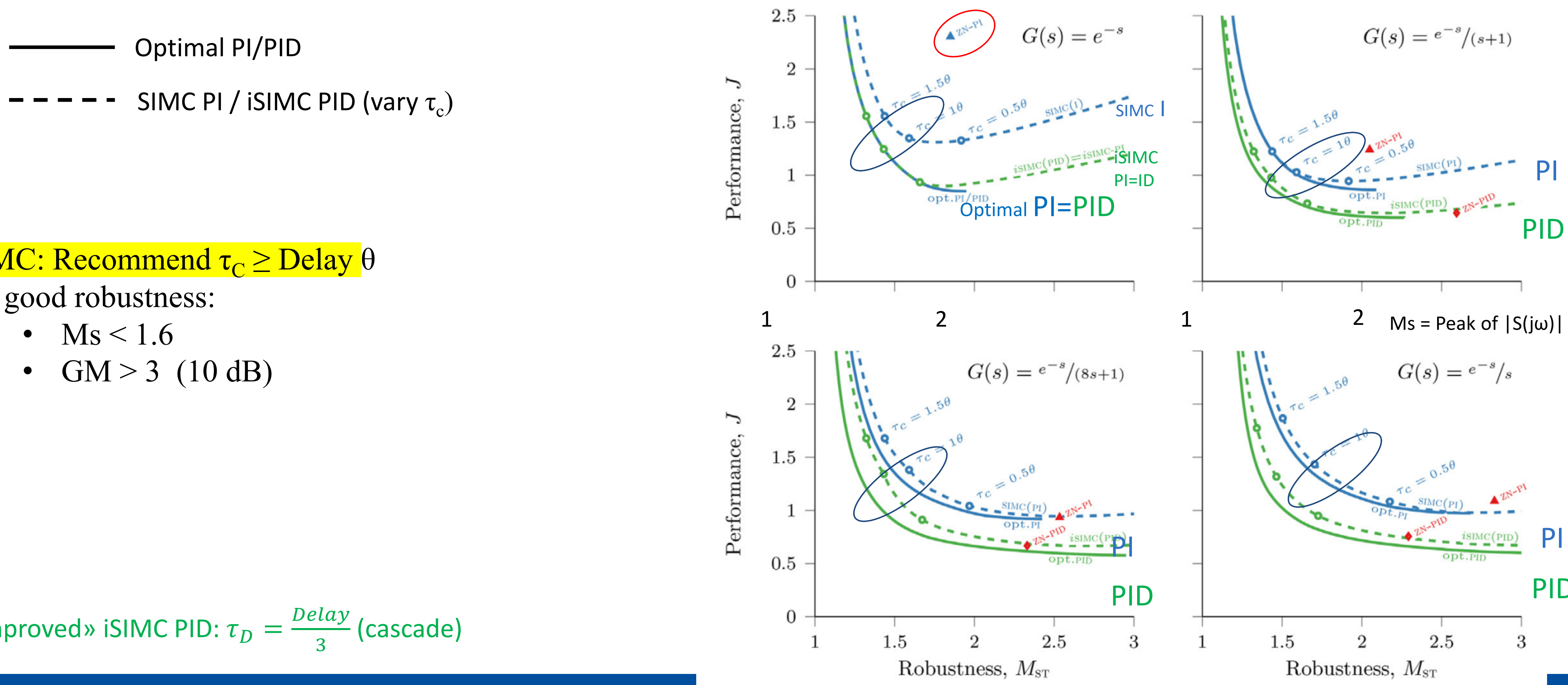


Ms = Peak of |S(jω)|

SIMC-tuning is almost Pareto-optimal!

Trade-off between robustness (Ms) and performance (J=IAE for setpoint and disturbance)

C. Grimholt, S. Skogestad / Journal of Process Control 70 (2018) 36–46



SIMC: Recommend $\tau_C \geq$ Delay θ
for good robustness:

- Ms < 1.6
- GM > 3 (10 dB)

«Improved» iSIMC PID: $\tau_D = \frac{Delay}{3}$ (cascade)

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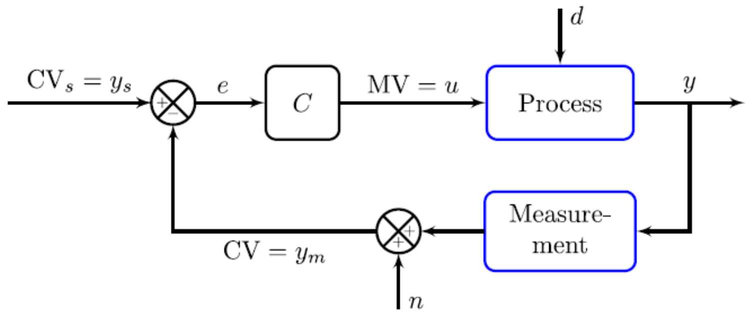
PID control: Most important is choice of pairing



MV-CV Pairing. Two main pairing rules:

1. ***"Pair-close rule" :*** *The MV should have a large, fast, and direct effect on the CV.*
2. ***"Input saturation rule":*** *Pair a MV that may saturate with a CV that can be .given up (when the MV saturates).*
   – *Exception: Have extra MV so we use MV-MV switching (e.g., split range control)*

Additional rule for interactive systems:

3. ***" RGA-rule".*** *Avoid pairing on negative steady-state RGA-element.*

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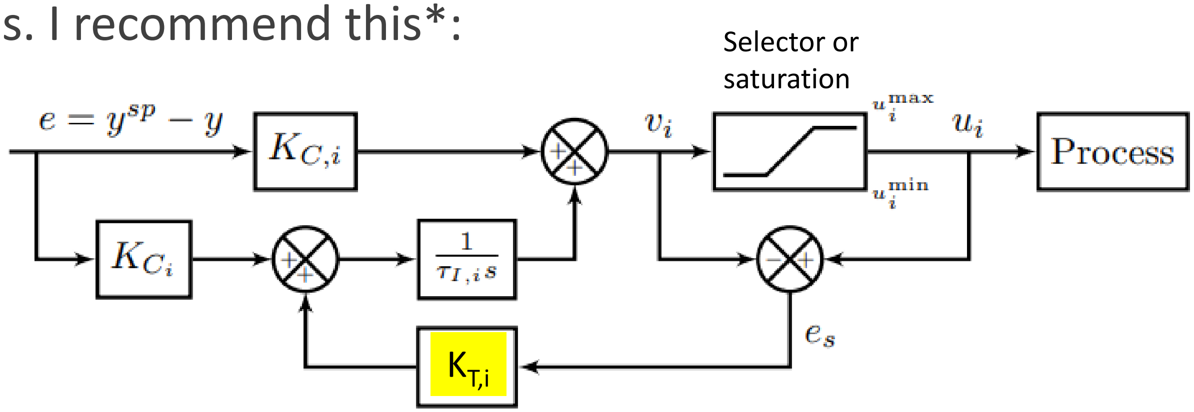
Anti-windup is important!

- All controllers with I action need anti-windup to «stop integration» during periods when the controller output (v_i) is not affecting the process:
  - Controller is disconnected (e.g., because of selector)
  - Physical MV u_i is saturated
- Many approaches. I recommend this*:



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

*Åström, K. J., & Hägglund, T. (1988). *Automatic tuning of PID controllers*. ISA.

Future

- SIMC Pareto-optimal ⇒ Little value in deriving new PID tuning rules
- BUT: A lot more work needed on **Advanced PID architectures** (ARC)

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

- Hint 1: According to Sigurd Skogestad
- Hint 2: All were in use around 1940

SOLUTION

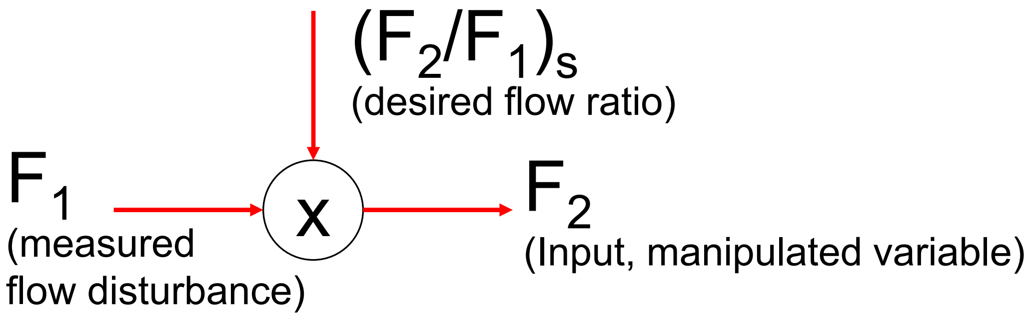
1. PID controller, in particular, I-action
2. Cascade control
3. Ratio control

Ratio control

Special case of to feedforward, but don't need model, just process insight.
Always use for mixing streams

- Note: Disturbance needs to be a flow (or more generally an extensive variable)

Use multiplication block (x):



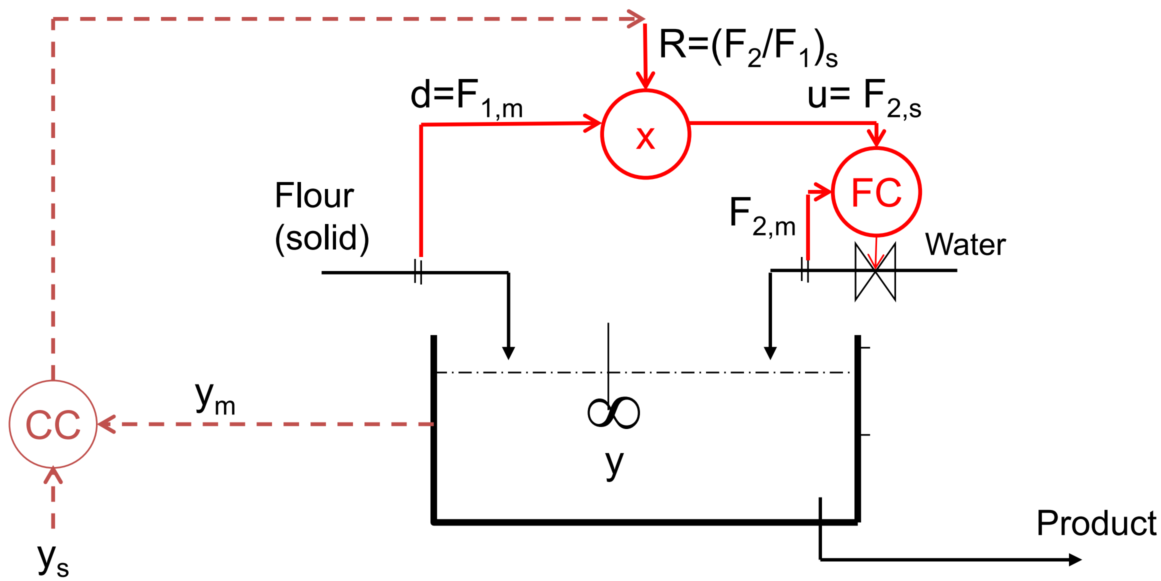
"Measure disturbance ($d=F_1$) and adjust input ($u=F_2$) such that ratio is at given value $(F_2/F_1)_s$"

Usually: Combine ratio (feedforward) with feedback

***Example cake baking**: Use recipe (ratio control = feedforward), but a good cook adjusts the ratio to get desired result (feedback)*

EXAMPLE: MIXING PROCESS

RATIO CONTROL with outer feedback (to adjust ratio setpoint)



Feedback correction («trim»)

Want to control: Composition y

(or any intensive quality variable, like viscicity, ρ or T*)

* S. Skogestad. *Ratio control: Theoretical basis and practical implementation*, Journal of process control, 2026.

Constraints

- **Active constraint**: Optimally at limiting value (max or min) in current situation.
  - current value of disturbances, parameters, and prices.
- Want to use Control to switch between changing active constraints

- *«OK, you may handle many things with PIDs - but for constraints you need MPC.»*

- No, this is an academic myth

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QUIZ, part 2
Three more important inventions of process control?

- Hint: All related to constraints

SOLUTION

1. Anti-windup on I-action *(MV constraint)*
2. A. Split-range/ B. Split-parallel control *(MV-MV constraint switching)*
3. MAX/MIN-Selectors ("Override") *(CV-CV constraint switching)*

Fact: Most constraints can be handled nicely using PIDs plus these three inventions

MV = manipulated variable
CV = controlled variable

How design ARC elements?

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

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

- Academia: Very little work
  - I feel alone

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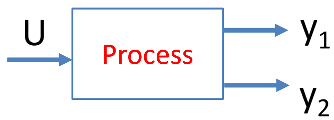
Review article

Advanced control using decomposition and simple elements

Sigurd Skogestad

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

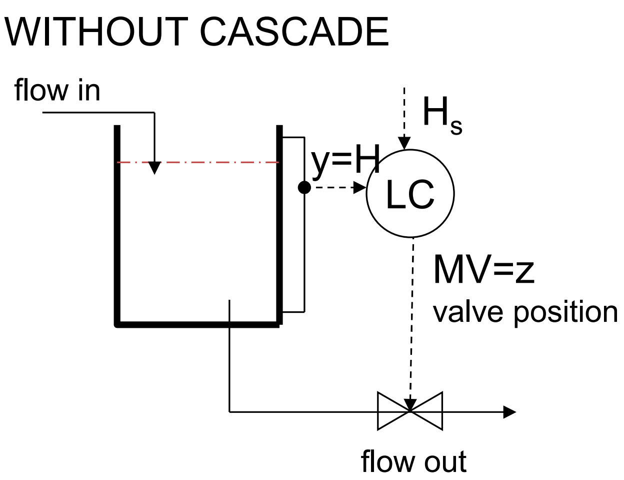
E1. Cascade control

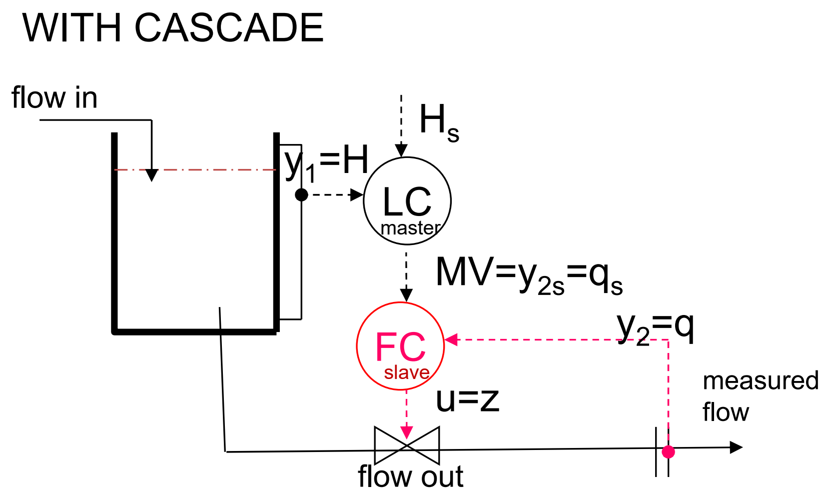


y1=primary output (given setpoint)
y2=secondary output (adjustable setpoint)

Idea: **make use of extra "local" output measurement (y_2)**
Implementation: Controller ("master") gives setpoint to another controller ("slave")

- **Example: Flow controller on valve (very common!)**





Tuning cascade control

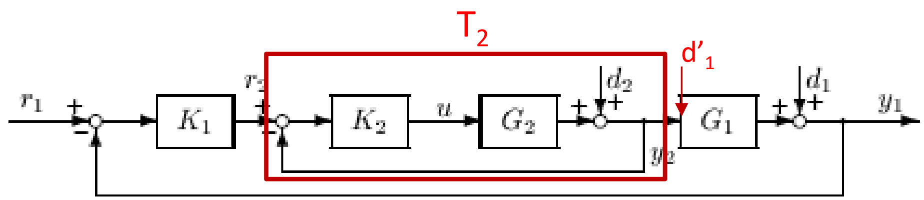


Figure 10.11: Common case of cascade control where the primary output y_1 depends directly on the extra measurement y_2

First tune fast inner controller K_2 ("slave")

Design K_2 based on model G_2

Select τ_{c2} based on effective delay θ_2 in G_2

Nonlinearity: Gain variations (in G_2) translate into variations in actual time constant τ_{C2}

Then with slave closed, tune slower outer controller K_1 ("master"):

Transfer function for inner loop (from y_{2s} to y_2): $T_2 = G_2 K_2/(1+G_2 K_2) \approx 1 \cdot e^{-\theta_2 s}/(\tau_{c2}s+1) \approx e^{-(\theta_2+\tau_{c2})s}$

Design K_1 based on model $G_1'=T_2 * G_1$

Typical choice: $\tau_{c1} = \sigma\, \tau_{c2}$ where time scale separation $\sigma = 4\ to\ 10$.

Time scale separation is needed for cascade control to work well

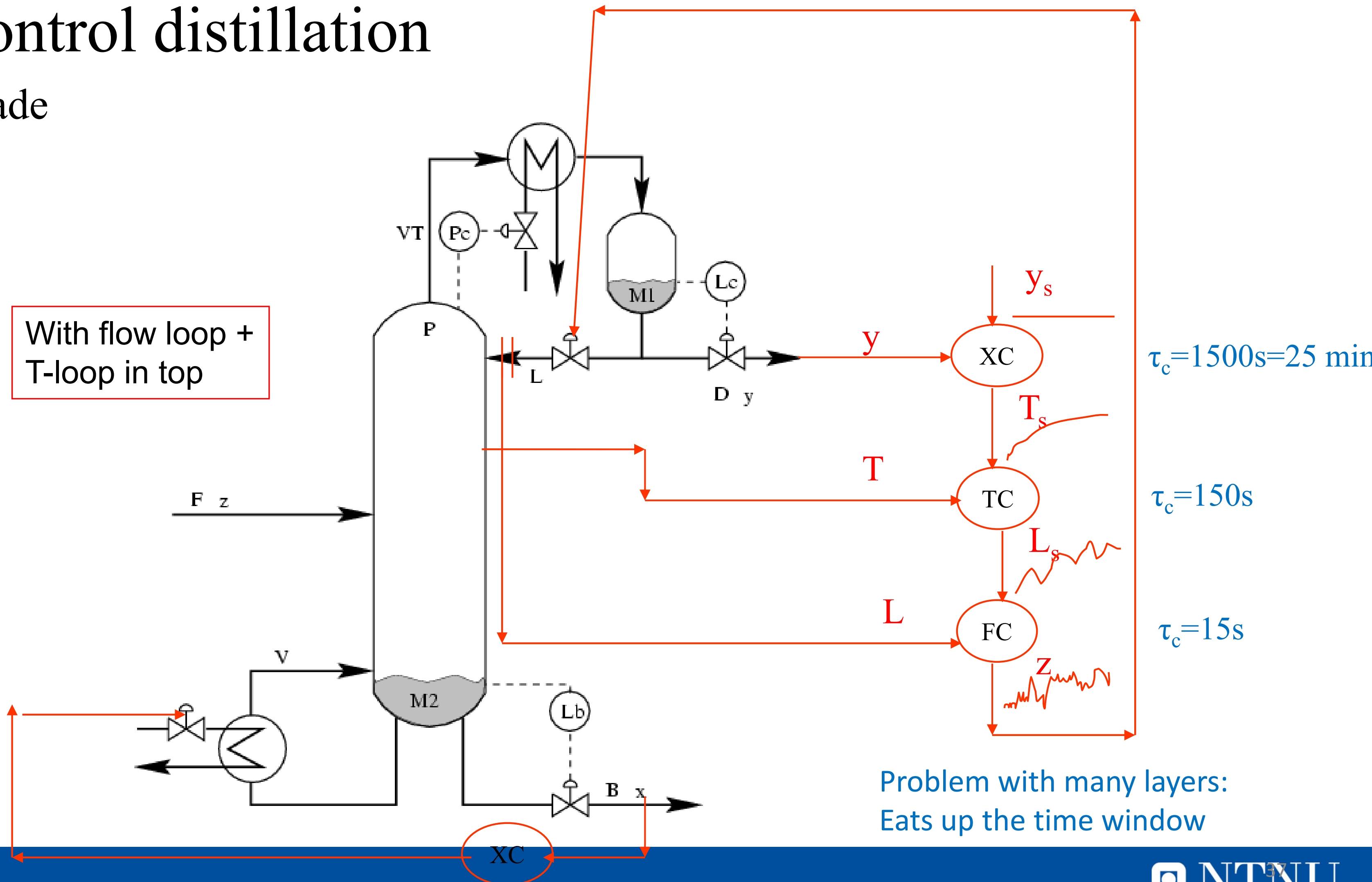
- Inner loop (slave) should be at least 4 times* faster than the outer loop (master)
  - This is to make the two loops (and tuning) independent.
  - Otherwise, the slave and master loops may start interacting
    - The fast slave loop is able to correct for local disturbances, but the outer loop does not «know» this and if it's too fast it may start «fighting» with the slave loop.
- Often recommend 10 times faster, $\sigma \equiv \frac{\tau_{c1}}{\tau_{c2}} = 10.$
  - A high σ is robust to gain variations (in both inner and outer loop)
  - The reason for the upper value (σ =10) is to avoid that control gets too slow, especially if we have many layers

- Use tuining rule where τ_c is specified (SIMC, Lambda-tuning..)

* Shinskey (Controlling multivariable processes, ISA, 1981, p.12)

Cascade control distillation

3 layers of cascade



Benefits of cascade control

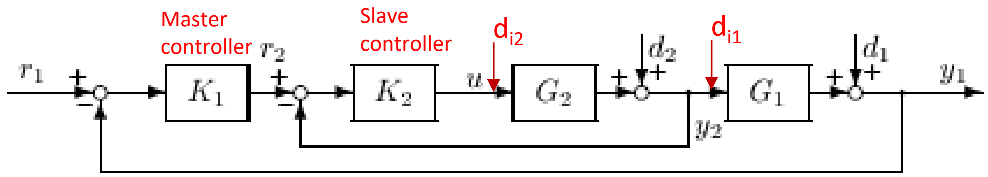


Figure 10.11: Common case of cascade control where the primary output y_1 depends directly on the extra measurement y_2

Use cascade control when:

1. **Significant disturbances d_2 and d_{i2} inside slave loop** **(and y_2 can be controlled faster than y_1)**
2. **Integrating dynamics** (including slow dynamics or unstable) in both G_1 and G_2, (because without cascade a double integrating plant G_1G_2 is difficult to control)
3. **The plant G_2 is nonlinear or varies with time or is uncertain.**
4. **Ratio control: Need FC to Set the flowrate (r_2) rather than valve position (u)**
5. **Measurement delay (or infrequent measurement) for y_1**

Shinskey (1967)

The principal advantages of cascade control are these:

1. Disturbances arising within the secondary loop are corrected by the secondary controller before they can influence the primary variable.

2. Phase lag existing in the secondary part of the process is reduced measurably by the secondary loop. This improves the speed of response of the primary loop.

3. Gain variations in the secondary part of the process are overcome within its own loop.

4. The secondary loop permits an exact manipulation of the flow of mass or energy by the primary controller.

Special common case ("series cascade")

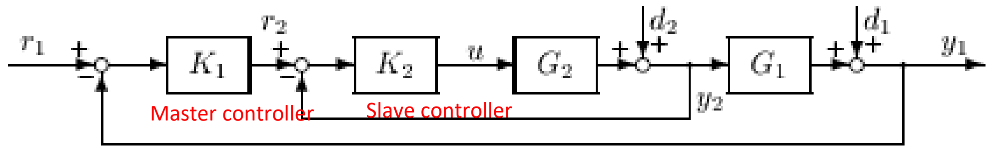
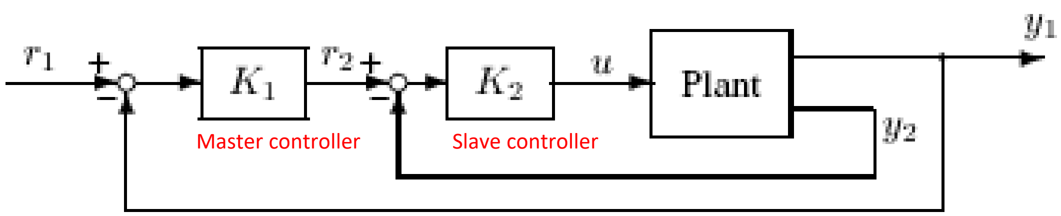


Figure 10.11: Common case of cascade control where the primary output y_1 depends directly on the extra measurement y_2

Always helpful

General case ("parallel cascade")



(a) Extra measurements y_2 (conventional cascade control)

Not always helpful

Constraints

- **Active constraint**: Optimally at limiting value (max or min) in current situation.
  - current value of disturbances, parameters, and prices.
- Control: Use to switch between changing active constraints

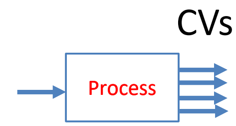
- *«OK, you may handle many things with PIDs - but for constraints you need MPC.»*

- No, this is an academic myth

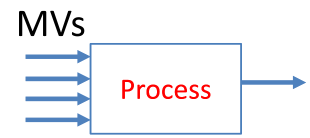
- Fact: Most constraints can be handled nicely using PIDs
  + split range/parallel control (for MV constraint switching)
  + selectors (overrides) (for CV constraint swithcing)
  + anti-windup on I-action

Constraint switching cases

- CV-CV switching («override»)
  - Control one CV at a time

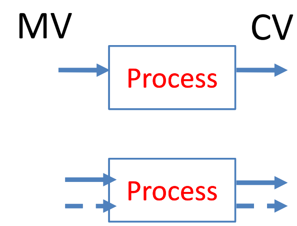


- MV-MV switching («split range»)
  - Use one MV at a time



- MV-CV switching
  - MV saturates so must give up CV

  1. Simple («do nothing»)
  2. Complex (need another MV);
     Combine MV-MV and CV-CV



MV = manipulated variable
CV = controlled variable

MV-MV switching

- Need several MVs to cover whole <u>steady-state</u> range (because primary MV may saturate)*
- Note that we only want to use one MV at the time.

Three solutions:

Alt.1 Split-range control (one controller) (E5)

Alt.2 Several controllers with different setpoints (E6) = Split parallel control

Alt.3 Split VPC (E7)

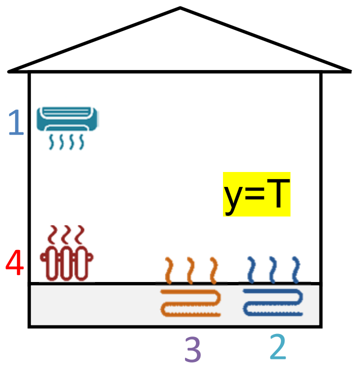
Which is best? It depends on the case!

Example MV-MV switching

- Break and gas pedal in a car
- Use only one at a time,
- «manual split range control»

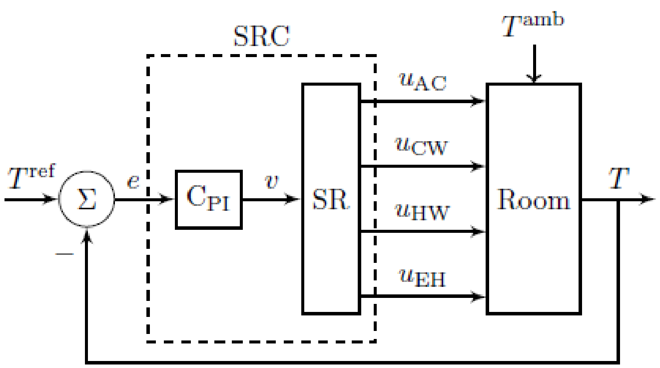
MV-MV switching

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

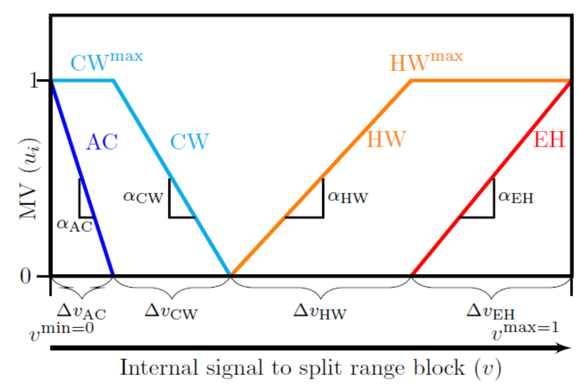


MVs (two for summer and two for winter):

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



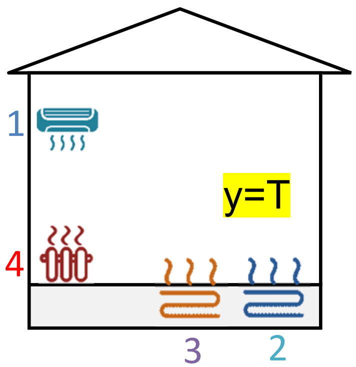
SR-block:

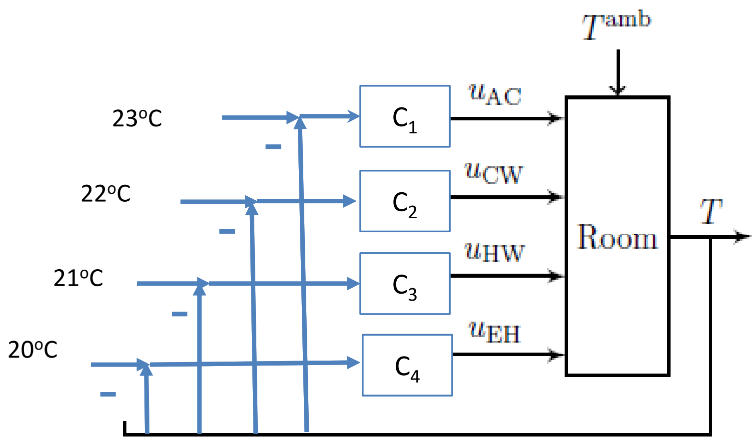


C_{PI} – same controller for all inputs (one integral time)
But get different gains by adjusting slopes α in SR-block

E6: MV-MV switching

Alternative: Multipliple Controllers with different setpoints (E6)
= Split parallel control





Advantage SPC (relative to SRC):
- One controller for each MV
- Simple: Switching by feedback

Disadvantage SPC:
- Different setpoints
  - may fix with outer cascade
- May get fighting («limbo») with small setpoint separation

But Advantage:
- Different setpoints (energy savings)
- «Limbo» may improve control of y

«Split parallel control – A little known control structure», K. Forsman, M. Adlouni, S. Skogestad, IFAC Dycops conference, 2025 (and JPC 2026)

Simulation Room temperature

- Dashed lines: Split range control (E5)
- Solid lines: Split parallel control with different setpoints (E6)

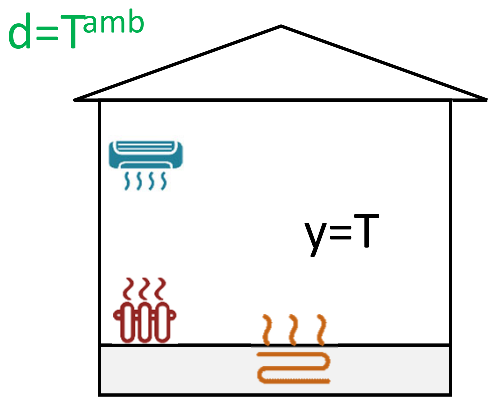
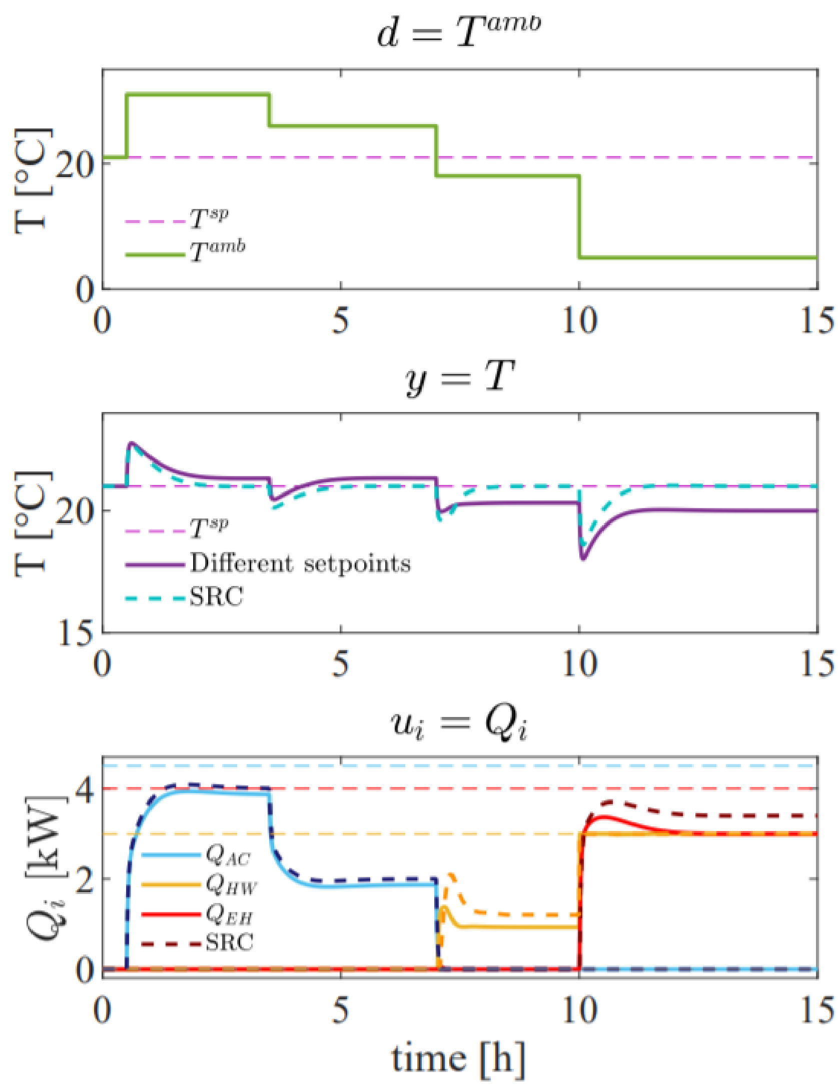


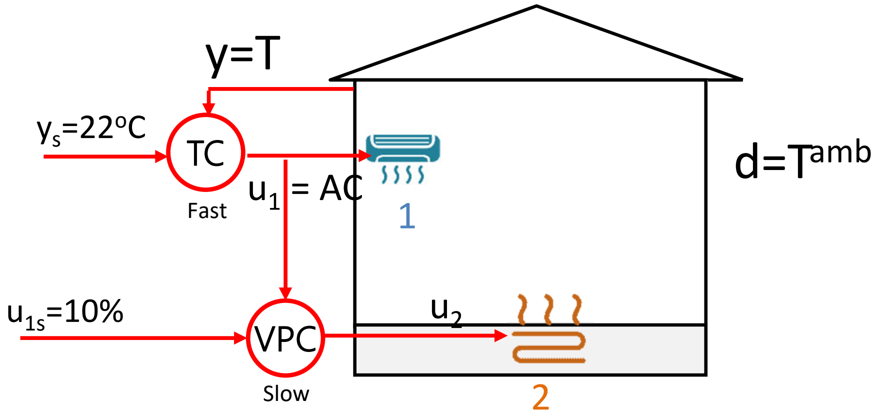
Table 1. Ranges for available inputs (u_k).

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



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)

Split VPC for MV-MV switching
Example: Room heating with fast cooling (AC) and slow heating



MVs:
1. u_1= AC (cooling with fan, fast)
2. u_2 = HW (hot water in floor, slow)

u_2= Floor heating with Hot water (VPC): used in winter (summer: u_2=0%).

Advantage: Temperature is always controlled by fast cooling (u_1=AC)
Economic disadvantage: Cooling u_1 is used also in winter (about 10% load)

Summary MV-MV switching

- Need several MVs to cover whole steady-state range (because primary MV may saturate)*
- Note that we only want to use one MV at the time.

Alt.1 Split-range control (one controller) (E5)

- Advantage: Easy to understand because SR-block shows clearly sequence of MVs
- Disdvantages: (1) Need same tunings (integral time) for all MVs . (2) May not work well if MV-limits inside SR-block change with time, so: Not good for MV-CV switching

Alt.2 Split-parallel control (with different setpoints) (E6)

- Advantages: 1. Simple to implement, do not need to keep track of MVs. 2. Can have independent tunings. .
- Disadvantages: Temporary loss of control during switching. Setpoint varies (which can be turned into an advantage in some cases)

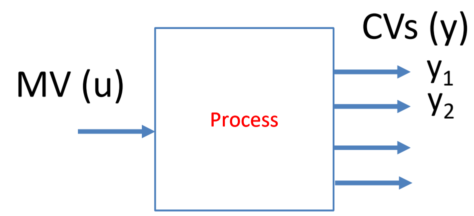
Alt.3 Split VPC (E7)

- Advantage: Always use "primary" MV for control of CV (avoids repairing of loops)
- Disadvantages: Gives some loss, because primary MV always must be used (cannot go to zero).

Which is best? It depends on the case!

***Optimal Operation with Changing Active Constraint Regions using Classical Advanced Control, Adriana Reyes-Lua Cristina Zotica, Sigurd Skogestad, Adchem Conference, Shenyang, China. July 2018** ,

CV-CV switching ("override")



- Only one input (MV) controls many outputs (CVs)
  - Change in active constraint
  - Example, Adaptive cruise control:
    - Control car speed (y_1) - but give up if too small distance (y_2) to car in front.
- Use max- or min-selectors (note: switching on controller outputs = inputs u!)

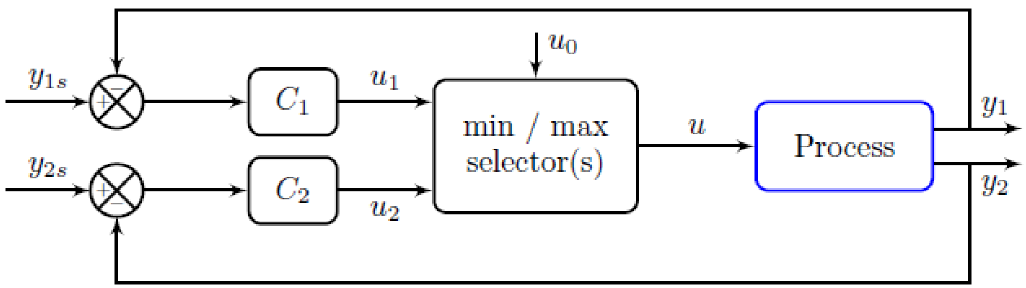
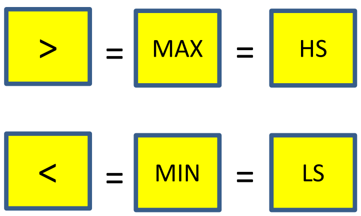
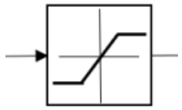


Figure 17: CV-CV switching with selector on MV (input u).



Sometimes the selector is on the setpoint, and the max/min-block may be replaced by a saturation element



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

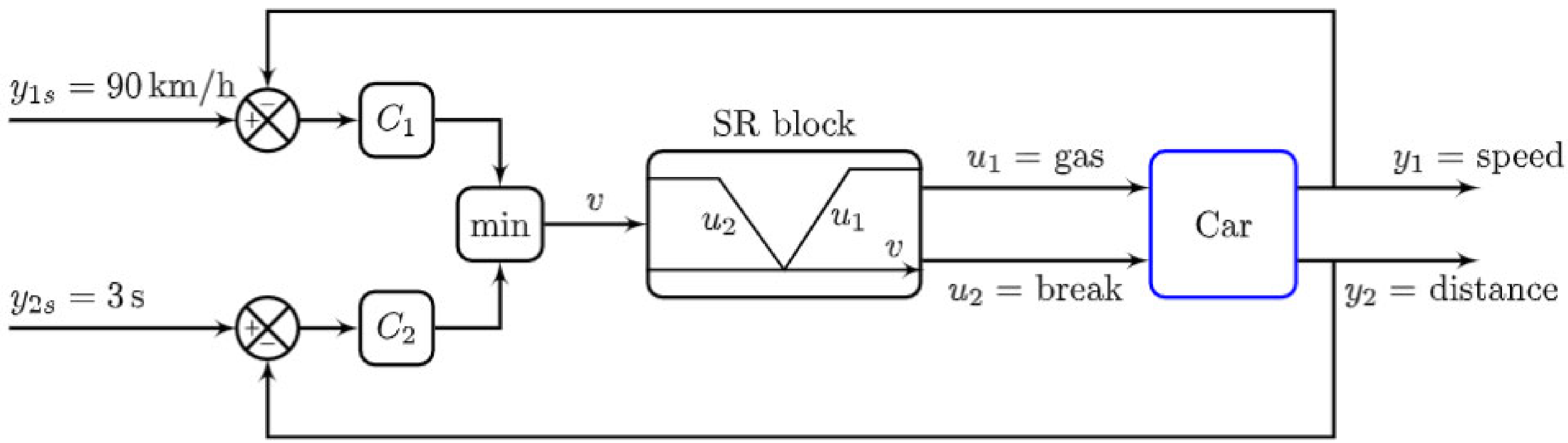


Fig. 31. Adaptive cruise control with selector and split range control.

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

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CV-CV switching

Furnace control with safety constraint

Input (MV)

u = Fuel gas flowrate

Output (CV)

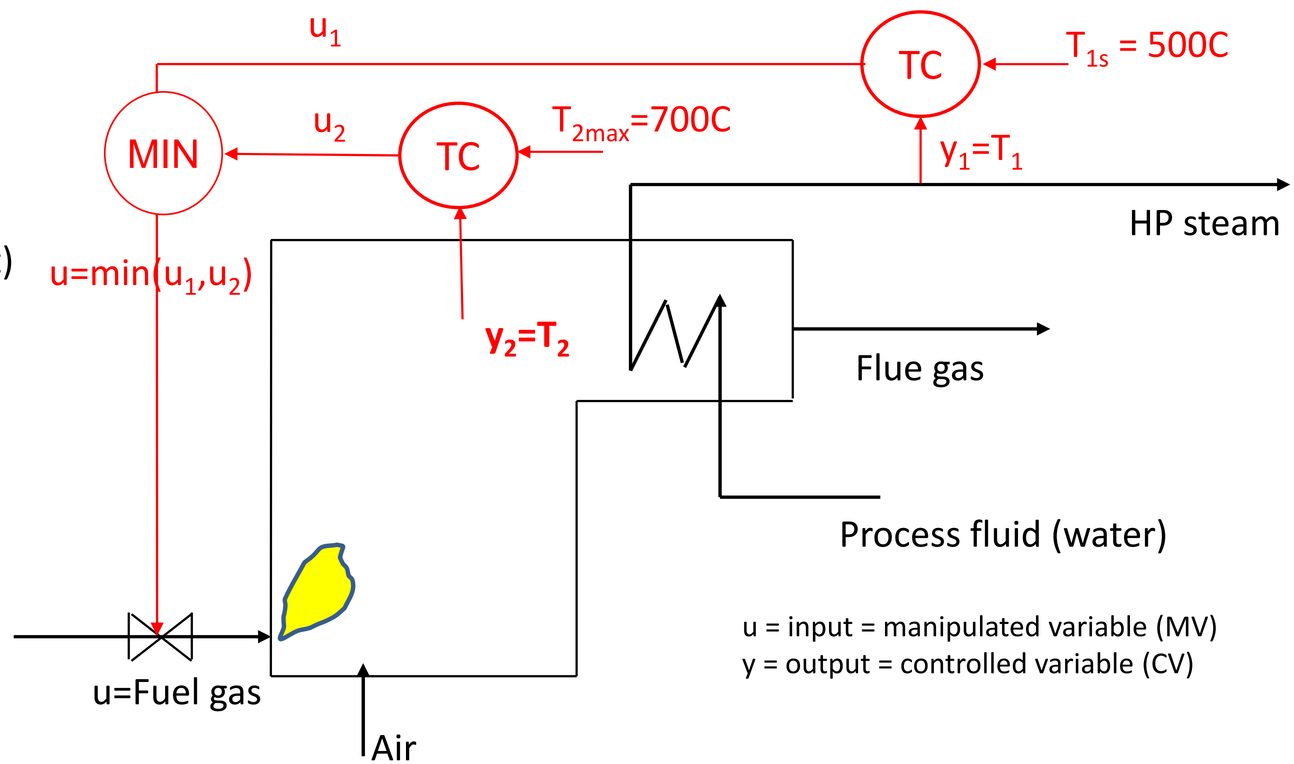
y_1 = process temperature T_1

(desired setpoint or max constraint)

y_2 = furnace temperature T_2

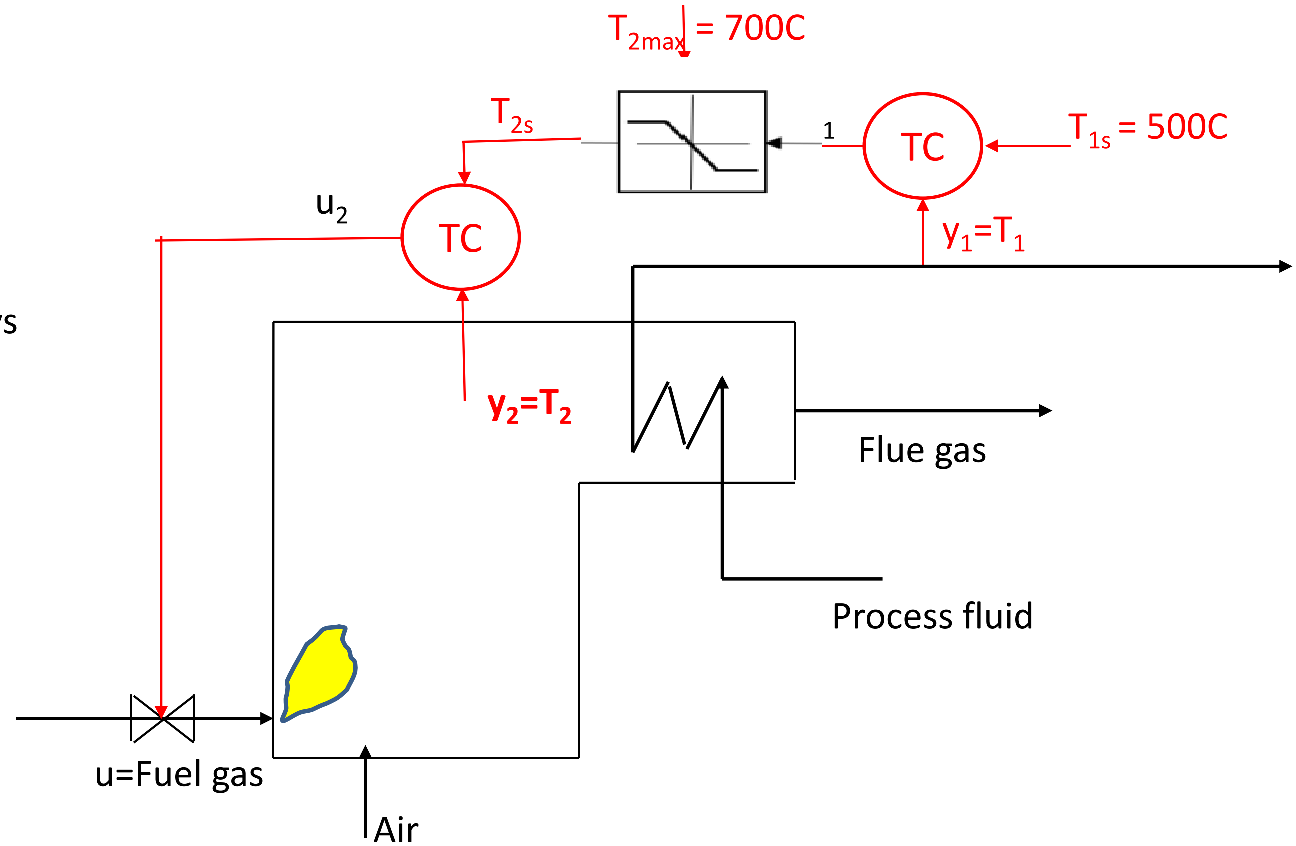
(T2max= 700C)

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



Furnace control with cascade (selector on CV-sp)

Comparison
The cascade solution is less general but it may be better in this case.
Why better? Inner T2-loop is fast and always active and may improve control of T1.



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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)

MV-CV switching (because reach constraint on MV)

- **Simple CV-MV switching**
  - Don't need to do anything if we followed the *Input saturation rule:*
  - *"Pair a MV that may saturate with a CV that can be given up (when the MV saturates)"*

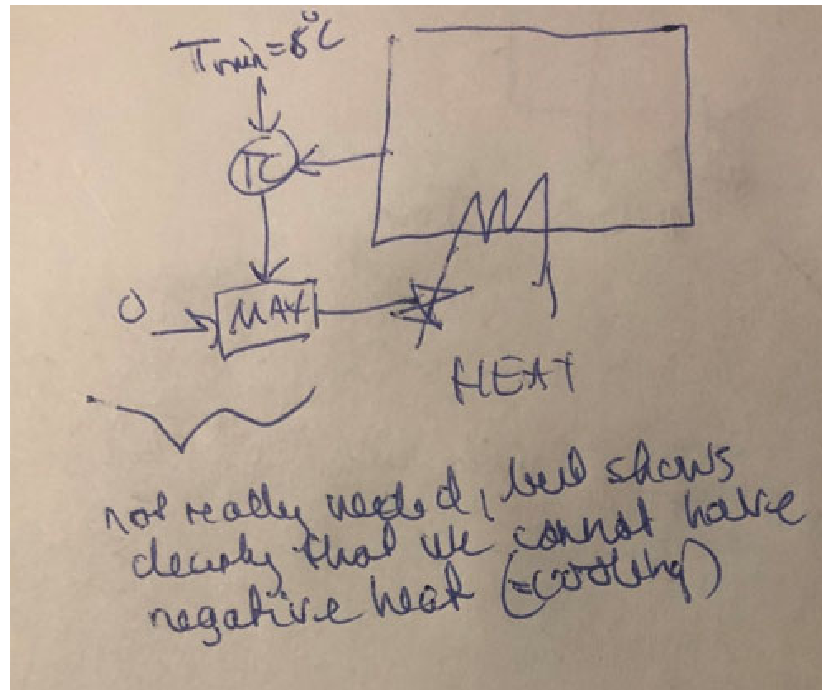
Simple MV-CV switching

Example: Avoid freezing in cabin

$$\text{Minimize } u \text{ (heating), subject to}$$
$$T \geq T_{min}$$
$$u \geq 0$$

Keep CV=T>T_{min} = 8C in cabin in winter by using MV=heating

If it's hot outside (>8C), then the heat will go to zero (MV=Q=0), but this does not matter as the constraint is over-satisfied.



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

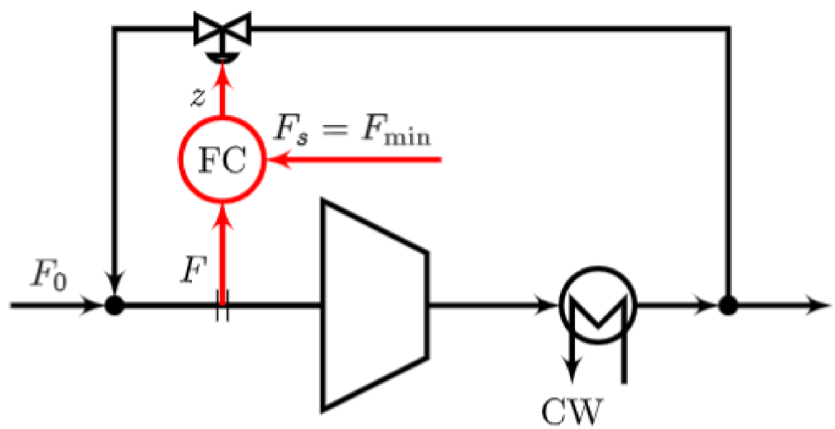


Fig. 32. Flowsheet of anti-surge control of compressor or pump (CW = cooling water). This is an example of simple MV-CV switching: When MV=z (valve position) reaches its minimum constraint ($z = 0$) we can stop controlling CV=F at $F_s = F_{min}$, that is, we do not need to do anything except for adding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

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MV-CV-CV switching

Example: Compressor with max-constraint on F_0

(in addition to the min-constraint on F)

Minimize u (recycle), subject to

MV = u = z ≥ 0

$CV_1 = F \geq F_{min}$

$CV_2 = F_0 \leq F_{0,max}$

- Both CV-constraints are satisfied by a large z
  ⇒ Max-selector for CV-CV
- When we reach MV-constraint (z=0) both constraints are oversatisfied
  ⇒ Simple MV-CV switching

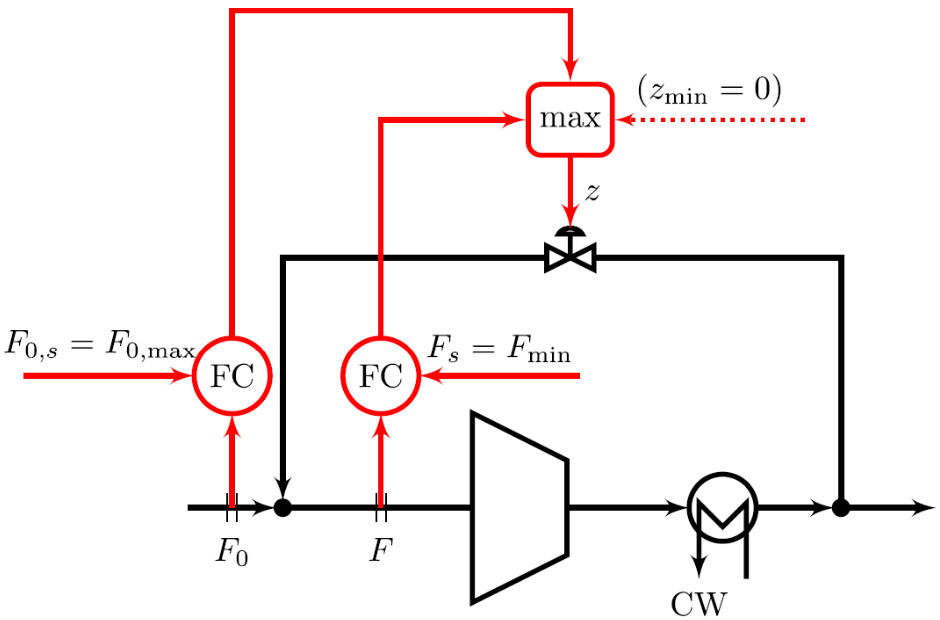


Fig. 33. Anti-surge compressor control with two CV constraints. This is an example of simple MV-CV-CV switching. MV = z, $CV_1 = F$, $CV_2 = F_0$ (all potentially active constraints).

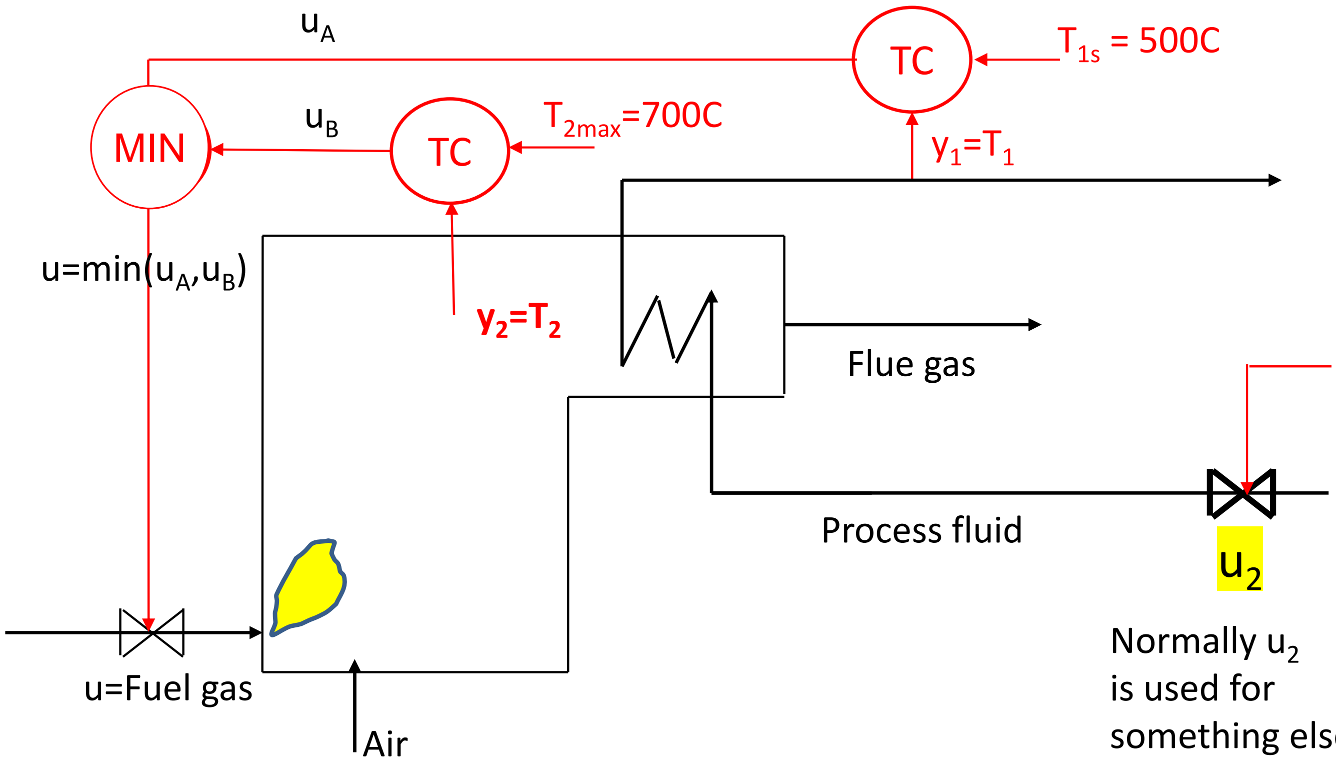
D. Complex MV-CV switching

- Didn’t follow input saturation rule
- Requires a repairing of loops
- Must combine MV-MV switching (3 alternatives) with CV-CV swithing (selector)

Furnace control : Cannot give up control of y_1=T_1. What to do?

Inputs (MV)
- u = Fuel gas flowrate
- u_2 = Process flowrate

Output (CV)
- y_1 = process temperature T_1 (with desired setpoint)



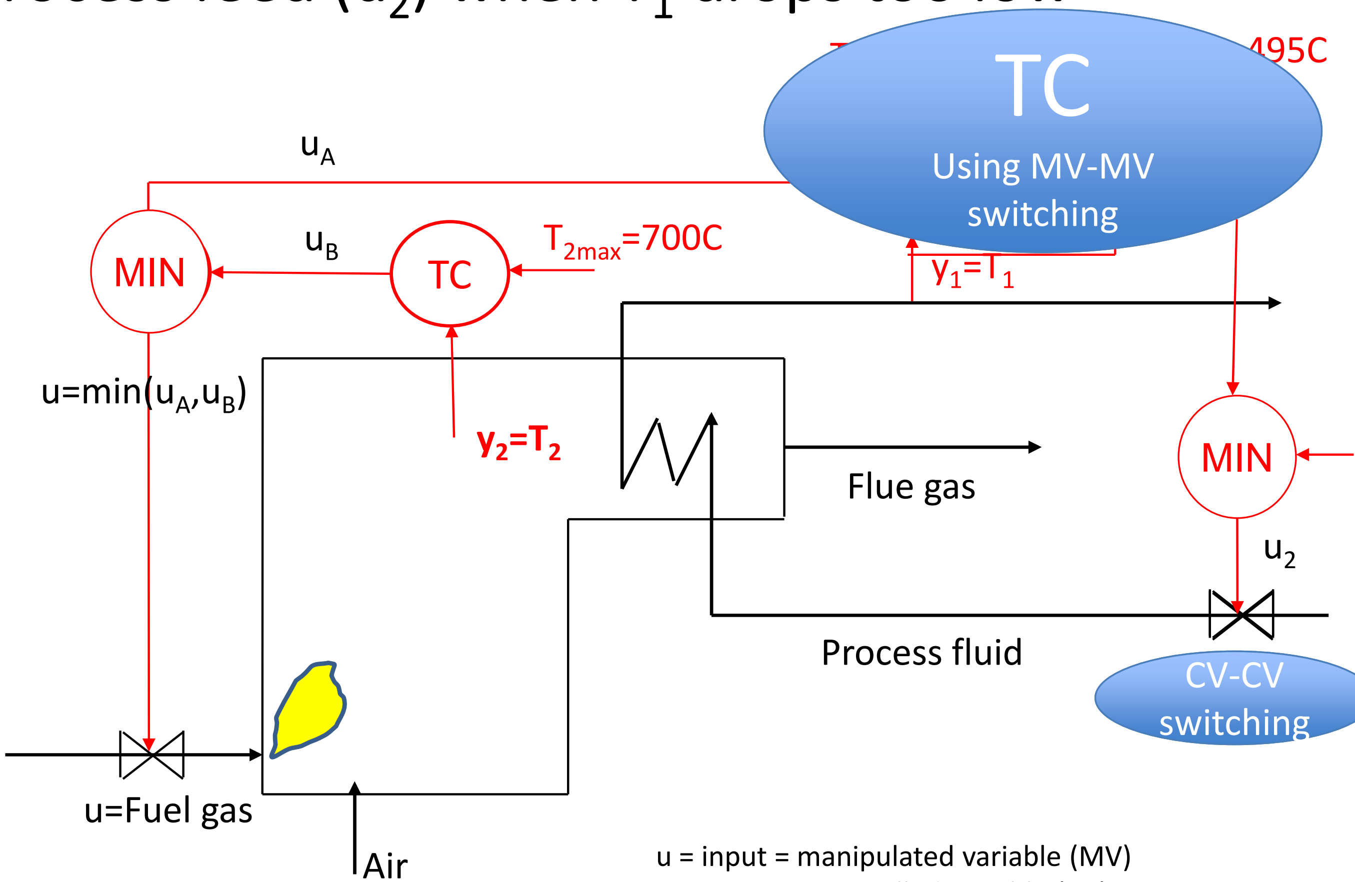
Complex MV-CV switching

Cannot give up controlling T_1

Solution: Cut back on process feed (u_2) when T_1 drops too low

Inputs (MV)
- u = Fuel gas flowrate
- u_2 = Process flowrate

Output (CV)
- y_1 = process temperature (with desired setpoint)



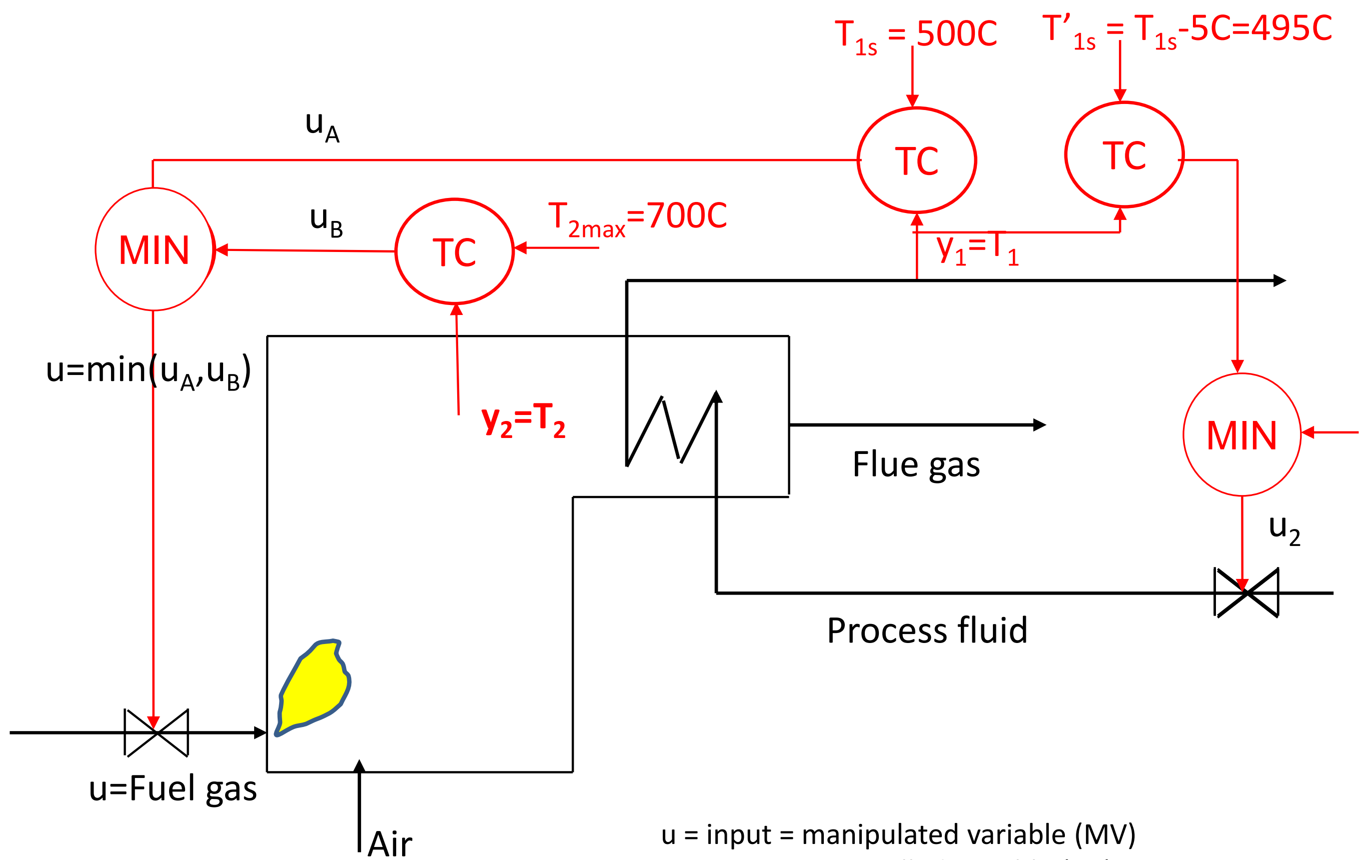
u = input = manipulated variable (MV)
y = output = controlled variable (CV)

Desired: $T_1 = T_{1s}, T_2 \leq T_{2\max}$

Use Alt. 2: Split parallell control (two setpoints)

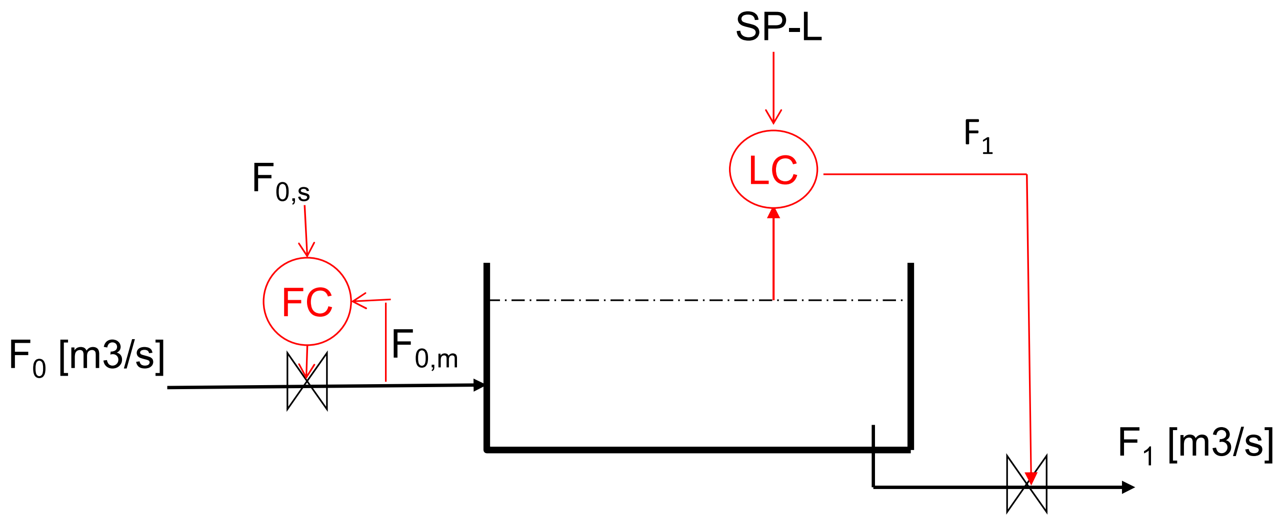
Inputs (MV)
- u = Fuel gas flowrate
- u2 = Process flowrate

Output (CV)
- y_1 = process temperature (with desired setpoint)



u = input = manipulated variable (MV)
y = output = controlled variable (CV)

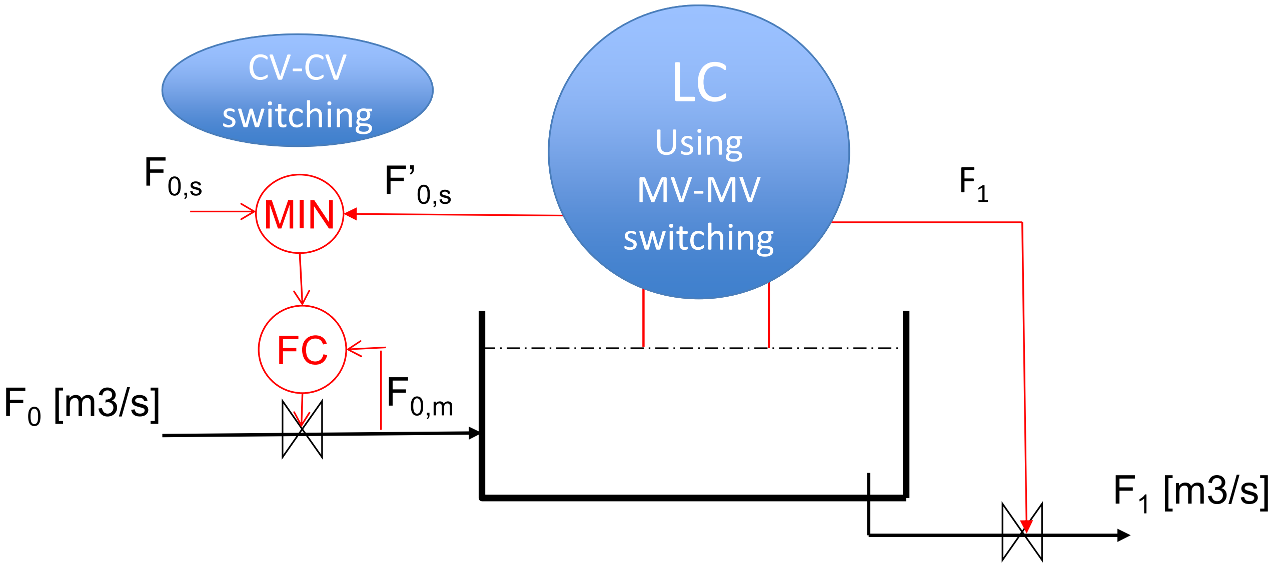
Example: Level control



What should we do if bottleneck at F1 (fully open valve, z1=1)?

Solution: Complex MV-CV switching

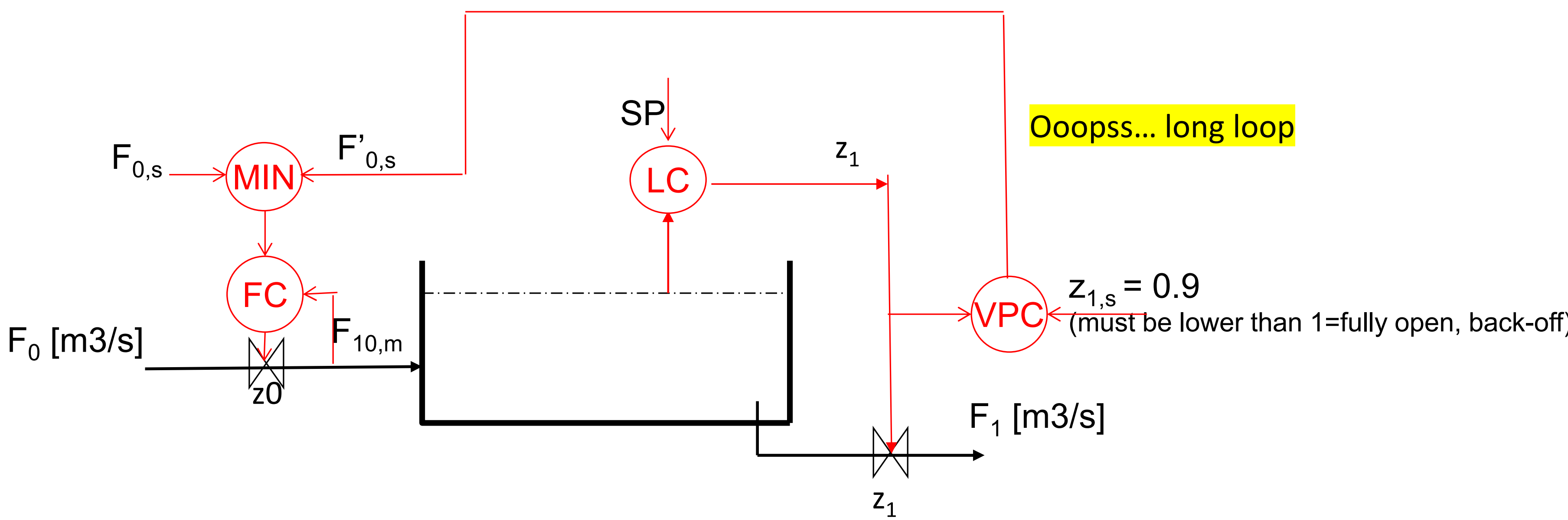
"Bidirectional inventory control"



Three alternatives for MV-MV switching

1. Split range control
2. **Split parallel control (two setpoints)**
3. Split VPC

Alt. 3 MV-MV switching: Split VPC

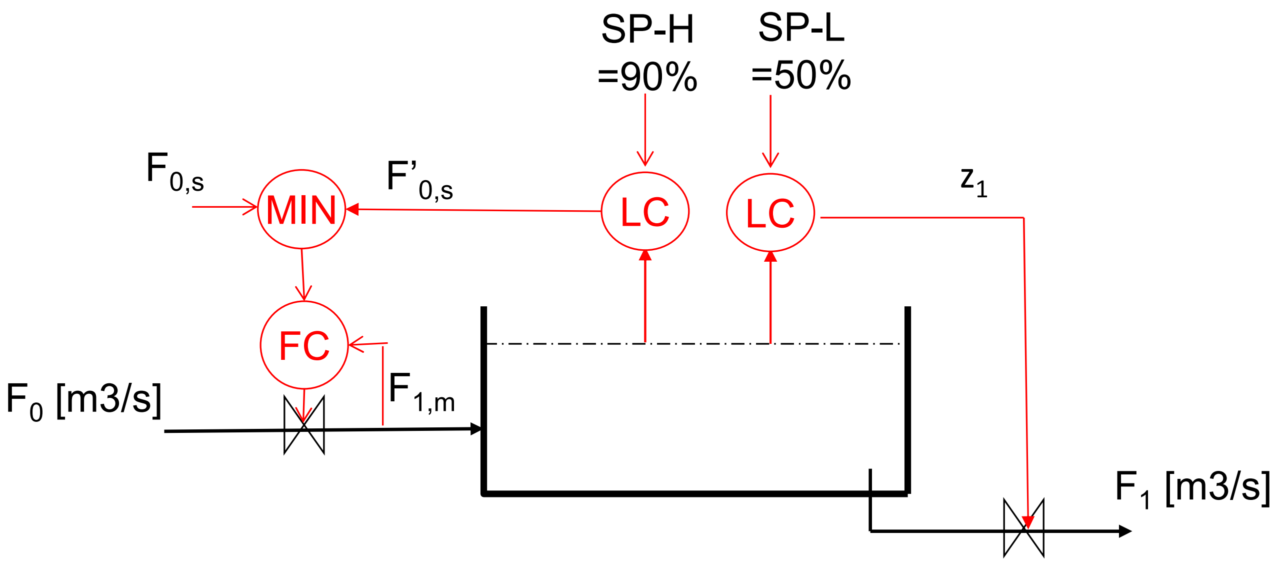


VPC: "reduce inflow (F_0) if outflow valve (z_1) approaches fully open"

«Long loop»: Only works indirectly through another loop (here: LC)

Bidirectional inventory control

Alt. 2 MV-MV switching: Split parallel control (recommended)



SP-L = low level setpoint

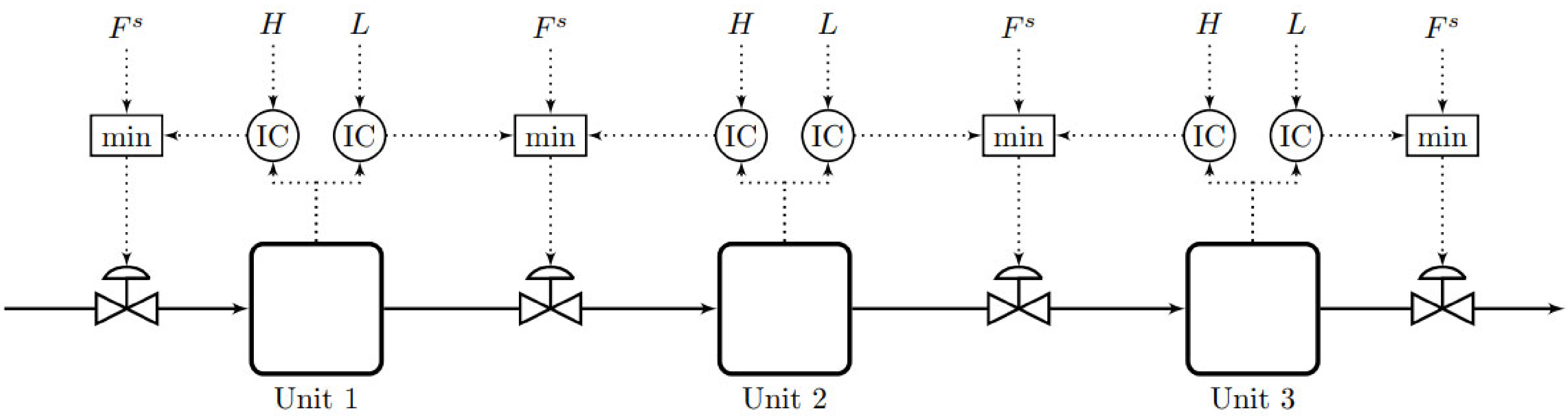
SP-H = high level setpoint

Extra benefit: Use of two setpoints is good for using buffer dynamically!!

Inventories in series . Very smart selector strategy based on **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

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IFAC presentation, 26 Aug. 2026

Advanced PID architectures for tracking changing active constraints

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Extended version (with simulations) of paper at IFAC World Congress, August 2026, Korea. This version (with also separator case study I): 13 August 2026

Abstract: Advanced regulatory control (ARC), also known as advanced PID architectures, is a simple and robust way of controlling processes with changing and possibly conflicting constraints, where it previously was believed - at least in academia - that model-based solutions, such as MPC, were the only effective solution. To illustrate this, ARC is applied in two case studies. The first is a gas-liquid separation process, in which selectors and split-parallel control are combined to achieve bidirectional inventory control in which the throughput manipulator moves automatically to the most optimal position. The second case study is on keeping acceptable air quality (CO2-level) and temperature in a room (in this case, a barn for cows). The CO2 and temperature constraints can be conflicting, leading to a hierarchical switching network of PID controllers.

Keywords: process control, PID control, decentralized control, selectors, hybrid control

1. INTRODUCTION

An active constraint is a constraint which, in the current situation, should be kept at its limiting (maximum or minimum) value to achieve optimal (economic) operation at steady state. The "current situation" is defined by the current value of disturbances, parameters, and prices in the cost function. To achieve optimal economic operation, the most important is usually tracking and controlling active constraints (Maarleveld and Rijnsdorp, 1970).

For about 40 years, since the 1980s, there has been a myth, at least in the academic community, that this requires model-based schemes, such as model predictive control (MPC) and real-time optimization (RTO). Of course, academic researchers had some knowledge about industrial schemes for dealing with constraints, including selectors and split-range control, but such "override" schemes (e.g., as described in Maarleveld and Rijnsdorp (1970) for an industrial distillation column) were thought to be *ad hoc*, old-fashioned and not optimal, and not worth the effort of more detailed studies. Until about 10 years ago this was also my view.

Selectors. Consider the case where a single manipulated variable (MV) (input) should control more than one controlled variable (CV) (output). MAX- and MIN-selectors are simple logic elements (sometimes referred to as over- used to control a single CV. Only one MV should be used at a time, which is known as *MV-MV switching*. The traditional solution is to use split-range control, but this requires the user to specify the MV values (say, 0% and 100% in the simple case of valve saturation) where switching should occur. This is avoided with split-parallel control where each MV has its own controller with a different CV setpoint. Here, the switching occurs by feedback when an MV saturates (or is taken over by another controller) and control of the CV is lost.

These and other ARC schemes were discussed in a recent paper (Skogestad, 2023). The objective of the present paper is to extend these findings (with some new rules) and show with two case studies, the power of selectors and split-parallel control.

2. CASE STUDY I: GAS-LIQUID SEPARATOR

2.1 General about inventory control and TPM

Maintaining throughput and managing and controlling inventories (e.g., levels) are of vital importance for process operation. An important concept is the throughput manipulator (TPM), which is the variable that sets the throughput (production rate) for the entire process. Most processes have only one TPM, which also sets the feed rate(s) and product rate(s). Additional feeds are usually

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MV-CV switching

Separator case study I

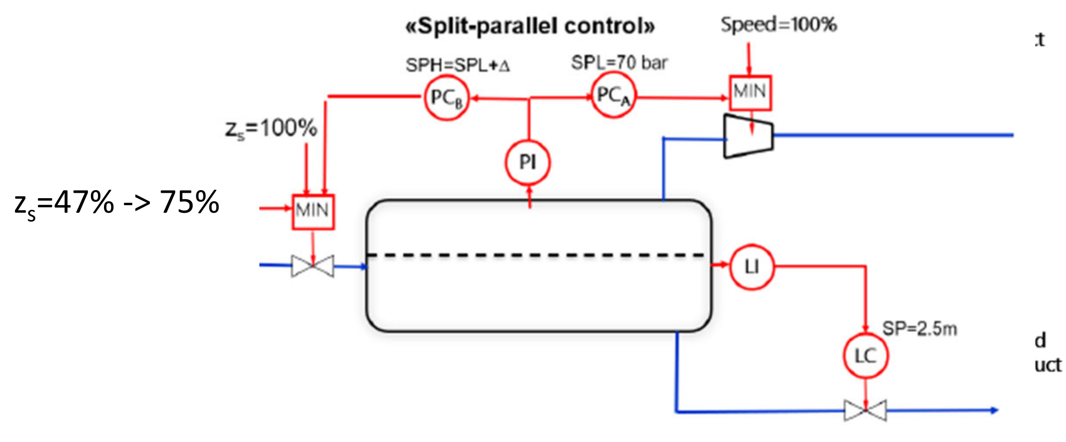


Fig. 4. Bidirectional control scheme where TPM moves automatically to the bottleneck (feed valve or compressor) to maximize throughput.

Δ = 1 bar (small setpoint separation)

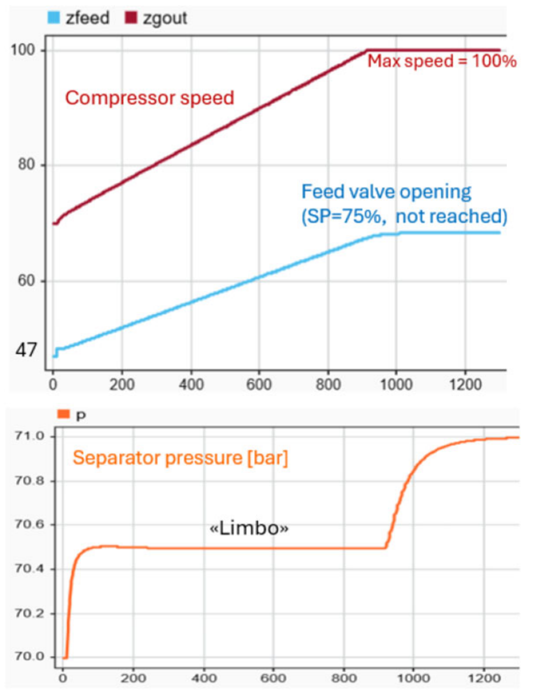


Fig. 14. Separator with SPC (Figure 4): Dynamic simulation of setpoint increase in feed valve opening from 47% to 75%.

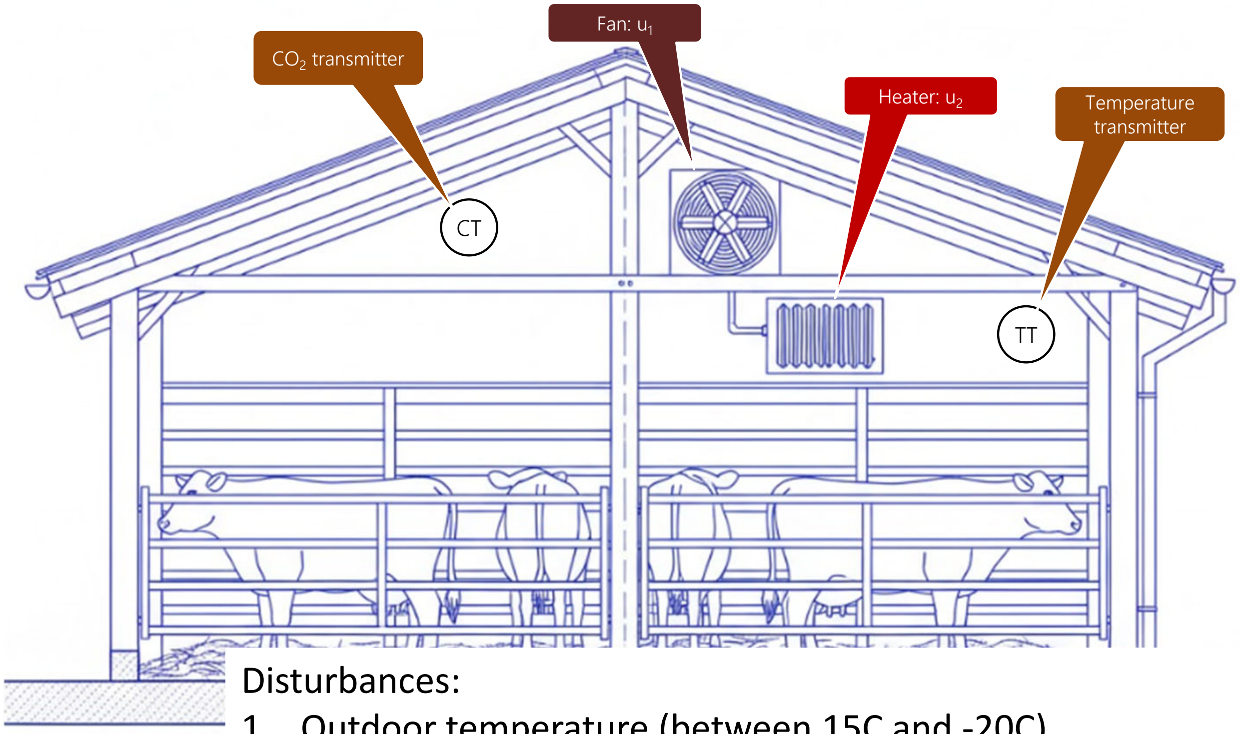
Cow case study II (Norway, outdoor 15C to -20C)

Happy cows
- y1=C (CO2) less than 1000 ppm
- y2=T between 5C and 20C
- Not too much draft or noise from fan
  - 50% fan speed is good

Unhappy (cold) cows
- y2=T less than 0C

Even more unhappy (sick) cows
- y1=C above 3000 ppm (poor air quality)

MVs
- u1=fan (cheap)
- u2=heater (expensive)



Disturbances:
1. Outdoor temperature (between 15C and -20C)
2. Number of cows (they generate CO2 and heat)
   - Because of heat from cows it's always colder outside

NOMINAL CASE

Happy cows

- y1=C (CO2) less than 1000 ppm
- y2=T between 5C and 20C
- Not too much draft or noise from fan
    - 50% fan speed is good

Unhappy cows

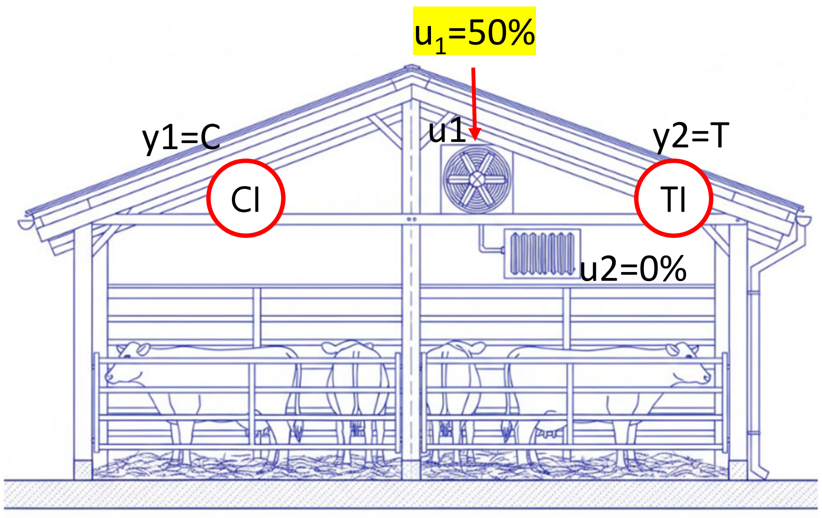
- y2=T less than 0C

Even more unhappy

- y1=C above 3000 ppm (poor air quality)

MVs

- u1=fan (cheap)
- u2=heater (expensive)



Many cows: Lower CO_2 (less than 1000ppm) is satisfied by large fan speed ⇒ MAX-selector

Happy cows

- y1=C (CO_2) less than 1000 ppm
- y2=T between 5C and 20C
- Not too much draft or noise from fan
    - 50% fan speed is good

Unhappy cows

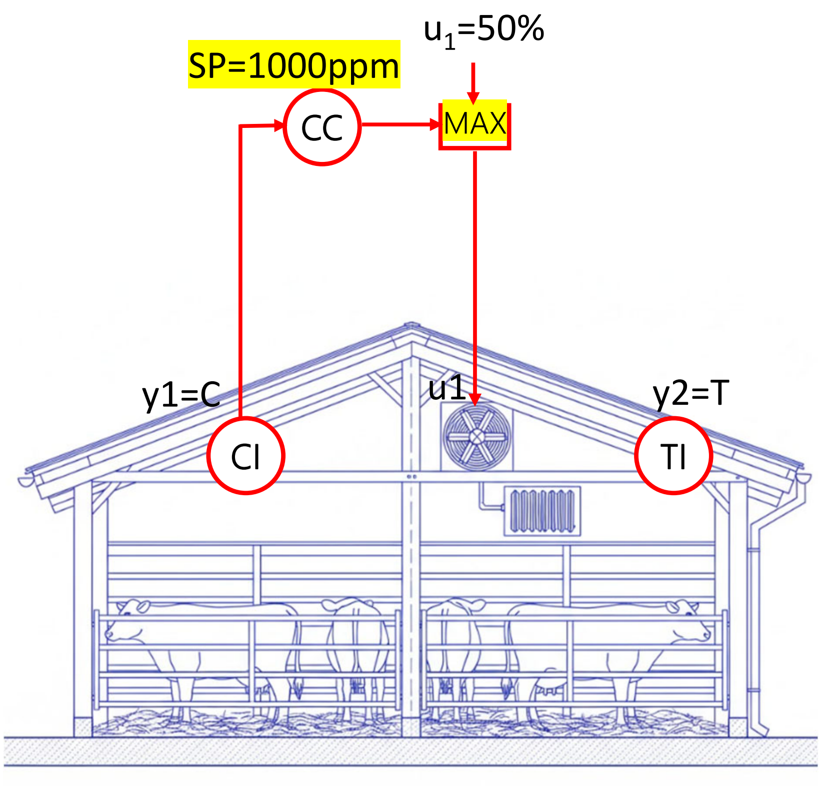
- y2=T less than 0C

Even more unhappy

- y1=C above 3000 ppm (poor air quality)

MVs

- u1=fan (cheap)
- u2=heater (expensive)



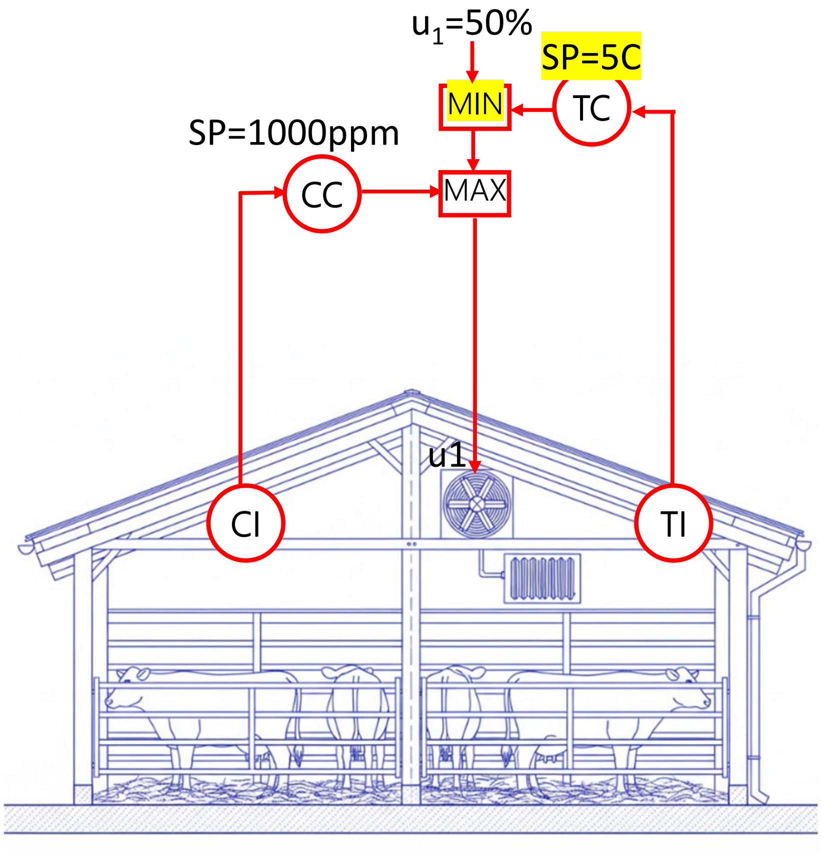
Winter: T (inside) may drop to 5C.
Control: Higher T (above 5C) is satisfied by small fan speed ⇒ MIN-selector

Happy cows
- y1=C (CO2) less than 1000 ppm
- y2=T between 5C and 20C
- Not too much draft or noise from fan
  - 50% fan speed is good

Unhappy cows
- y2=T less than 0C

Even more unhappy
- y1=C above 3000 ppm (poor air quality)

MVs
- u1=fan (cheap)
- u2=heater (expensive)



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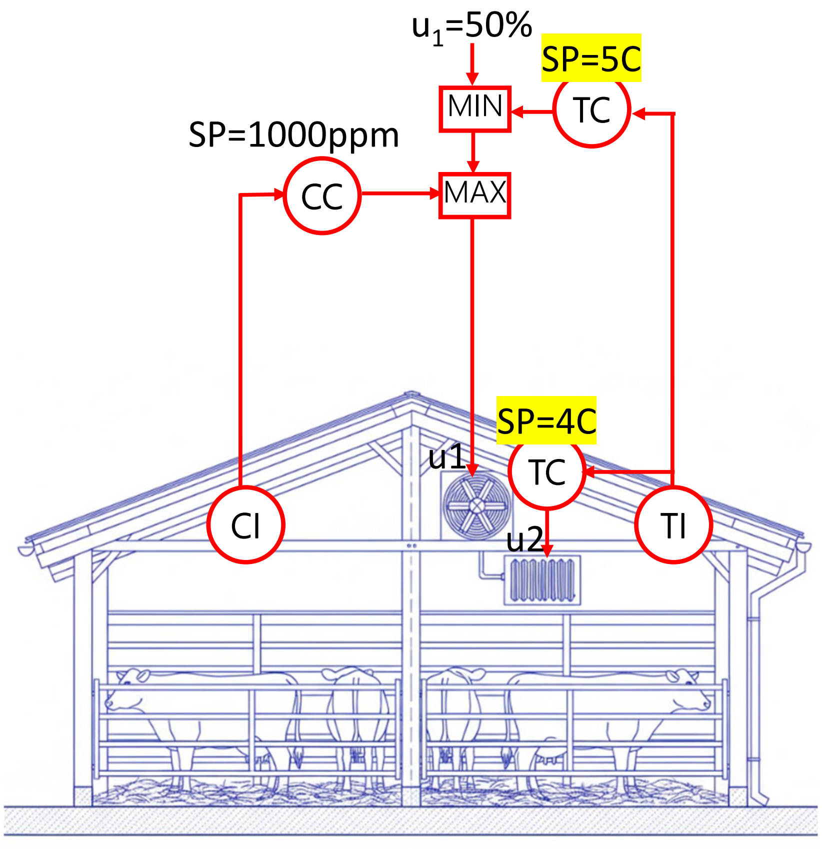
Winter (colder): Cannot reduce fan more because CO2 > 1000 ppm.
Start heater. Use split parallell control (could use split range control instead)

Happy cows
- y1=C (CO2) less than 1000 ppm
- y2=T between 5C and 20C
- Not too much draft or noise from fan
    - 50% fan speed is good

Unhappy cows
- y2=T less than 0C

Even more unhappy
- y1=C above 3000 ppm (poor air quality)

MVs
- u1=fan (cheap)
- u2=heater (expensive)



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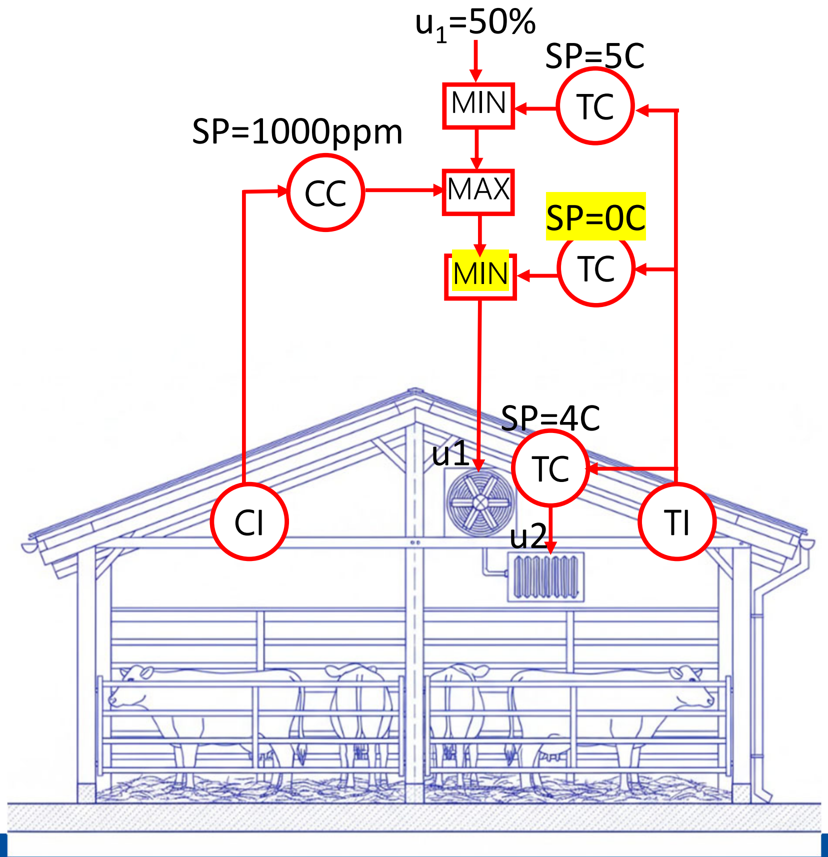
Winter (even colder outide): Heater at max and T drops to 0C.
To avoid freezing unhappy cows: reduce fan speed and accept CO2 > 1000 ppm
Recall: Higher T (above 0C) is satisfied by small fan speed ⇒ MIN-selector

Happy cows
- y1=C (CO2) less than 1000 ppm
- y2=T between 5C and 20C
- Not too much draft or noise from fan
  - 50% fan speed is good

Unhappy cows
- y2=T less than 0C

Even more unhappy
- y1=C above 3000 ppm (poor air quality)

MVs
- u1=fan (cheap)
- u2=heater (expensive)



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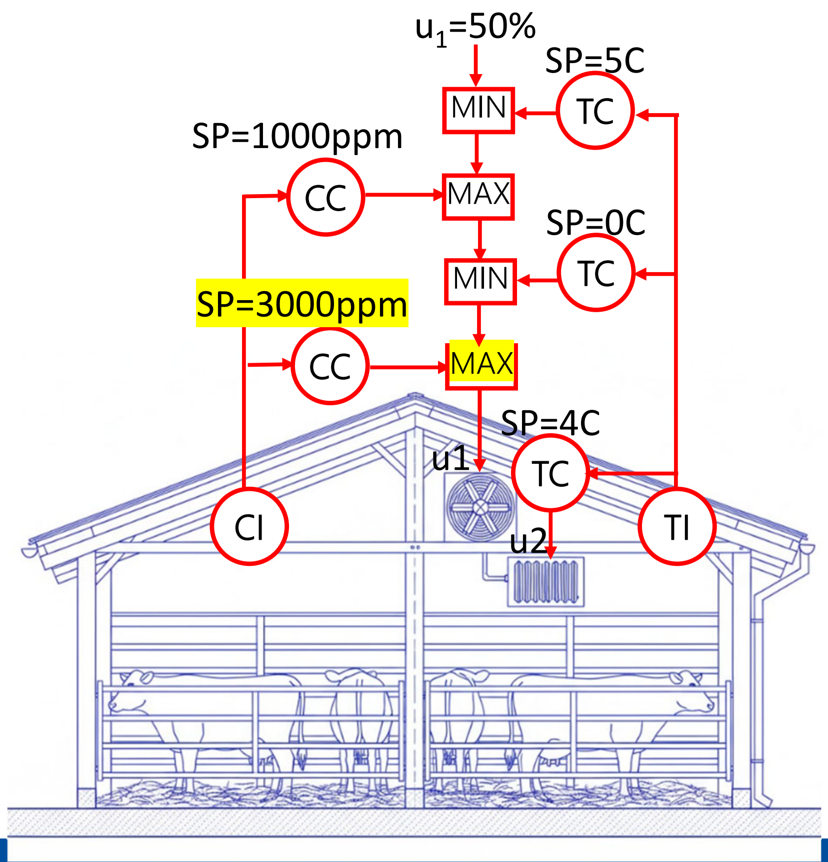
Winter (even colder outside): CO_2 reaches unacepptable level (3000ppm)
Recall: Lower CO_2 is satisfied by large fan speed ⇒ MAX-selector

Happy cows
- y1=C (CO_2) less than 1000 ppm
- y2=T between 5C and 20C
- Not too much draft or noise from fan
  - 50% fan speed is good

Unhappy cows
- y2=T less than 0C

Even more unhappy
- y1=C above 3000 ppm (poor air quality)

MVs
- u1=fan (cheap)
- u2=heater (expensive)



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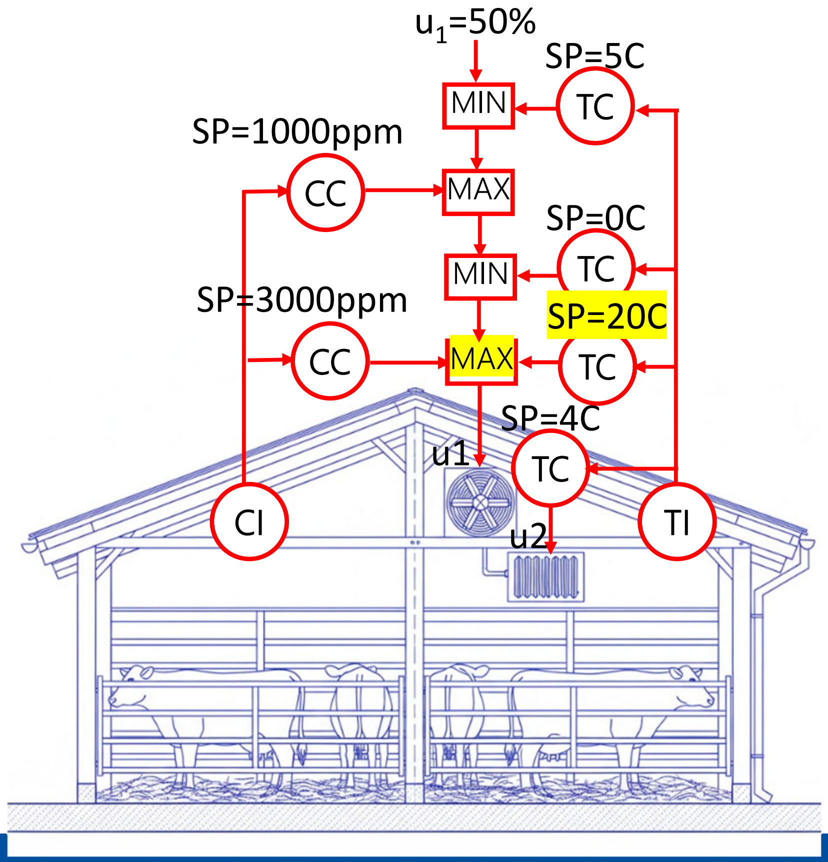
Lower T (below 20C) is satisfied by large fan speed ⇒ MAX-selector
(Consistent with large fan speed (MAX-selector) to keep CO2 < 3000ppm!)

Happy cows
- y1=C (CO2) less than 1000 ppm
- y2=T between 5C and 20C
- Not too much draft or noise from fan
    - 50% fan speed is good

Unhappy cows
- y2=T less than 0C

Even more unhappy
- y1=C above 3000 ppm (poor air quality)

MVs
- u1=fan (cheap)
- u2=heater (expensive)



FINAL OPTIMAL CONTROL STRUCTURE:
4 TCs, 2 CCs, 4 selectors

If extreme cold: Use blankets for cows
If very hot: Spray water on cows

Happy cows

- y1=C (CO2) less than 1000 ppm
- y2=T between 5C and 20C
- Not too much draft or noise from fan
    - 50% fan speed is good

Unhappy cows

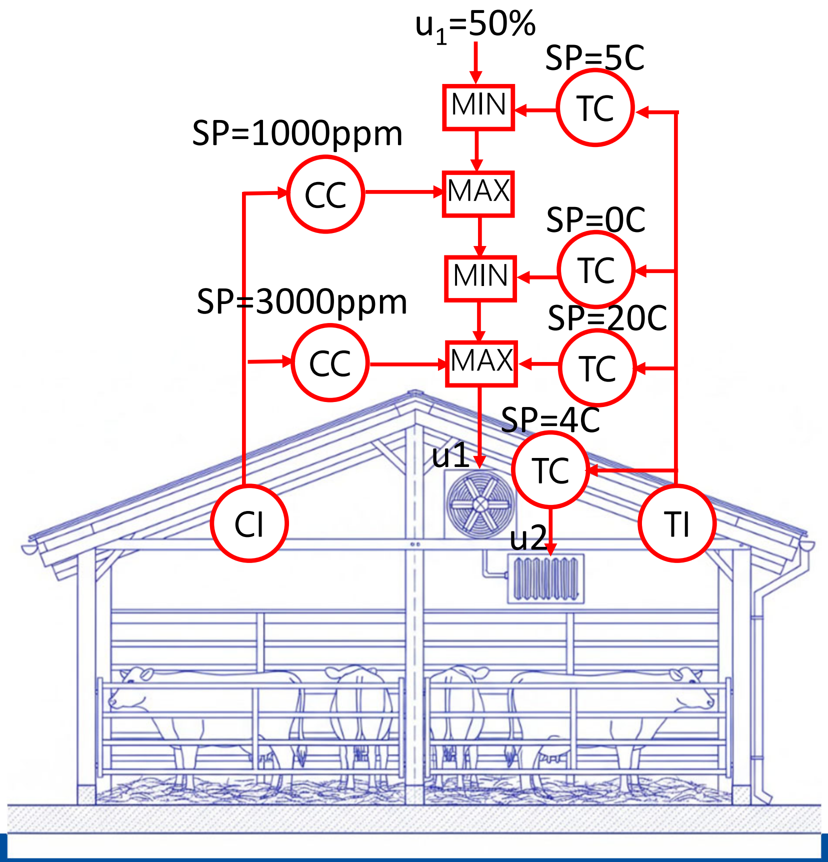
- y2=T less than 0C

Even more unhappy

- y1=C above 3000 ppm (poor air quality)

MVs

- u1=fan (cheap)
- u2=heater (expensive)



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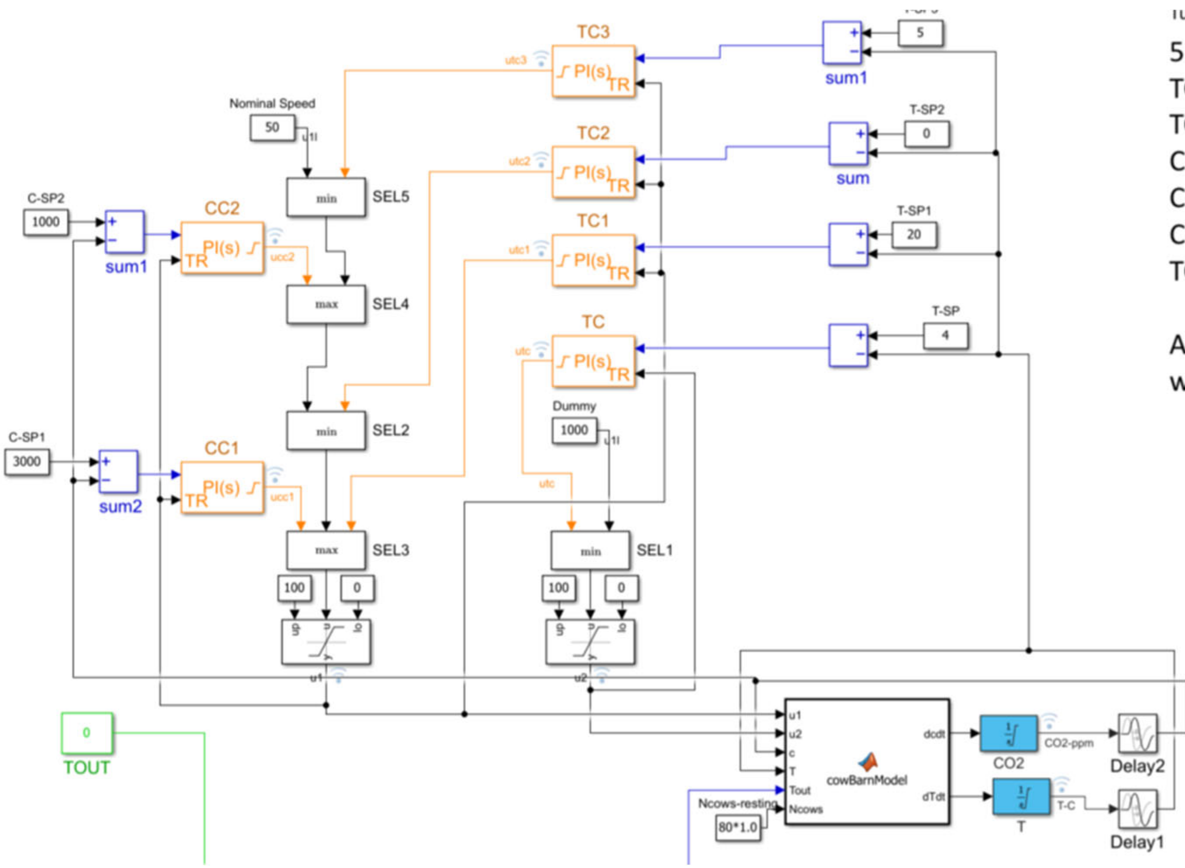
SIMULINK Simulation



Tunings

5 Controllers using u1=fan

TC3, TC1. Kc=-10, taui=350

TC2. Kc=-10/3, taui=350*3

CC2. Kc=-0.1, taui=350

CC1. Kc=-0.1/5, taui=350*5

Controller using u2=heat

TC. Kc=22, taui=350

All controllers use anti-windup with tracking: tauT=taiu

Table 2. Parameters for cow case study II.

Symbol	Value	Description
V	$3\,000\,\mathrm{m}^3$	Air volume of the barn
N_{cows} (nominal)	80	Cow occupancy
$G_{\mathrm{CO_2}}$	$5 \times 10^{-5}\ \frac{m^3/s}{cow}$	Per-cow CO_2 generation
Q_{cow}	$1\,000\ \mathrm{W/cow}$	Per-cow metabolic heat
$q_{\max}, q_{\min}$	$15,\ 0.1\ \mathrm{m}^3/\mathrm{s}$	Fan airflow at 100 % and 0 %
$Q_{\mathrm{h,max}}$	$50\ 000\ \ \mathrm{W}$	Heater rating at 100 %
UA	$2\ 000\ \mathrm{W/K}$	Barn heat-loss coefficient
c_{out}	420 ppm	Outdoor CO_2 concentration

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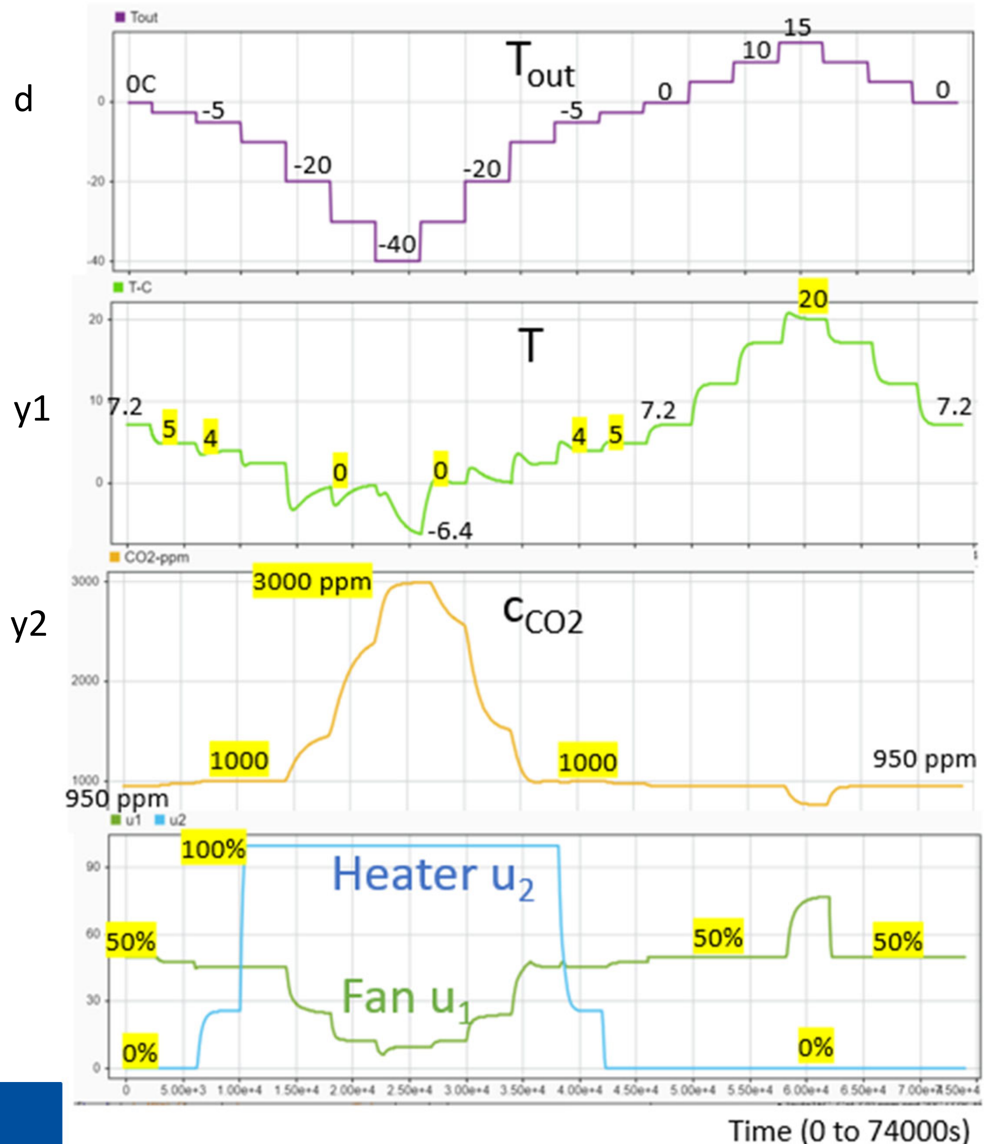

In yellow:
Active constraint/setpoint
(always 2 active)

y1(T): 20C, 5C, 4C, 0C
y2(CO2): 1000 ppm, 3000 ppm
u1(Fan): 50%
u2(Heater): 0%, 100%

Concusion:
Common ARC elements

E1. Cascade control

- Have Extra output (state) measurements

E2. Ratio control

- “Feedforward” for mixing process

E12. Decoupling elements

- Have interactive process

E13. Linearization elements / Adaptive gain

- Have Nonlinear process

E5-E7. Split-range control (or split-parallel control or split-VPC)

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

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

- Have extra inputs dynamically

E4. Selectors

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


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Review article
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Department of Chemical Enginee
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Review
Transformed inputs for linearization, decoupling and feedforward control
Sigurd Skogestad [a,*], Cristina Zotică [a], Nicholas Alsop [b]

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

ARC = Advanced regulatory control
VPC = Valve position control

CV = Controlled variable
MV = manipulated variable

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**) /AI

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

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Review article

Advanced control using decomposition and simple elements

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ARTICLE INFO

Keywords:
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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.

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   2.1. What is control? 6

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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