Advanced process control

Sigurd Skogestad

2025

- ADVANCED PROCESS CONTROL (APC) module 2025
- Need to register in TKP4555 (module) or KP8115 (PhD)
- Ask for office space in K4, 2nd floor

- An introduction to the module is given
- **Tuesday 19 August 2024 10:00-12:00 in room K4.205 (2nd floor in building K4)**
- The rest of the semester the lectures/exercises are planned to be (I hope this is OK for everyone)
  - **Tuesdays 14:15-17:00 in room K4.205 (some days in K4-438)**
- Lectures: Sigurd Skogestad (Professor)
- Exercises: Jonas Fraihat (PhD student)
- Course contents:
- • **Learning outcome:** the students will be able to design REAL plantwide control systems.

- • **Content:**
  - – Control structure design for complete chemical plants.
  - – Optimal economic operation
  - – Selection of economic controlled variables
  - * Active constraints
  - * Self-optimizing control
  - * Gradients as self-optimizing variables.
  - – Advanced regulatory control ("advanced PID control" = decomposition of the control systrem)
  - – Consistent inventory control.
  - – Tuning of PID controllers.
  - – Multivariable control
  - * Decentralized control and RGA.
  - * MPC (when should it be used)
  - – Real-time optimization (RTO)
  - * Feedback implementions

- • **Teaching activities:** Lectures, exercises, computer simulation.
- • **Course material**: Copies from scientific papers and books including
- * New paper (2023) on "Advanced control using decompotion and simple elements": https://folk.ntnu.no/skoge/publications/2023/skogestad-advanced-regulatory-control_arc/
- * Chapter 10 in Skogestad and Postlethwaite, "Multivariable Feedback Control, Wiley, 2010: https://folk.ntnu.no/skoge/book/ps/
- See also here for more information: https://folk.ntnu.no/skoge/vgprosessregulering/

Course information

- Lectures + industrial guest lectures (TBD)
  - Time: Tuesdays 14-16 (K4.205)
- 6 exercises + help sessions
  - Sessions : Tuesday 16-17 (K4.205) ++
- Exercises count 20% of the grade of the module

Course Summary

This course is about how to operate and control complete chemical plants

- Part 0: Introduction and motivation
- Part 1 : Plantwide control
- Part 2 : Self-optimizing control
- Part 3 : PID tuning (SIMC rules)
- Part 4 : Inventory control and TPM
- Part 5: Supervisory control and switching
- Part 6 : ARC elements
- Part 7: More on switching and regulatory control
- Part 8: Nonlinear steady-state models: Transformed inputs, RTO and Extremum seeking control.

Course Plan 2025

Lecturer: Sigurd Skogestad (skoge@ntnu.no)
Exercises: Jonas Fraihat
First lecture (week 34): Tuesday 19 August 10:00-12:00 in K4-205
Normal Lecture time: Tuesdays 09-12 in K4-205
Exercise: usually 11-12 the day of the lectures
Note, that the days of the lecture may change.

Week/Date	Lecture	Exercise
Week 34 / 19.08.	0. Introduction 1. Plant-wide control procedure	Exercise 1 out (2 weeks)
Week 35 / 26.08.	1. Continue 2. Self-optimizing control	
Week 36 / 02.09.	2. Self-optimizing control, continmue	Exercise 1 deadline Exercise 2 out (1 week)
Week 37 / 09.09.	3. Controller tuning	Exercise 2 deadline Exercise 3 out (1 week)
Week 38 / 16.09.	4. Inventory control and TPM 5. Supervisory control / switching	Exercise 3 deadline Exercise 4 out (2 weeks)
Week 39 / 23.10.	6. ARC elements	
Week 40 / 30.09.	7. More on switching and regulatory control	Exercise 4 deadline Exercise 5 out (2 weeks)
Week 41 / 07.10.	8. Transformed inputs / Guest lecture MPC	
Week 42 / 14.10.	9. Guest lecture Forsman	Exercise 5 deadline Exercise 6 out (2 weeks)
Week 43 / 21.10.		
Week 44 / 28.10.		Exercise 6 deadline

There will be two guest lectures given by

- TBA: MPC application in Equinor
- Krister Forsman: Advanced process control in Perstorp

Part 0: Introduction and motivation

"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]
- 05 Jan. 2023: Tutorial paper on "Transformed inputs for linearization, decoupling and feedforward control" published in JPC. [paper]
- 13 June 2022: Plenary talk on "Putting optimization into the control layer using the magic of feedback control", at ESCAPE-32 conference, Toulouse, France [slides]
- 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...

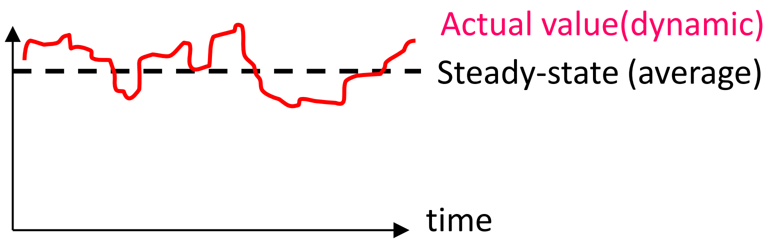
- Book: S. Skogestad and I. Postlethwaite: MULTIVARIABLE FEEDBACK CONTROL-Analysis and design. Wiley (1996; 2005)
- Book: S. Skogestad: CHEMICAL AND ENERGY PROCESS ENGINEERING CRC Press (Taylor&Francis Group) (Aug. 2008)
- Bok: S. Skogestad: PROSESSTEKNIKK- Masse- og energibalanser Tapir (2000; 2003;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

Why do we need control?

- *Operation*



In practice steady state doesn't happen:

- Feed changes
- Startup
- Operator changes
- Failures
- …..

"Disturbances" (d's)

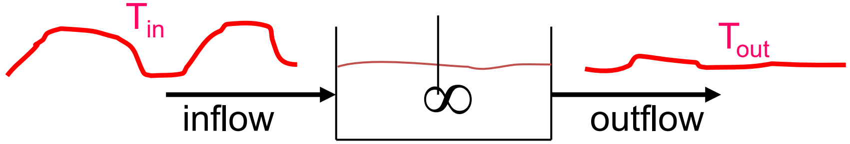
- Control is needed to reduce the effect of **disturbances**
- 30% of investment costs are typically for instrumentation and control

Countermeasures to disturbances

I. Eliminate/reduce the disturbance

(a) Design process so it is insensitive to disturbances

- Example: Use buffer tank to dampen disturbances



(b) Detect and remove source of disturbances

- "Statistical process control"
- Example: Detect and eliminate variations in feed composition

Countermeasures to disturbances

II. Process control

Do something (usually manipulate a valve) to

counteract the **effect** of the disturbance

(a) Manual control: Need operator

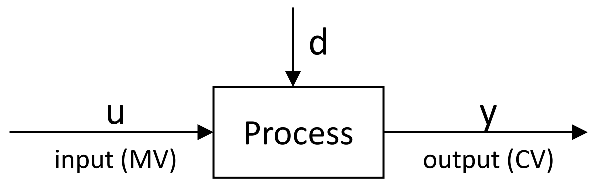
(b) Automatic control: Need measurement + automatic valve + computer

Goals automatic control:

- Smaller variations
  - » more consistent quality
  - » More optimal
- Smaller losses (environment)
- Lower costs
- More production

Industry: Still large potential for improvements!

Classification of variables



Independent variables ("the cause"):

(a) Inputs (MV, u): Variables we can adjust (valves)

(b) Disturbances (DV, d): Variables outside our control

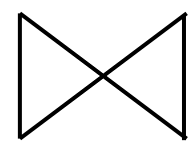
Dependent (output) variables ("the effect or result"):

(c) Primary outputs (CVs, y_1): Variables we want to keep at a given setpoint

(d) Secondary outputs (y_2): Extra measurements that we may use to improve control

Inputs for control (MVs)

- Usually in process control: Inputs (MVs) are valves.
  - Physical input is valve position (z), but we often simplify and say that flowrate (q) is input



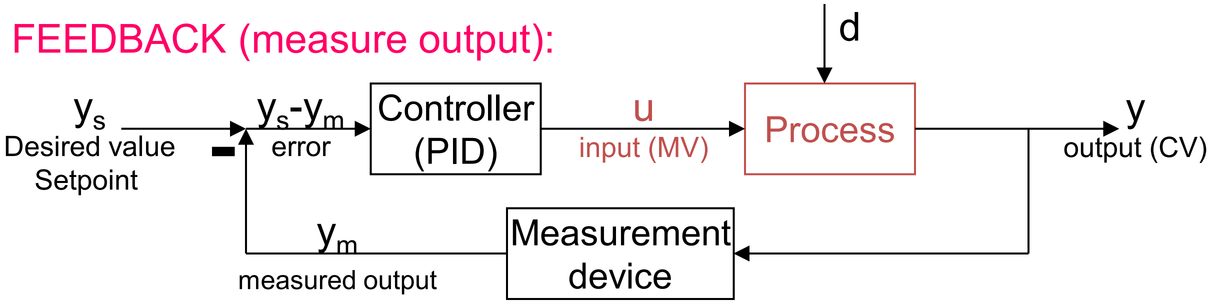
- Valve equation: $q\left(\frac{m^3}{s}\right) = c_v f(z)\sqrt{\Delta p/\rho}$
  - Δp = pressure drop over valve (disturbance; can be counteracted with flow controller)

We use two kind of diagrams

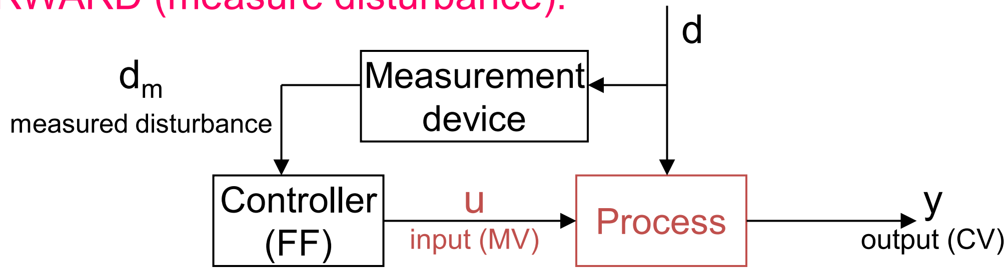
- **Block diagram (information)**
  - Used by control engineers
- **Flowsheet (piping & instrumentation diagram, P&ID)**
  - Used by process engineers

BLOCK DIAGRAMS

FEEDBACK (measure output):



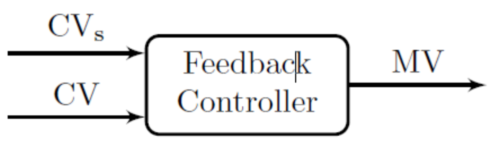
FEEDFORWARD (measure disturbance):



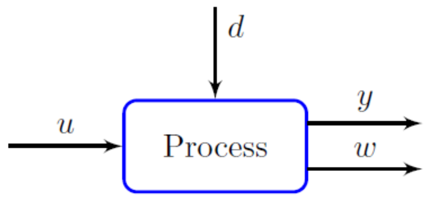
- All lines: Signals (information)
- Blocks: controllers and process
- Do not confuse block diagram (lines are signals) with flowsheet (lines are flows); see below

BLOCK DIAGRAMS

Two DOF feedback controller



MV = manipulated variable (controller output)
CV = controlled variable (controller input)



u = process input (adjustable)
d = disturbance (non-adjustable)
y = process output (with setpoint y_s)
w = extra measured process variable (state x, y_2)

Simple Combination:

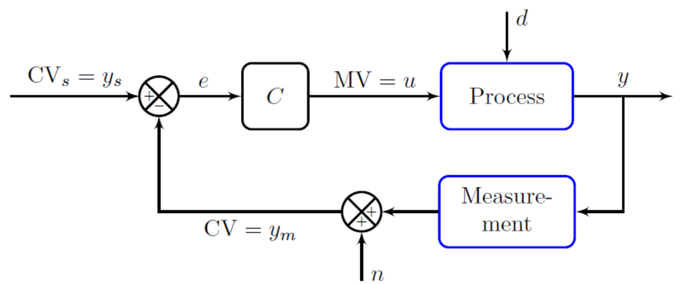


Figure 3: Block diagram of common "one degree-of-freedom" negative feedback control system.

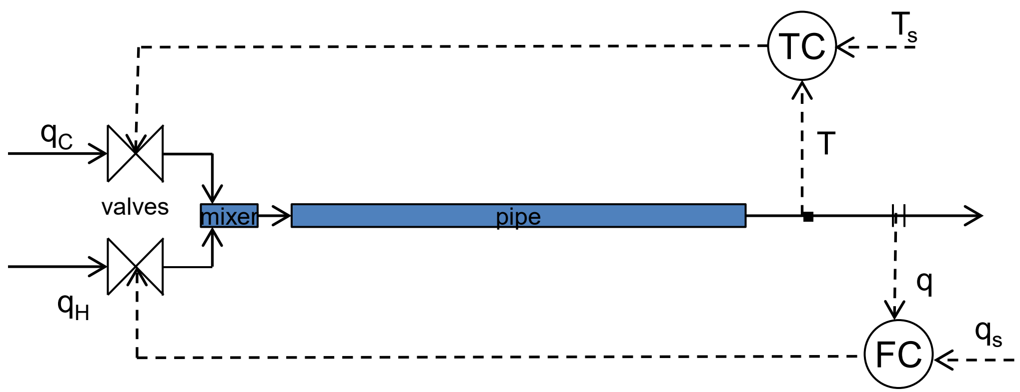
$$e = y_s - y_m$$

Block diagrams: All lines are signals (information)

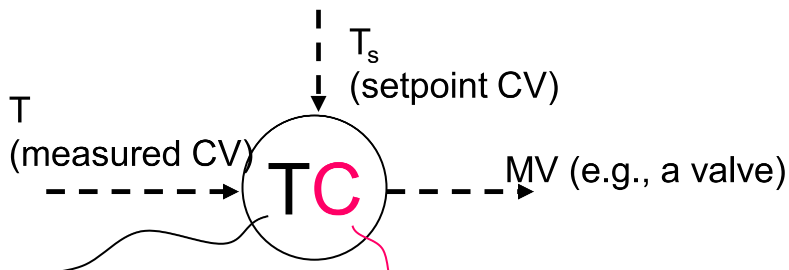
Piping and instrumentation diagram (P&ID) (flowsheet)

- Solid lines: mass flow (streams)
- Dashed/colored lines: signals (control)

Example: Shower



Notation feedback controllers (Flowsheet = P&ID)

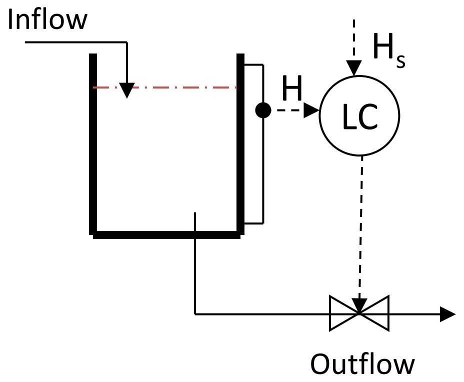


1st letter: Controlled variable (CV)
- **T: temperature**
- **F: flow**
- **L: level**
- **P: pressure**
- DP: differential pressure (Δp)
- A: Analyzer (composition)
- C: composition
- X: quality (composition)
- H: enthalpy/energy

…

2nd letter:
- C: controller
- I: indicator (measurement)
- T: transmitter (measurement)

Example flowsheet: Level control



INPUT (u): OUTFLOW (Input for control!)
OUTPUT (y): LEVEL
DISTURBANCE (d): INFLOW

SIMC* PID tuning rule (2001,2003)

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.

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

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

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

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

$$\tau_D = \tau_2$$

Tuning parameter:

$$\tau_c \geq \theta$$

λ

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*

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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.

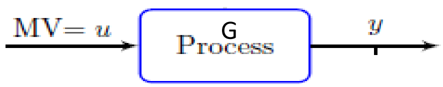
*SIMC = Simple/Skogestad IMC

Scaling

- Variables (u=MV,y=CV,d=DV) are usually scaled
  - Simplifies interpretation and tuning
  - Make sure that valves actually go to fully open or closed
- Linear systems (block diagrams): Variables usually «shifted» with 0 as nominal value, Typical scaled range: -1 to 1
  - where -1 is min. value, 1 is max. value
- Industry (flowsheets): Usually scaled 0-1 or 0-100%.
  - where 0 is min. value (e.g., closed valve)
  - 1 or 100% is max. value (fully open valve)
  - Usually no nominal value

What is best? Feedback or feedforward?

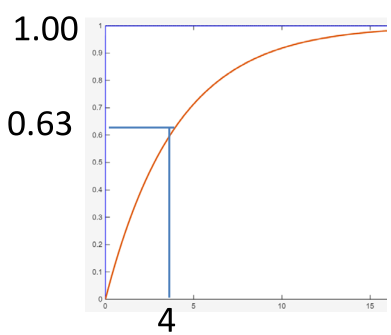
Example: Feedback vs. feedforward for setpoint control of uncertain process



y = G(s) u

$$G(s) = \frac{k}{\tau s + 1}, \quad k = 3,\ \tau = 6 \qquad \text{(B.2)}$$

$$\boxed{\text{Desired response : } \quad y = \frac{1}{\tau_c s + 1} y_s = \frac{1}{4s+1} y_s}$$



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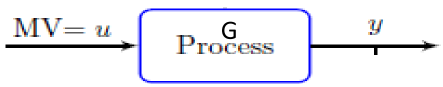

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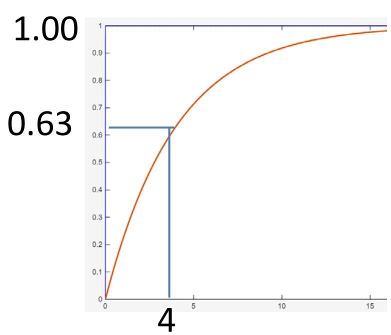


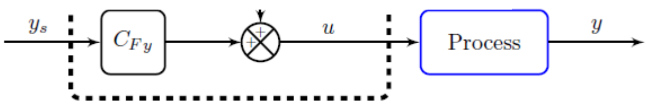


Figure A.42: Block diagram of feedforward control system with linear combination of feedforward from measured disturbance (d) and setpoint (y_s) (E14).

Feedforward solution. We use feedforward from the setpoint (Fig. A.42):

$$u = C_{Fy}(s) y_s$$

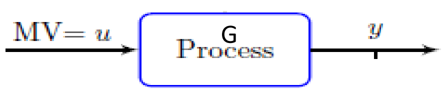
where we choose

$$C_{Fy}(s) = \frac{1}{\tau_c s + 1} G(s)^{-1} = \frac{1}{k} \frac{\tau s + 1}{\tau_c s + 1} = \frac{1}{3} \frac{6s+1}{4s+1} \qquad \text{(B.3)}$$

The output response becomes as desired,

$$y = \frac{1}{4s+1} y_s \qquad \text{(B.4)}$$

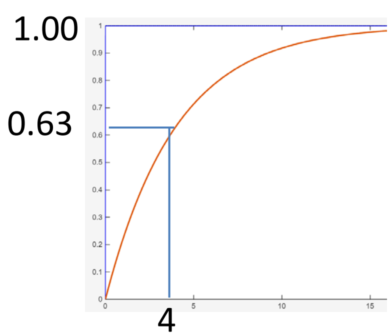
Example: Feedback vs. feedforward for setpoint control of uncertain process



y = G(s) u

$$G(s) = \frac{k}{\tau s + 1}, \quad k = 3, \ \tau = 6 \qquad \text{(B.2)}$$

$$\boxed{\text{Desired response :} \quad y = \frac{1}{\tau_c s + 1} y_s = \frac{1}{4s+1} y_s}$$



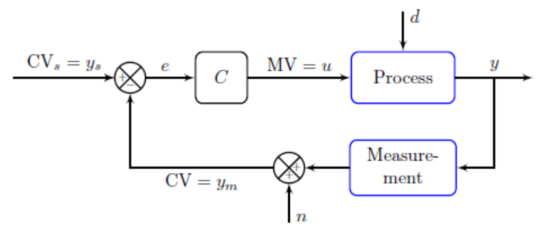


Figure 3: Block diagram of common "one degree-of-freedom" negative feedback control system.

Feedback solution. We use a one degree-of-freedom feedback controller (Fig. 3) acting on the error signal $e = y_s - y$:

$u = C(s)(y_s - y)$

We choose a PI-controller with $K_c = 0.5$ and $\tau_I = \tau = 6$ (using the SIMC PI-rule with $\boxed{\tau_c = 4}$ see Appendix C.2):

$$C(s) = K_c \left(1 + \frac{1}{\tau_I s}\right) = 0.5 \frac{6s+1}{6s} \qquad \text{(B.5)}$$

Note that we have selected $\tau_I = \tau = 6$, which implies that the zero dynamics in the PI-controller C, cancel the pole dynamics of the process G. The closed-loop response becomes as desired:

$$y = \frac{1}{\tau_c s + 1} y_s = \frac{1}{4s+1} y_s \qquad \text{(B.6)}$$

Proof. $y = T(s) y_s$ where $T = L/(1+L)$ and $L = GC = kK_c/(\tau_I s) = 0.25/s$. So $T = \frac{0.25/s}{1+0.25/s} = \frac{1}{4s+1}$.

Thus, we have two fundamentally different solutions that give the same nominal response, both in terms of the process input $u(t)$ (not shown) and the process output $y(t)$ (black solid curve in Fig. B.43).

- But what happens if the process changes?
  - Consider a gain change so that the model is wrong
    - Change process gain from k=3 to k'=4.5

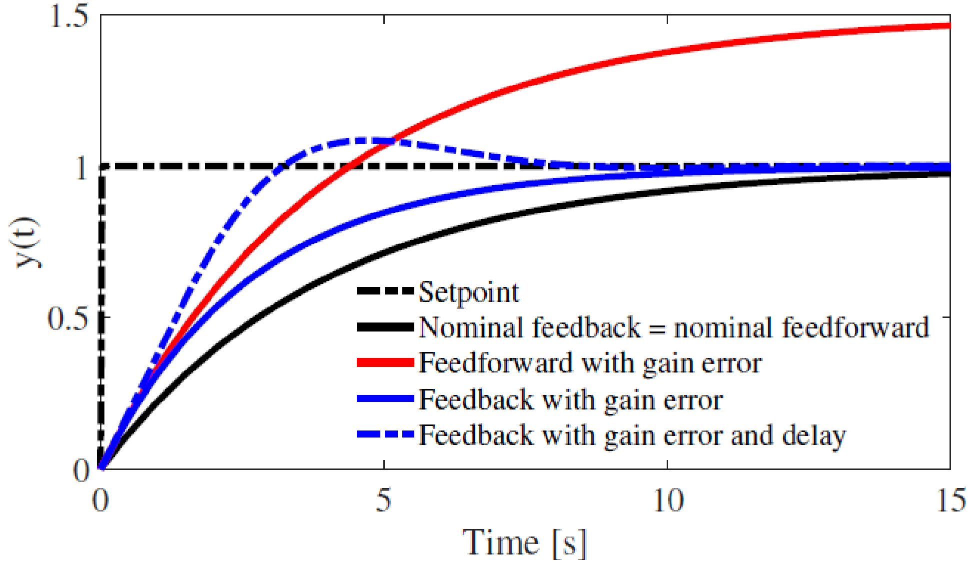


Figure B.43: Setpoint response for process (B.2) demonstrating the advantage of feedback control for handling model error.

Gain error (feedback and feedforward): From k=3 to k'=4.5
Time delay (feedback): From $\theta = 0 \; to \; \theta = 1.5$

Conclusion: Feedback-based solutions (PID) are generally more robust then feedforward-based solutions (MPC)

FEEDBACK

+ Self-correcting with negative feedback (keeps adjusting until $y=y_s$ at steady state)
+ Do not need good model (but must know process sign!)
- May give instability if controller overreacts
- Need good and fast measurement of output

MAIN ENEMY OF FEEDBACK: TIME DELAY
(in process or in measurement of y)

FEEDFORWARD

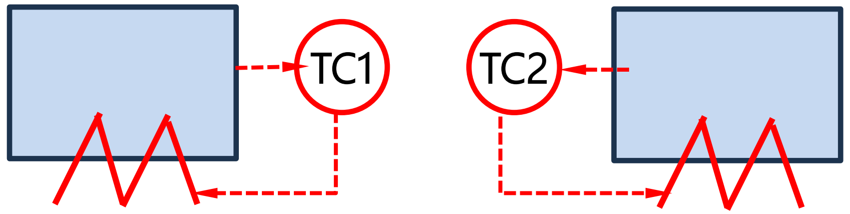
+ Consider when large time delay (in process or in measurement of y)
+ May react before damage is done
- Need good model
- Sensitive to changes and errors
- Works only for known and measured disturbances

USUALLY COMBINED WITH FEEDBACK

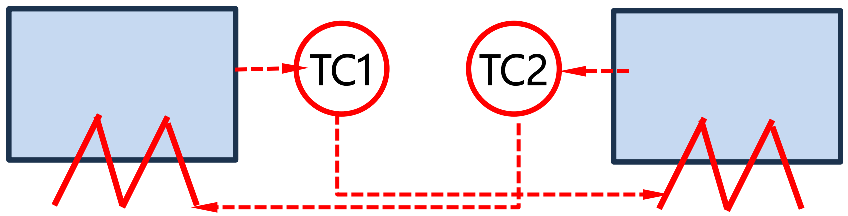
NOTE

- «That something works doesn't mean that it couldn't be much better or simpler (PID), or even both better and simpler at the same time».
- Example: Sensor in wrong room

1. PI control of room temperature in two different buildings:



2. Ooopss….. Can it work? (by retuning TC1 and TC2)



NOTE

- «That something works doesn't mean that it couldn't be much better or simpler (PID), or even both better and simpler at the same time».
- Example: Sensor in wrong room

MULTIVARIABLE FEEDBACK CONTROL
Analysis and Design

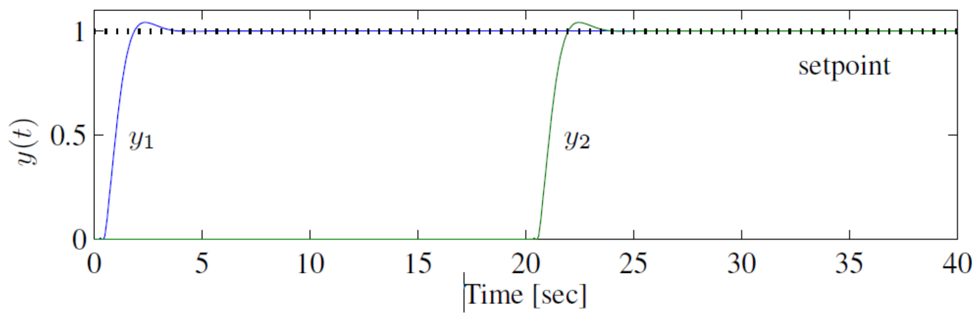
Second Edition

Sigurd Skogestad
Norwegian University of Science and Technology

Ian Postlethwaite
University of Leicester

$$G_1 = \begin{bmatrix} g_{11} & g_{12} \\ g_{21} & g_{22} \end{bmatrix} = \begin{bmatrix} 1 & 0 \\ 0 & 1 \end{bmatrix}$$

+ delay θ=0.5



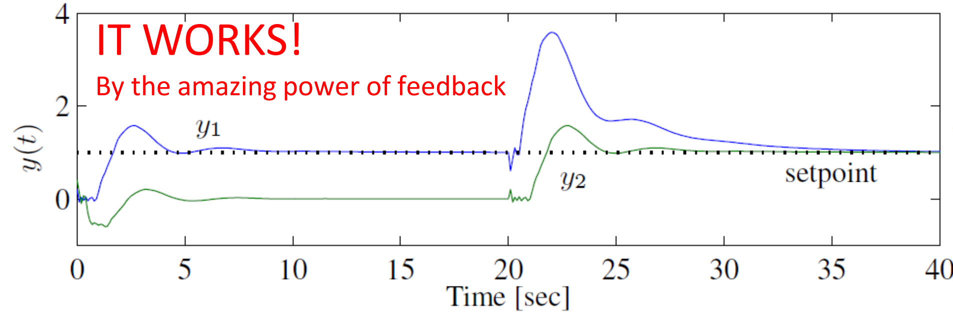
(a) Diagonal pairing; controller (10.51) with $\tau_1 = \tau_2 = 1$

$$K = \begin{bmatrix} \frac{1}{\tau_1 s} & 0 \\ 0 & \frac{1}{\tau_2 s} \end{bmatrix}$$

(SIMC tuning, $\tau_c = \theta = 0.5$))

$$G_2 = \begin{bmatrix} g_{12} & g_{11} \\ g_{22} & g_{21} \end{bmatrix} = \begin{bmatrix} 0 & 1 \\ 1 & 0 \end{bmatrix}$$

+ delay θ=0.5



(b) Off-diagonal pairing; plant (10.53) and controller (10.54)

$$K^*(s) = \begin{bmatrix} \frac{-(0.5s+0.1)}{s} & 0 \\ 0 & \frac{(0.5s+2)}{s} \end{bmatrix}$$

$$G_{\text{sim}} = Ge^{-0.5s}.$$

Figure 10.15: Decentralized control of diagonal plant (10.50)

Process systems engineering

- About getting the big picture without loosing the details

- Studying and combining different length scales (nm to km)
- … different time scales

- Also: … combining physical models and data

- Classical solution: Make «boxes»

- Get rid of the boxes? Tempting, but it's not a good idea

- This course: Break the optimization into smaller pieces (boxes) and make use of process control for solving

Course overview

Part 1: Plantwide control

Introduction to plantwide control (what should we really control?)

Introduction

- Two main objectives for control: Longer-term economics (CV1) and shorter-term stability (CV2)
- Regulatory (basic) and supervisory (advanced) control layer

Optimal operation (economics)

- Define cost J and constraints
- Active constraints (as a function of disturbances)
- Selection of economic controlled variables (CV1). Self-optimizing variables.

Part 2: Self-optimizing control theory

- Active constraints

- Remaining Unconstrained degrees of freedom:

  - Ideal CV1 = Gradient (J_u)
  - Nullspace method
  - Exact local method
  - Link to other approaches
    - Gradient estimation methods
  - Examples

Part 3: PID tuning

PID controller tuning: It pays off to be systematic!

- Derivation SIMC PID tuning rules
  - Controller gain, Integral time, derivative time

- Obtaining first-order plus delay models
  - Open-loop step response
  - From detailed model (half rule)
  - From closed-loop setpoint response

- Special topics
  - Integrating processes (level control)
  - Other special processes and examples
  - When do we need derivative action?
  - Near-optimality of SIMC PID tuning rules

- Examples

Part 4: Regulatory («stabilizing») control

Inventory (level) control structure

- Location of throughput manipulator
- Consistency and radiating rule

Structure of regulatory control layer (PID)

- Selection of controlled variables (CV2) and pairing with manipulated variables (MV2)
- Main rule: Control drifting variables and "pair close"

Summary: Sigurd’s rules for plantwide control

Part 5

- Putting optimization into the control layer
- Supervisory control and switching

Part 6: Advanced control elements

Advanced control layer

- Design based on simple elements:
  - Ratio control
  - Cascade control
  - Selectors
  - Input resetting (valve position control)
  - Split range control
  - Decouplers (including phsically based)
  - When should these elements be used?
- When do we use MPC instead?

Case studies

- Example: Distillation column control
- Example: Plantwide control of complete plant Recycle processes: How to avoid snowballing

- Part 7: More on switching and inventory control
- Part 8: Transformed inputs