Lynkurs Prosessregulering
(Crash course process control)

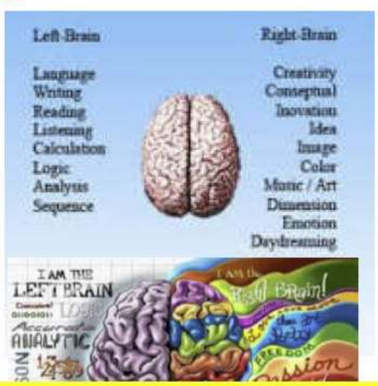
Sigurd Skogestad
Institutt for kjemisk prosessteknologi
Rom K4-211
skoge@ntnu.no

TKP4140. Process control .

In this course you must use both sides of your brain – and try to connect them in the end!

2. Then we switch to the left side:
Modelling - mathematics,
- Laplace - …….

1. We start towards the right side (first two weeks
Control structures – concepts- intuition



It will be a lot of work, but it will be fun, and the goal is that you in the end can design real control systems that work well!

Course for 3rd year students:

- Pensum (syllabus): Lectures/exercises

  Literature (see www.nt.ntnu.no/users/skoge/prosessregulering_lynkurs):
  - 1. Nybraaten og Svendsen, "Kort innføring i prosessregulering" (1986)
    (det kan synes gammelt, men det står faktisk mye bra her)
  - 2. 12 sider fra F. Haugen, "Anvendt reguleringsteknikk", 1992
  - 3. S. Skogestad, "Prosessteknikk", 2./3. utgave, Tapir: Kap. 11.3 (tidsrespons) og 11.8/11.6 (prosessregulering)
  - 3. S. Skogestad, "Chemical and Energy Process Engineering", CRC Press, 2009: Ch. 11.3 (Dynamic analysis and time response) + Ch. 11.6 (Process control)
  - Slides

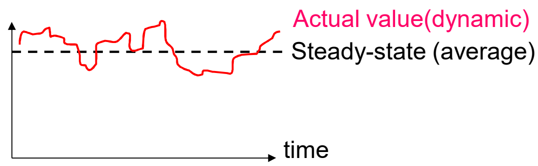
- Forelesningsplan
- F1: Oversikt over regulering, forover- og tilbakekobling, dusjeksempel
- F2: Klassifisering av variable, prosedyre for utforming av reguleringssystem
- F3: Eksempler
- F4: Prosessdynamikk, tidskonstant, dødtid, PID-regulering,
- F5/F6: Stabilitet, Tuning PID, Forsøk, Eksempler,

More information (literature, old exams, etc.):
- www.nt.ntnu.no/users/skoge/prosessregulering_lynkurs

English	Norsk	English	Norsk
Control	regulering		
Operation	drift	Loop	Sløyfe
Measurement	måling	Valve	Ventil
Disturbance DV	forstyrrelse	Gain	Forsterkning
Manipulated var. (MV) = input	Pådrag = inngang	Time delay = dead time (θ)	Tidsforsinkelse = dødtid (θ)
Controlled variab. (CV) = output	Regulert variabel = utgang		
Feedback	Tilbakekobling		
Feedforward	Foroverkobling		
Controller	Regulator		

Why control?

- Until now: *Design* of process. Assume steady-state
- Now: *Operation*



In practice never steady-state:

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

"Disturbances" (d's)

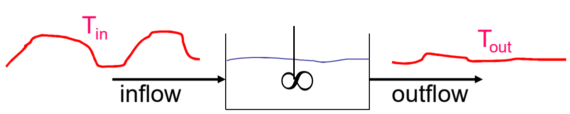
- Control is needed to reduce the effect of disturbances – remain at steady state
- 30% of investment costs are typically for instrumentation and control

Countermeasures to disturbances (I)

I. Reduce/Eliminate the disturbance

(a) Design process so it is insensitive to disturbances

- Example: Use buffertank to dampen disturbances



(b) Detect and remove source of disturbances

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

Countermeasures to disturbances (II)

II. Counteract the disturbance: Process control *(prosessregulering)*

Do something (usually manipulate valve) to **counteract** the **effect** of the disturbances

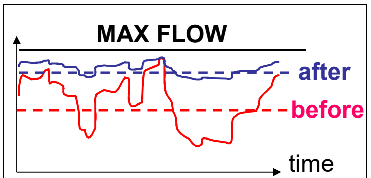
(a) Manual control: Need operator

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

Goals automatic control:

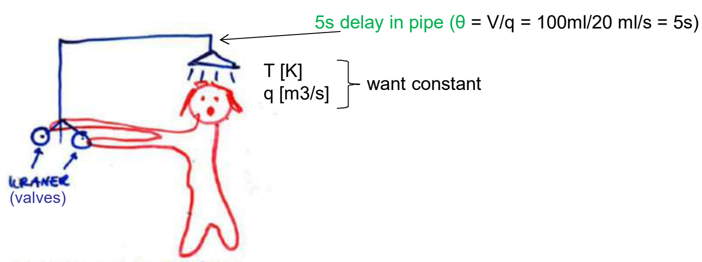
- Smaller variations
  - more consistent quality
  - More optimal ("squeeze and shift")
- Smaller losses (environment)
- Lower costs
- More production

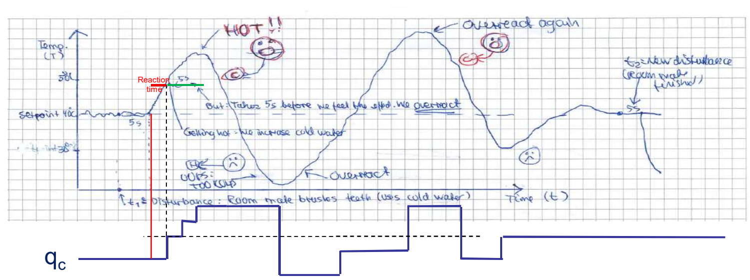
Industry: Still large potential for improvements!



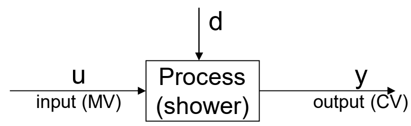
By improving control and squeezing the variations we can shift the setpoint (average) closer to the constraint and increase production

Example: Control of shower temperature





Classification of variables



Independent variables ("the cause"):

- **(a) Inputs (MV, u):** Variables we can adjust (valves)
- **(b) Disturbances (DV, d):** Variables outside our control

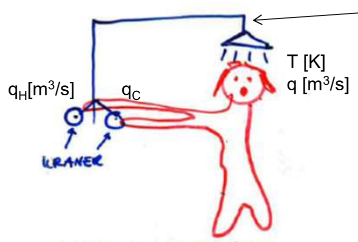
Dependent (output) variables ("the effect or result"):

- (c) **Primary outputs (CVs, y):** Variables we want to keep at a given setpoint
- (d) Internal variables in dynamic model ("states") (x)

MV = manipulated vartiable (input u)
CV = controlled variable (output y)
DV = disturbance variable (d)

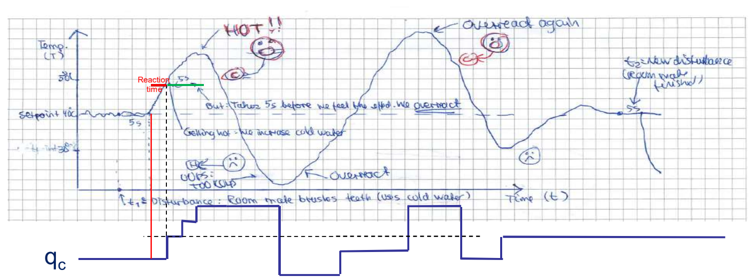
Example: Control of shower temperature

5s delay: $\theta = V/q$ = 100ml/20 ml/s = 5s



Control objective. MVs, CVs, DVs

1. Control objective
   Keep temperature (y_1=T) a given setpoint
   Keep flow (y_2=q) ("pressure") at given setpoint
2. Classify variables
   MVs (u) = q_H, q_C (strictly speaking, valve positions z_H, z_C)
   CVs (y) = T, q
   DVs (d) = q_H, q_C (strictly speaking, upstream pressure which gives "uncontrolled" flow changes)

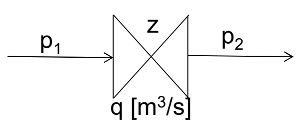


Inputs for control (MVs)

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

Valve equation: $q[m^3/s] = C_v f(z)\sqrt{DP/\rho}$

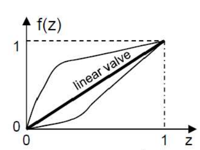
Valve equation



Valve equation: $q[m^3/s] = C_v f(z)\sqrt{DP/\rho}$

$z \in [0,1]$ - valve opening (adjustable), $z = 0$: closed. $z = 1$: fully open.
$q[m^3/s]$ - volumetric flowrate
$DP = p_1 - p_2$ [N/m^2] - pressure drop over valve (Typical value: $DP = 0.1$ bar)

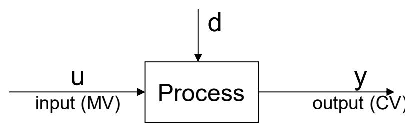
$C_v[m^2] = C_d A$ - valve coeffisient
C_d - dimensionless valve constant (Typical value: $C_d = 1$)
$A[m^2]$ = valve cross-sectional area
$f(z) \in [0,1]$ - valve characteristic (Linear valve has $f(z) = z$)
ρ [kg/m^3] - fluid density (e.g., 1000 kg/m^3 for water; 1.19 kg/m^3 for air at 1 bar/25C)



Comment. Mass flowrate: $w[kg/s] = \rho q = C_v f(z)\sqrt{\rho \cdot DP}$

Control

- Use inputs (MVs, u) to counteract the effect of the disturbances (DVs, d) such that the outputs (CVs, y) are kept close to their setpoints (y_s)



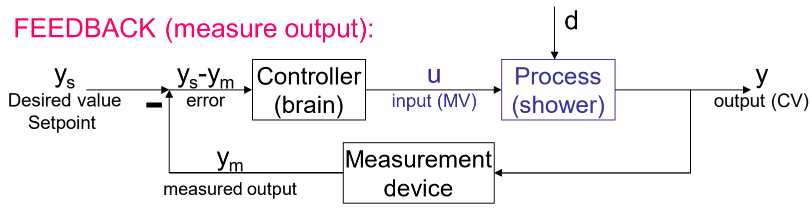
Two fundamental control principles

- **Feedback**: **Measure the result** (= controlled variable CV; output y) and keep adjusting the manipulated variable (MV; input u) until the results is OK
  - Example: Measure the temperature T (CV) and adjust the flow of cold water (MV)

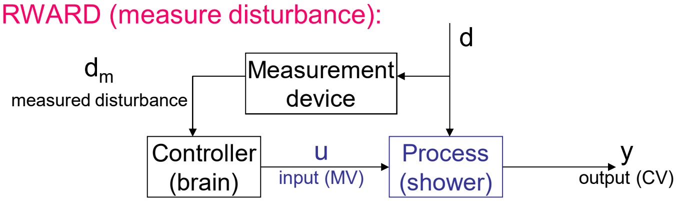
- **Feedforward**: **Measure the cause** (= disturbance d; DV) and based on a prediction (model!) make a "forward" adjustment of the MV (input u) to (hopefully) counteract its effect on the result (output y)
  - Example: Room mate (disturbance d) says "I am tapping cold water" - and you know your friend so well (model) that you can make the correct increase in your cold water (MV) to counteract d.
  - NOT VERY REALISTIC FOR SHOWER EXAMPLE
  - BUT a good example of feedforward is coming in time to lecture!

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

FEEDBACK

+ Self-correcting with negative feedback (keeps adjusting until $y=y_s$ at steady state)
+ Do not need model (but most 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 output measurement)

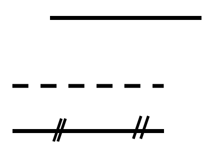
FEEDFORWARD

+ Good when large time delay (in process or output measurement)
+ Reacts before damage is done
- Need good model
- Sensitive to changes and errors
- Works only for known and measured disturbances

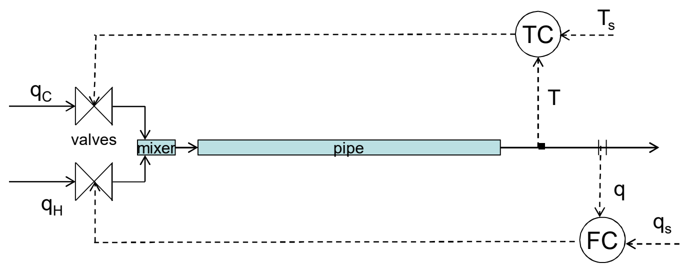
USUALLY COMBINED WITH FEEDBACK

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

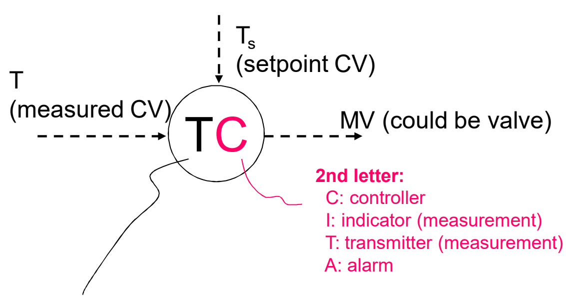
- Solid lines: mass flow (streams)
- Dashed lines: signals (control)



Example: Shower



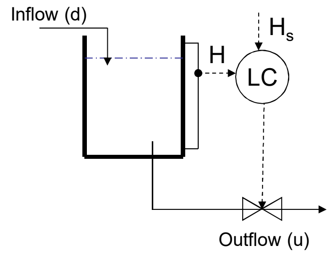
Notation feedback controllers (P&ID)



1st letter: Controlled variable (CV) = What we are trying to control (keep constant)

T: temperature
F: flow
L: level
P: pressure
DP: differential pressure (Δp)
A: Analyzer (composition)
C: composition
X: quality (composition)
H: enthalpy/energy

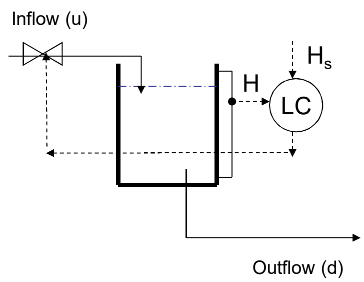
Example: Level control



CLASSIFICATION OF VARIABLES FOR CONTROL (MV, CV. DV):

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

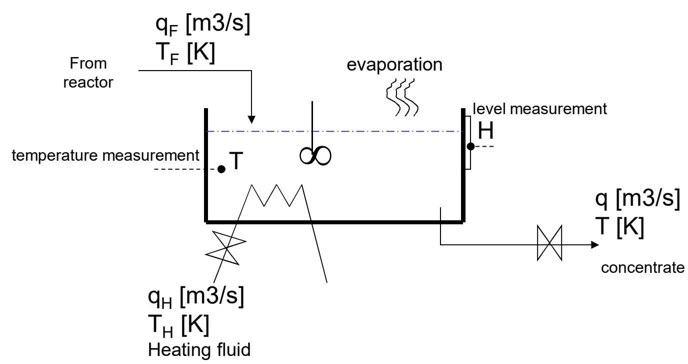
Level control when product rate is given (less common)



CLASSIFICATION OF VARIABLES FOR CONTROL (MV, CV. DV):

INPUT (u, MV): INFLOW
OUTPUT (y, CV): LEVEL
DISTURBANCE (d, DV): OUTFLOW

Example: Evaporator with heating



1. Control objective
   - Keep level H at desired value
   - Keep temperature T at desired value
2. Classify variables (CVs, MVs, important DVs)
3. Process matrix (from MVs to CVs)
4. Suggest pairings and put control loops on the flowsheet

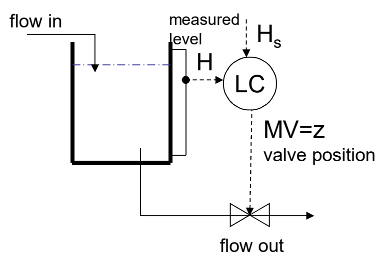
Most important control structures

1. Feedback control
2. Cascade control
3. Ratio control (special case of feedforward)

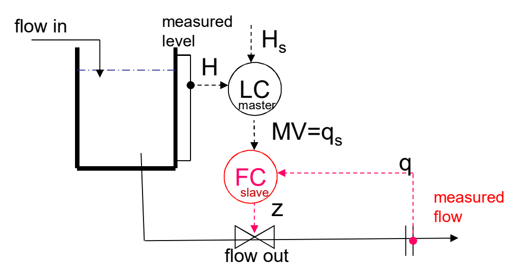
Cascade control

- **Controller ("master") gives setpoint to another controller ("slave")**
  - Without cascade: "Master" controller directly adjusts u (input, MV) to control y
  - With cascade: Local "slave" controller uses u to control "extra"/fast measurement (y').
    "Master" controller adjusts setpoint y'_s.
- **Example: Flow controller on valve (very common!)**
  - y = level H in tank (or could be temperature etc.)
  - u = valve position (z)
  - y' = flowrate q through valve

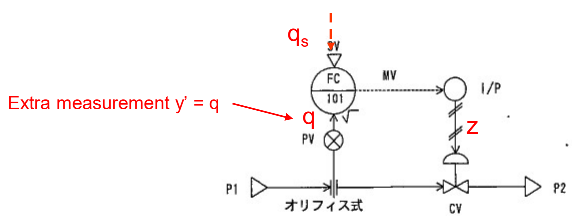
WITHOUT CASCADE



WITH CASCADE

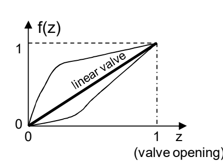


What are the benefits of adding a flow controller (inner cascade)?

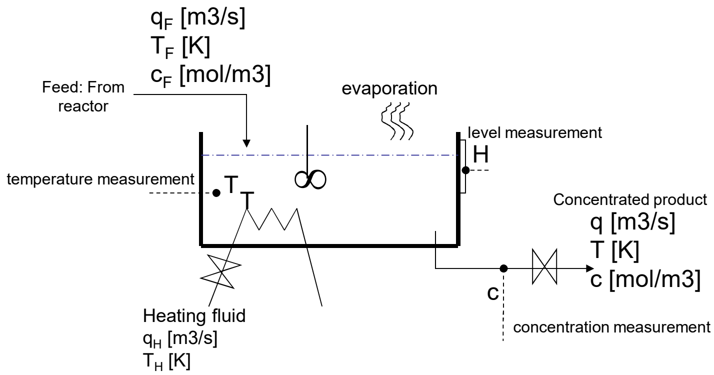


$$\text{Flow rate: } q = C_v f(z)\sqrt{\frac{p_1 - p_2}{\rho}} \qquad [\mathrm{m}^3/\mathrm{s}]$$

1. Counteracts nonlinearity in valve, f(z)
   - With fast flow control we can assume $q = q_s$
2. Eliminates effect of disturbances in p1 and p2
   (FC reacts faster than outer level loop)



Example: Evaporator with heating



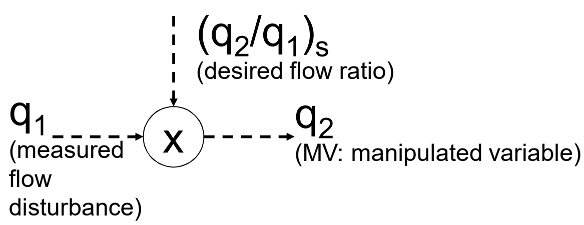
Control objectives
- Keep level H at desired value
- NEW: Keep composition c at desired value

BUT: Composition measurement has large delay + unreliable
Suggest control structure based on cascade control

Ratio control (most common case of feedforward)

Example: Process with two feeds q_1(d) and q_2 (u), where ratio should be constant.

Use multiplication block (x):



"Measure disturbance ($d=q_1$) and adjust input ($u=q_2$) such that ratio is at given value $(q_2/q_1)_s$"

Usually: Combine ratio (feedforward) with feedback

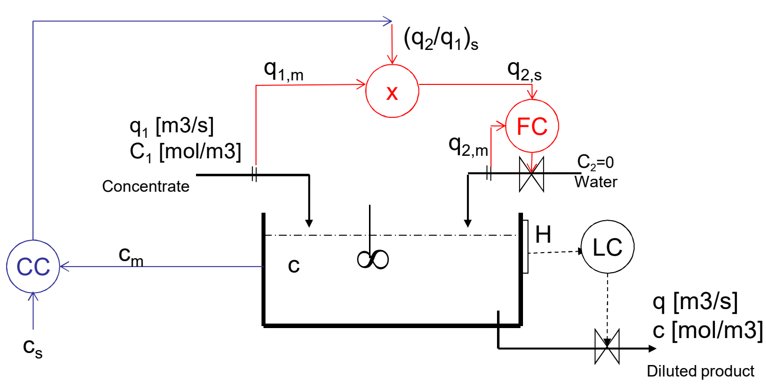
- Adjust $(q1/q2)_s$ based on feedback from process, for example, composition controller.
  - This is a special case of cascade control

  – ***Example cake baking***: *Use recipe (ratio control = feedforward), but adjust ratio if result is not as desired (feedback)*

  – ***Example evaporator:*** *Fix ratio q_H/q_F (and use feedback from T to fine tune ratio)*

EXAMPLE: MIXING PROCESS

RATIO CONTROL with outer cascade (to adjust ratio setpoint)



Procedure for design of control system

1. Define control objective (why control?)
2. Classify variables
   - MVs (u)
   - Disturbances (d)
   - CVs (y)

   \+ measurements
3. Process description
   - Flow sheet
   - Process matrix
     - Qualitative: with 0, +, -, (+)*, (-)*
       Quantitative: transfer matrix (see later courses)
4. Control structure
   - Feedforward / feedback
   - Pairing of variables (avoid pairing on 0!)
   - Cascade loops (MV from one controller (master) is setpoint for another (slave))
   - Put on process & instrumentation diagram (P&ID)
5. Control algorithm
   - On/off
   - PID (proportional-integral-derivative)
   - Model based (MPC)
6. Implementation
   - Today: Normally computer + connect measurements and valves (actuators)

Process matrix

	Input 1	input2
Output 1	+	-
Output 2	0	+

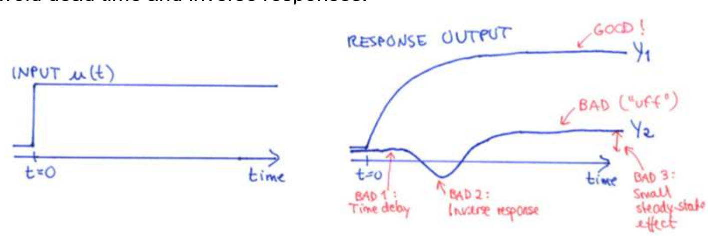
Process engineer (YOU):
- Responsible for items 1- 4
- The most important is process understanding

*(has some effect, but too small for control)

Rules for pairing of variables and choice of control structure

Main rule: “Pair close”

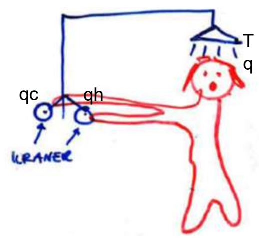
1. The response (from input to output) should be fast, large and in one direction. Avoid dead time and inverse responses!



2. The input (MV) should preferably affect only one output (to avoid interaction between the loops; may use process matrix)
3. Try to avoid input saturation (valve fully open or closed) in “basic” control loops for level and pressure
4. The measurement of the output y should be fast and accurate. It should be located close to the input (MV) and to important disturbances.
   - Use extra measurements y’ and cascade control if this is not satisfied
5. The system should be simple
   - Avoid too many feedforward and cascade loops
6. “Obvious” loops (for example, for level and pressure) should be closed first, before you spend too much time on deriving process matrices etc.

Example: Shower

1. Define control objective (why control?)
   - CVs: Control temperature T and flow q
2. Classify variables
   - MVs (u): qc, qh
   - Disturbances (d): Focus on main
   - CVs (y): T, q
3. Process description
   - Flow sheet
   - Process matrix
4. Control structure
   - Pairing of variables (Alt.1, Alt.2)
   - Multivariable (Alt 3)



	Input 1 qc	Input2 qh
Output 1 T	-	+
Output 2 q	+	+

In this case the process matrix has no 0's ⇒ Interactive, so pairing is not obvious!

Multivariable control ("decoupling") is used in practice:
One handle for total flow (qh+qc), one for ratio (qh/qc)

3x3 pairing example

Pairing: Choose one pairing from each row/column. Avoid pairing on 0's

Outputs \ Inputs	u_1	u_2	u_3
y_1	(+)	+	+
y_2	0	+	(-)
y_3	0	(+)	0

Conclusion:

$y_1 \leftrightarrow u_1$

$y_2 \leftrightarrow u_3$

$y_3 \leftrightarrow u_2$

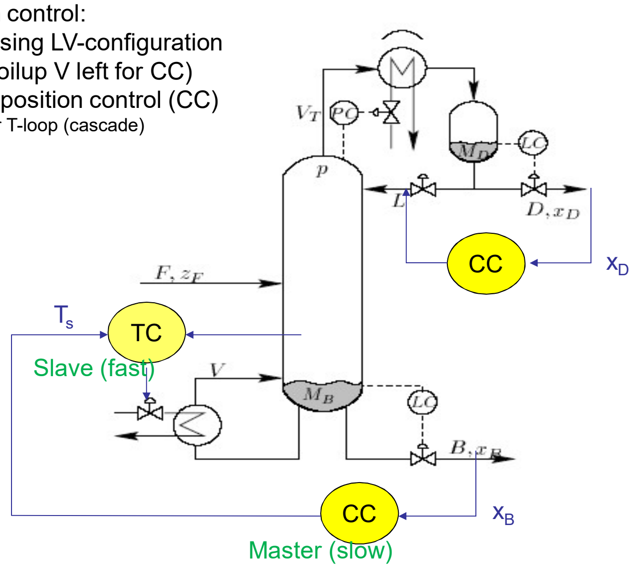
Example: Distillation

- Here: Given feed (i.e., feedrate is disturbance)

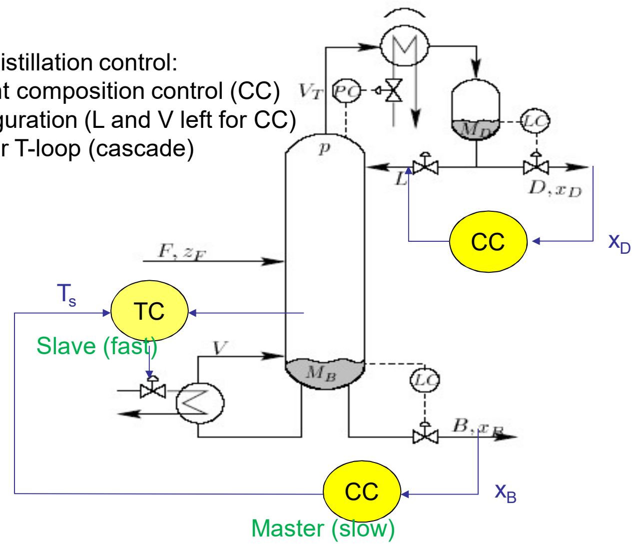
1. Objective: "Stabilize" column + keep compositions in top and bottom constant
   - But compositions measurements delayed + unreliable
2. Classify variables
3. Process description
   - Flowsheet
   - Process matrix
4. Control structure: Stabilize column "profile" using sensitive temperature measurement.

Typical distillation control:

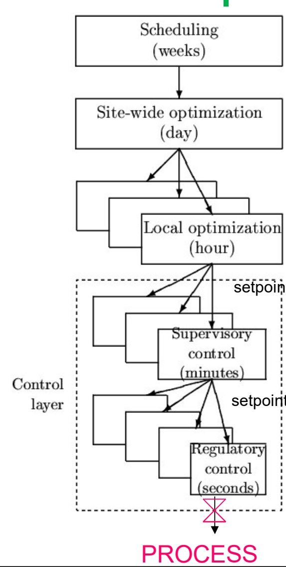
- Level control using LV-configuration (reflux L and boilup V left for CC)
- Two-point composition control (CC)
  - with inner T-loop (cascade)



Typical distillation control:
Two-point composition control (CC)
LV-configuration (L and V left for CC)
with inner T-loop (cascade)



Control hierarchy based on "time scale separation"



MPC or cascade/feedforward/…
(slower advanced and multivariable control)

PID (fast "regulatory" control)

MPC = model predictive control
PID = proportional+integral+derivative control

Inventory control rule (TPM)

- Inventory control: Usually control of level and pressure
- LocateTPM: Find out where throughput is set
- Rule («to keep things flowing»): Inventory control must be radiating around TPM

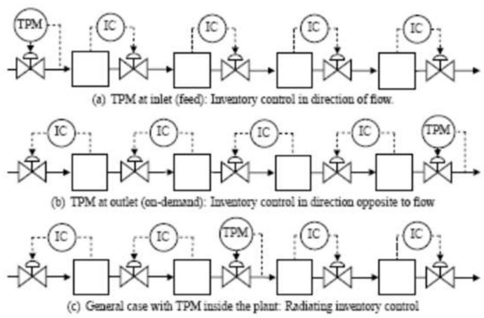
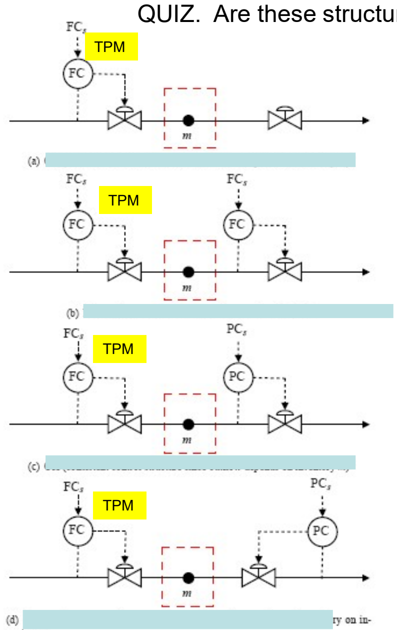


Figure 2.6: Self-consistency requires a radiating inventory control around a fixed flow (TPM)

QUIZ. Are these structures workable? Yes or No?



Quiz 2. Workable? Yes or No