Part 3: Regulatory («stabilizing») control

Outline

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

Procedure

- Skogestad procedure for control structure design

  I Top Down

  - Step S1: Define operational objective (cost) and constraints
  - Step S2: Identify degrees of freedom and optimize operation for disturbances
  - Step S3: Implementation of optimal operation
    - What to control ? (primary CV's)
    - Active constraints
    - Self-optimizing variables for unconstrained, c=Hy
  - Step S4: Where set the production rate? (Inventory control)

  II Bottom Up

  - Step S5: Regulatory control: What more to control (secondary CV's) ?
  - Step S6: Supervisory control
  - Step S7: Real-time optimization

Step S4. Where set production rate?

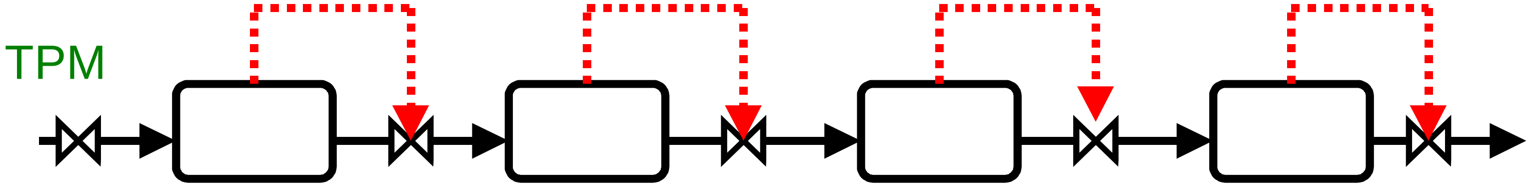
- Very important decision that determines the structure of the rest of the inventory control system!
- May also have important economic implications
- Link between **Top-down** (economics) and **Bottom-up** (stabilization) parts
  - Inventory control is the most important part of stabilizing control

- "Throughput manipulator" (TPM)
  = MV for controlling throughput (production rate, network flow)
- Where set the production rate = Where locate the TPM?
  - Traditionally: At the feed
  - For maximum production (with small backoff): At the bottleneck

TPM and link to inventory control

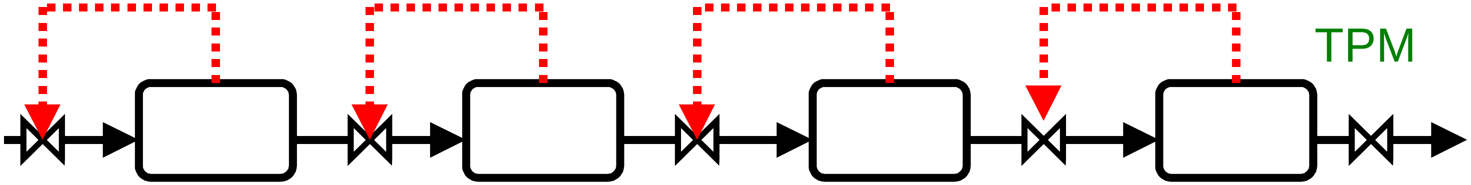
- Liquid inventory: Level control (LC)
  - Sometimes pressure control (PC)
- Gas inventory: Pressure control (PC)
- Component inventory: Composition control (CC, XC, AC)

Production rate set at inlet : Inventory control in direction of flow*



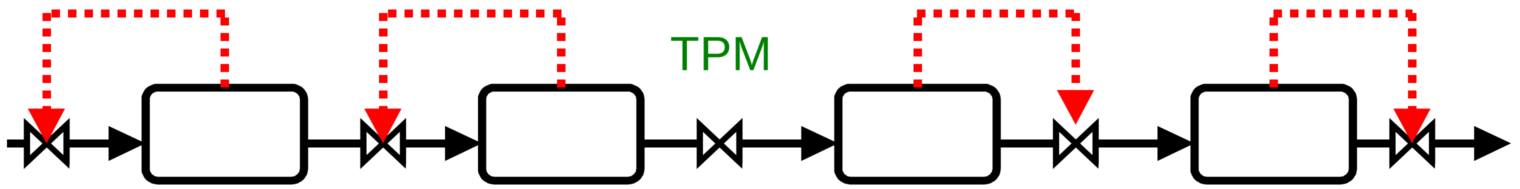
* Required to get “local-consistent” inventory control

Production rate set at outlet: Inventory control opposite flow*



* Required to get "local-consistent" inventory control

Production rate set inside process*



* Required to get “local-consistent” inventory control

General: “Need radiating inventory control around TPM” (Georgakis)

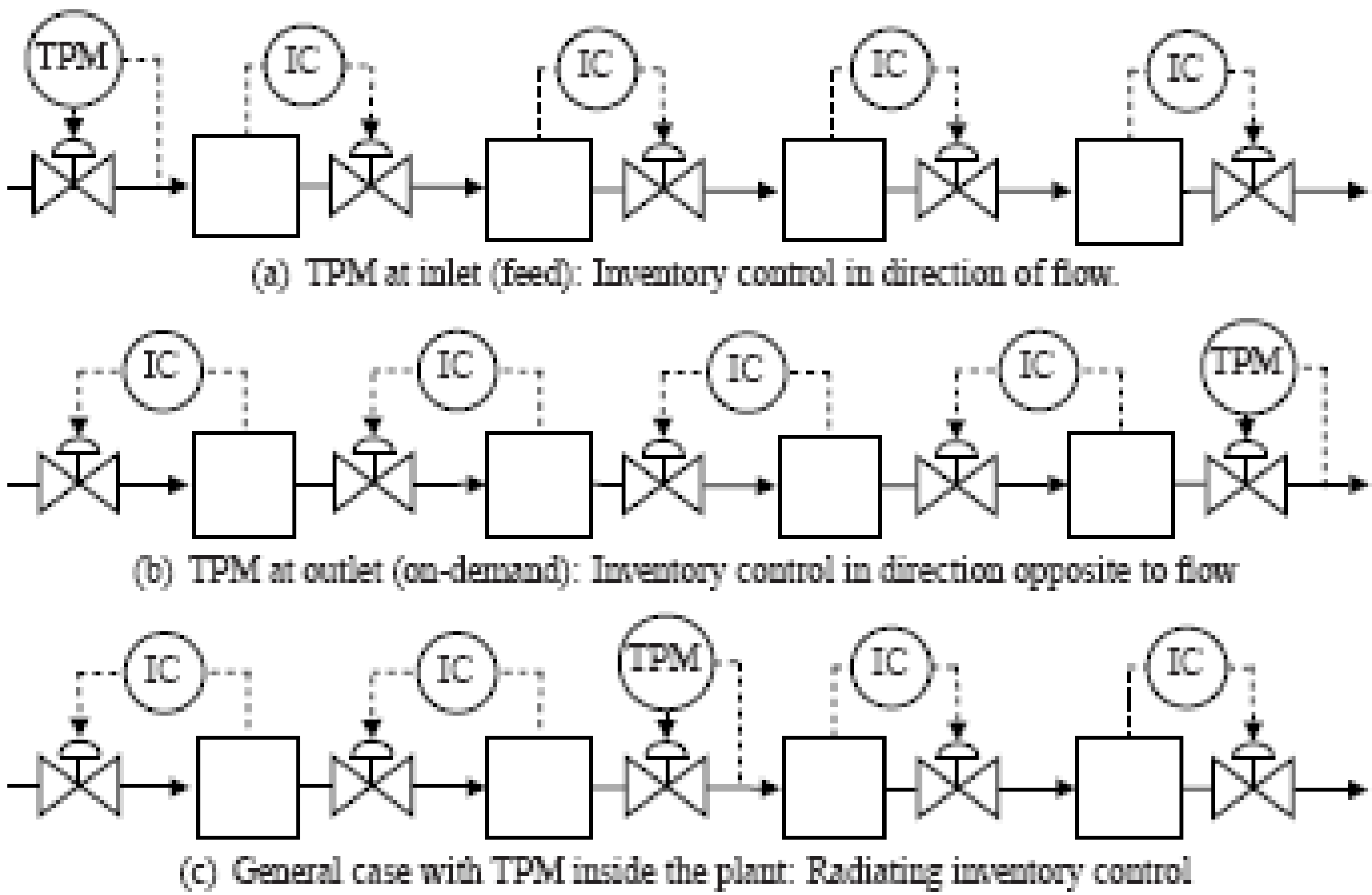


Figure 2.6: Self-consistency requires a radiating inventory control around a fixed flow (TPM)

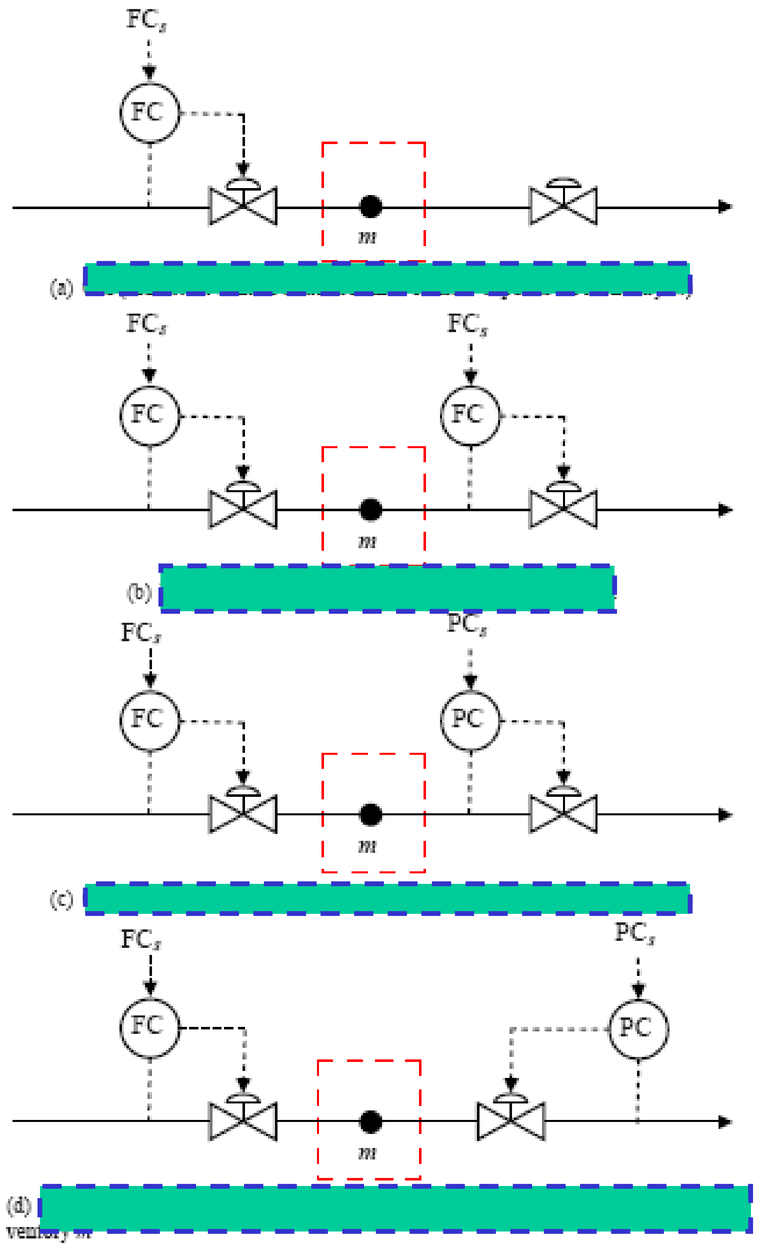
Consistency of inventory control

- **Consistency** (required property):

An inventory control system is said to be **consistent** *if the steady-state mass balances (total, components and phases) are satisfied for any part of the process, including the individual units and the overall plant.*

QUIZ 1

CONSISTENT?



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

Rule: Controlling pressure at inlet or outlet gives indirect flow control

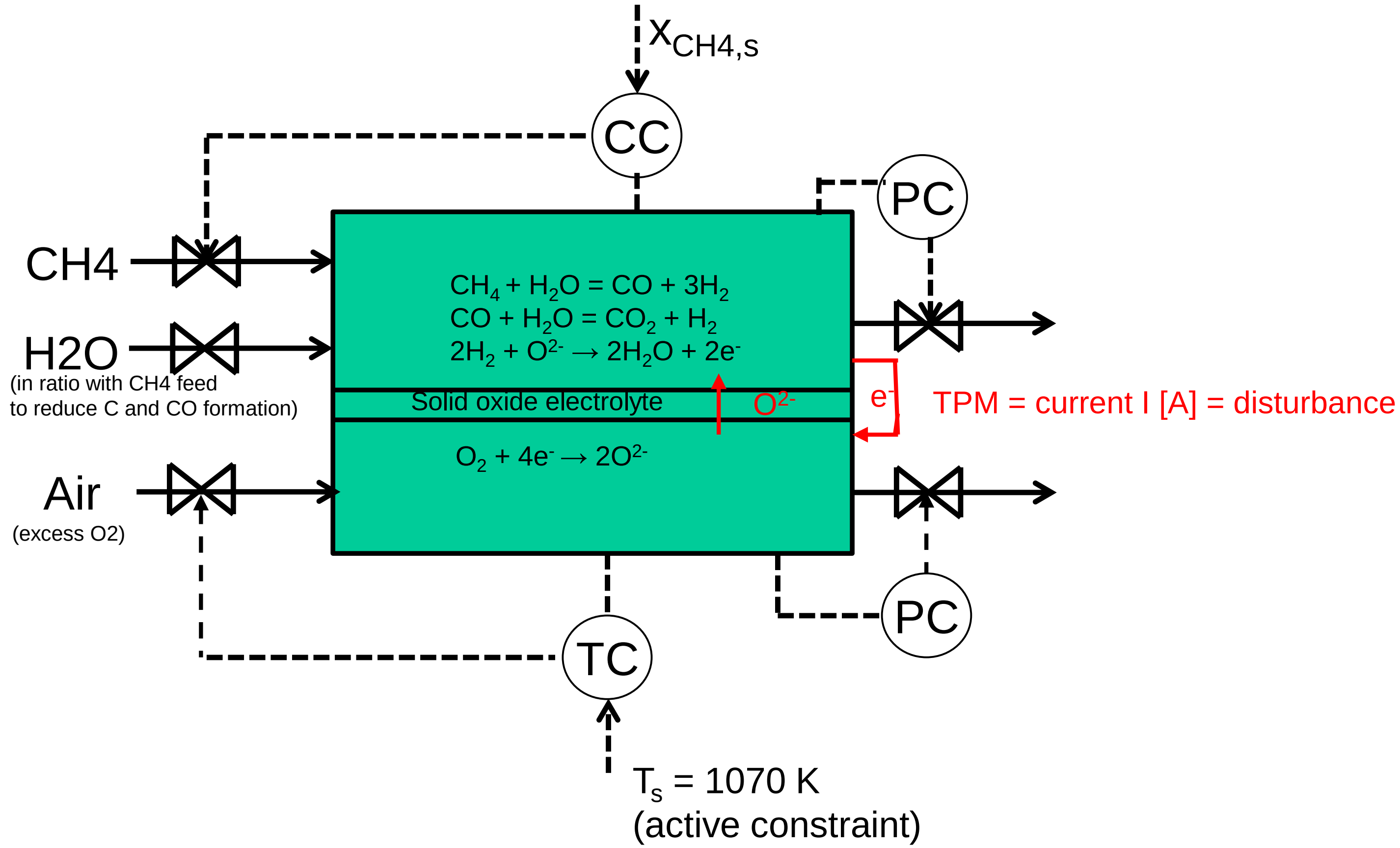
(because of pressure boundary condition)

LOCATION OF SENSORS

- Location flow sensor (before or after valve or pump): Does not matter from consistency point of view
  – Locate to get best flow measurement
    - Before pump: Beware of cavitation
    - After pump: Beware of noisy measurement

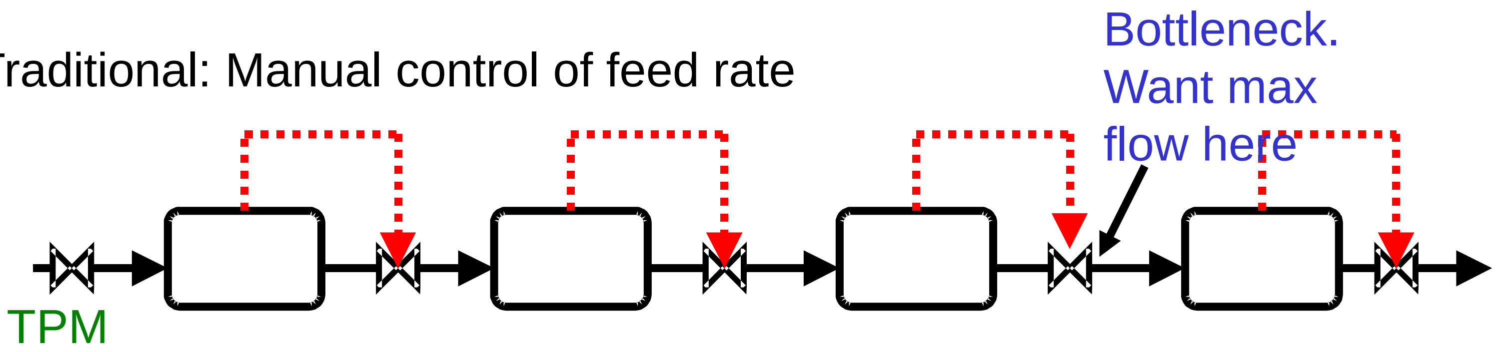
- Location of pressure sensor (before or after valve, pump or compressor): Important from consistency point of view

Example: Solid oxide fuel cell

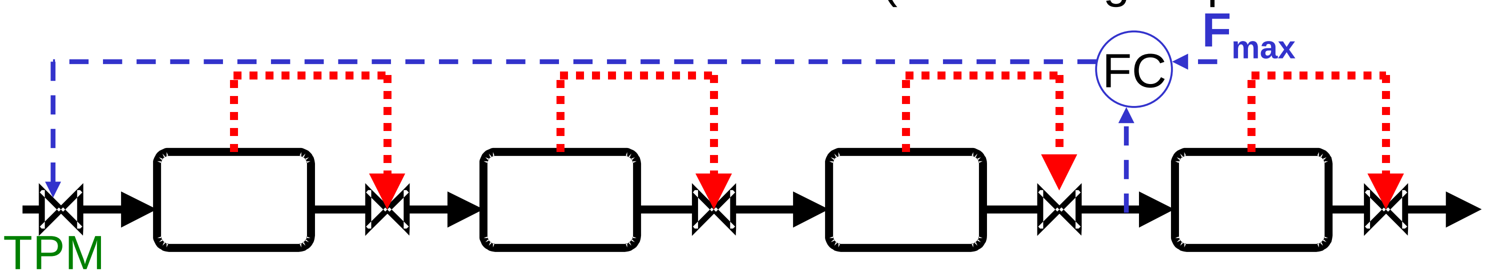


Single-loop alternatives for bottleneck control

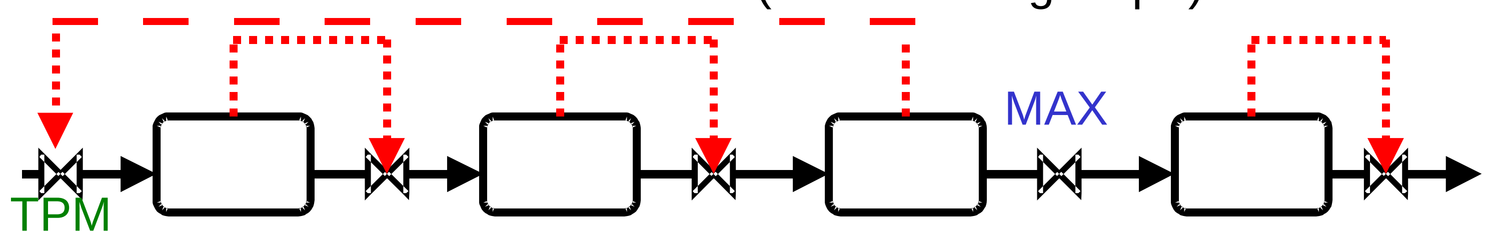
Traditional: Manual control of feed rate



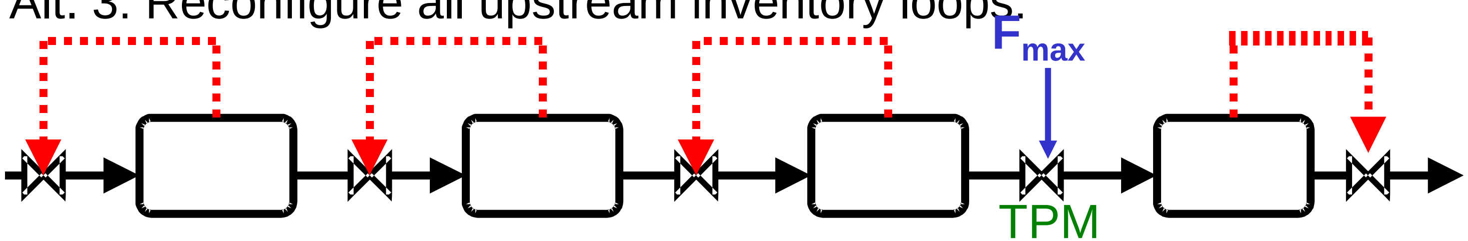
Alt.1. Feedrate controls bottleneck flow (VPC: “long loop” with backoff…):



Alt. 2: Feedrate controls lost task (another “long loop”):

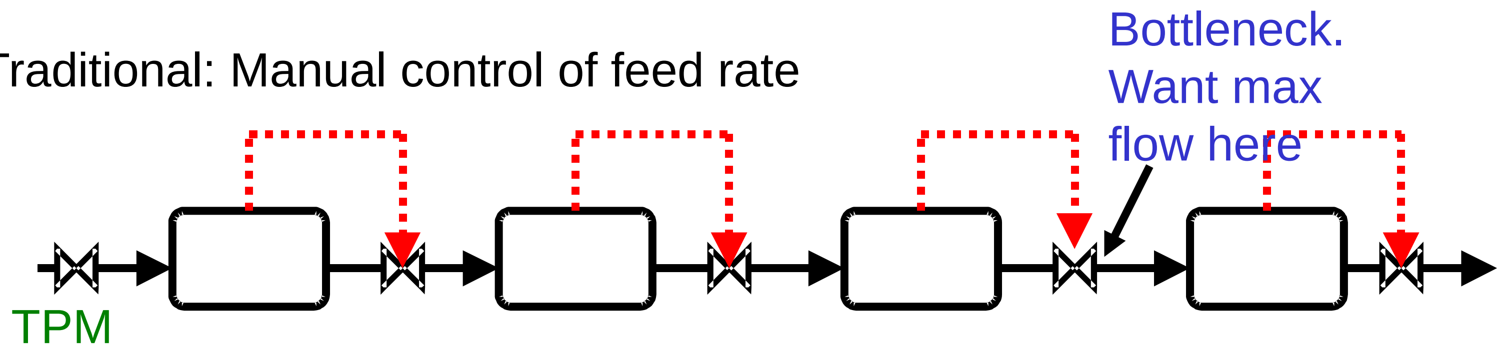


Alt. 3: Reconfigure all upstream inventory loops:

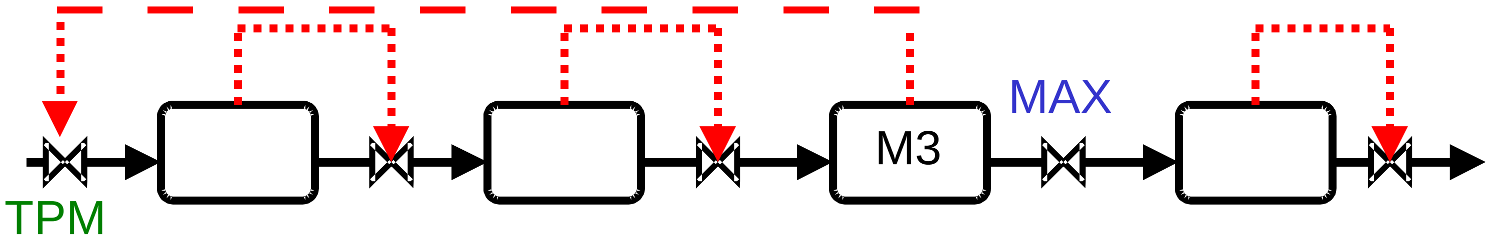


Single-loop alternatives for bottleneck control

Traditional: Manual control of feed rate



Alt. 2: Feedrate controls level upstream bottleneck:



Comment on Alt. 2 where feed controls M3. «Long loop», so slow.
Can work if M3 is large,
Rule: Can keep TPM at feed if surge volume (M3) before bottleneck is large

Where should we place TPM?

- TPM = MV used to control throughput
- Traditionally: TPM = Main feed valve (or pump/compressor)
  - Gives inventory control "in direction of flow"

Consider moving TPM if:

1. There is an important CV that could otherwise not be well controlled
   - Dynamic reasons
   - Special case: Max. production important: Locate TPM at process bottleneck* !
     - TPM can then be used to achieve tight bottleneck control (= achieve max. production)
     - Economics: Max. production is very favorable in "sellers marked"
2. If placing it at the feed may yield infeasible operation ("overfeeding")
   - If "snowballing" is a problem (accumulation in recycle loop), then consider placing TPM inside recycle loop

BUT: Avoid a variable that may (optimally) saturate as TPM (unless it is at bottleneck)

- Reason: To keep controlling CV=throughput, we would need to reconfigure (move TPM)**

*Bottleneck: Last constraint to become active as we increase throughput -> TPM must be used for bottleneck control

**Sigurd's general pairing rule (to reduce need for reassigning loops): "Pair MV that may (optimally) saturate with CV that may be given up"

Conclusion TPM (production rate manipulator)

- Think carefully about where to place it!
- Difficult to undo later

Session 5: Design of regulatory control layer

Outline

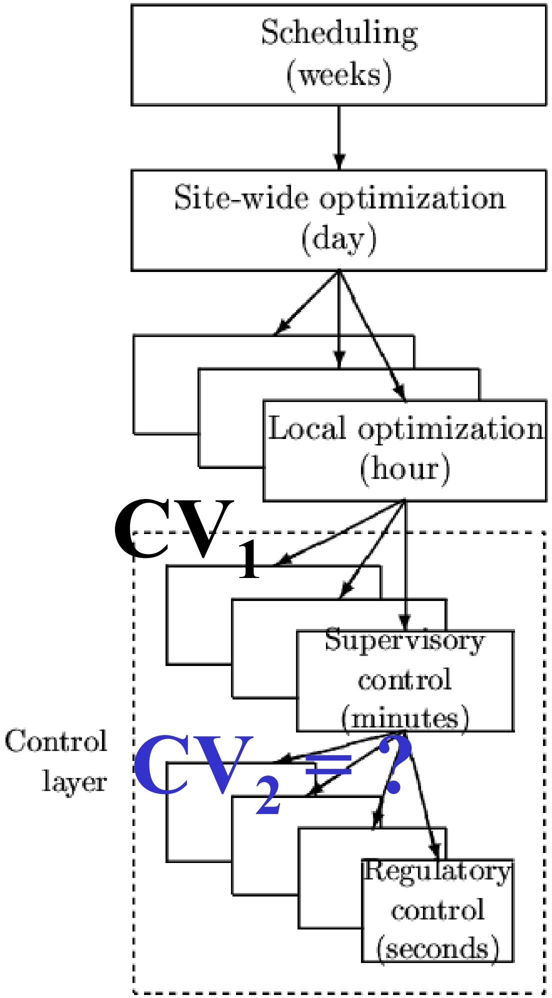
- Skogestad procedure for control structure design

  I Top Down
  - Step S1: Define operational objective (cost) and constraints
  - Step S2: Identify degrees of freedom and optimize operation for disturbances
  - Step S3: Implementation of optimal operation
    - What to control ? (primary CV's) (self-optimizing control)
  - Step S4: Where set the production rate? (Inventory control)

  II Bottom Up
  - Step S5: Regulatory control: What more to control (secondary CV's) ?
    - Distillation example
  - Step S6: Supervisory control
  - Step S7: Real-time optimization

Step 5. Regulatory control layer

- *Purpose*: "Stabilize" the plant using a simple control configuration (usually: local SISO PID controllers + simple cascades)
- Enable manual operation (by operators)
- Main structural decisions:
  - What more should we control? (secondary cv's, CV_2, use of extra measurements)
  - Pairing with manipulated variables (mv's u_2)



Structure of regulatory control layer (PID)

Main decisions:

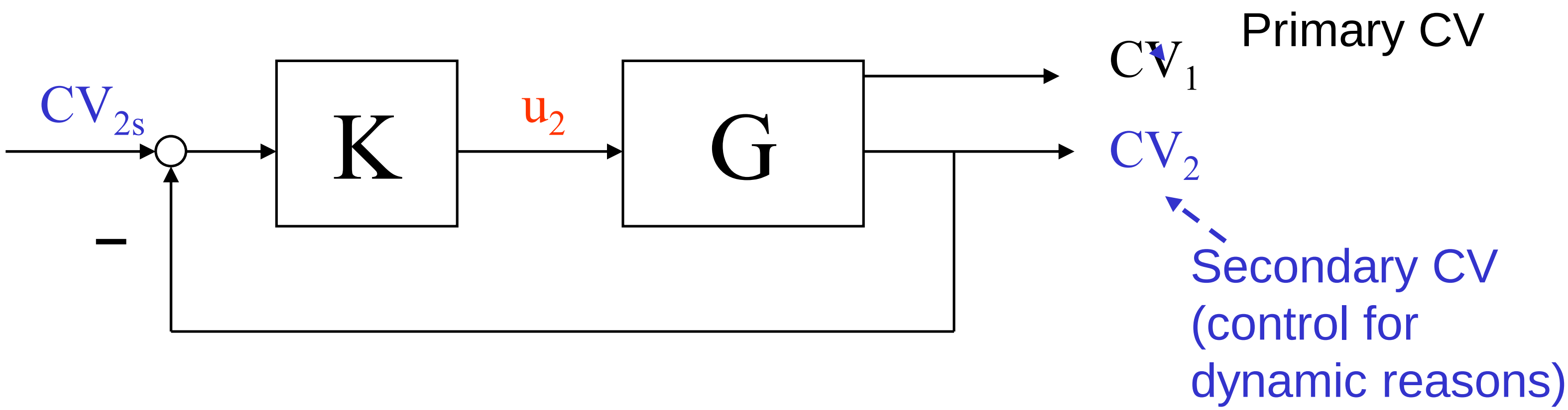
1. Selection of controlled variables (CV2)
2. Pairing with manipulated variables (MV2)

Main rules:

1. **Control drifting variables**
2. **«Pair close"**

CV2
↕
MV2

Stabilizing control: Use inputs $MV_2=u_2$ to control "drifting" variables CV_2



Key decision: Choice of CV_2 (controlled variable)

Also important: Choice of MV2=u_2 ("pairing")

Process control: Typical «drifting» variables (CV2) are

- Liquid inventories (level)
- Vapor inventories (pressure)
- Some temperatures (reactor, distillation column profile)

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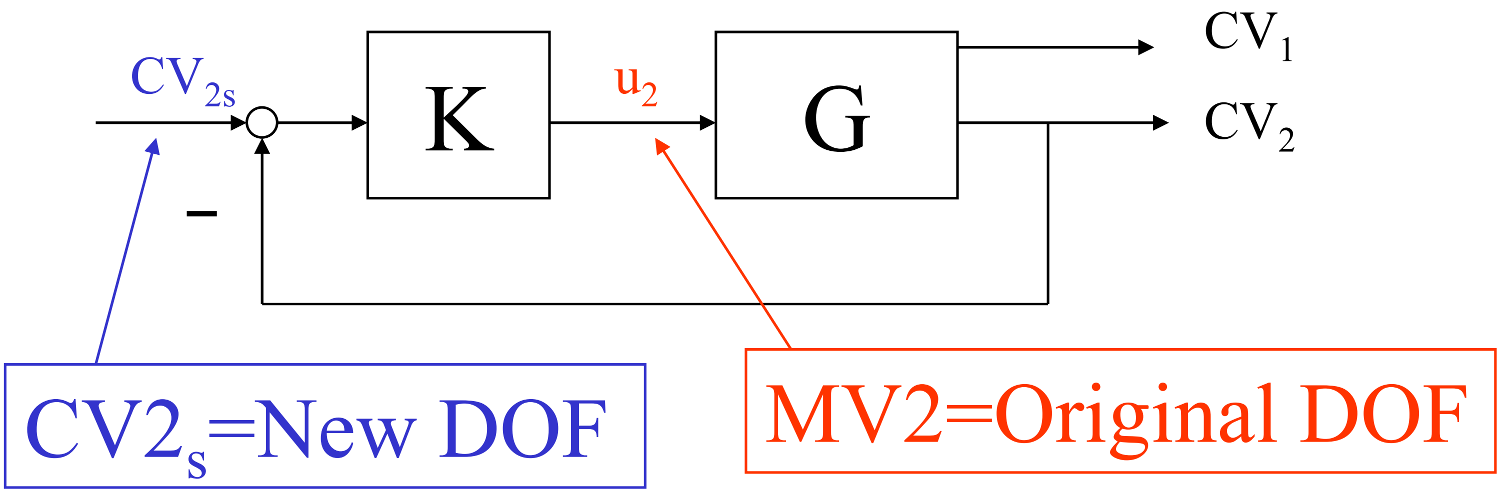
Degrees of freedom unchanged

- No degrees of freedom lost as setpoints y_{2s} replace inputs u_2 as new degrees of freedom for control of y_1

Cascade control:



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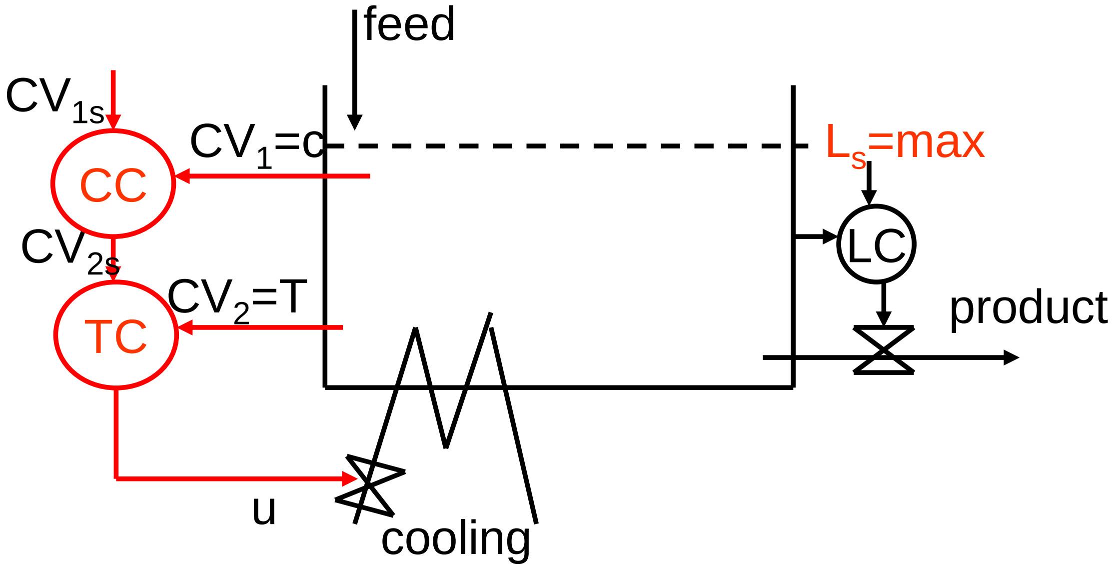
Objectives regulatory control layer

1. Allow for manual operation
2. Simple decentralized (local) PID controllers that can be tuned on-line
3. Take care of "fast" control
4. Track setpoint changes from the layer above
5. Local disturbance rejection
6. Stabilization (mathematical sense)
7. Avoid "drift" (due to disturbances) so system stays in "linear region"
   - "stabilization" (practical sense)
8. Allow for "slow" control in layer above (supervisory control)
9. Make control problem easy as seen from layer above
10. Use "easy" and "robust" measurements (pressure, temperature)
11. Simple structure
12. Contribute to overall economic objective ("indirect" control)
13. Should not need to be changed during operation

Example: Exothermic reactor (unstable)

Active constraints (economics):
Product composition c + level (max)

- u = cooling flow (q)
- CV_1 = composition (c)
- CV_2 = temperature (T)



"Control CV2 that stabilizes the plant (stops drifting)"

In practice, control:

1. **Levels** (inventory liquid)
2. **Pressures** (inventory gas/vapor) (note: some pressures may be left floating)
3. **Inventories of components** that may accumulate/deplete inside plant
   - E.g., amine/water depletes in recycle loop in CO2 capture plant
   - E.g., butanol accumulates in methanol-water distillation column
   - E.g., inert N2 accumulates in ammonia reactor recycle
4. **Reactor temperature**
5. **Distillation column profile** (one temperature inside column)
   - Stripper/absorber profile does <u>not</u> generally need to be stabilized

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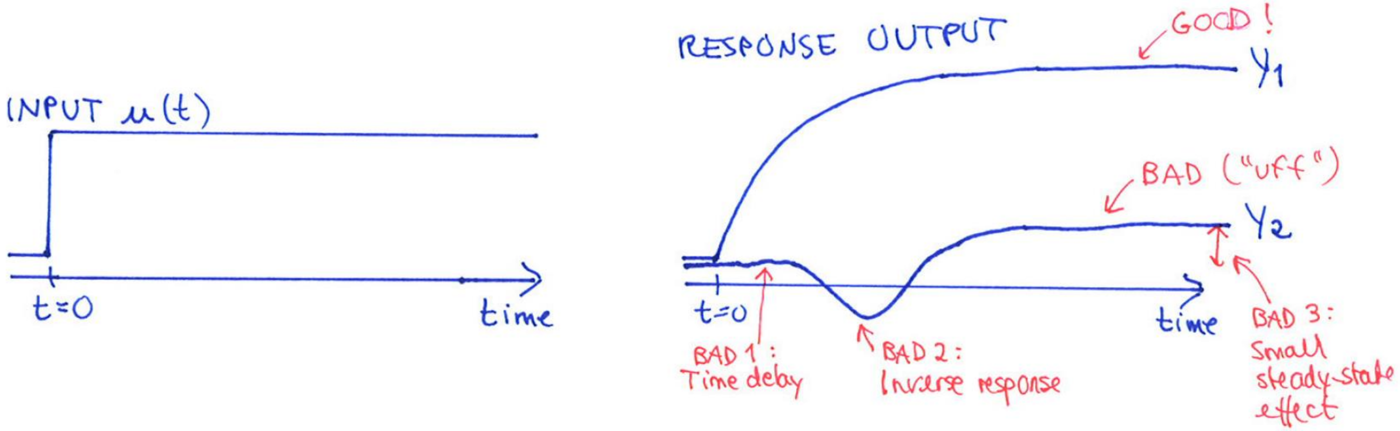
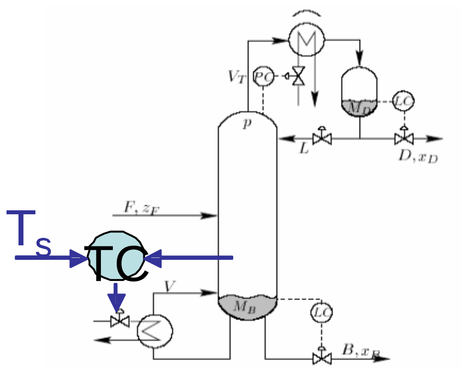
Main rule: "Pair close"



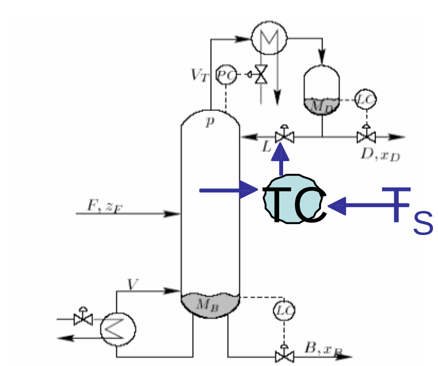
The response (from input to output) should be fast, large and in one direction. Avoid dead time and inverse responses!

Sigurd's pairing rule for regulatory layer: "Avoid using MVs that may optimally saturate (at steady state) to control CV2s"

- Main reason: Minimizes need for reassigning loops



(a) Normal: Control T using V



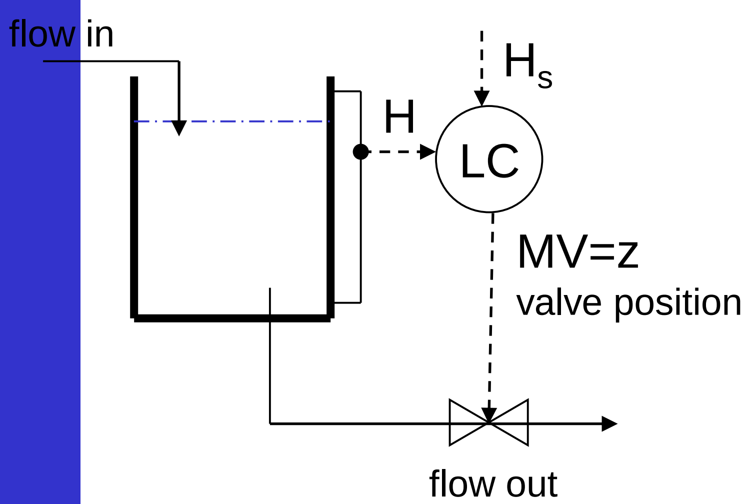
(b) If V may saturate: Use L

- Important: Always feasible (and optimal) to give up a CV when optimal MV saturation occurs.
  - Proof (DOF analysis): When one MV disappears (saturates), then we have one less optimal CV.
- Failing to follow this rule: Need some "fix" when MV saturates to remain optimal, like
  - reconfiguration (logic)
  - backoff (loss of optimality)
- BUT: Rule may be in conflict with other criteria
  - Dynamics ("pair close" rule)
  - Interactions ("avoid negative steady-state RGA" rule)
  - If conflict: Use reconfiguration (logic) or go for multivariable constraint control (MPC which may provide "built-in" logic)

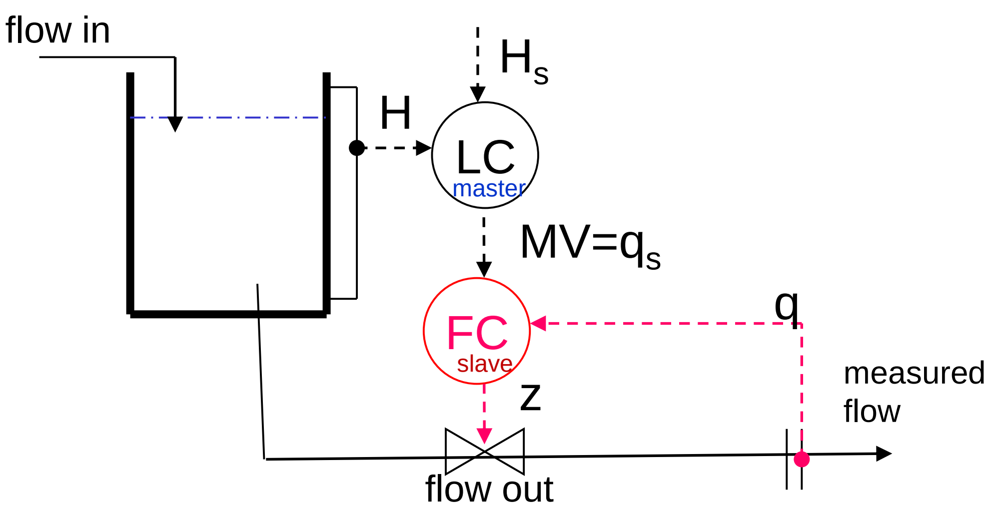
Cascade control in regulatory layer

- **May be helpful to reduce nonlinearity and improve disturbance rejection**
- **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

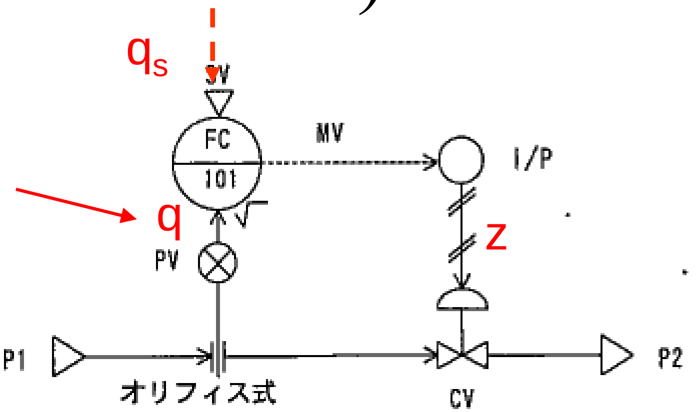


WITH CASCADE

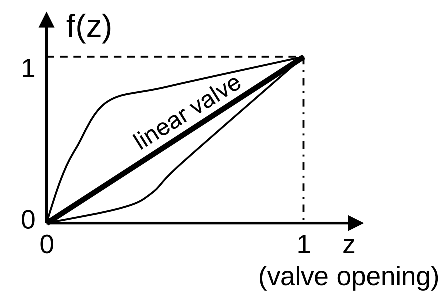


What are the benefits of adding a flow controller (inner cascade)?

Extra measurement y' = q



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



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)

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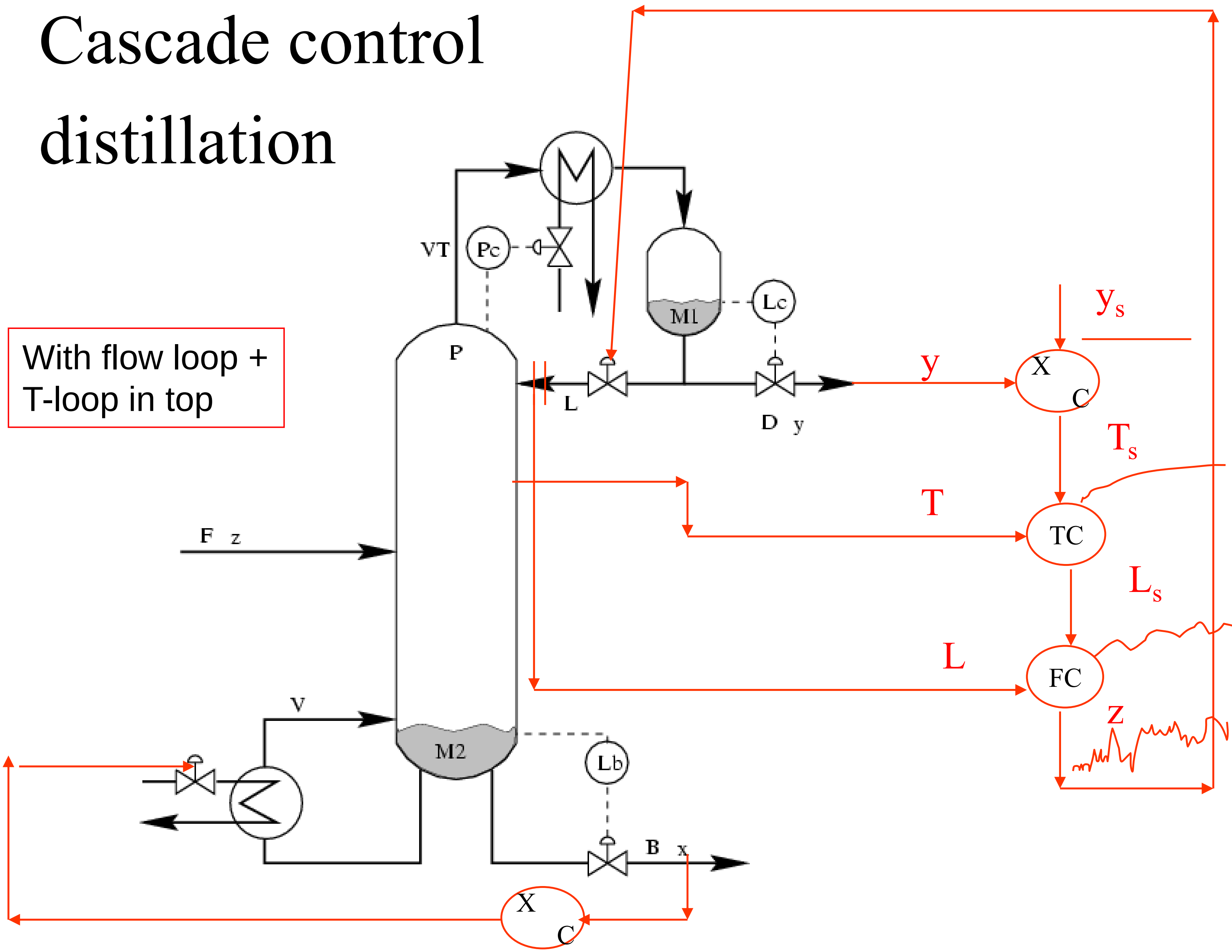
Counteracting nonlinearity using cascade control: Process gain variation -> Time constant variation

Proof:

- Slave controller with $u = z$ (valve position) and $y=q$ (flow)
- Nonlinear valve with varying gain k: $G = k / (\tau s+1)$
- PI-controller with gain K_c and integral time $\tau_I = \tau$.
- With slave (flow) controller: Transfer function from y_s to y (for master loop):
- $T = L/(1+L) = 1/(\tau_{CL}\ s + 1)$
  - where $\tau_{CL} = \tau/(k\ K_c)$
- **So variation in k translates into variation in τ_{CL}**
- In practise this gives a variation in the effective time delay in the master loop
  - Low gain k for valve gives large effective time delay (bad)

Cascade control distillation

With flow loop + T-loop in top



Hierarchical/cascade control: Time scale separation

- With a "reasonable" time scale separation between the layers

  (typically by a factor 5 or more in terms of closed-loop response time)

  we have the following advantages:

1. The stability and performance of the lower (faster) layer (involving y_2) is not much influenced by the presence of the upper (slow) layers (involving y_1)

   Reason: The frequency of the "disturbance" from the upper layer is well inside the bandwidth of the lower layers

2. With the lower (faster) layer in place, the stability and performance of the upper (slower) layers do not depend much on the specific controller settings used in the lower layers

   Reason: The lower layers only effect frequencies outside the bandwidth of the upper layers

Summary: Rules for plantwide control

- Here we present a set of simple rules for economic plantwide control to facilitate a close-to-optimal control structure design in cases where the optimization of the plant model is not possible.
- the rules may be conflicting in some cases and in such cases, human reasoning is strongly advised.

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Rule 1: Control the active constraints.

- In general, process optimization is required to determine the active constraints, but in many cases these can be identified based on a good process knowledge and engineering insight. Here is one useful rule:
- Rule 1A: The purity constraint of the valuable product is always active and should be controlled.
- This follows, because we want to maximize the amount of valuable product and avoid product “give away” (Jacobsen and Skogestad, 2011). Thus, we should always control the purity of the valuable product at its specification.
- For “cheap” products we may want to overpurify (purity constraint may not be active) because this may reduce the loss of a more valuable component.
- In other cases, we must rely on our process knowledge and engineering insight. For reactors with simple kinetics, we usually find that, the reaction and conversion rates are maximized by operating at maximum temperature and maximum volume (liquid phase reactor). For gas phase reactor, high pressure may increase the reaction rate, but this must be balanced against the compression costs.

Rule 2: (for remaining unconstrained steady-state degrees of freedom, if any):
Control the "self-optimizing" variables.

- This choice is usually not obvious, as there may be several alternatives, so this rule is in itself not very helpful. The ideal self-optimizing variable, at least, if it can be measured accurately, is the gradient of the cost function. Ju, which should be zero for any disturbance. Unfortunately, it is rarely possible to measure this variable directly and the "self-optimizing" variable may be viewed as an estimate of the gradient Ju

The two main properties of a good "self-optimizing" (CV1=c=Hy) variable are:

1. Its optimal value is insensitive to disturbances (such that the optimal sensitivity $dc_{opt}/dd = F^c$ HF = is small)
2. It is sensitive to the plant inputs (so the process gain dc/du = G = HG^y is large).

The following rule shows how to combine the two desired properties:

- Rule 2A: Select the set CV1=c such that the ratio $G^{-1}F^c$ is minimized.
- This rule is often called the "maximum scaled gain rule".

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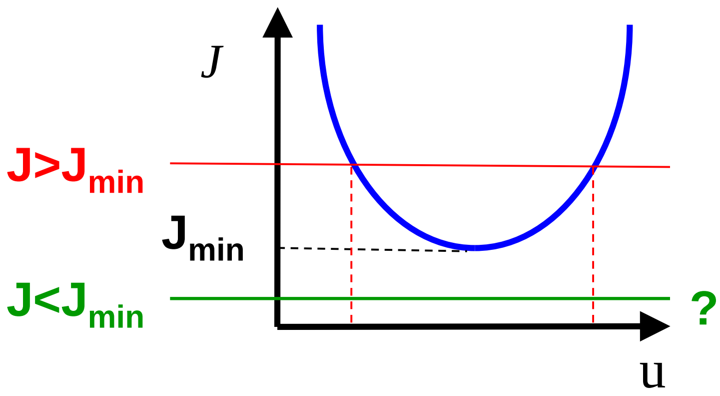
Rule 3: (for remaining unconstrained steady-state degrees of freedom, if any):

Never try to control the cost function J

(or any other variable that reaches a maximum or minimum at the optimum)



- First, the cost function J has no sensitivity to the plant inputs at the optimal point and so G = 0 which violates Rule 2A.
- Second, if we specify J lower than its optimal value, then clearly, the
- operation will be infeasible
- Also, specifying J higher than its optimal value is problematic, as we have multiplicity of solutions. As mentioned above, rather controlling the cost J, we should control its gradient, Ju.

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Rule 4: Locate the TPM close to the process bottleneck

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- The justification for this rule is to take advantage of the large economic benefits of maximizing production in times when product prices are high relative to feed and energy costs (Mode 2). To maximize the production rate, one needs to achieve tight control of the active constraints, in particular, of the bottleneck, which is defined as the last constraint to become active when increasing the throughput rate (Jagtap et al., 2013).

Rule 5: (for processes with recycle) Locate the TPM inside the recycle loop.

- The point is to avoid “overfeeding” the recycle loop which may easily occur if we operate close to the throughput where “snowballing” in the recycle loop occurs. This is a restatement of Luyben’s rule “Fix a Flow in Every Recycle Loop” (Luyben et al., 1997). From this perspective, snowballing can be thought of as the dynamic consequence of operating close to a bottleneck which is within a recycle system.
- In many cases, the process bottleneck is located inside the recycle loop and Rules 4 and 5 give the same result.

Rule 6: Arrange the inventory control loops (for level, pressures, etc.) around the TPM location according to the radiation rule.

- The radiation rule (Price et al., 1994), says that, the inventory loops upstream of the TPM location must be arranged opposite of flow direction. For flow downstream of TPM location it must be arranged in the same direction. This ensures “local consistency” i.e. all inventories are controlled by their local in or outflows.

Rule 7: Select "sensitive/drifting" variables as controlled variables CV2 for regulatory control.

- This will generally include inventories (levels and pressures), plus certain other drifting (integrating) variables, for example,
  - a reactor temperature
  - a sensitive temperature/composition in a distillation column.
- This ensures "stable operation, as seen from an operator's point of view.
- Some component inventories may also need to be controlled, especially for recycle systems. For example, according to "Down's drill" one must make sure that all component inventories are "self-regulated" by flows out of the system or by removal by reactions, otherwise their composition may need to be controlled (Luyben, 1999).

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Rule 8: Economically important active constraints (a subset of CV1), should be selected as controlled variables CV2 in the regulatory layer.

- Economic variables, CV1, are generally controlled in the supervisory layer. Moving them to the faster regulatory layer may ensure tighter control with a smaller backoff.

- **Backoff: difference between the actual average value (setpoint) and the optimal value (constraint).**

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Rule 9: "Pair-close" rule: The pairings should be selected such that, effective delays and loop interactions are minimal.

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Rule 10: : Avoid using MVs that may optimally saturate (at steady state) to control CVs in the regulatory layer (CV2)

- The reason is that we want to avoid re-configuring the regulatory control layer. To follow this rule, one needs to consider also other regions of operation than the nominal, for example, operating at maximum capacity (Mode 2) where we usually have more active constraints.

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Rule 11: MVs that may optimally saturate (at steady state) should be paired with the subset of CV1 that may be given up.

- This rule applies for cases when we use decentralized control in the supervisory layer and we want to avoid reconfiguration of loops.
- The rule follows because when a MV optimally saturates, then there will be one less degree of freedom, so there will be a CV1 which may be given up without any economic loss. The rule should be considered together with rule 10.
- Example: Gives correct answer for the process where we want to control flow and have p>pmin: Pair the valve (MV) with CV1 (p) which may be given up.

Plantwide control. Main references

- The following paper summarizes the procedure:
  - S. Skogestad, ``Control structure design for complete chemical plants'', *Computers and Chemical Engineering*, **28** (1-2), 219-234 (2004).
- There are many approaches to plantwide control as discussed in the following review paper:
  - T. Larsson and S. Skogestad, ``Plantwide control: A review and a new design procedure'' *Modeling, Identification and Control*, **21**, 209-240 (2000).
- The following paper updates the procedure:
  - S. Skogestad, ``Economic plantwide control'', Book chapter in V. Kariwala and V.P. Rangaiah (Eds), *Plant-Wide Control: Recent Developments and Applications*'', Wiley (2012).
- More information:

http://www.nt.ntnu.no/users/skoge/plantwide

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