ECONOMIC PLANTWIDE CONTROL: Control structure design for complete processing plants

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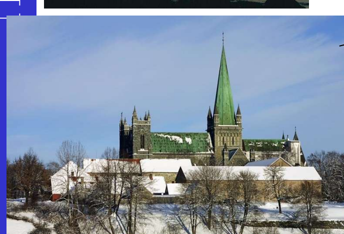
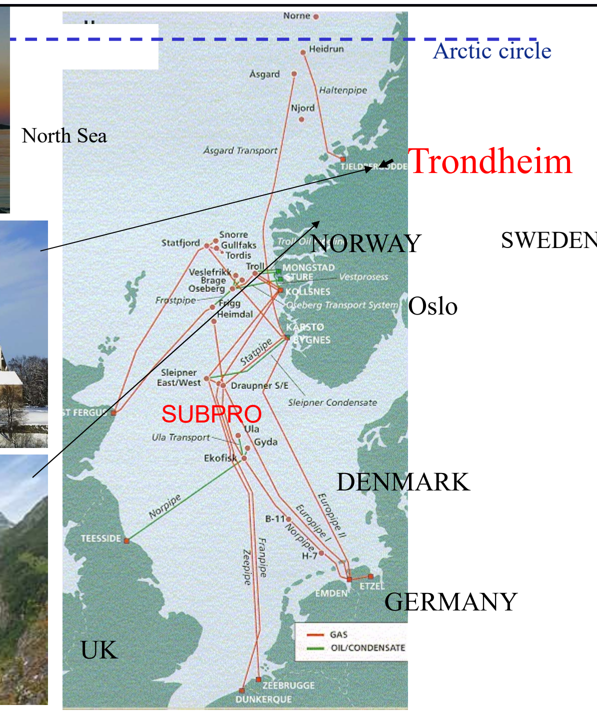
Adconip, Taiwan, May 2017





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Outline

- Paradigm: Based on time scale separation
- Plantwide control procedure: Based on economics
- Example: Runner
- Selection of primary controlled variables (CV_1=H y)
  - Optimal is gradient, $CV_1=J_u$ with setpoint=0
  - General CV_1=Hy. Nullspace and exact local method
- Throughput manipulator (TPM) location
- Examples
- Conclusion

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ECONOMIC PLANTWIDE CONTROL: Control structure design for complete processing plants

- **Sigurd Skogestad** , Department of Chemical Engineering, Norwegian University of Science and Technology (NTNU), Trondheim, Norway

- Abstract: A chemical plant may have thousands of measurements and control loops. By the term *plantwide control* it is not meant the tuning and behavior of each of these loops, but rather the *control philosophy* of the overall plant with emphasis on the *structural decisions*. In practice, the control system is usually divided into several layers, separated by time scale: scheduling (weeks) , site-wide optimization (day), local optimization (hour), supervisory and economic control (minutes) and regulatory control (seconds). Such a hierchical (cascade) decomposition with layers operating on different time scale is used in the control of all real (complex) systems including biological systems and airplanes, so the issues in this section are not limited to process control. In the talk the most important issues are discussed, especially related to the choice of "self-optimizing" variables that provide the link the control layers. Examples are given for optimal operation of a runner and distillation columns.

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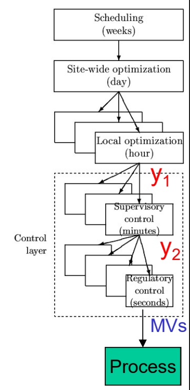
Outline of the plantwide control procedure



I Top Down

- **Step 1:** Define optimal operation
- **Step 2:** Optimize for expected disturbances
  - Find active constraints
- **Step 3**: Select primary controlled variables $c=y_1$ (CVs)
  - Self-optimizing variables
- **Step 4:** Where locate the throughput manipulator?

II Bottom Up

- **Step 5**: Regulatory / stabilizing control (PID layer)
  - What more to control (y_2)?
  - Pairing of inputs and outputs
- **Step 6:** Supervisory control (MPC layer)
- **Step 7:** Real-time optimization (Do we need it?)

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

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

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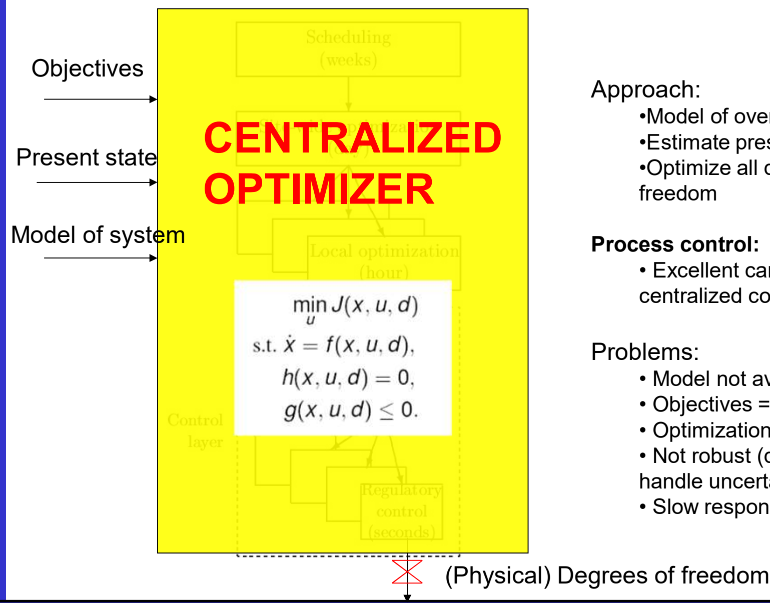
In theory: Optimal control and operation



Approach:
- Model of overall system
- Estimate present state
- Optimize all degrees of freedom

Process control:
- Excellent candidate for centralized control

Problems:
- Model not available
- Objectives = ?
- Optimization complex
- Not robust (difficult to handle uncertainty)
- Slow response time

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Practice: Engineering systems

- Most (all?) large-scale engineering systems are controlled using hierarchies of quite simple controllers
  - Large-scale chemical plant (refinery)
  - Commercial aircraft
- 100's of loops
- Simple components:

  PI-control + selectors + cascade + nonlinear fixes + some feedforward

Same in biological systems

But: Not well understood

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

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

Previous work on plantwide control:

•Page Buckley (1964) - *Chapter on "Overall process control"* (still industrial practice)
•Greg Shinskey (1967) – *process control systems*
•Alan Foss (1973) - *control system structure*
•Bill Luyben et al. (1975-) – *case studies ; "snowball effect"*
•George Stephanopoulos and Manfred Morari (1980) – *synthesis of control structures for chemical processes*
•Ruel Shinnar (1981-) - *"dominant variables"*
•Jim Downs (1991) - *Tennessee Eastman challenge problem*
•Larsson and Skogestad (2000): *Review of plantwide control*

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Main objectives 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)

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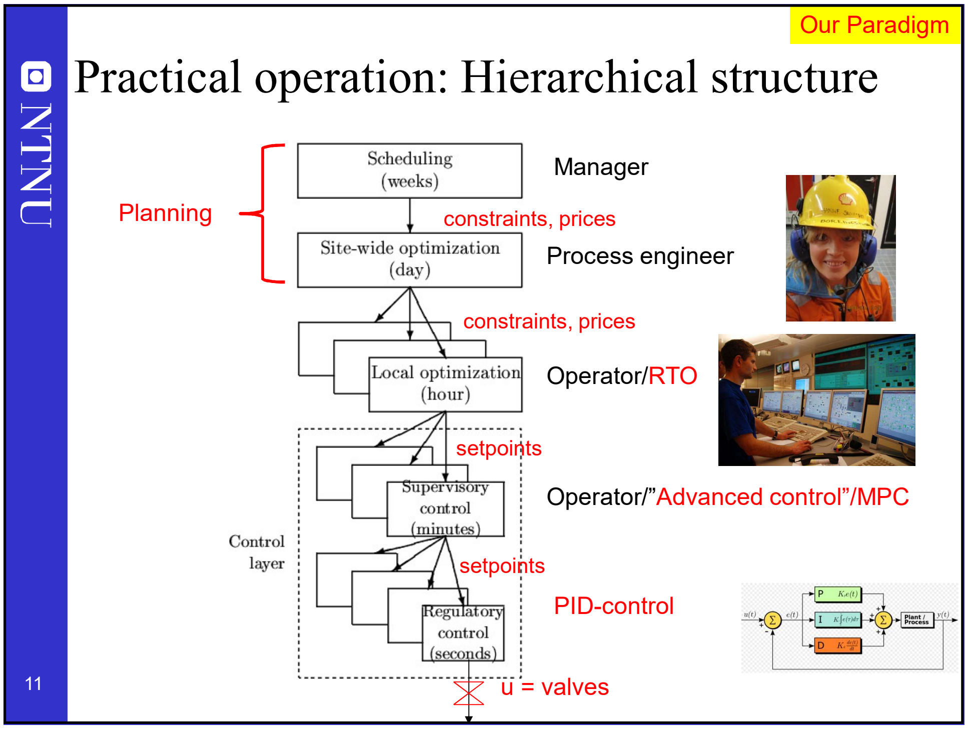
Our Paradigm
Practical operation: Hierarchical structure
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Planning
Scheduling (weeks)
Manager
constraints, prices
Site-wide optimization (day)
Process engineer
constraints, prices
Local optimization (hour)
Operator/RTO
setpoints
Supervisory control (minutes)
Operator/"Advanced control"/MPC
Control layer
setpoints
Regulatory control (seconds)
PID-control
u = valves
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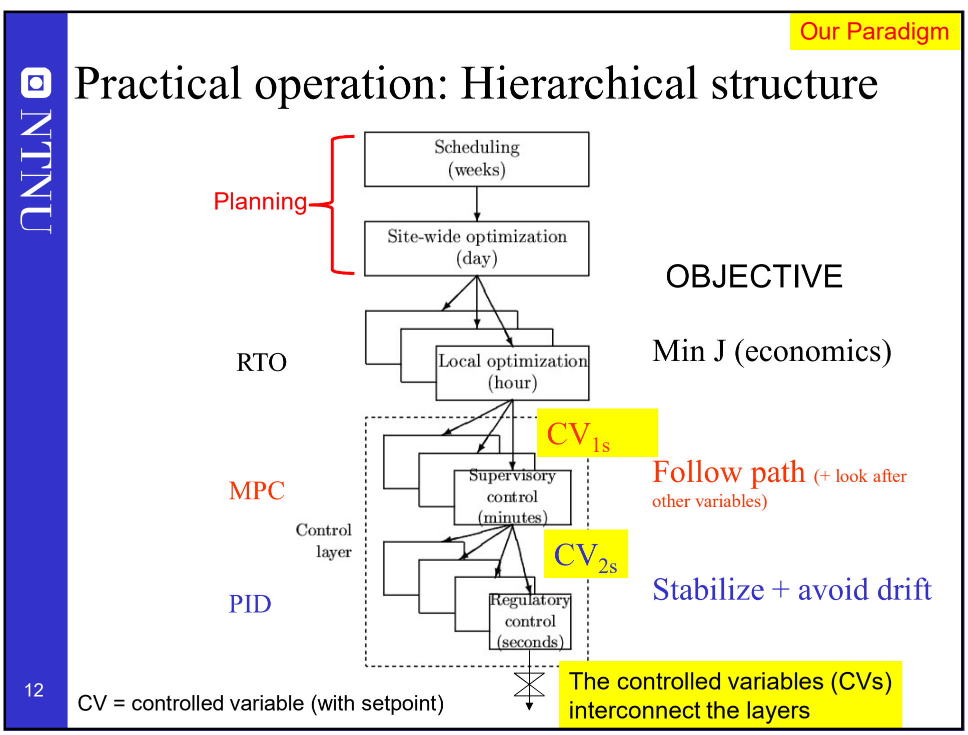
Our Paradigm
Practical operation: Hierarchical structure
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Planning
Scheduling (weeks)
Site-wide optimization (day)
OBJECTIVE
RTO
Local optimization (hour)
Min J (economics)
CV_{1s}
MPC
Supervisory control (minutes)
Follow path (+ look after other variables)
Control layer
CV_{2s}
PID
Regulatory control (seconds)
Stabilize + avoid drift
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CV = controlled variable (with setpoint)
The controlled variables (CVs) interconnect the layers

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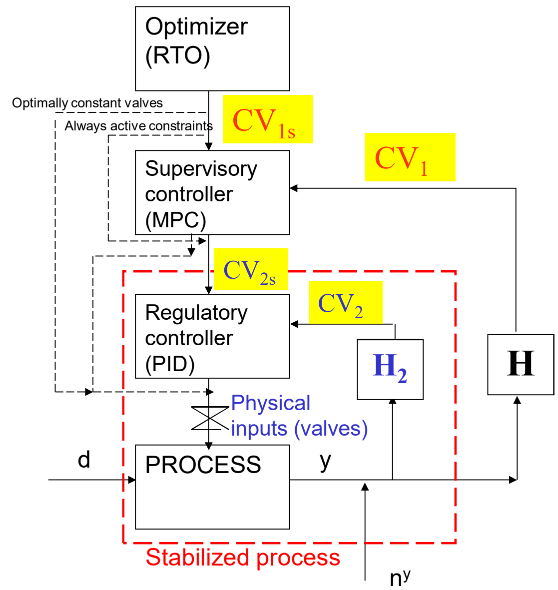

Degrees of freedom for optimization (usually steady-state DOFs), MVopt = CV1s
Degrees of freedom for supervisory control, MV1=CV2s + unused valves
Physical degrees of freedom for stabilizing control, MV2 = valves (dynamic process inputs)

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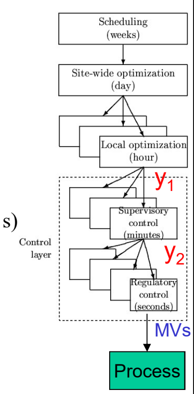
Control structure design procedure

I Top Down (mainly steady-state economics, y_1)

- **Step 1: Define** operational objectives (optimal operation)
  - Cost function J (to be minimized)
  - Operational constraints
- **Step 2:** Identify degrees of freedom (MVs) and **optimize** for expected disturbances
  - Identify **Active constraints**
- **Step 3**: **Select** primary "economic" controlled variables $c=y_1$ (CV1s)
  - **Self-optimizing variables (find H)**
- **Step 4:** Where locate the **throughput manipulator (TPM)?**



II Bottom Up (dynamics, y_2)

- **Step 5**: **Regulatory** / stabilizing control (PID layer)
  - What more to control (y_2; local CV2s)? **Find H_2**
  - Pairing of inputs and outputs
- **Step 6: Supervisory** control (MPC layer)
- **Step 7: Real-time optimization** (Do we need it?)

S. Skogestad, ``Control structure design for complete chemical plants'',
Computers and Chemical Engineering, **28** (1-2), 219-234 (2004).

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Step 1. Define optimal operation (economics)

- What are we going to use our degrees of freedom u (MVs) for?
- Define scalar cost function J(u,x,d)
  - u: degrees of freedom (usually steady-state)
  - d: disturbances
  - x: states (internal variables)

  Typical cost function:

  J = cost feed + cost energy – value products

- Optimize operation with respect to u for given d (usually steady-state):

  $\min_u \; J(u,x,d)$

  subject to:

  Model equations: $f(u,x,d) = 0$

  Operational constraints: $g(u,x,d) < 0$

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Step S2. Optimize

(a) Identify degrees of freedom

(b) Optimize for expected disturbances

- Need good model, usually steady-state
- Optimization is time consuming! But it is offline
- Main goal: Identify ACTIVE CONSTRAINTS
- A good engineer can often guess the active constraints

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Step S3: Implementation of optimal operation

- Have found the optimal way of operation. How should it be implemented?
- **What to control ?** (CV_1).
  1. Active constraints
  2. Self-optimizing variables (for unconstrained degrees of freedom)

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Optimal operation - Runner

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Optimal operation of runner

– Cost to be minimized, J=T
– One degree of freedom (u=power)
– What should we control?

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Optimal operation - Runner

1. Optimal operation of Sprinter

- 100m. J=T
- Active constraint control:
  - Maximum speed (”no thinking required”)
  - CV = power (at max)

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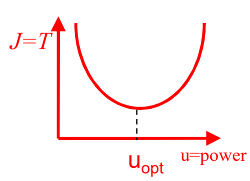
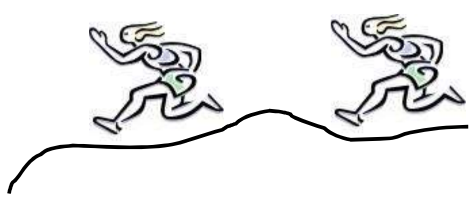
Optimal operation - Runner

2. Optimal operation of Marathon runner

- 40 km. J=T
- What should we control? CV=?
- Unconstrained optimum





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Optimal operation - Runner

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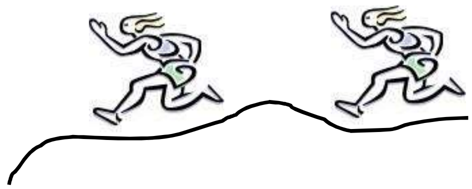
Self-optimizing control: Marathon (40 km)

- Any self-optimizing variable (to control at constant setpoint)?
  - c_1 = distance to leader of race
  - c_2 = speed
  - c_3 = heart rate
  - c_4 = level of lactate in muscles



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Optimal operation - Runner

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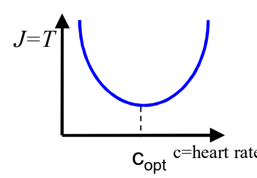
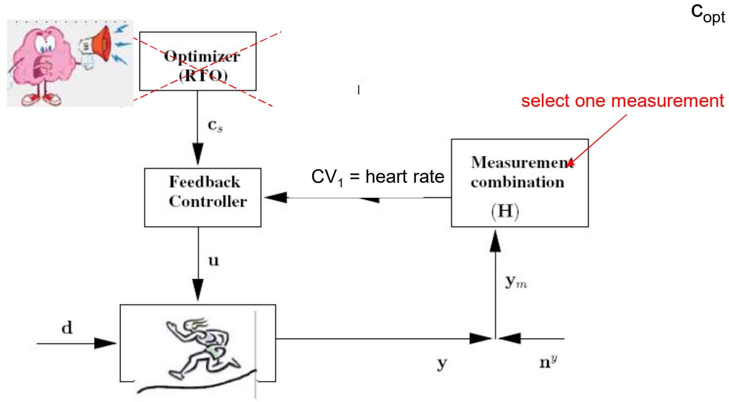
Conclusion Marathon runner





- CV = heart rate is good “self-optimizing” variable
- Simple and robust implementation
- Disturbances are indirectly handled by keeping a constant heart rate
- May have infrequent adjustment of setpoint (c_s)

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Summary Step 3.
What should we control (CV_1)?

Selection of primary controlled variables $c = CV_1$

1. **Control active constraints!**
2. **Unconstrained variables: Control self-optimizing variables!**

- Old idea (Morari *et al.*, 1980):

"We want to find a function c of the process variables which when held constant, leads automatically to the optimal adjustments of the manipulated variables, and with it, the optimal operating conditions."

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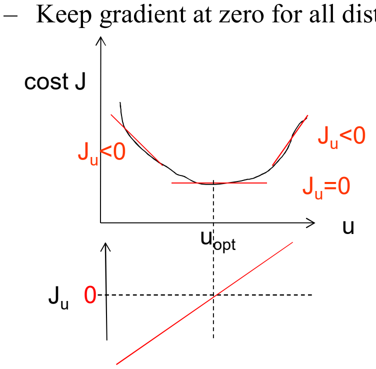
Unconstrained degrees of freedom

The ideal "self-optimizing" variable is the gradient, J_u

$$c = \partial J/\partial u = J_u$$

– Keep gradient at zero for all disturbances ($c = J_u = 0$)



Problem: Usually no measurement of gradient

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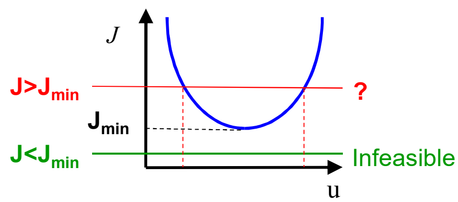
Never try to control the cost function J

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



- Better: control its gradient, Ju, or an associated “self-optimizing” variable.

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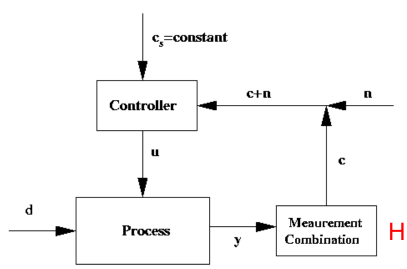
Ideal: c = J_u

In practise, use available measurements: c = H y.

Task: Select H!



- Single measurements:

$$\mathbf{c} = \mathbf{H}\mathbf{y} \qquad \mathbf{H} = \begin{bmatrix} 1 & 0 & 0 & 0 \\ 0 & 1 & 0 & 0 \end{bmatrix}$$

- Combinations of measurements:

$$\mathbf{c} = \mathbf{H}\mathbf{y} \qquad \mathbf{H} = \begin{bmatrix} h_{11} & h_{12} & h_{13} & h_{14} \\ h_{21} & h_{22} & h_{23} & h_{24} \end{bmatrix}$$

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Combinations of measurements, c= Hy Nullspace method for H (Alstad): HF=0 where F=dy_{opt}/dd

$$J_u(u,d) = \underbrace{J_u(u_{opt}(d),d)}_{=0} + J_{uu}\cdot(u-u_{opt})$$

$u - u_{opt} = (HG^y)^{-1}(c - c_{opt})$

Here: $c - c_{opt} = \Delta c - \Delta c_{opt}$

where we have introduced deviation variables around a nominal optimal point (c^*, d^*) (where $c^* = c_{opt}(d^*)$)

Assume perfect control of c (no noise): $\Delta c = 0$

Optimal change: $\Delta c_{opt} = H\Delta y_{opt} = HF\Delta d$

Gives: $J_u = -J_{uu}(HG^y)^{-1}HF\Delta d$

$\Rightarrow HF = 0$ gives $J_u = 0$ for any disturbance Δd

Gives J_u=0

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- Proof. Appendix B in: Jäschke and Skogestad, "NCO tracking and self-optimizing control in the context of real-time optimization", *Journal of Process Control*, 1407-1416 (2011)

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Example. Nullspace Method for Marathon runner

u = power, d = slope [degrees]

y_1 = hr [beat/min], y_2 = v [m/s]

c = Hy, H = [h_1 h_2]]

F = dy_{opt}/dd = [0.25 -0.2]'

HF = 0 -> $h_1 f_1 + h_2 f_2 = 0.25 h_1 - 0.2 h_2 = 0$

Choose $h_1 = 1$ -> $h_2 = 0.25/0.2 = 1.25$

Conclusion: **c = hr + 1.25 v**

Control **c = constant** -> hr increases when v decreases (OK uphill!)

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With measurement noise

“Exact local method”

“=0” in nullspace method (no noise)

$$\min_H \| J_{uu}^{1/2} (HG^y)^{-1} H [F W_d \quad W_{n^y}] \|_2$$

“Minimize” in Maximum gain rule
(maximize $S_1 \, G \, J_{uu}^{-1/2}$, $G=HG^y$)

“Scaling” S_1

Analytical solution:
$H = G^{yT}(YY^T)^{-1}$ where $Y = [FW_d \quad W_{n^y}]$

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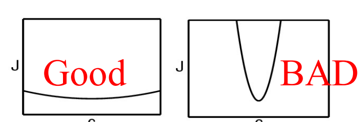
In practice: What variable c=Hy should we control? (for self-optimizing control)

$$\min_H \| J_{uu}^{1/2} (HG^y)^{-1} H [F W_d \quad W_{n^y}] \|_2$$

1. **The *optimal value* of c should be *insensitive* to disturbances**
   - Small HF = dc_{opt}/dd
2. **The *value* of c should be *sensitive* to the inputs** (“**maximum gain rule**”)
   - Large G = HG^y = dc/du
   - Equivalent: Want flat optimum



(b) Flat optimum: Implementation easy

(c) Sharp optimum: Sensitive to implementation erros

Note: Must also find optimal setpoint for c=CV_1

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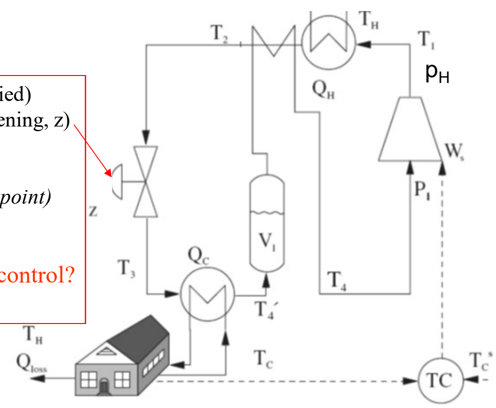
Example: CO2 refrigeration cycle

$J = W_s$ (work supplied)
DOF = u (valve opening, z)
Main disturbances:

$d_1 = T_H$
$d_2 = T_{Cs}$ *(setpoint)*
$d_3 = UA_{loss}$

What should we control?



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CO2 refrigeration cycle

Step 1. One (remaining) degree of freedom (u=z)

Step 2. Objective function. $J = W_s$ (compressor work)

Step 3. Optimize operation for disturbances ($d_1=T_C$, $d_2=T_H$, $d_3=UA$)

- Optimum always unconstrained

Step 4. Implementation of optimal operation

- No good single measurements (all give large losses):
  - p_h, T_h, z, …
- Nullspace method: Need to combine $n_u+n_d=1+3=4$ measurements to have zero disturbance loss
- Simpler: Try combining two measurements. Exact local method:
  - $c = h_1 p_h + h_2 T_h = p_h + k T_h$; $k = -8.53$ bar/K
- Nonlinear evaluation of loss: OK!

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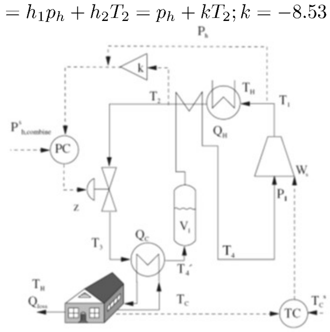
CO2 cycle: Maximum gain rule

Linear "maximum gain" analysis of controlled variables for CO_2 case

Variable (y)	Nom.	$G = \frac{\Delta y}{\Delta u}$	$\lvert\Delta y_{opt}(d_i)\rvert$ d_1 (T_H)	d_2 (T_C)	d_3 (UA_{loss})	$\lvert\Delta y_{opt}\rvert$	n	Span y $\lvert\Delta y_{opt}\rvert + n$	$\lvert G'\rvert = \frac{\lvert G\rvert}{\text{span}y}$
P_h/T_2' (bar °C^{-1})	0.32	−0.291	0.140	−0.047	0.093	0.174	0.0033	0.177	0.25
P_h (bar)	97.61	−78.85	48.3	−15.5	31.0	59.4	1.0	60.4	1.31
T_2' (°C)	35.5	36.7	16.27	−2.93	7.64	18.21	1	19.2	1.91
$T_2' - T_H$ (°C)	3.62	24	4.10	−1.92	5.00	6.75	1.5	8.25	2.91
z	0.34	1	0.15	−0.04	0.18	0.24	0.05	0.29	3.45
V_1 (m^3)	0.07	0.03	−0.02	0.005	−0.03	0.006	0.001	0.007	4.77
T_2 (°C)	25.5	60.14	8.37	0.90	3.18	9.00	1	10.0	6.02
$P_{h,combine}$ (bar)	97.61	−592.0	−23.1	−23.1	3.91	33.0	9.53	42.5	13.9
m_{gco} (kg)	4.83	−11.18	0.151	−0.136	0.119	0.235	0.44	0.675	16.55

Nullspace method: $c = p_{h,\text{combine}} = h_1 p_h + h_2 T_2 = p_h + kT_2; k = -8.53$ bar/K



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J.B. Jensen, S. Skogestad / Computers and Chemical Engineering 31 (2007) 1590–1601

CV=Measurement combination

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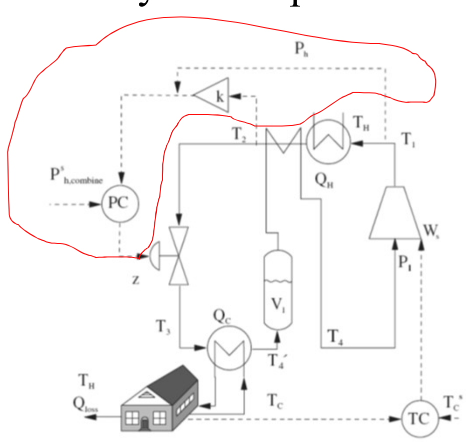
Refrigeration cycle: Proposed control structure



CV1= Room temperature
CV2= "temperature-corrected high CO2 pressure"

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Step 4. Where set production rate?

- Where locale the **TPM (throughput manipulator)?**
  - The ”gas pedal” of the process
- Very important!
- Determines structure of remaining inventory (level) control system
- Suggestion: Set production rate at (dynamic) bottleneck
- Link between **Top-down** and **Bottom-up** parts

- NOTE: TPM location is a dynamic issue.
  Link to economics: Better control of active constraints (reduce backoff)

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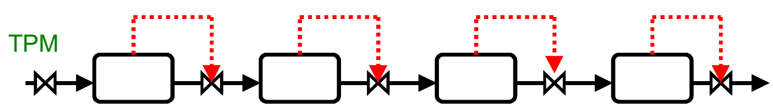
Production rate set at inlet : Inventory control in direction of flow*



* Required to get “local-consistent” inventory control

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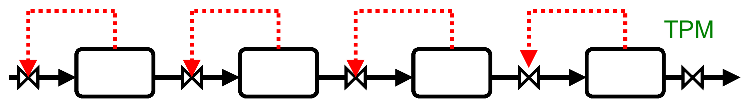
Production rate set at outlet: Inventory control opposite flow



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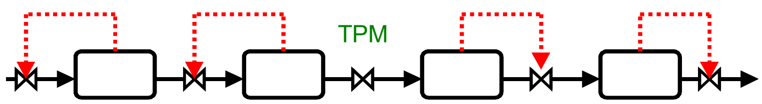
Production rate set inside process



General: "Need radiating inventory control around TPM" (Georgakis)

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QUIZ

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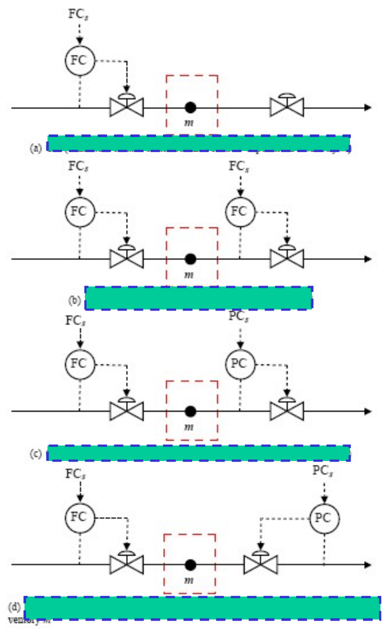
CONSISTENT?



YES

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

NO

YES

NO

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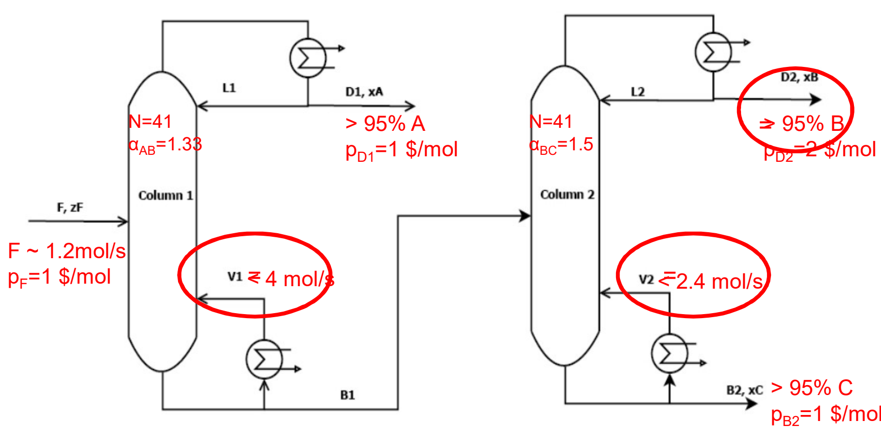
Operation of Distillation columns in series

- Cost (J) = - Profit = $p_F F + p_V(V_1+V_2) - p_{D1}D_1 - p_{D2}D_2 - p_{B2}B_2$
- Prices: $p_F=p_{D1}=P_{B2}=1$ \$/mol, $p_{D2}=2$ \$/mol, Energy p_V= 0-0.2 \$/mol (varies)
- With given feed and pressures: 4 steady-state DOFs.
- Here: 5 constraints (3 products > 95% + 2 capacity constraints on V)



DOF = Degree Of Freedom
Ref.: M.G. Jacobsen and S. Skogestad (2011)

QUIZ: What are the expected active constraints?
1. Always. 2. For low energy prices.

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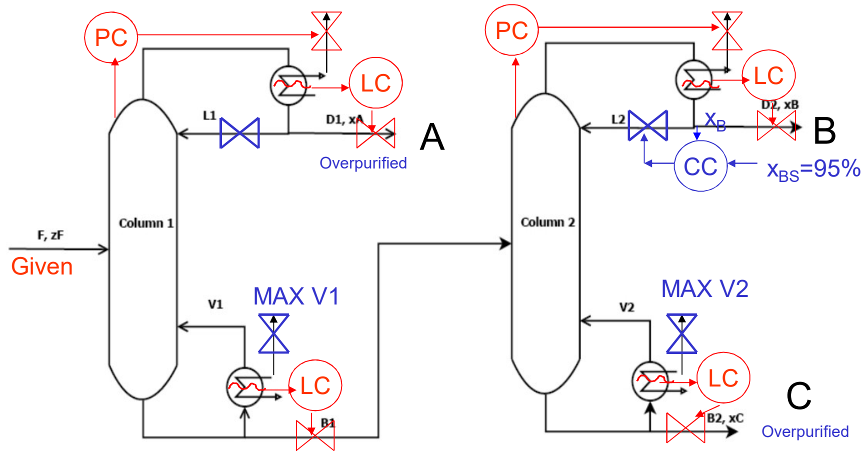
Example.

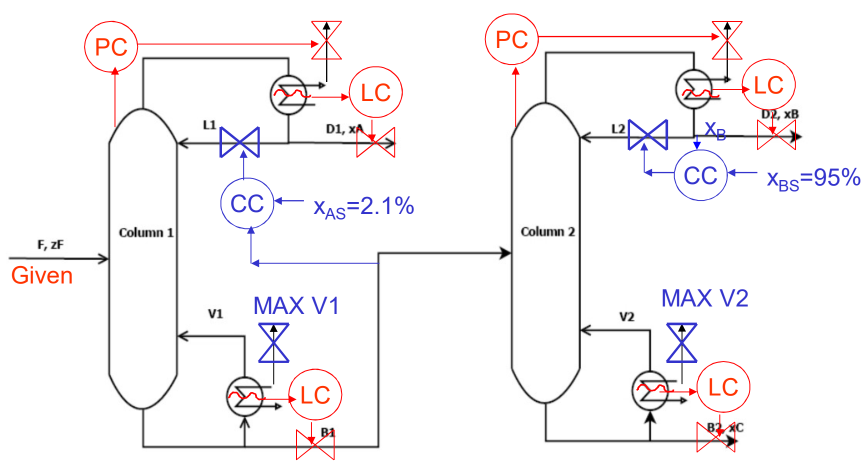
Control of Distillation columns. Cheap energy



Overpurified: To avoid loss of valuable product B

Solution.

Control of Distillation columns. Cheap energy



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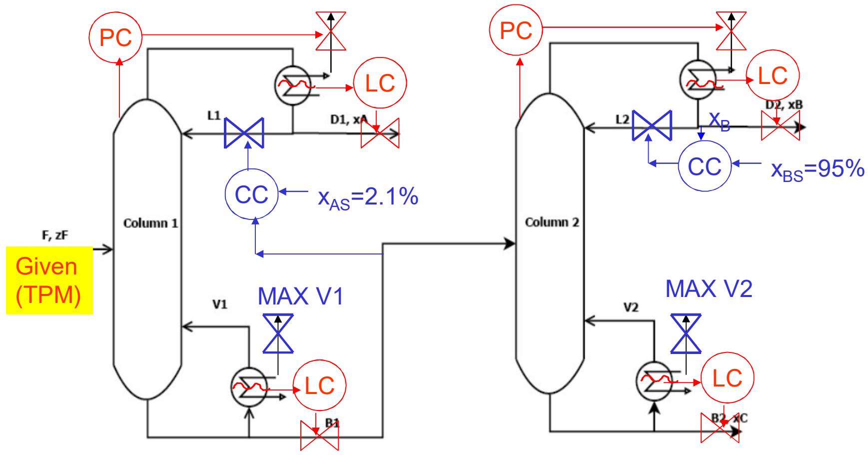
What happens if we increase the federate?



Is this control structure still OK??

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TPM

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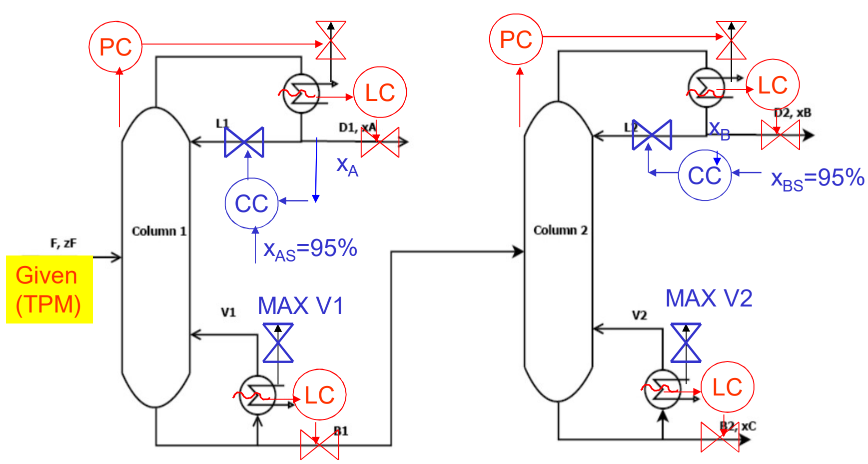
Increase federate: Reach x_A-constraint



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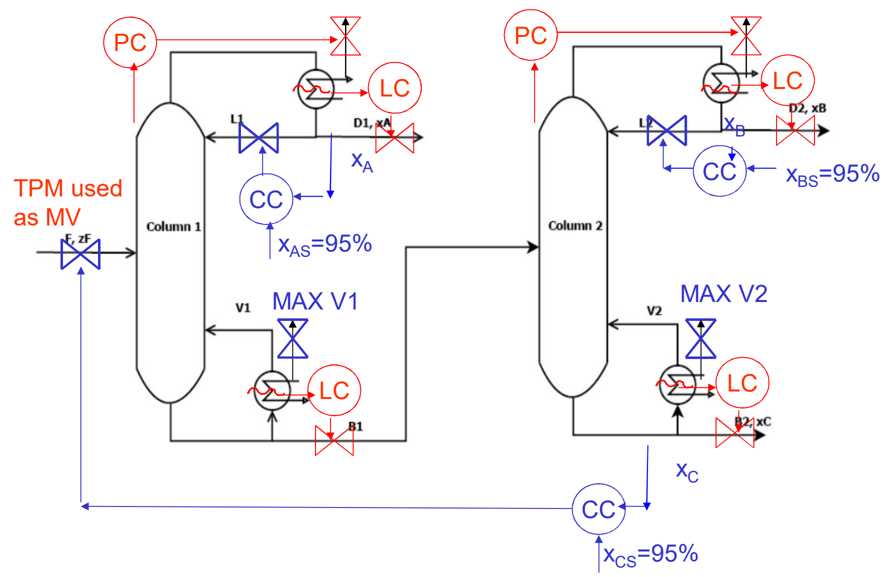
TPM

Increase federate further: Reach also x_C-constraint (Bottleneck)



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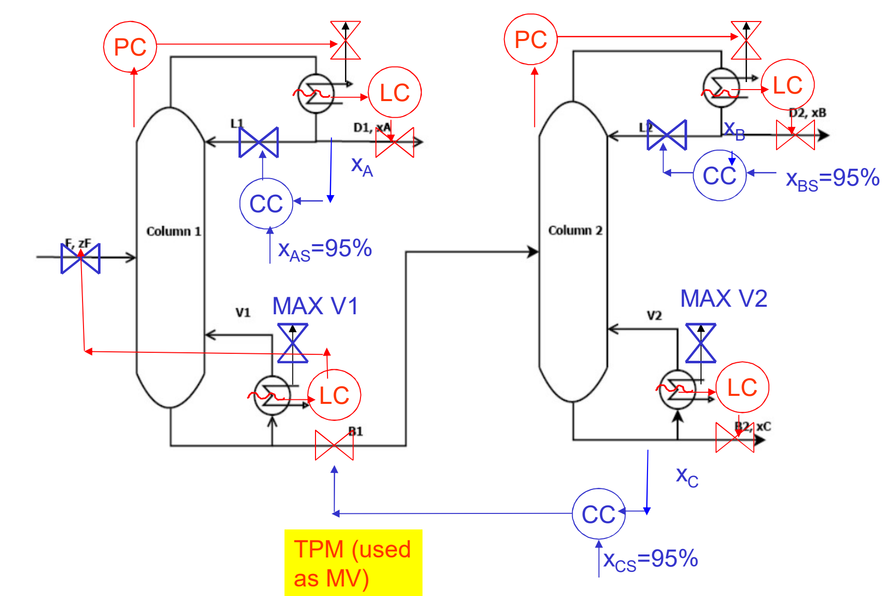
Move TPM

Move TPM to F2 (closer to bottleneck) and rearrange level loop



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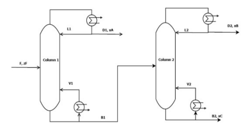
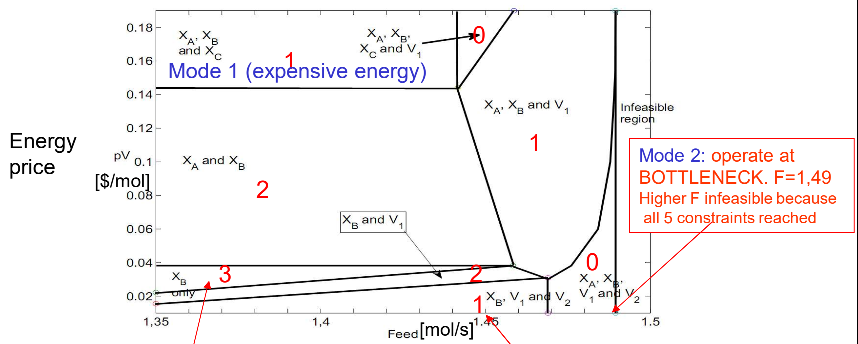
Distillation example: Not so simple

Active constraint regions for two distillation columns in series





Mode 1, Cheap energy: 3 active constraints -> 1 remaining unconstrained DOF (L_1) -> Need to find 1 additional CVs (“self-optimizing”)

More expensive energy: Only 1 active constraint (xB) ->3 remaining unconstrained DOFs -> Need to find 3 additional CVs (“self-optimizing”)

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CV = Controlled Variable

How many active constraints regions?

- Maximum: 2^{n_c}

n_c = number of constraints

BUT there are usually fewer in practice

- Certain constraints are always active (reduces effective n_c)
- Only n_u can be active at a given time
  - n_u = number of MVs (inputs)
- Certain constraints combinations are not possibe
  - For example, max and min on the same variable (e.g. flow)
- Certain regions are not reached by the assumed disturbance set

Distillation
$n_c = 5$
$2^5 = 32$

x_B always active
2^4 = 16
-1 = 15

In practice = 8

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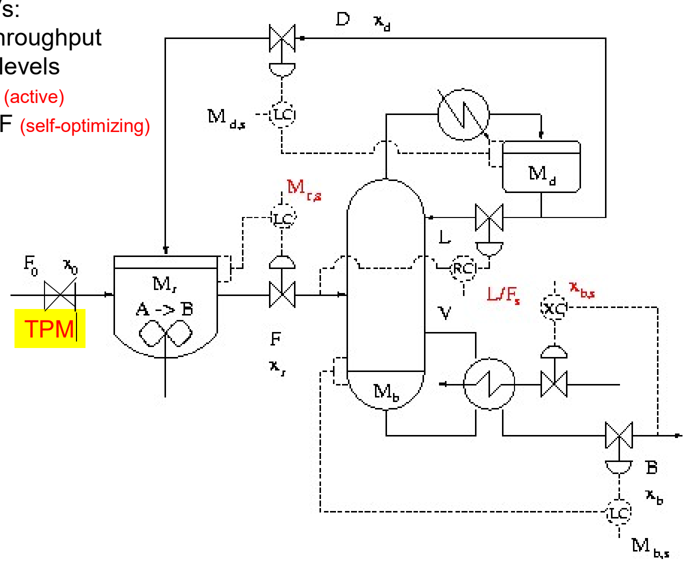
Reactor-recycle process
1. Given throughput: Minimize energy V

6 MVs

6 CVs:

- Throughput
- 3 levels
- x_b (active)
- L/F (self-optimizing)



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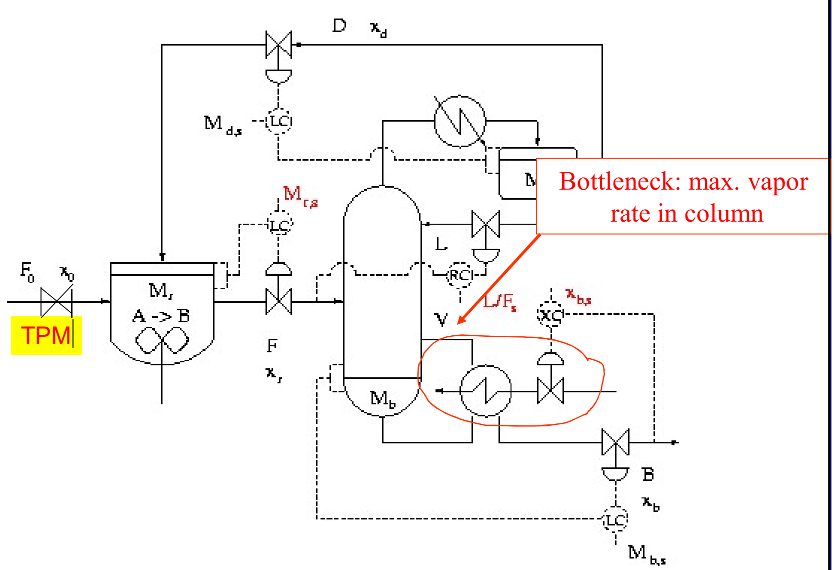
Reactor-recycle process:
2. Maximize throughput: reach bottleneck in column



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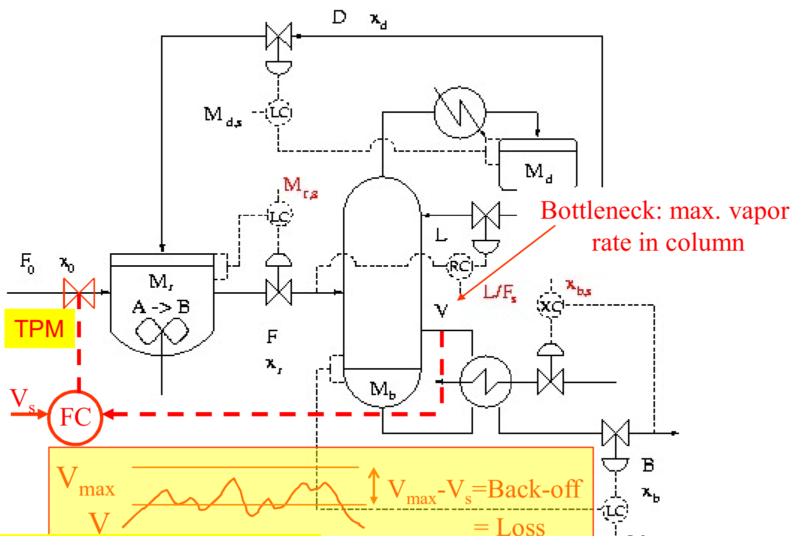
Reactor-recycle process with max. throughput

Alt.A: Feedrate controls bottleneck flow



Get "long loop": Need back-off in V

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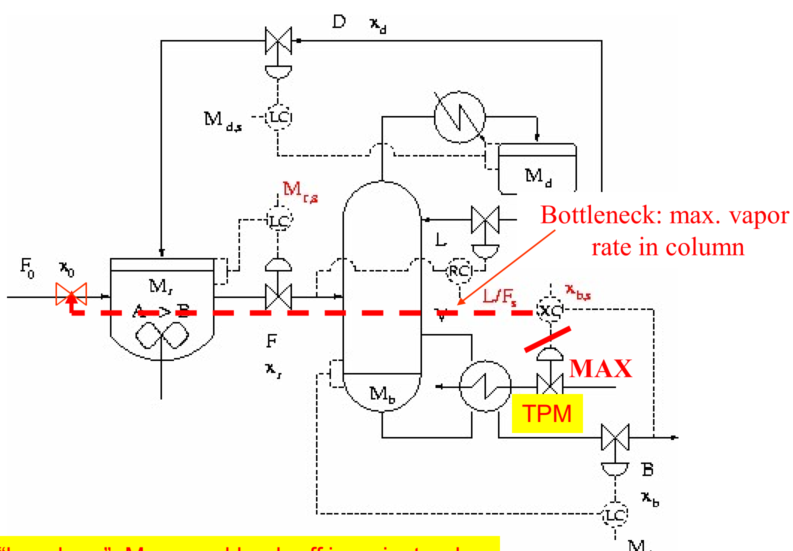
Reactor-recycle process with max. throughput:

Alt. B TPM to bottleneck. Feedrate for lost task (x_B)



Get "long loop": May need back-off in x_B instead…

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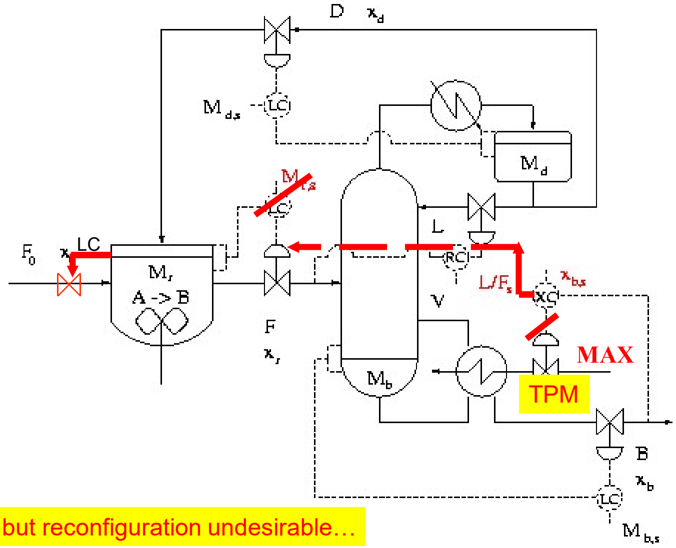
Reactor-recycle process with max. feedrate:

Alt. C: Best economically: Move TPM to bottleneck (MAX) + Reconfigure upstream loops



OK, but reconfiguration undesirable…

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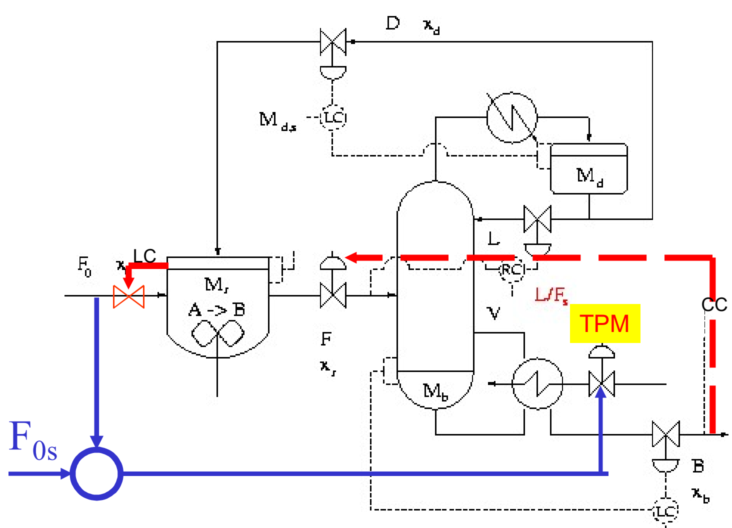
Reactor-recycle process:

Alt.C': Move TPM + reconfigure (permanently!)



For cases with given feedrate: Get "long loop" but no associated loss

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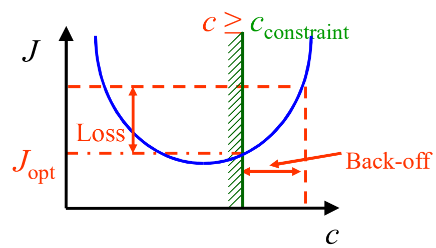
Backoff

Back-off: distance to active constraint to guarantee feasibility



a) If constraint can be violated dynamically (only average matters)

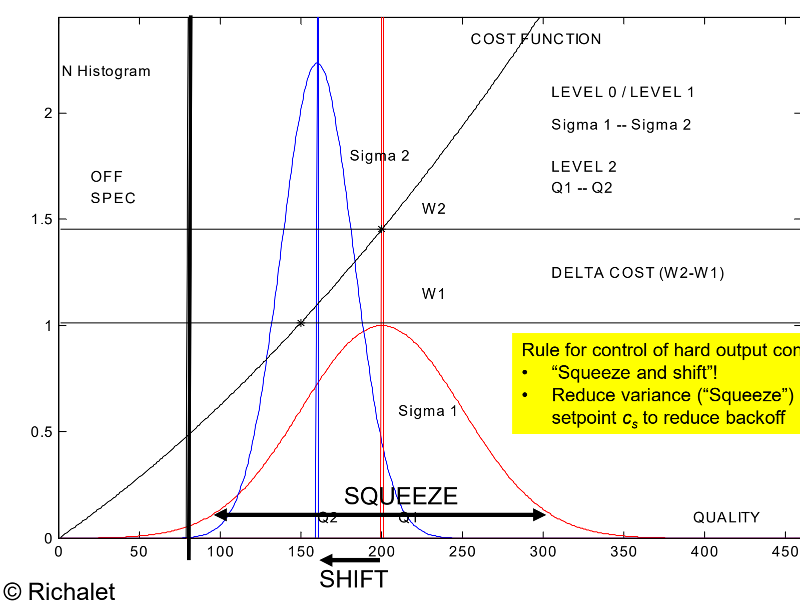
- Required Back-off = "measurement bias" (steady-state measurement error for c)

b) If constraint cannot be violated dynamically ("hard constraint")

- Required Back-off = "measurement bias" + maximum dynamic control error

Want tight control of hard output constraints to reduce the back-off. "Squeeze and shift"-rule

Hard Constraints: «SQUEEZE AND SHIFT»



Rule for control of hard output constraints:

- "Squeeze and shift"!
- Reduce variance ("Squeeze") and "shift" setpoint c_s to reduce backoff

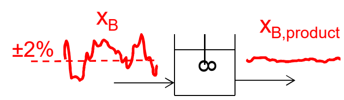
© Richalet

CV = Active constraint

Example back-off.
x_B = purity product > 95% (min.)

- D_1 directly to customer (hard constraint)
  - Measurement error (bias): 1%
  - Control error (variation due to poor control): 2%
  - Backoff = 1% + 2% = 3%
  - Setpoint x_{Bs}= 95 + 3% = 98% (to be safe)
  - Can reduce backoff with better control ("squeeze and shift")

- D_1 to large mixing tank (soft constraint)
  - Measurement error (bias): 1%
  - Backoff = 1%
  - Setpoint x_{Bs}= 95 + 1% = 96% (to be safe)
  - Do not need to include control error because it averages out in tank



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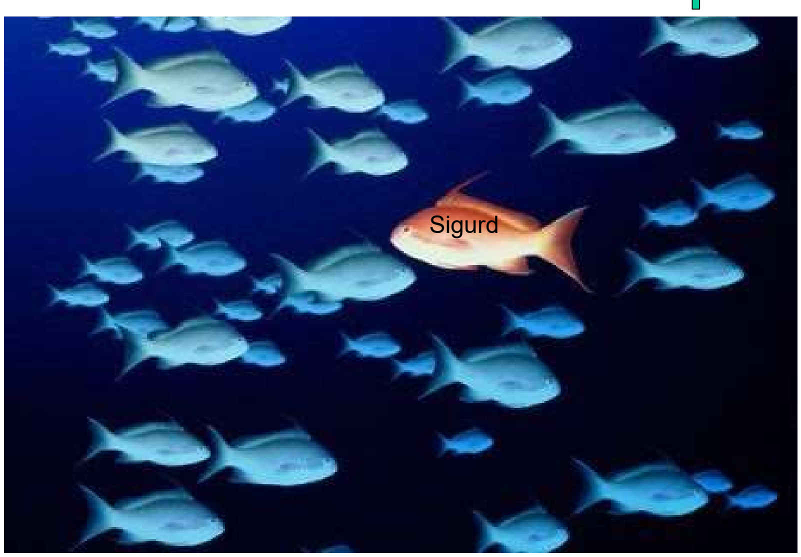
Academic process control community fish pond

Optimal centralized Solution (EMPC)



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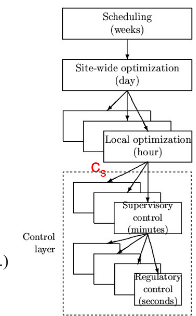
Conclusion: Systematic procedure for plantwide control

- **Start "top-down" with economics:**
  - Step 1: Define operational objectives and identify degrees of freeedom
  - Step 2: Optimize steady-state operation.
  - Step 3A: Identify active constraints = primary CVs c.
  - Step 3B: Remaining unconstrained DOFs: Self-optimizing CVs c.
  - Step 4: Where to set the throughput (often best: at bottleneck)
- **Regulatory control I: Move mass through the plant:**
  - Step 5A: Propose "local-consistent" inventory (level) control structure.
- **Regulatory control II: "Bottom-up" stabilization of the plant**
  - Step 5B: Control variables to stop "drift" (sensitive temperatures, pressures, ...)
    - Pair variables to avoid interaction and saturation
- **Finally: Make link between "top-down" and "bottom up".**
  - Step 6: "Advanced/supervisory control" system (MPC):
    - CVs: Active constraints and self-optimizing economic variables +
    - look after variables in layer below (e.g., avoid saturation)
    - MVs: Setpoints to regulatory control layer.
    - Coordinates within units and possibly between units



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

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Summary and 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).
- Another paper:
  - S. Skogestad "Plantwide control: the search for the self-optimizing control structure'", *J. Proc. Control*, 10, 487-507 (2000).
- More information:

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

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