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PLANTWIDE CONTROL

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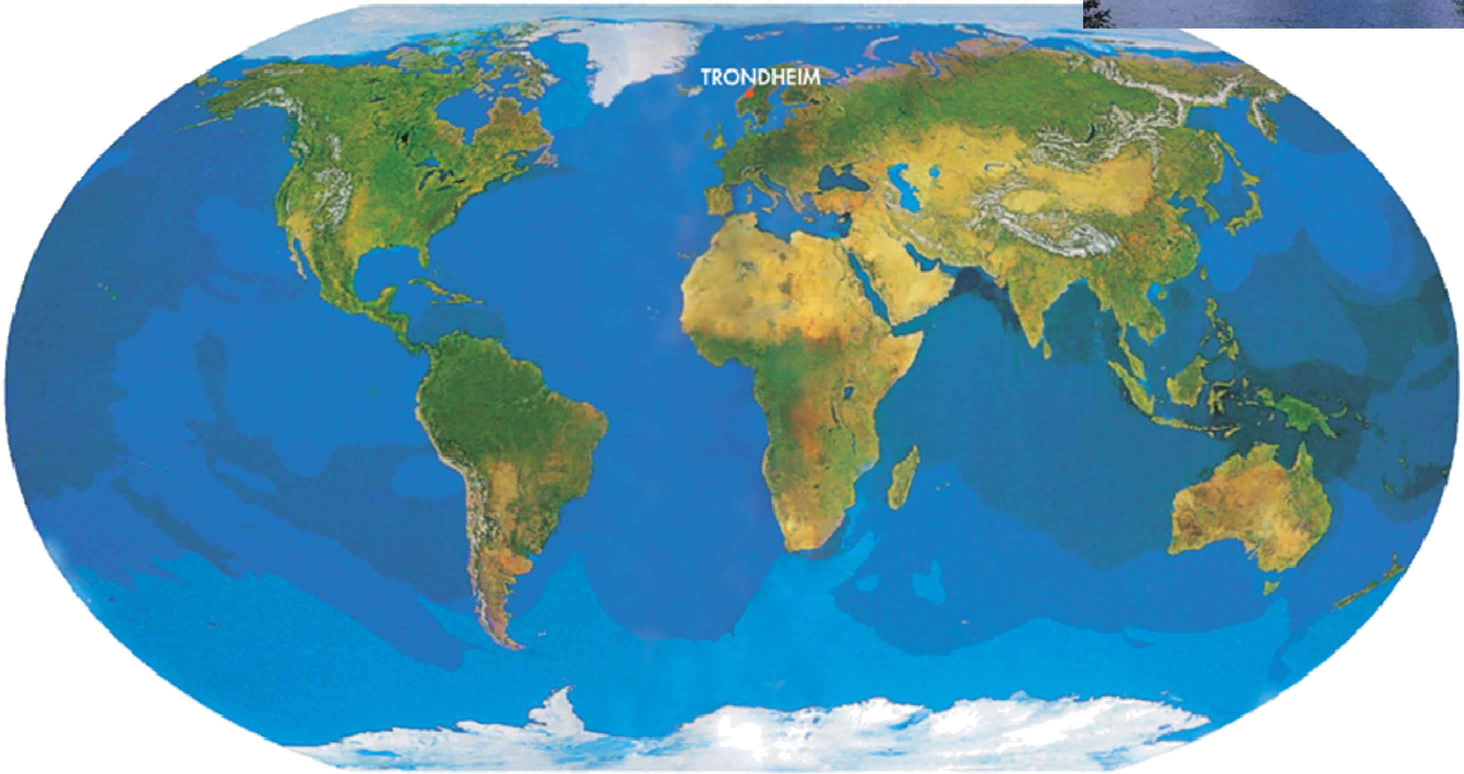

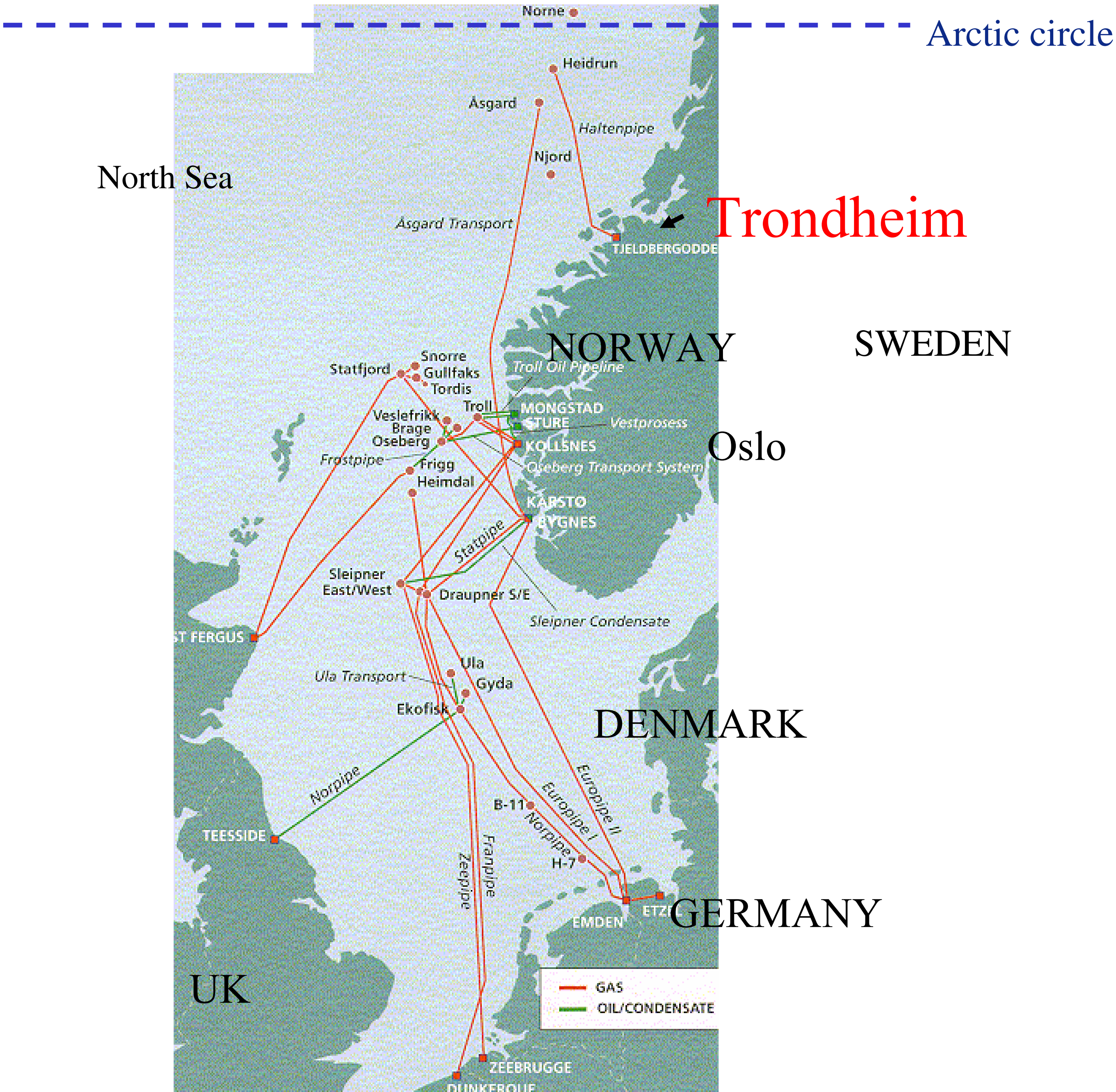
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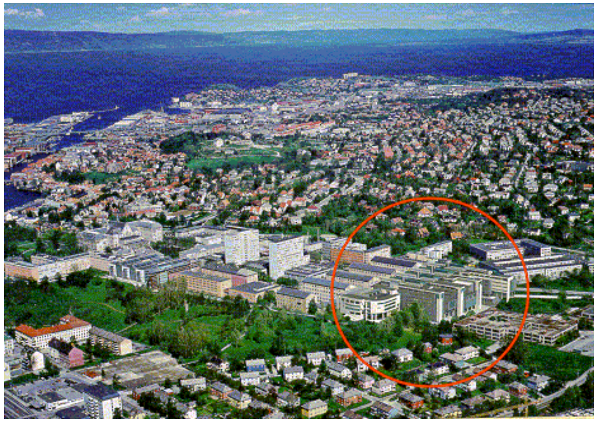

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Outline

- Control structure design (plantwide control)
- A procedure for control structure design

  I Top Down
  - Step 1: Degrees of freedom
  - Step 2: Operational objectives (optimal operation)
  - Step 3: What to control ? (primary CV's) (self-optimizing control)
  - Step 4: Where set production rate?

  II Bottom Up
  - Step 5: Regulatory control: What more to control (secondary CV's) ?
  - Step 6: Supervisory control
  - Step 7: Real-time optimization
- Case study: HDA
- ++ Simple PID tuning rules

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Main message

- 1. Control for economics (Top-down steady-state arguments)
  - Primary controlled variables $c = y_1$
- 2. Control for stabilization (Bottom-up; regulatory PID control)
  - Secondary controlled variables y_2 ("inner cascade loops")

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

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

- 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.*

- Carl Nett (1989):

Minimize control system complexity subject to the achievement of accuracy specifications in the face of uncertainty.

"Plantwide control" = "Control structure design for complete chemical plant"

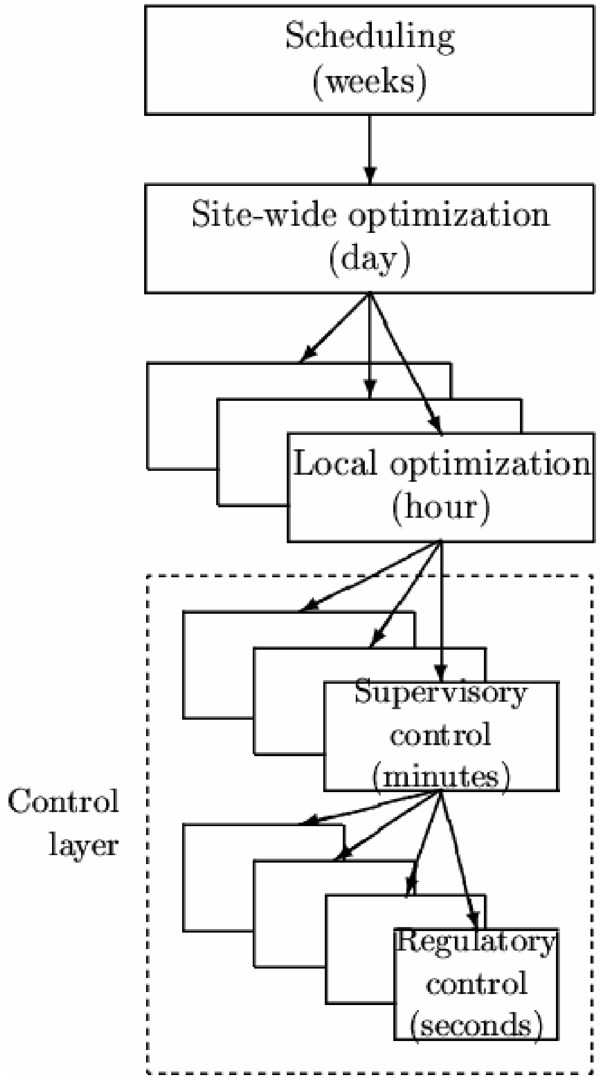
- *Not* the tuning and behavior of each control loop,
- But rather the *control philosophy* of the overall plant with emphasis on the ***structural decisions***:
  - *Selection of controlled variables ("outputs")*
  - *Selection of manipulated variables ("inputs")*
  - *Selection of (extra) measurements*
  - *Selection of control* ***configuration*** (structure of overall controller that interconnects the controlled, manipulated and measured variables)
  - *Selection of controller type* (PID, decoupler, MPC, LQG etc.).

Main simplification: Hierarchical structure

RTO

MPC

PID



Need to define objectives and identify main issues for each layer

Regulatory control (seconds)

- *Purpose*: “Stabilize” the plant by controlling selected ‘’secondary’’ variables (y_2) such that the plant does not drift too far away from its desired operation

- Use simple single-loop PI(D) controllers

- *Status:* Many loops poorly tuned
  - Most common setting: K_c=1, $_I$=1 min (default)
  - Even wrong sign of gain K_c ….

Regulatory control……...

- *Trend:* Can do better! Carefully go through plant and retune important loops using standardized tuning procedure

- Exists many tuning rules, including Skogestad (SIMC) rules:
  - K_c = (1/k) ($_1$/ [$_c$ +]) $_I$ = min ($_1$, 4[$_c$ +]), Typical: $_c$=
  - "Probably the best simple PID tuning rules in the world" © Carlsberg

- *Outstanding structural issue*: What loops to close, that is, which variables (y_2) to control?

Supervisory control (minutes)

- *Purpose*: Keep primary controlled variables ($c=y_1$) at desired values, using as degrees of freedom the setpoints y_{2s} for the regulatory layer.

- *Status:* Many different "advanced" controllers, including feedforward, decouplers, overrides, cascades, selectors, Smith Predictors, etc.

- *Issues:*
  - Which variables to control may change due to change of "active constraints"
  - Interactions and "pairing"

Supervisory control……

- *Trend:* Model predictive control (MPC) used as unifying tool.
  - Linear multivariable models with input constraints
  - Tuning (modelling) is time-consuming and expensive
- *Issue:* When use MPC and when use simpler single-loop decentralized controllers ?
  - MPC is preferred if active constraints (“bottleneck”) change.
  - Avoids logic for reconfiguration of loops
- *Outstanding structural issue*:
  - What primary variables $c=y_1$ to control?

Local optimization (hour)

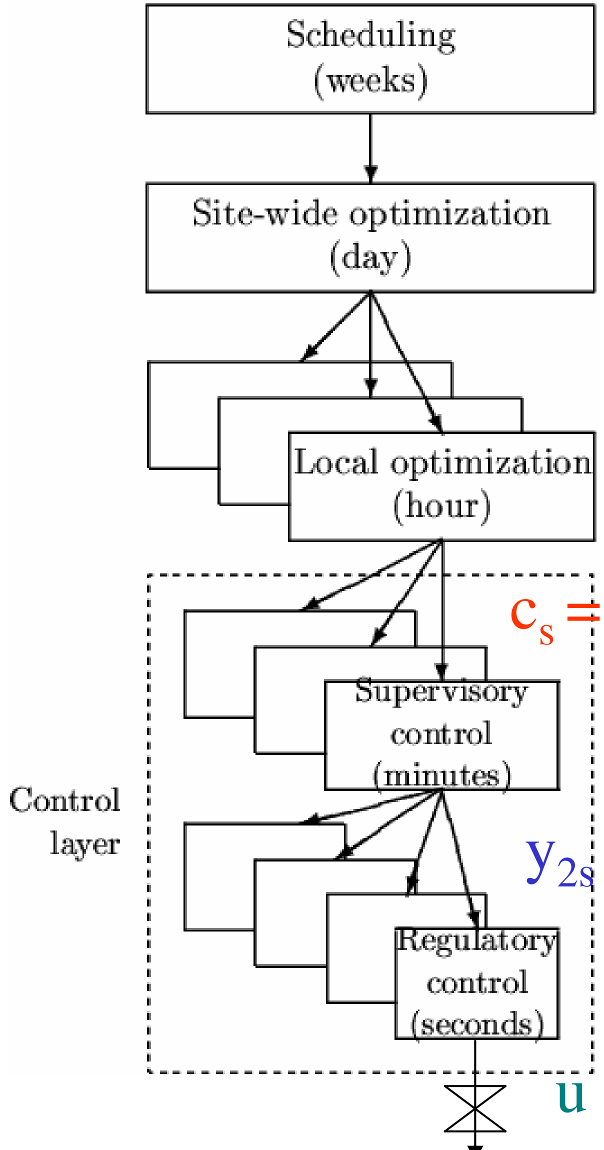
- *Purpose*: Minimize cost function J and:
  - Identify active constraints
  - Recompute optimal setpoints y_{1s} for the controlled variables

- *Status:* Done manually by clever operators and engineers

- *Trend:* Real-time optimization (RTO) based on detailed nonlinear steady-state model

- *Issues:*
  - Optimization not reliable.
  - Need nonlinear steady-state model
  - Modelling is time-consuming and expensive

Objectives of layers: MV's and CV's



RTO

Min J (economics); MV=y_{1s}

$c_s = y_{1s}$

MPC

CV=y_1; MV=y_{2s}

y_{2s}

PID

CV=y_2; MV=u

u (valves)

Stepwise procedure plantwide control

I. TOP-DOWN

Step 1. DEGREES OF FREEDOM
Step 2. OPERATIONAL OBJECTIVES
Step 3. WHAT TO CONTROL? (primary CV's $c=y_1$)
Step 4. PRODUCTION RATE

II. BOTTOM-UP (structure control system):

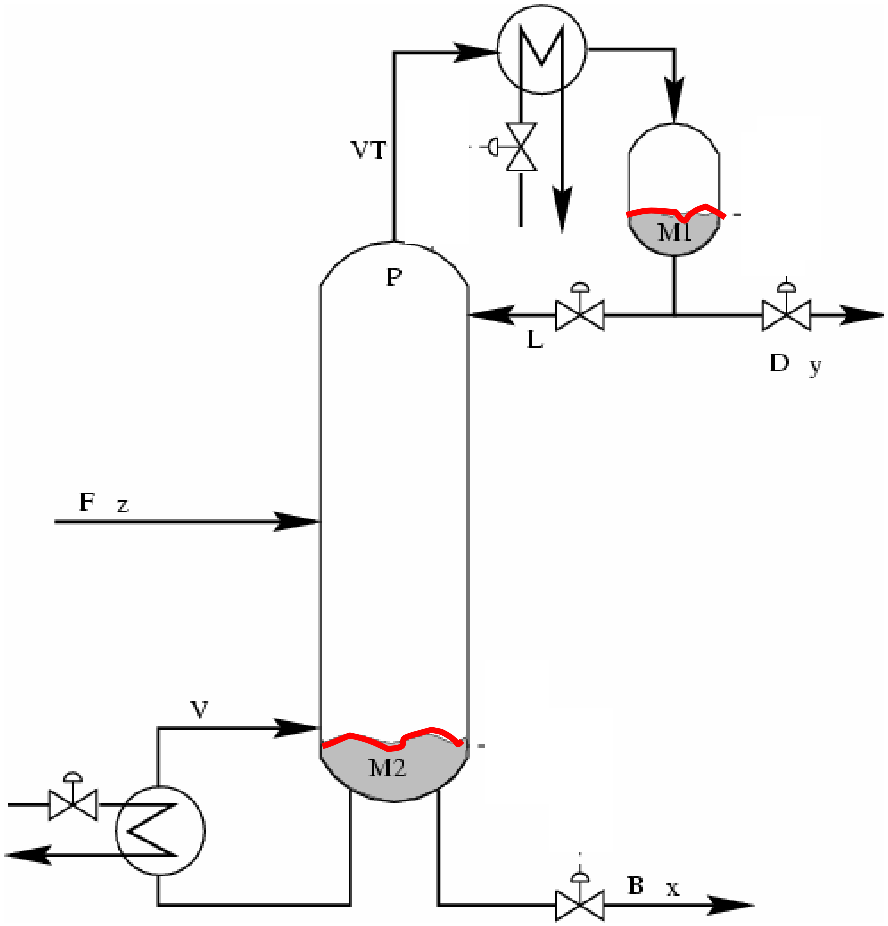
Step 5. REGULATORY CONTROL LAYER (PID)
"Stabilization"
What more to control? (secondary CV's y_2)
Step 6. SUPERVISORY CONTROL LAYER (MPC)
Decentralization
Step 7. OPTIMIZATION LAYER (RTO)
Can we do without it?

Steady-state degrees of freedom (N_{ss}): Typical number for some process units

- each external feedstream: 1 (feedrate)
- splitter: n-1 (split fractions) where n is the number of exit streams
- mixer: 0
- compressor, turbine, pump: 1 (work)
- adiabatic flash tank: 0*
- liquid phase reactor: 1 (holdup-volume reactant)
- gas phase reactor: 0*
- heat exchanger: 1 (duty or net area)
- distillation column excluding heat exchangers: 0* + number of sidestreams
- pressure* : add 1DOF at each extra place you set pressure (using an extra valve, compressor or pump!). Could be for adiabatic flash tank, gas phase reactor, distillation column

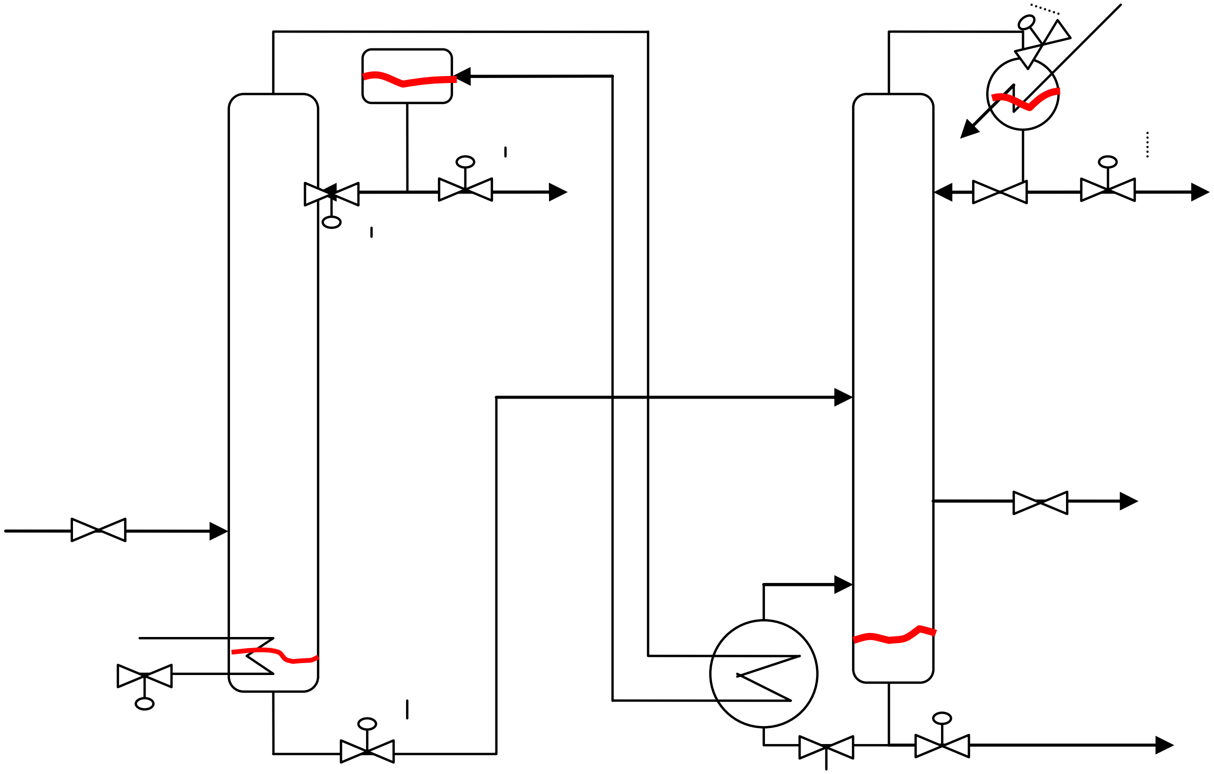
* Pressure is normally assumed to be given by the surrounding process and is then not a degree of freedom

Distillation column with given feed and pressure



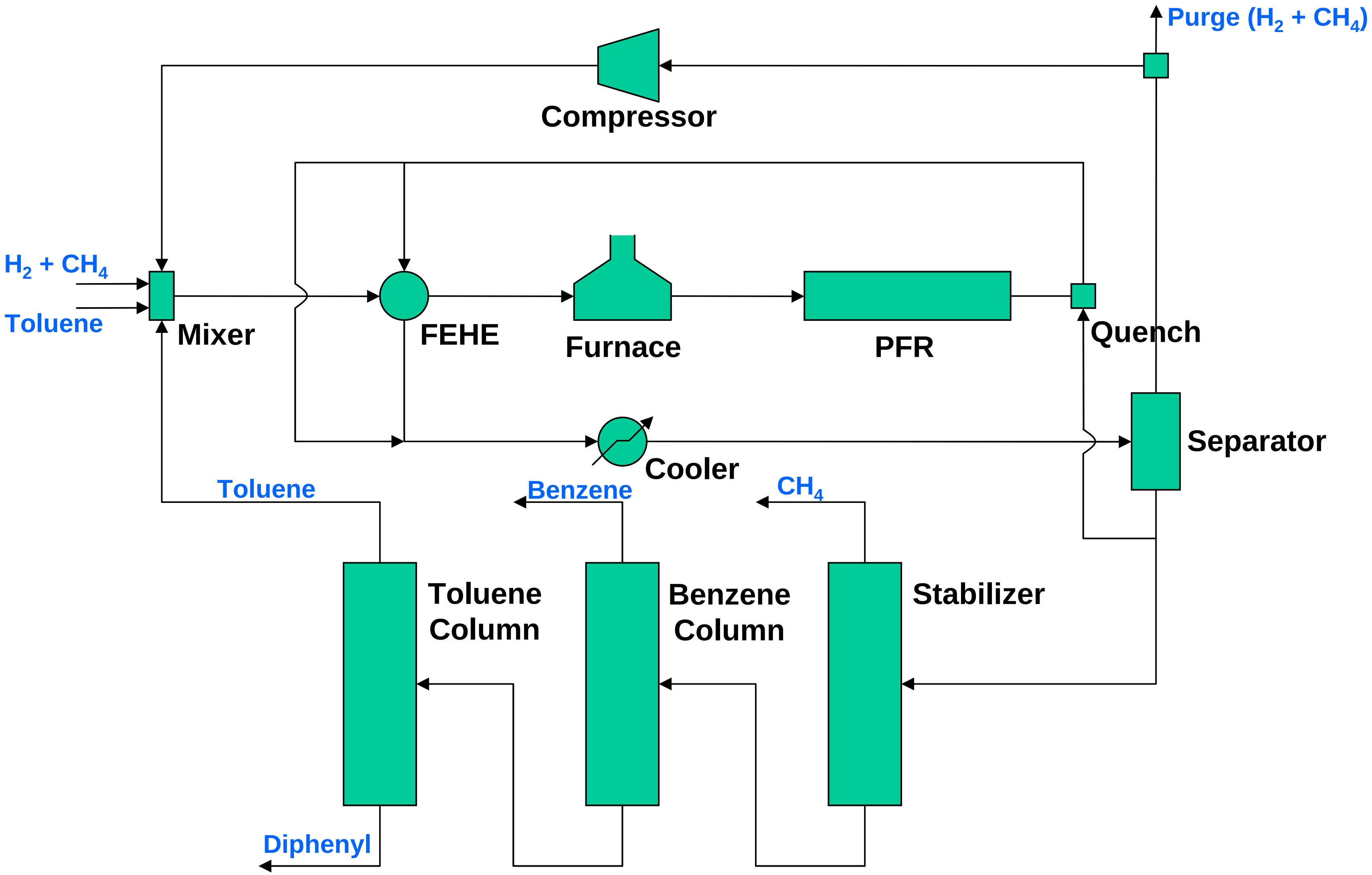
"Typical number",
N_{ss}= 0 (distillation) + 2*1 (heat exchangers) = 2

Heat-integrated distillation process



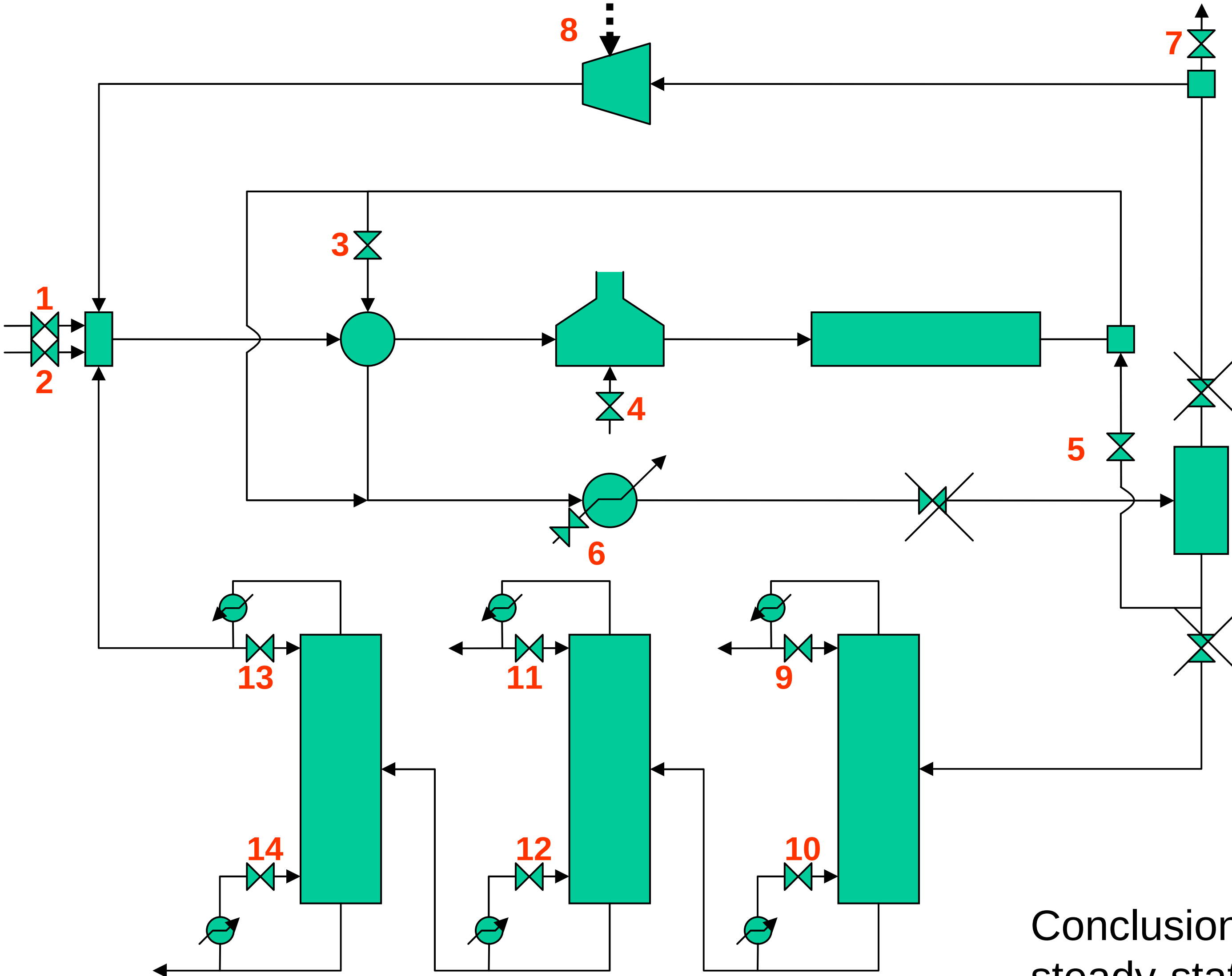
Typical number, N_{ss} = 1 (feed) + 2*0 (columns) + 2*1 (column pressures) + 1 (sidestream) + 3 (hex) = 7

HDA process



Purge (H_2 + CH_4)
Compressor
H_2 + CH_4
Toluene
Mixer
FEHE
Furnace
PFR
Quench
Separator
Cooler
Toluene
Benzene
CH_4
Toluene Column
Benzene Column
Stabilizer
Diphenyl

HDA process: steady-state degrees of freedom



feed:1.2

hex: 3, 4, 6

splitter 5, 7

compressor: 8

distillation: rest

Assume given column pressures

Conclusion: 14 steady-state DOFs

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Optimal operation (economics)

- What are we going to use our degrees of freedom for?
- Define scalar cost function $J(u_0,x,d)$
  - u_0: degrees of freedom
  - d: disturbances
  - x: states (internal variables)

  Typical cost function:

  J = cost feed + cost energy – value products

- Optimal operation for given d:

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

subject to:

Model equations:	$f(u,x,d) = 0$
Operational constraints:	$g(u,x,d) < 0$

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Optimal operation

minimize J = cost feed + cost energy – value products

Two main cases (modes) depending on marked conditions:

1. Given feed

   Amount of products is then usually indirectly given and J = cost energy.
   Optimal operation is then usually *unconstrained:*

   "maximize efficiency (energy)"

   Control: Operate at optimal trade-off (not obvious how to do and what to control)

2. Feed free

   Products usually much more valuable than feed + energy costs small.
   Optimal operation is then usually *constrained:*

   "maximize production"

   Control: Operate at bottleneck ("obvious")

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Implementation of optimal operation

- Optimal operation for given d^*:

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

subject to:

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

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

$$\rightarrow u_{opt}(d^*)$$

Problem: Usally cannot keep u_{opt} constant because disturbances d change

How should be adjust the degrees of freedom (u)?

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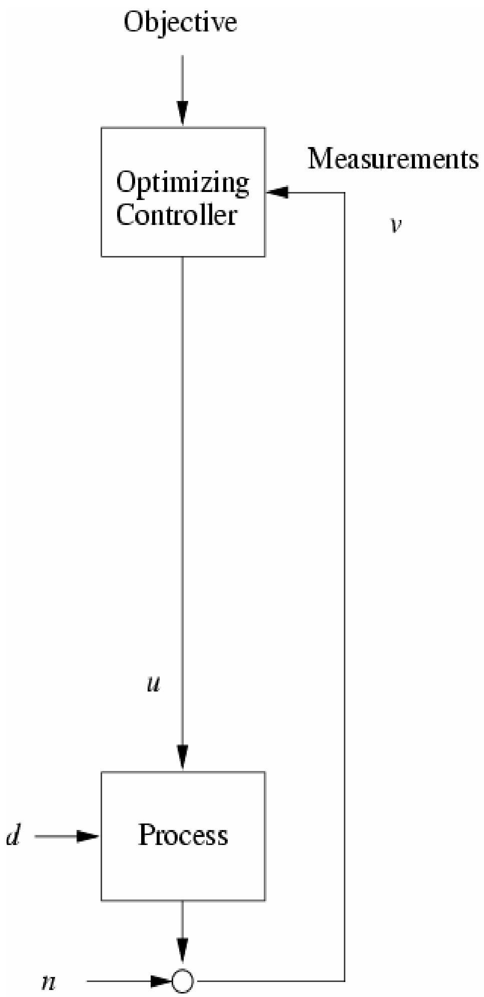
Implementation of optimal operation (Cannot keep u_{0opt} constant)

"Obvious" solution: Optimizing control

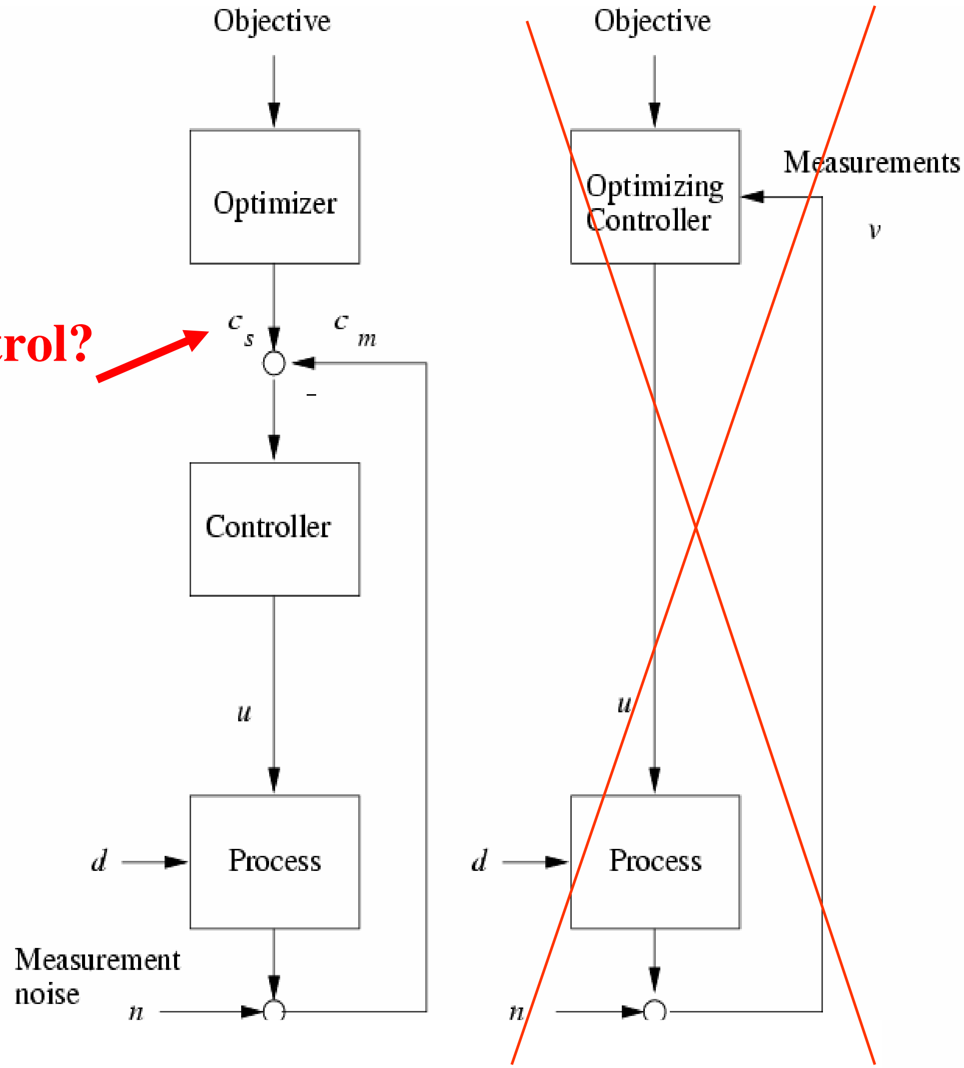
Estimate d from measurements and recompute $u_{opt}(d)$

Problem: Too complicated (requires detailed model *and* description of uncertainty)

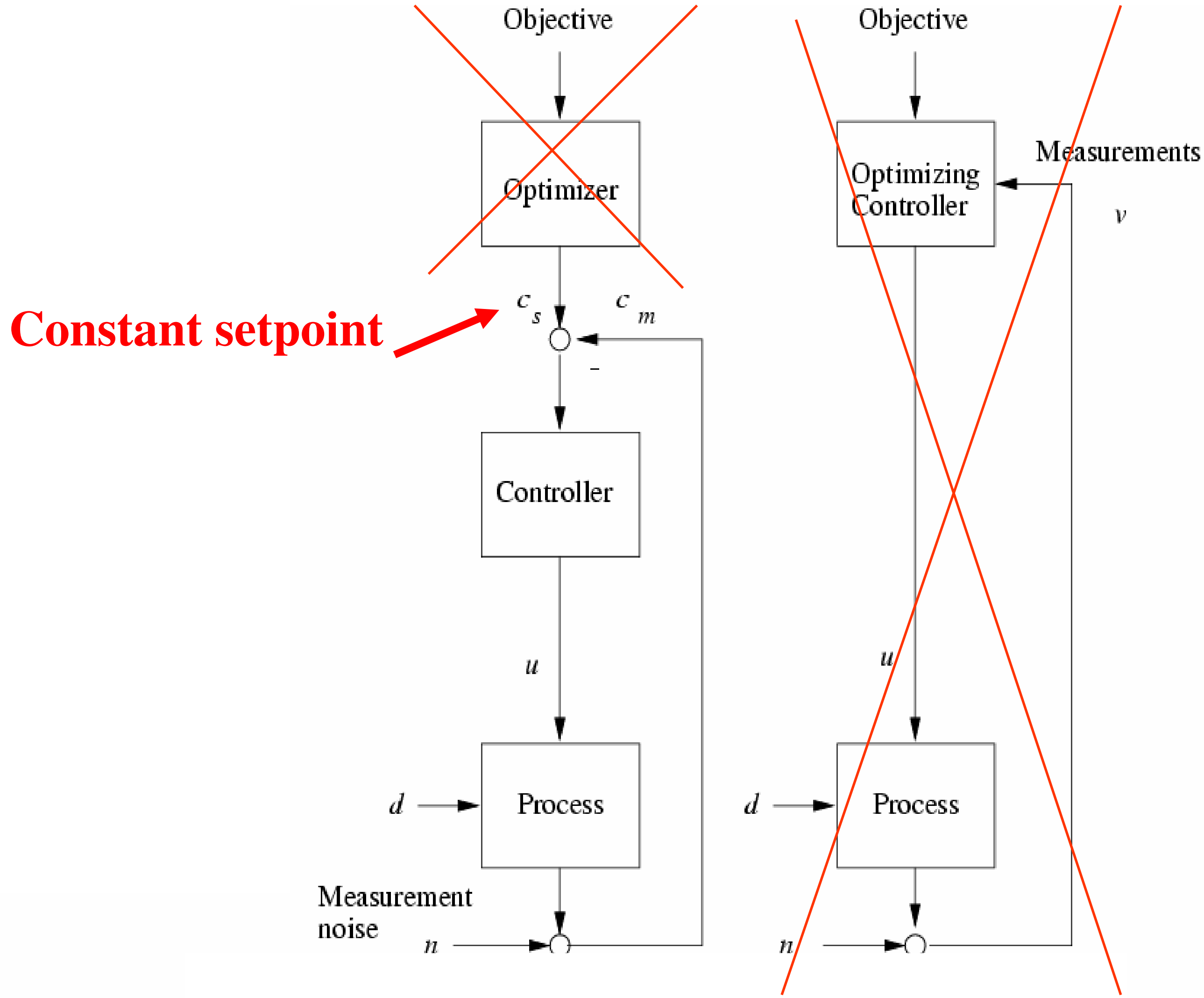


In practice: Hierarchical decomposition with separate layers

What should we control?



Self-optimizing control: When constant setpoints is OK



Unconstrained variables:

Self-optimizing control

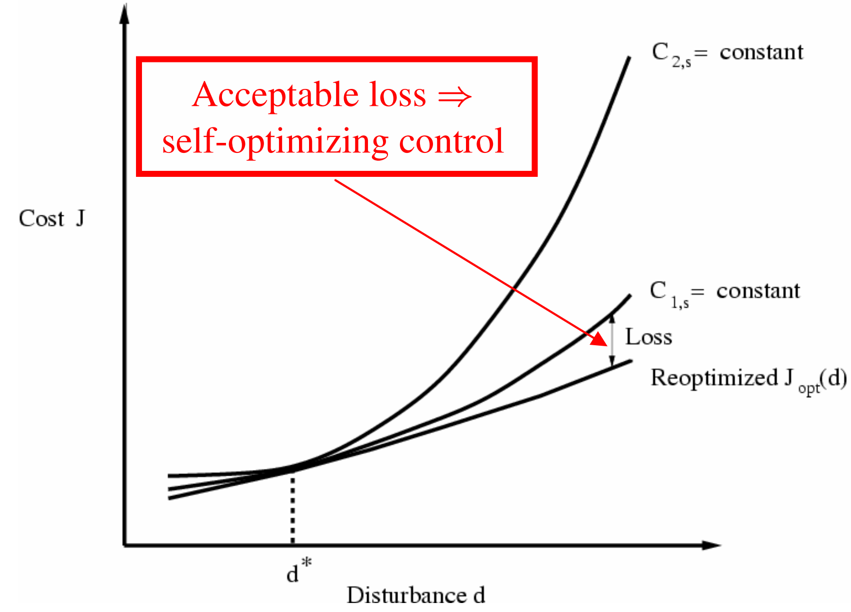
- *Self-optimizing control:*

  Constant setpoints c_s give

  ”near-optimal operation”

  (= acceptable loss L for expected disturbances d and implementation errors n)

$$L(d) = J(c_s + n, d) - J_{opt}(d)$$



What c's should we control?

- Optimal solution is usually at constraints, that is, most of the degrees of freedom are used to satisfy "active constraints", g(u,d) = 0

- CONTROL ACTIVE CONSTRAINTS!
  - c_s = value of active constraint
  - Implementation of active constraints is usually simple.

- WHAT MORE SHOULD WE CONTROL?
  - Find "self-optimizing" variables c for remaining unconstrained degrees of freedom u.

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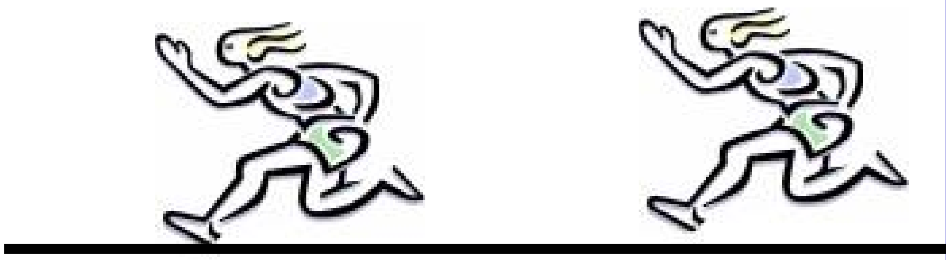
What should we control? – Sprinter

- Optimal operation of Sprinter (100 m), J=T
  - One input: ”power/speed”
  - **Active constraint control**:
    - Maximum speed (”no thinking required”)



What should we control? – Marathon

- Optimal operation of Marathon runner, J=T
  - No active constraints
  - Any self-optimizing variable c (to control at constant setpoint)?

Self-optimizing Control – Marathon

- Optimal operation of Marathon runner, J=T
  - Any self-optimizing variable c (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

Unconstrained variables: What should we control?

- *Intuition:* "Dominant variables" (Shinnar)
- Is there any systematic procedure?

Candidate controlled variables

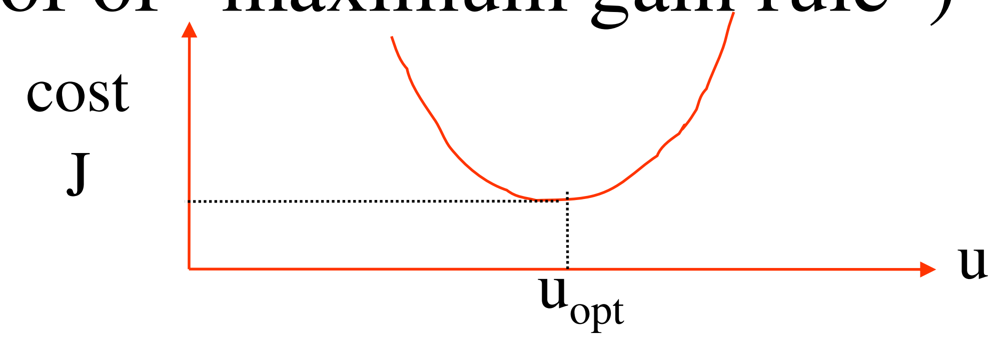
- **We are looking for some "magic" variables c to control..... What properties do they have?'**
- Intuitively 1: Should have small <u>optimal</u> range delta c_{opt}
  - since we are going to keep them constant!
- Intuitively 2: Should have small "implementation error" n

} span(c)

- Intuitively 3: Should be sensitive to inputs u (remaining unconstrained degrees of freedom), that is, the gain G_0 from u to c should be large
  - G_0: (unscaled) gain from u to c
  - large gain gives flat optimum in c
  - Charlie Moore (1980's): Maximize minimum singular value when selecting temperature locations for distillation

- Will show shortly: Can combine everything into the "maximum gain rule":
  - Maximize scaled gain $G = G_o / span(c)$

Mathematic local analysis (Proof of "maximum gain rule")



$$\text{Loss} = J(u,d) - J_{\text{opt}}(d) = \underbrace{\left(\frac{\partial J}{\partial u}\right)^T_{\text{opt}}}_{=0}(u - u_{\text{opt}}(d))$$

$$+\frac{1}{2}(u - u_{\text{opt}}(d))^T \underbrace{\left(\frac{\partial^2 J}{\partial u^2}\right)_{\text{opt}}}_{=J_{uu}} (u - u_{\text{opt(d)}}) + \cdots$$

$$\approx \frac{1}{2}\left(c - c_{\text{opt}}(d)\right)^T G_0^{-T} J_{uu} G_0^{-1} (c - c_{\text{opt}}(d))$$

$$\leq \frac{1}{2}\, G^{-T} J_{uu} G^{-1}$$

where G_0 = unscaled gain ($\Delta c = G_0 \Delta u$)

G = scaled gain $= \frac{|G_0|}{span(c)}$;

span (c) = max. expected $c - c_{\text{opt}}(d)$

Minimum singular value of scaled gain

Maximum gain rule *(Skogestad and Postlethwaite, 1996):*
Look for variables that maximize the scaled gain _(G)
(minimum singular value of the appropriately scaled
steady-state gain matrix G *from u to c)*

$$\text{Loss} \approx \frac{\bar{\sigma}(J_{uu})}{2} \cdot \frac{1}{\underline{\sigma}(G)^2}; \qquad G = \frac{G_0}{span(c)}$$

__(G) is called the Morari Resiliency index (MRI) by Luyben

Detailed proof: I.J. Halvorsen, S. Skogestad, J.C. Morud and V. Alstad,
``Optimal selection of controlled variables'', *Ind. Eng. Chem. Res.*, **42** (14), 3273-3284 (2003).

Improved minimum singular value rule for ill-conditioned plants

$$\text{Maximize } \underline{\sigma}(G J_{uu}^{-1/2})$$

G: Scaled gain matrix (as before)
J_{uu}: Hessian for effect of u's on cost

Problem: J_{uu} can be difficult to obtain

Improved rule has been used successfully for distillation

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Maximum gain rule for scalar system

$$\text{Loss} \approx \frac{|J_{uu}|}{2} \cdot \frac{1}{|G|^2}$$

Scaled steady-state gain from u to c:

$$|G| = \frac{|G_0|}{|c_{\mathrm{opt}}| + |n|} = \frac{|G_0|}{\text{optimal span } c}$$

Summary unconstrained degrees of freedom:

Looking for "magic" variables to keep at constant setpoints.

How can we find them systematically?

Candidates

$c = y_i$: Single measurements, e.g. pressure, temperature, composition

$c = \frac{y_i}{y_j}$: Combinations of measurements (e.g flow ratios)

A. Start with: Maximum gain (minimum singular value) rule:

(Scaled) gain $\underline{\sigma}(G)$ from u to y should be large

B. Then: "Brute force evaluation" of most promising alternatives.

Evaluate loss when the candidate variables c are kept constant.

In particular, may be problem with **feasibility**

C. More general candidates: Find optimal linear combination (matrix H):

$$c = h_1 y_1 + h_2 y_2 + \ldots + h_n y_n = Hy$$

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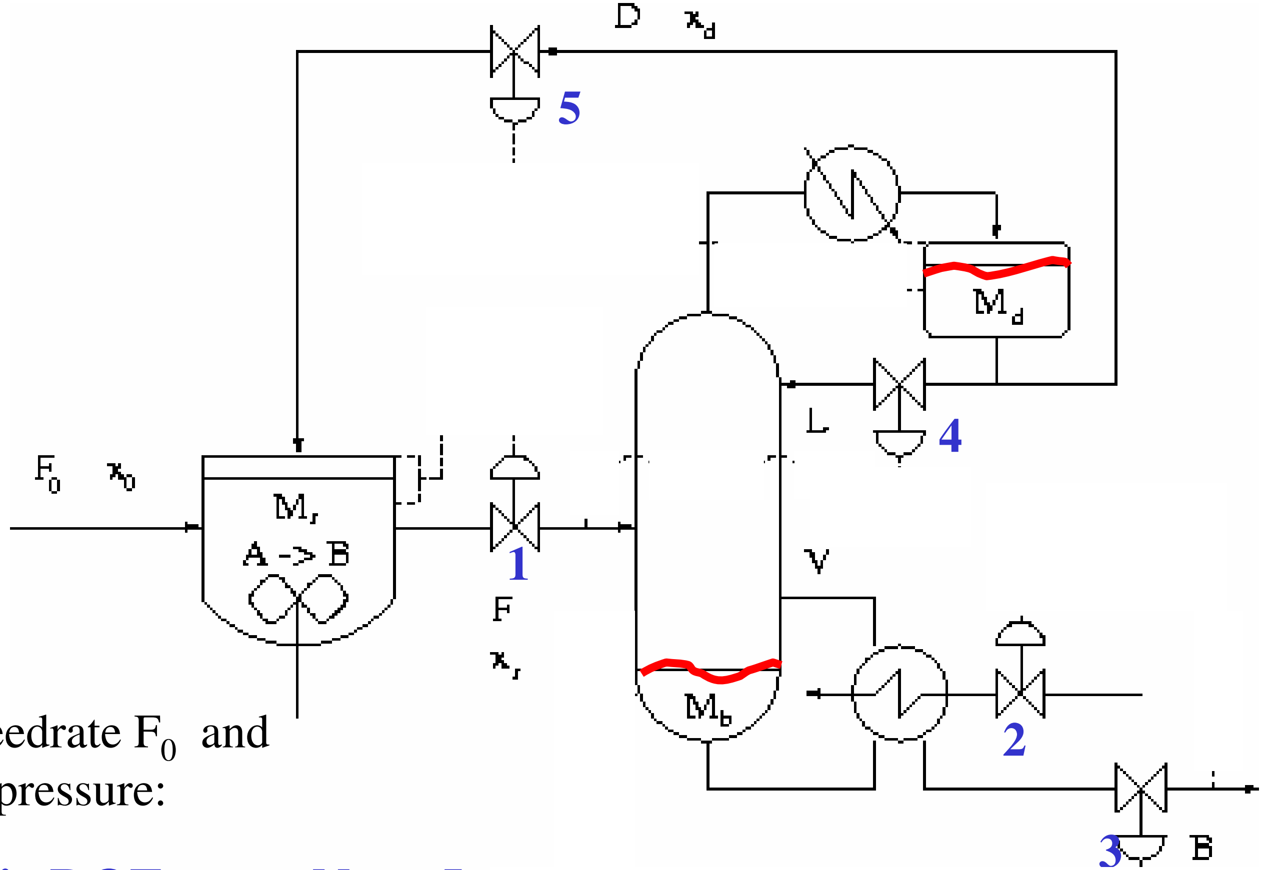
EXAMPLE: Recycle plant (Luyben, Yu, etc.)



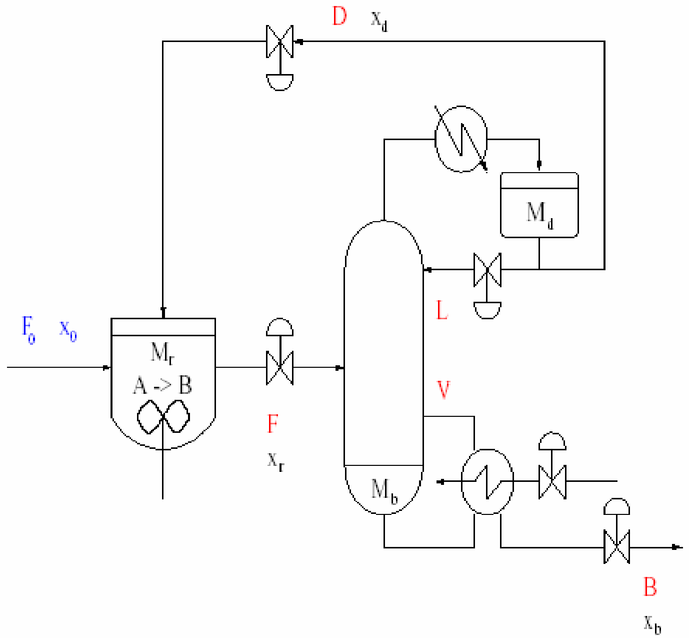
Given feedrate F_0 and column pressure:

Dynamic DOFs: $N_m = 5$

Column levels: $N_{0y} = 2$

Steady-state DOFs: $N_0 = 5 - 2 = 3$

Recycle plant: Optimal operation



Manipulated variables:

$$\mathbf{m}^T = [V\ L\ B\ D\ F]$$

Steady-state degrees of freedom: 3

Minimize costs

$$J = V$$

Constraints:

$$x_b \leq 0.015 \ \frac{\text{molA}}{\text{mol}} \quad \text{active}$$
$$M_r \leq 2800 \text{ kmol} \quad \text{active}$$
$$V \leq 5000 \text{ kmol/h}$$
$$Flows \geq 0 \text{ kmol/h}$$

Disturbances:

$$d^T = [F_0\ x_0] = [460 \pm 92 \ \frac{\text{kmol}}{\text{h}} \ 0.90 \pm 0.05$$

1 remaining unconstrained degree of freedom

A. Maximum gain rule: Steady-state gain

Rank	c	$\|G(0)\| \cdot 10^3$
1	x_D	13.1
2	L/F	8.9
3	D/L	7.7
4	D/V	5.8
5	V/L	4.5
6	B/L	4.1
7	V/F	4.0
8	B/D	3.3
9	L	3.0
10	B/F	2.6
11	D	2.6
12	F/F_0	2.5
13	D/F	2.5
14	F	1.9
15	B/V	0
15	V	0
15	x_r	0
15	B	0

Conventional:
Looks good

Luyben rule:
Not promising
economically

Gain from $u = L$ to c with active constraints (M_r and x_B) constant.
Scaling: $\Delta c_i = \max_d |c_{opt,i}(d) - c_{opt,i}(d^*)|) +$ implementation error.
$d = F_0, z_F$.

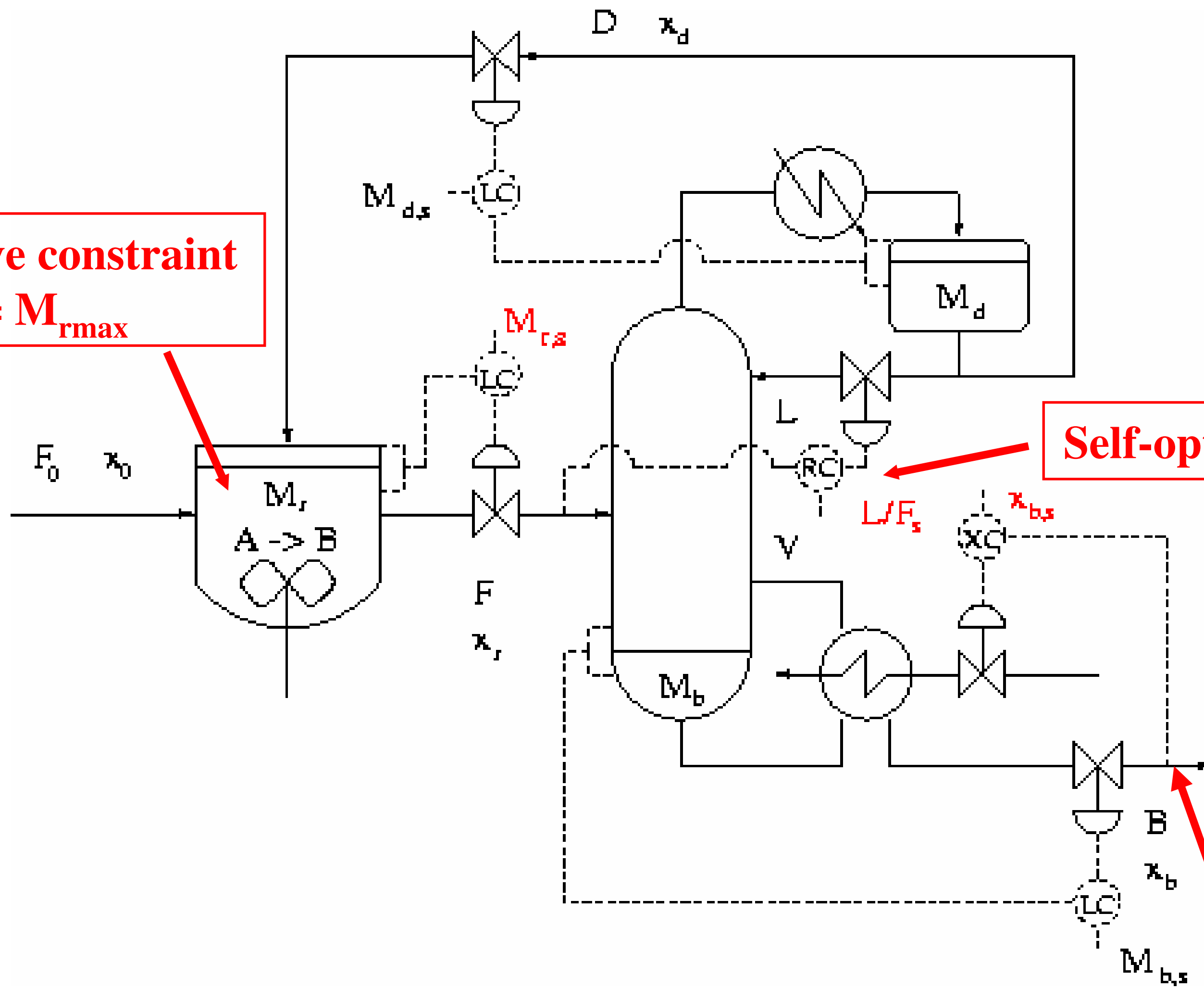
How did we find the gains in the Table?

1. Find nominal optimum
2. Find (unscaled) gain G_0 from input to candidate outputs: $c = G_0 \quad u$.
   - In this case only a single unconstrained input (DOF). Choose at u=L
   - Obtain gain G_0 numerically by making a small perturbation in u=L while adjusting the other inputs such that the active constraints are constant (bottom composition fixed in this case)

IMPORTANT!

3. Find the span for each candidate variable
   - For each disturbance d_i make a typical change and reoptimize to obtain the optimal ranges $c_{opt}(d_i)$
   - For each candidate output obtain (estimate) the control error (noise) n
   - $\text{span}(c) = \; _i \, | \; c_{opt}(d_i)| \; + n$
4. Obtain the scaled gain, $G = G_0 / \text{span}(c)$

Conclusion: Control of recycle plant



Active constraint
$M_r = M_{rmax}$

Self-optimizing

Active constraint
$x_B = x_{Bmin}$

L/F constant: Easier than “two-point” control

Assumption: Minimize energy (V)

Summary:
Procedure selection controlled variables

1. Define economics and operational constraints
2. Identify degrees of freedom and important disturbances
3. Optimize for various disturbances
4. Identify (and control) active constraints (off-line calculations)
   - May vary depending on operating region. For each operating region do step 5:
5. Identify “self-optimizing” controlled variables for remaining degrees of freedom
   1. (A) Identify promising (single) measurements from “maximize gain rule” (gain = minimum singular value)
      - (C) Possibly consider measurement combinations if no promising
   2. (B) “Brute force” evaluation of loss for promising alternatives
      - Necessary because “maximum gain rule” is local.
      - In particular: Look out for feasibility problems.
   3. Controllability evaluation for promising alternatives

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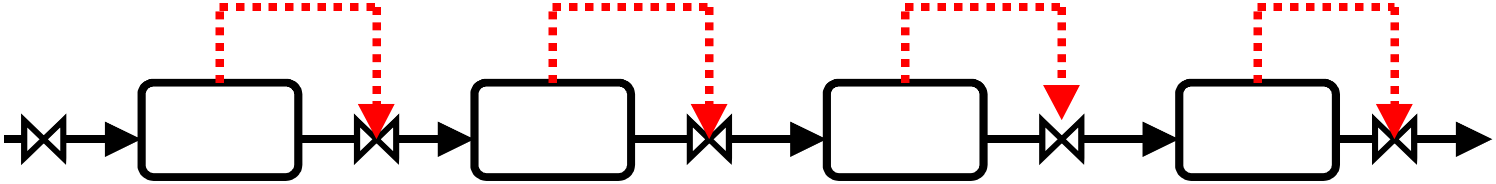
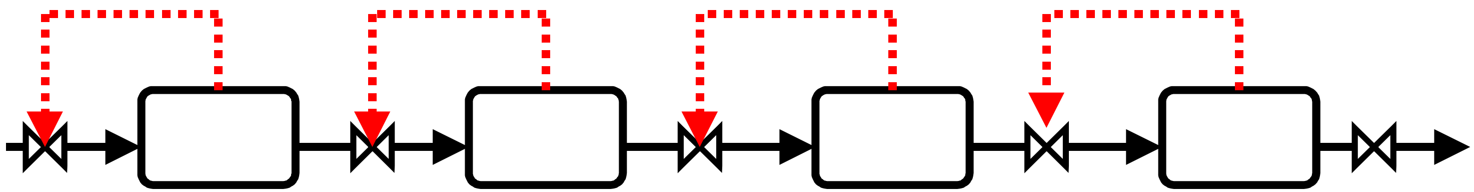
Step 4. Where set production rate?

- Very important!
- Determines structure of remaining inventory (level) control system
- Set production rate at (dynamic) bottleneck
- Link between **Top-down** and **Bottom-up** parts

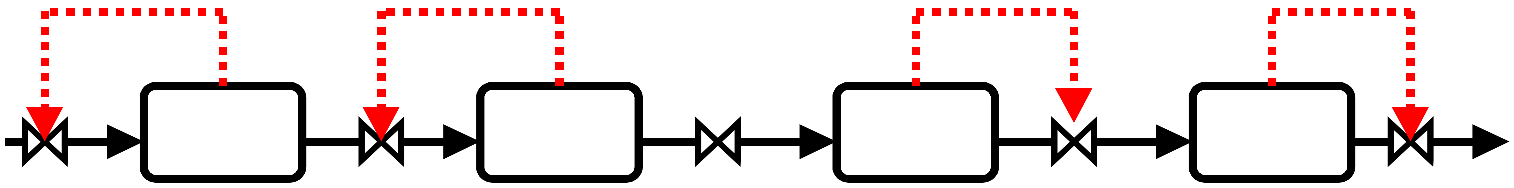
Production rate set at inlet :
Inventory control in direction of flow



Production rate set at outlet: Inventory control opposite flow



Production rate set inside process



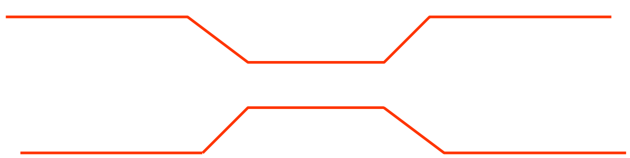
Where set the production rate?

- Very important decision that determines the structure of the rest of the control system!
- May also have important economic implications

Often optimal: Set production rate at bottleneck!

- "A bottleneck is an extensive variable that prevents an increase in the overall feed rate to the plant"
- If feed is cheap and available: Optimal to set production rate at bottleneck

- If the flow for some time is not at its maximum through the bottleneck, then this loss can never be recovered.



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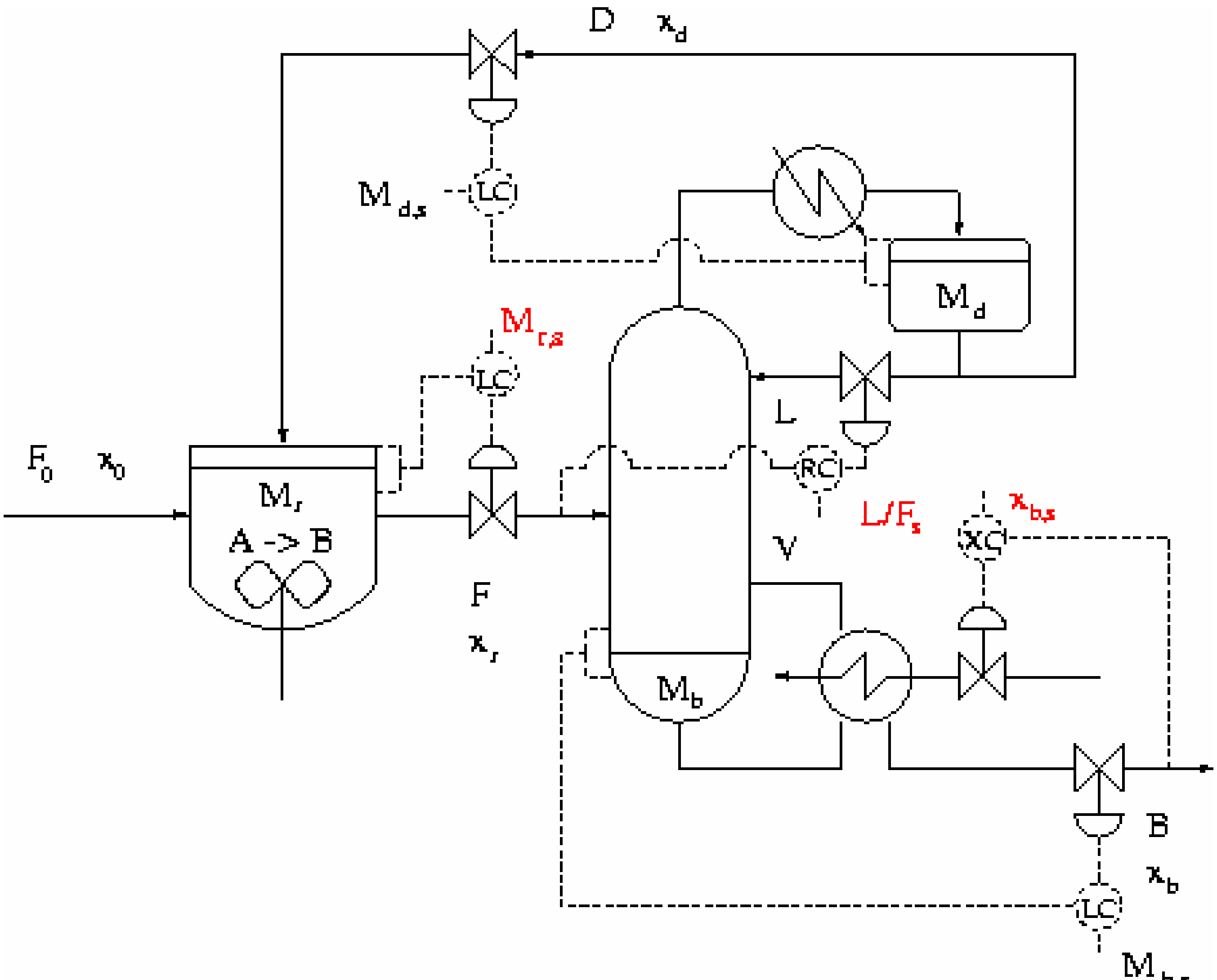
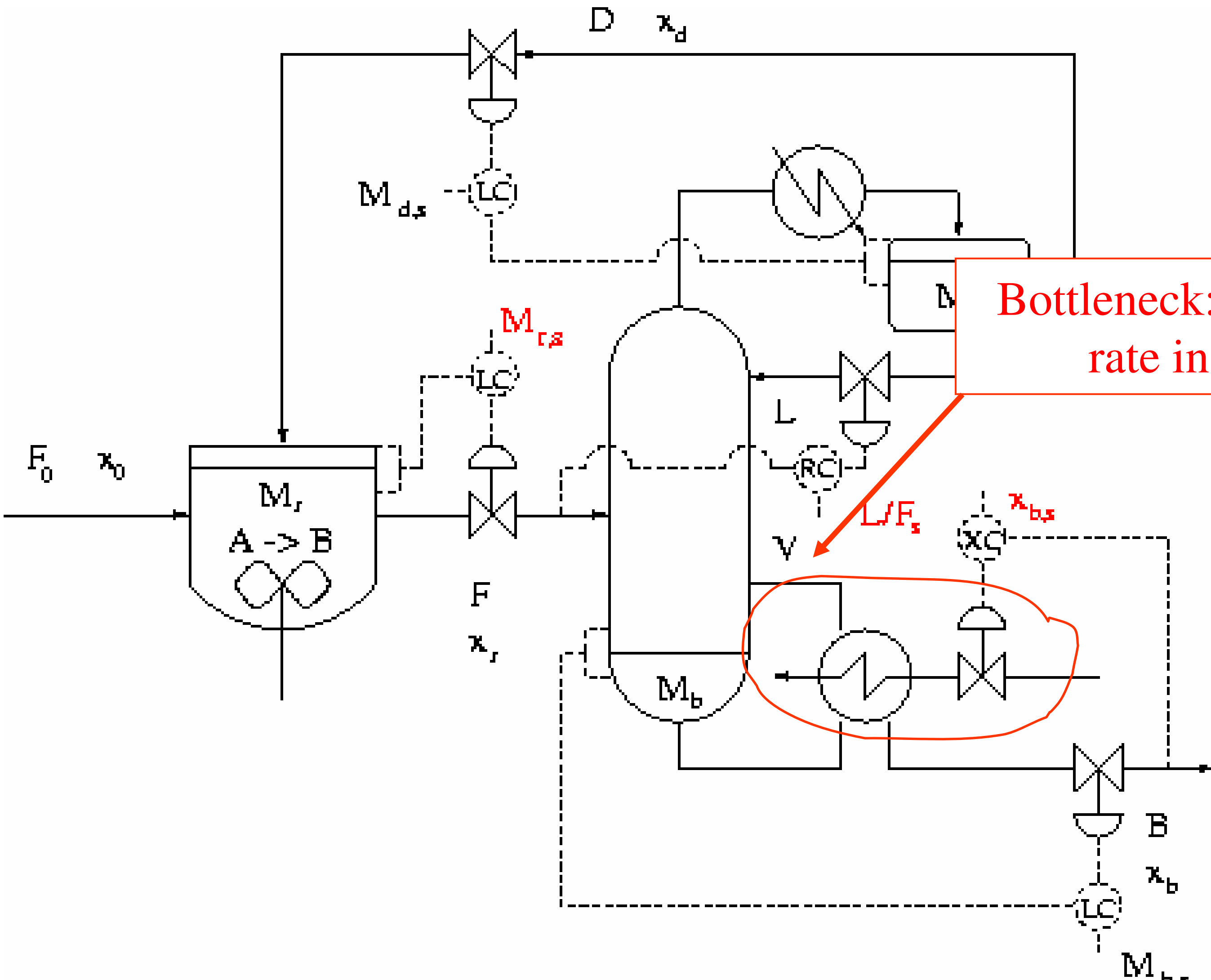
Reactor-recycle process:
Given feedrate with production rate set at inlet



Reactor-recycle process:
Want to maximize feedrate: reach bottleneck in column

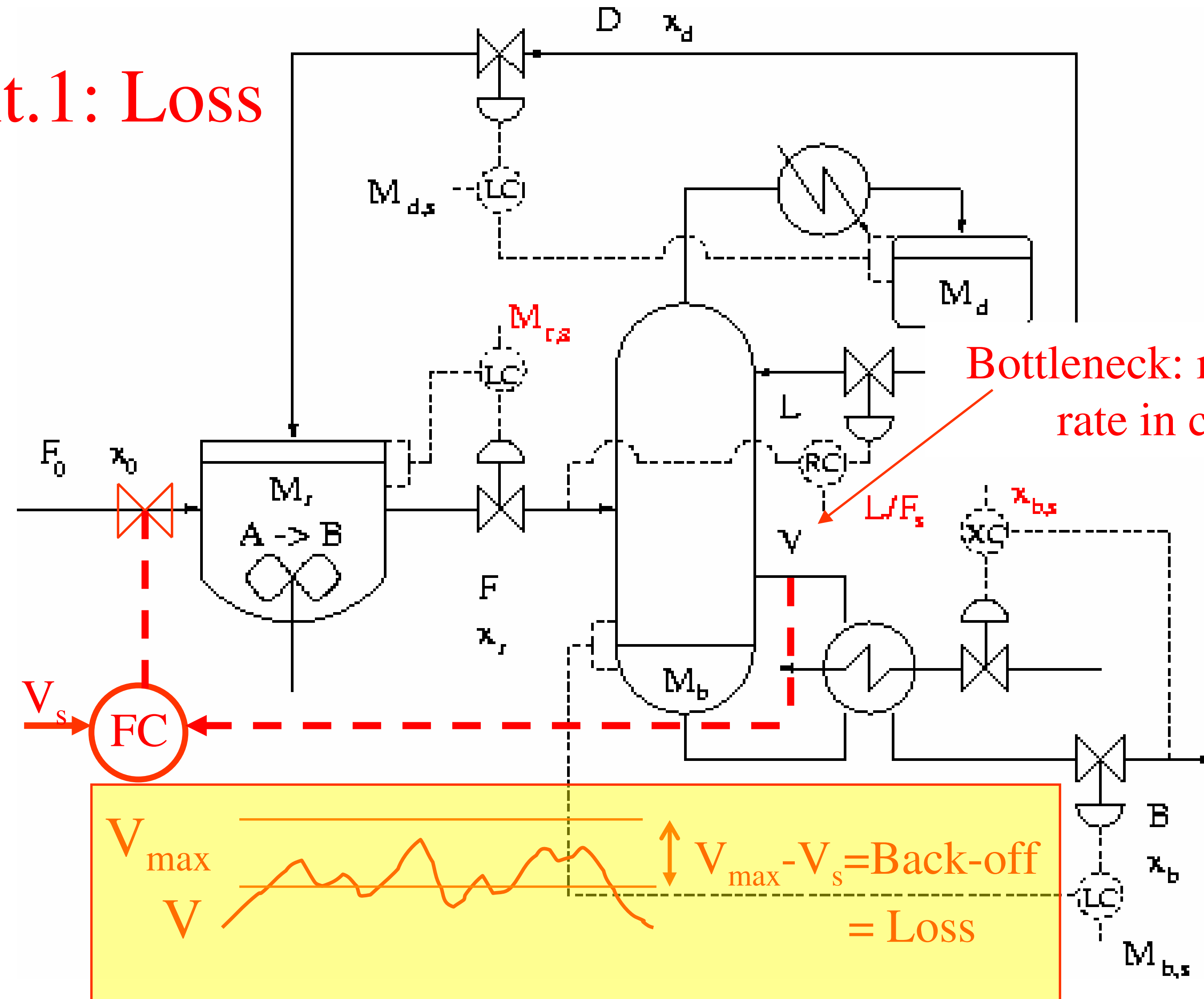


Bottleneck: max. vapor rate in column

Reactor-recycle process with production rate set at inlet

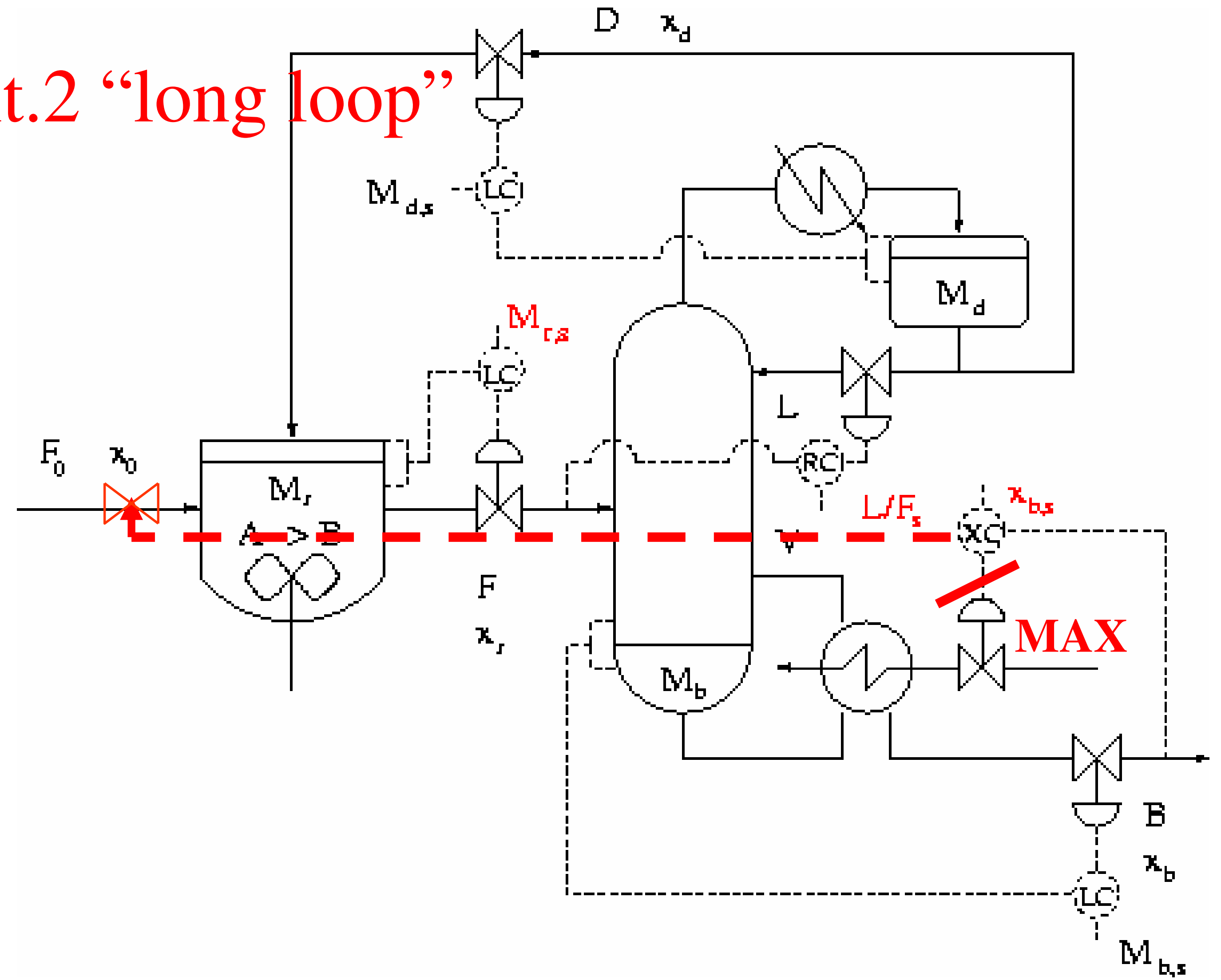
Want to maximize feedrate: reach bottleneck in column

Alt.1: Loss



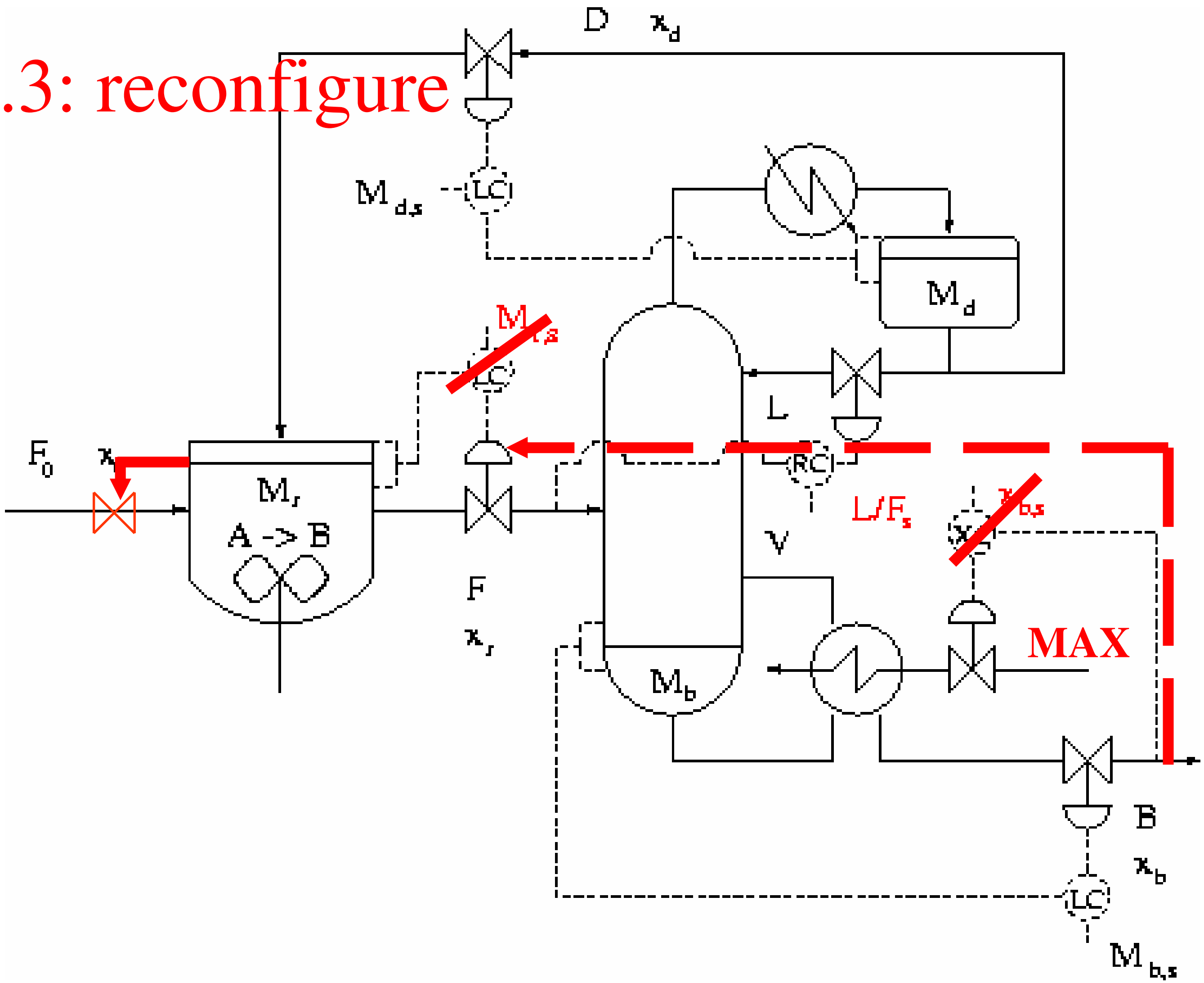
Reactor-recycle process with increased feedrate:
Optimal: Set production rate at bottleneck

Alt.2 "long loop"



Reactor-recycle process with increased feedrate: Optimal: Set production rate at bottleneck

Alt.3: reconfigure



Alt.4: Multivariable control (MPC)

- Can reduce loss
- BUT: Is generally placed on top of the regulatory control system (including level loops), so it still important where the production rate is set!

Conclusion production rate manipulator

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

BREAK

Outline

- Control structure design (plantwide control)
- A procedure for control structure design

  I Top Down

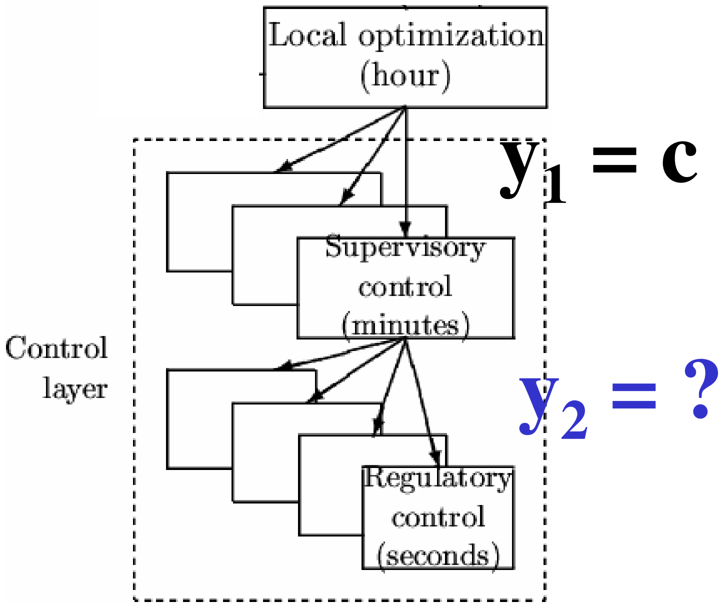
  - Step 1: Degrees of freedom
  - Step 2: Operational objectives (optimal operation)
  - Step 3: What to control ? (self-optimizing control)
  - Step 4: Where set production rate?

  II Bottom Up

  - Step 5: Regulatory control: What more to control ?
  - Step 6: Supervisory control
  - Step 7: Real-time optimization
- Case studies

Step 5. Regulatory control layer

- *Purpose*: “Stabilize” the plant using local SISO PID controllers
- Enable manual operation (by operators)
- Main structural issues:
  - What more should we control? (secondary cv’s, y_2)
  - Pairing with manipulated variables (mv’s u_2)



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

Implications for selection of y_2:

1. Control of y_2 "stabilizes the plant"
2. y_2 is easy to control (favorable dynamics)

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Rules for selecting y_2 (and u_2 to be paired with y_2)

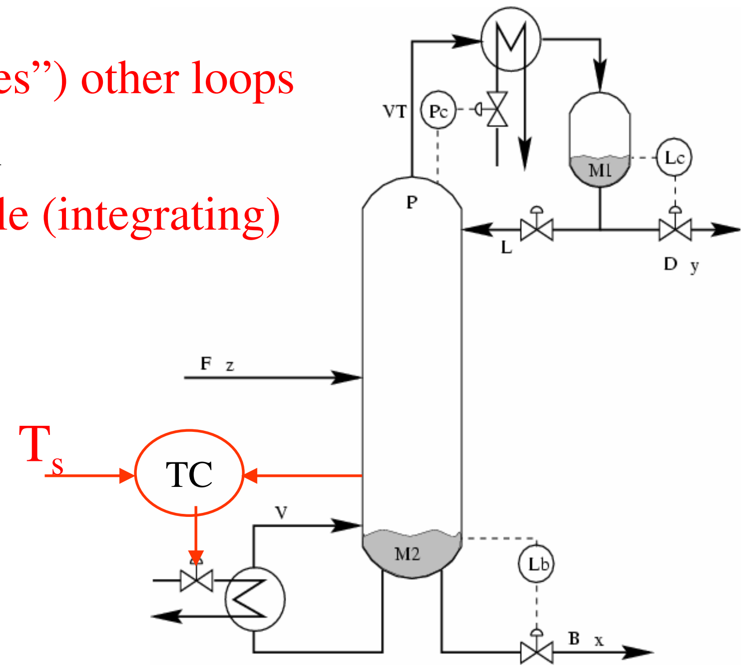
1. y_2 should be easy to measure
2. Control of y_2 stabilizes the plant
3. y_2 should have good controllability, that is, favorable dynamics for control
4. y_2 should be located "close" to a manipulated input (u_2) (follows from rule 3)
5. The (scaled) gain from u_2 to y_2 should be large
6. The effective delay from u_2 to y_2 should be small
7. Avoid using inputs u_2 that may saturate (should generally avoid saturation in lower layers)

Example: Distillation

- Primary controlled variable: $y_1 = c = x_D, x_B$ (compositions top, bottom)
- BUT: Delay in measurement of x + unreliable
- Regulatory control: For "stabilization" need control of (y_2):
  - Liquid level condenser (M_D)
  - Liquid level reboiler (M_B)

  } Unstable (Integrating) + No steady-state effect

  - Pressure (p) — Disturbs ("destabilizes") other loops
  - Holdup of light component in column (temperature profile) — Almost unstable (integrating)



T-loop in bottom

Outline

- Control structure design (plantwide control)
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  I Top Down
  - Step 1: Degrees of freedom
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  - Step 3: What to control ? (primary CV's) (self-optimizing control)
  - Step 4: Where set production rate?

  II Bottom Up
  - Step 5: Regulatory control: What more to control (secondary CV's) ?
  - Step 6: Supervisory control
  - Step 7: Real-time optimization
- Case studies

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Step 6. Supervisory control layer

- *Purpose*: Keep primary controlled outputs $c=y_1$ at optimal setpoints c_s
- Degrees of freedom: Setpoints y_{2s} in reg.control layer
- *Main structural issue:* Decentralized or multivariable?

Decentralized control (single-loop controllers)

Use for: Noninteracting process and no change in active constraints

+ Tuning may be done on-line
+ No or minimal model requirements
+ Easy to fix and change
- Need to determine pairing
- Performance loss compared to multivariable control
- Complicated logic required for reconfiguration when active constraints move

Multivariable control
(with explicit constraint handling = MPC)

Use for: Interacting process and changes in active constraints

\+ Easy handling of feedforward control

\+ Easy handling of changing constraints

- no need for logic
- smooth transition

\- Requires multivariable dynamic model

\- Tuning may be difficult

\- Less transparent

\- "Everything goes down at the same time"

Outline

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Step 7. Optimization layer (RTO)

- *Purpose:* Identify active constraints and compute optimal setpoints (to be implemented by supervisory control layer)
- *Main structural issue:* Do we need RTO? (or is process self-optimizing)

- RTO not needed when
  - Can "easily" identify change in active constraints (operating region)
  - For each operating region there exists self-optimizing var

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  - Step 5: Regulatory control: What more to control ?
  - Step 6: Supervisory control
  - Step 7: Real-time optimization
- Conclusion / References

Conclusion

Procedure plantwide control:

I. Top-down analysis to identify degrees of freedom and primary controlled variables (look for self-optimizing variables)

II. Bottom-up analysis to determine secondary controlled variables and structure of control system (pairing).

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More examples and case studies

- HDA process
- Cooling cycle
- Distillation (C3-splitter)
- Blending

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- … + more…..

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