

INTERACTIONS BETWEEN PROCESS DESIGN AND CONTROL

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(1)

USE OF CONTROLLABILITY ANALYSIS

1. SELECT BETWEEN DESIGN ALTERNATIVES
2. GIVE IDEAS FOR DESIGN CHANGES
3. SELECT CONTROL STRUCTURE:
A CONTROLLABLE STRUCTURE HAS GOOD
"SELF-REGULATION"

(2)

OUTLINE

1. CONTROL-
2. CONTROLLABILITY
3. SCALAR CASE (SISO)
4. PH-TANK EXAMPLE
5. MULTIVARIABLE CASE, RCA
6. DISTILLATION EXAMPLE
7. OTHER EXAMPLES: FCC, HEN
8. CONCLUSIONS

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CONTROL HIEARCHY

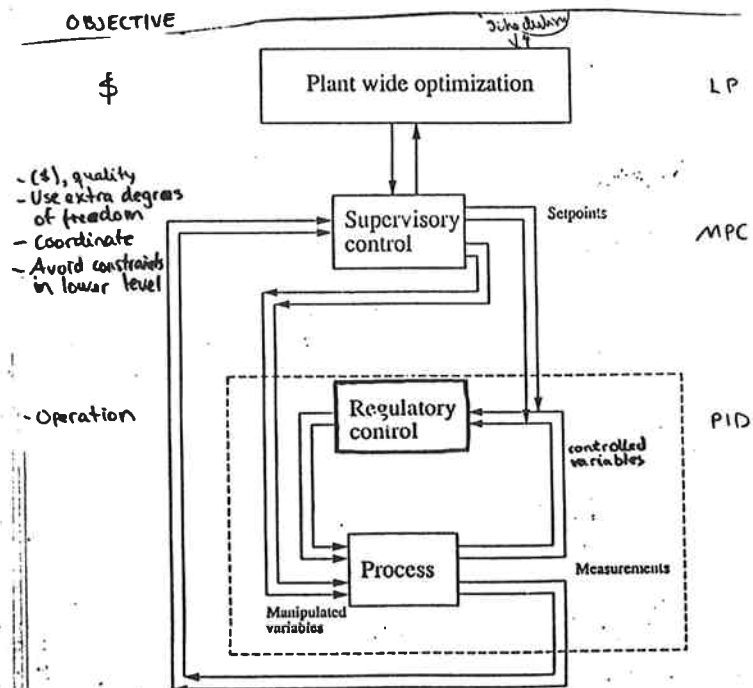


Figure 2: Schematic representation of a hierarchical control system.

- "control structure", "control configuration"
- time scale separation

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Regulatory control

Objectives:

- Control where fast control is needed.
- Make the control problem seem simple from the levels above.
- Provide access for higher levels through cascades.

GIVEN A PLANT:

- NEED TO UNDERSTAND THE PLANT BEFORE YOU START DOING CONTROL.
- HOW WELL CAN IT BE CONTROLLED?

CONTROLLABILITY:

- INHERENT CONTROL CHARACTERISTICS OF THE PLANT
- INDEPENDENT OF THE CONTROLLER
- ZIEGLER & NICHOLS, 1943:
"The ability of the process to achieve and maintain the desired equilibrium value"

CONTROLLABILITY:

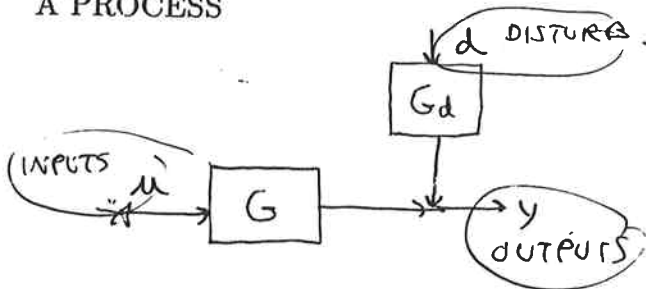
CAN ONLY BE AFFECTED BY "DESIGN" CHANGES:

- NEW EQUIPMENT
- NEW MEASUREMENTS
- NEW ACTUATORS
- { • NEW CONTROL OBJECTIVES
- { • NEW CONTROL STRUCTURE

Control structure selection decisions:

1. Selection of controlled variables. & measurements (y)
2. Selection of manipulated variables. ~~u~~ (u)
3. Pairing of controlled and manipulated variables. (y₁ ← u₁, y₂ ← u₂)

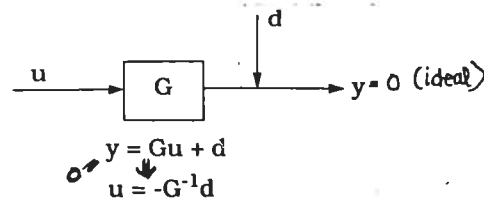
CONTROLLABILITY ANALYSIS OF A PROCESS



1. Obtain model(G, G_d) (Linearize in various operating points)
2. Scale variables(± 1)
3. Compute various controllability measures
4. Analyze, Compare
5. If not OK propose design changes

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WHAT LIMITS CONTROLLABILITY?



- "PERFECT CONTROL": CONTROLLER INVERTS PROCESS
- CONTROLLABILITY IS LIMITED WHEN THIS IS NOT POSSIBLE:
 - TIME DELAYS
 - INVERSE RESPONSES
 - CONSTRAINTS IN VALVES & EQUIPMENT
 - MODEL UNCERTAINTY & CHANGES IN OPER. POINT
 - DISTURBANCES
 - INSTABILITY
 - INTERACTIONS

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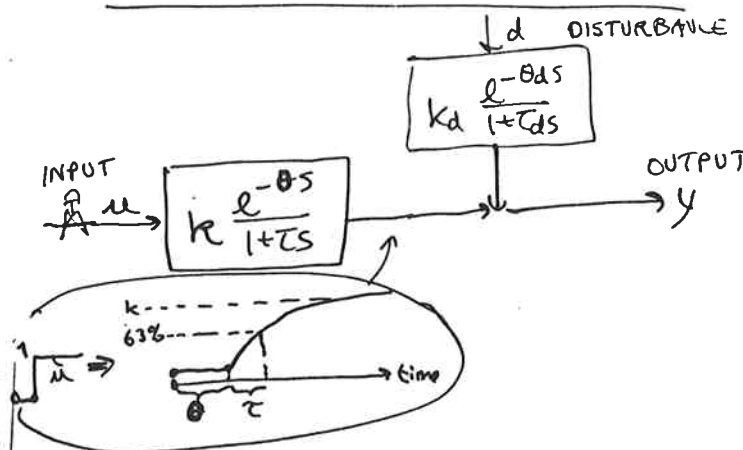
TOOLS FOR CONTROLLABILITY ANALYSIS

Mostly linear tools, frequency domain :

1. State controllability and observability (Kalman)
2. Functional controllability (Rosenbrock)
3. Dead time and inverse response (RHP-zeros)
4. Instability
5. Multivariable couplings and interactions, condition no., SVD, RGA
6. Sensitivity to disturbances, G_d
7. Input constrains, $G^{-1}G_d$
8. Sensitivity to model error, RGA
9. Specific tools for decentralized control: RGA, CLDG

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SISO CONTROLLABILITY ANALYSIS

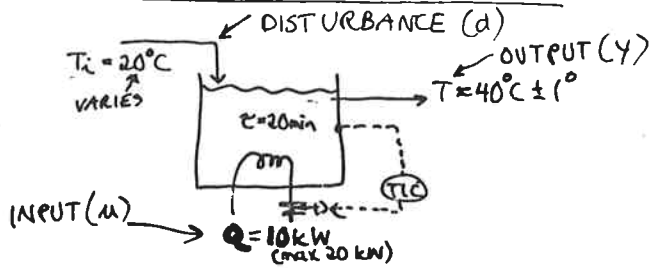


SCALED VARIABLES : $u = \pm 1$
 $d = \pm 1$
 $y = \pm 1$

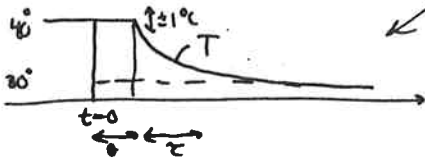
THEN: ACCEPTABLE CONTROLLABILITY :

1. SPEED OF RESPONSE : $\theta < \tau_d / k_d$
2. CONSTRAINTS : $k > k_d$ & $k/\tau > k_d/\tau_d$

Example. Heated tank.

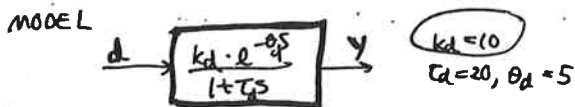


SUDDENLY T_i DROPS TO 10°C
NO CONTROL ($Q = 10\text{ kW}$):



MATHEMATICALLY:
SCALED VARIABLES

$d = 1$: EXPECTED DISTURBANCE $d = \Delta T_i / 10^\circ$
 $y = 1$: ALLOWED DEVIATION $y = \Delta T / 1^\circ$



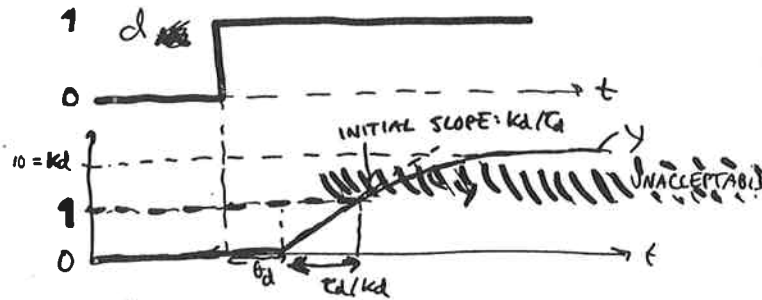
CONCLUSION: NEED CONTROL SINCE $k_d \geq 1$

$$\theta < \frac{\tau_d}{k_d}$$

- AVOID INPUTS WITH LARGE DELAY (avoid θ large)

- AVOID DISTURBANCES WITH A LARGE ^{GAIN} (k_d large) AND FAST EFFECT (τ_d small)

GENERALIZATION TO DYNAMICS (STEP RESPONSE)

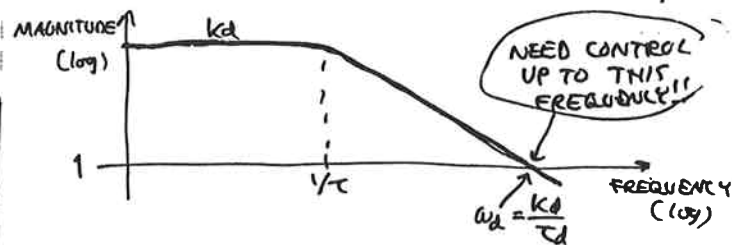


TO AVOID UNACCEPTABLE OUTPUT ($y > 1$):
MUST REACT WITHIN τ_d/k_d TIME UNITS

ex: $\tau_d = 20, k_d = 10 \Rightarrow \tau_d/k_d = 2 \text{ min}$

ACCEPTABLE CONTROLLABILITY:
THE DELAY θ IN THE INPUT CHANNEL (heater) MUST BE LESS: $\theta < \tau_d/k_d$

GENERALIZATION TO FREQUENCY DOMAIN



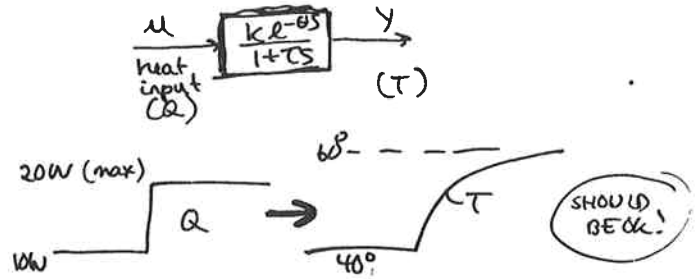
$\frac{k_d}{\tau_d}$ = SLOPE OF INITIAL RESPONSE
- IMPORTANT CONTROLLABILITY PARAMETER (WANT SMALL)

NEED: $\theta < \frac{1}{\omega_d}$

$$|g| > |g_d|$$

- PREFER INPUTS WITH LARGE (k large) AND FAST EFFECT (τ small)

INPUT CONSTRAINTS:



SCALED VARIABLES

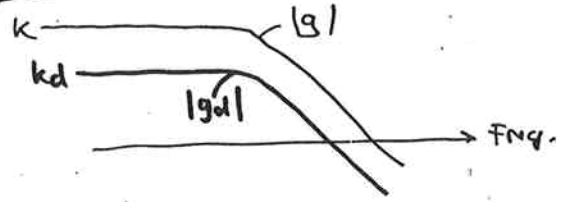
$$\left. \begin{aligned} u &= \Delta Q / 10 \quad \leftarrow \text{max change} \\ y &= \Delta T / 1 \end{aligned} \right\} \Rightarrow k = \frac{y}{u} = 20$$

- steady-state

TO AVOID INPUT CONSTRAINTS:

NEED $k > k_d$ ($20 > 10$ OK!)

GENERALIZATION



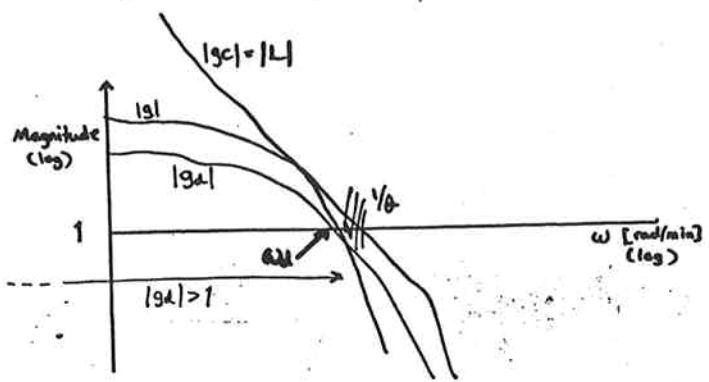
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SUMMARY SISO Plants

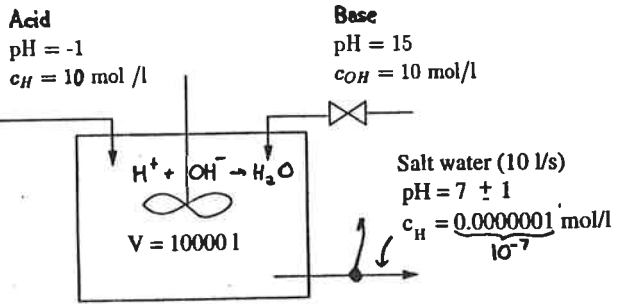
Consider frequencies where $|g_d| > 1$ (i.e. control is needed for acceptable performance).

Need:

1. $|g| > |g_d|$: Avoid input constraints ($|u| \leq 1$)
 \Downarrow
 $(|g^{-1} g_d| < 1)$
2. $|g_c| \approx |g_d|$: Acceptable performance ($|y| \leq 1$).



EXAMPLE. Neutralization process.



Let $y = c_H - 10^{-7}$ (difference from neutrality)
 $u = \text{Flow}_{\text{base}}$
 $d = \text{Flow}_{\text{acid}}$

Model with appropriate scalings $g = -g_d$
 $y = \frac{k_d}{1 + \tau s} u + d$

$k_d = 0.25 \cdot 10^7$ $\tau = V/q = 1000s$

EXTREMELY SENSITIVE TO DISTURBANCES.
 Frequency up to which feedback is needed

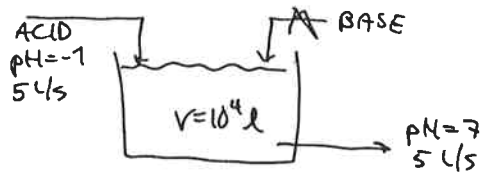
$\omega_B > \omega_d = \frac{0.25 \cdot 10^7}{\tau} = 2500 \text{ rad/s}$
 But delay is $\theta = 10s$ so bandwidth must be less than $\omega_B < 1/\theta = 0.1 \text{ rad/s}$

} INCOMPATIBLE
 => POOR CONTROLLABILITY

Conclusion: Process is impossible to control irrespective of controller design.

$\tau = 10s = 0.4ms$

PHYSICAL EXPLANATION:



- HOW MUCH ACID (d) DECREASES PH FROM 7 TO 6 ?

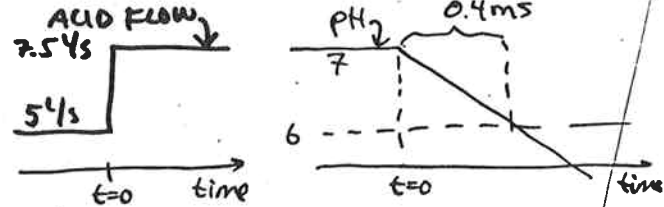
ANSWER:

1. MUST ADD H⁺ IONS: $\Delta C \cdot V = 10^{-2} \text{ mol H}^+$
 $10^{-6} \frac{\text{mol}}{\text{l}} \cdot 10^4 \text{ l}$

2. THIS CORRESPONDS TO: $10^{-2} \text{ mol} / 10^1 \frac{\text{mol}}{\text{l}} = 10^{-3} \text{ L ACID}$
 (pH=1)

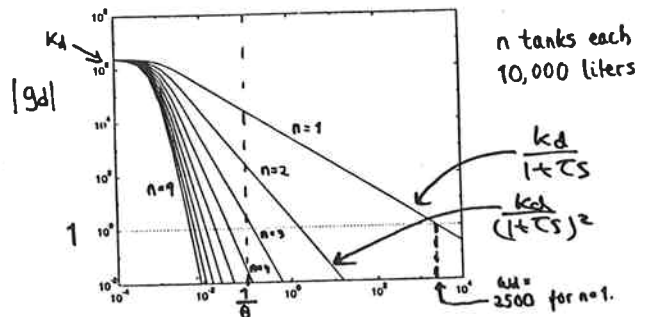
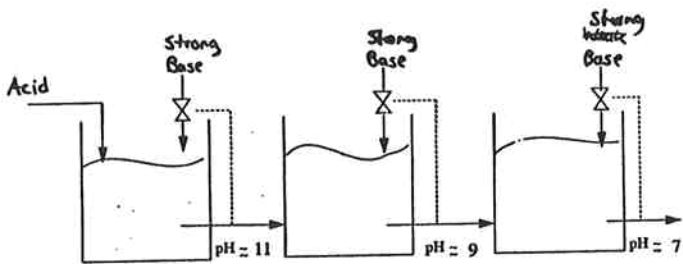
3. IF THE FEED INCREASES FROM 5 l/s TO 7.5 l/s THIS TAKES: $10^{-3} \text{ l} / 2.5 \text{ l/s} = 0.4 \cdot 10^{-3} \text{ s}$

4. GET:



IMPROVE CONTROLLABILITY BY REDESIGN OF PROCESS

- Use several similar tanks in series with gradual adjustment
- Similar to golf



With n tanks:

$$G_d(s) = \frac{k_d}{(1 + \tau s)^n}$$

$$\omega_d = k_d^{1/n} / \tau$$

Assume delay for control is about 10 s in each tank.

Get same controllability with:

- 3 tanks of about 13500 l each - 40.5 m³ total volume
- 4 tanks of about 4000 l each - 16.0 m³ total
- 5 tanks of about 1900 l each 9.5 m³ total
- 6 tanks of about 1160 l each - 7.0 m³ total
- 7 tanks of about 820 l each - 5.7 m³ total
- 8 tanks of about 630 l each - 5.0 m³ total

Minimum total volume:

16 tanks of about 251 l each - 4.02 m³ total

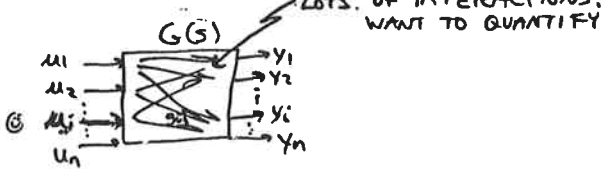
CONCLUSION: USE ~ 5 tanks.

$$k_d = \frac{C_d}{C_{y, \max}} \cdot \left(\frac{\Delta q_d}{q^*} \right)$$

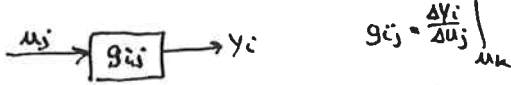
Annotations: pH feed, normalized flow change, 5 ± 2.5 l/s, pH = 7 ± 1 IN OUTLET, 10 l/s, 10⁻⁶, 10⁻⁷, PH=1 FOR ACID

RELATIVE GAIN ARRAY

MULTIVARIABLE PLANT

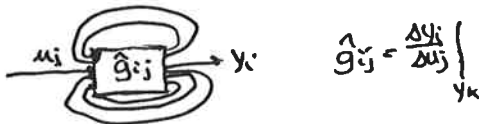


"YOUR" LOOP ALONE:



$$g_{ij} = \left. \frac{\Delta y_i}{\Delta u_j} \right|_{u_k}$$

"YOUR" LOOP WHEN THE OTHERS ARE CLOSED



$$\hat{g}_{ij} = \left. \frac{\Delta y_i}{\Delta u_j} \right|_{y_k}$$

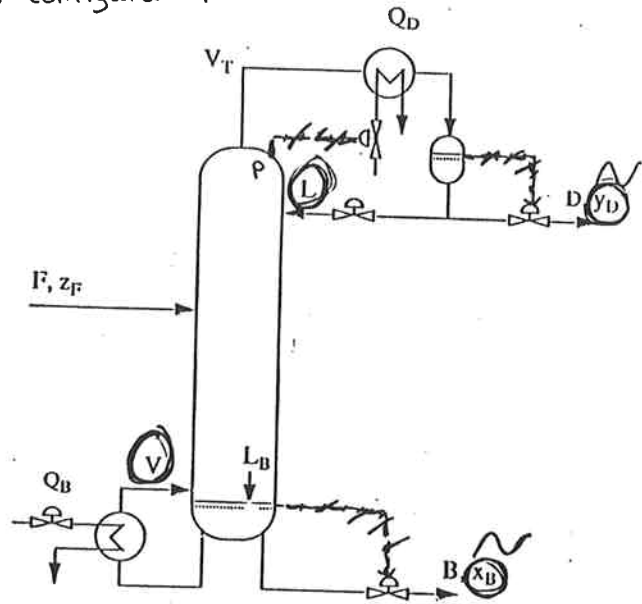
RGA: THE RATIO
 $= g_{ij} / \hat{g}_{ij}$ = "OPEN LOOP GAIN" / "GAIN WITH OTHER LOOPS CLOSED"

WANT ≈ 1 (NO INTERACTION)

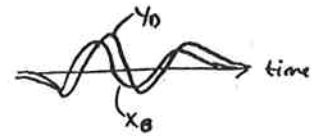
IN GENERAL: $RGA = G \times (G^{-1})^T$

MEASURE OF TWO-WAY INTERACTIONS

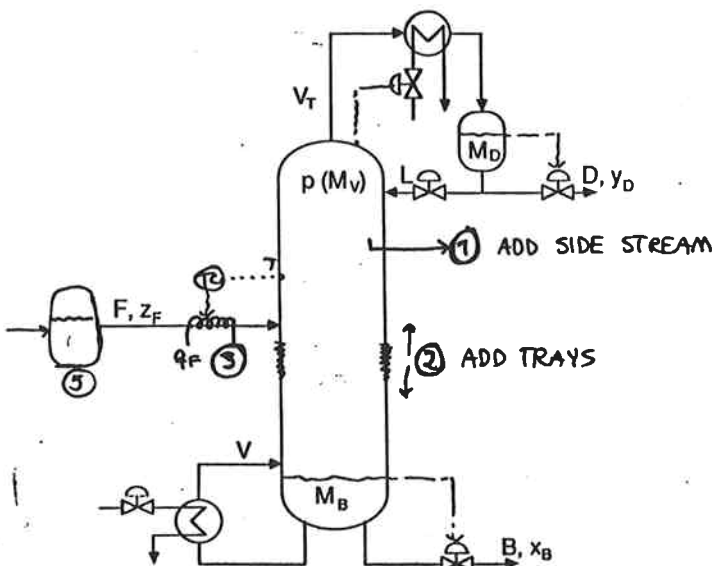
EXAMPLE: DISTILLATION LV-configuration:



MAIN PROBLEM: Strong interactions:



Design changes for improved control



- ④ INCREASE LIQUID HOLDUP ON TRAYS
- ⑤ BUFFER TANK FOR FEED

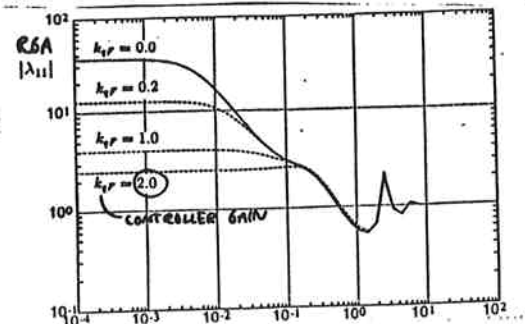
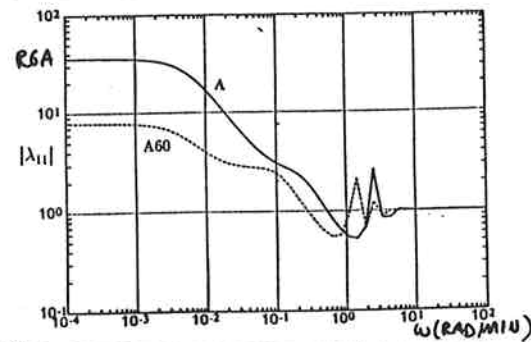
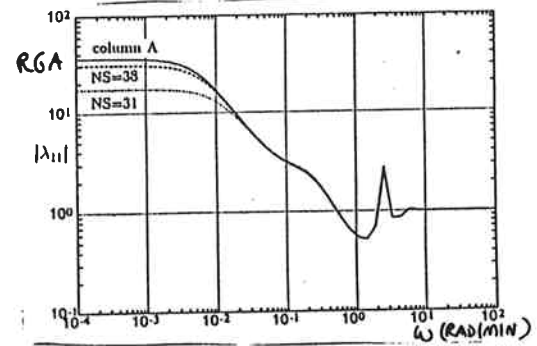
CONSIDER LV-CONFIGURATION

(Jacobsen and Skogestad, 1991c)

① SIDE STREAM

② ADD TRAYS

③ FEED PREHEATER



③ FEED PREHEATER

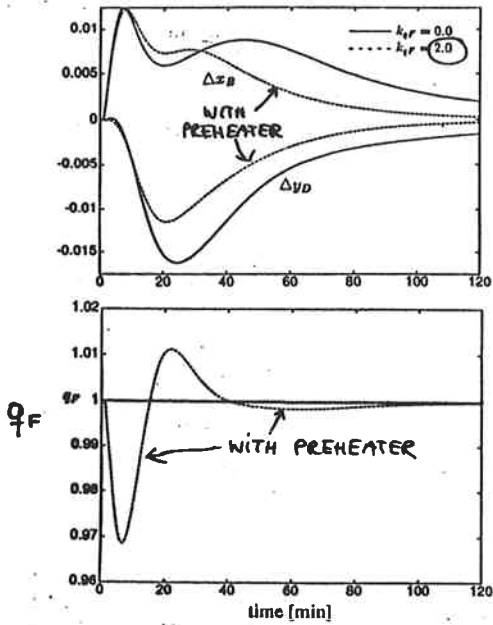
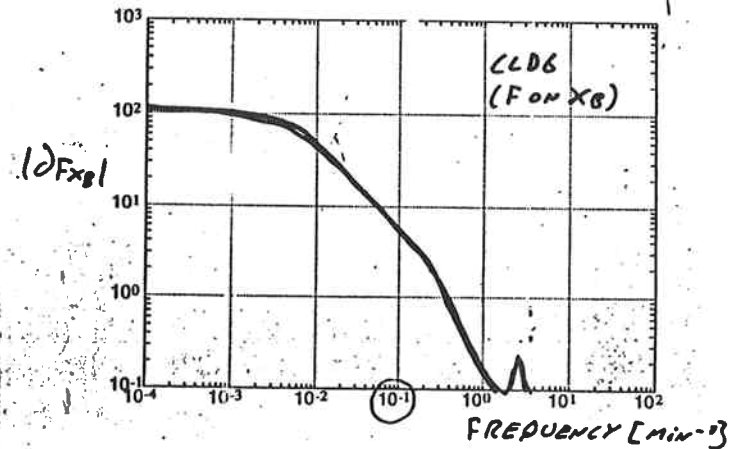
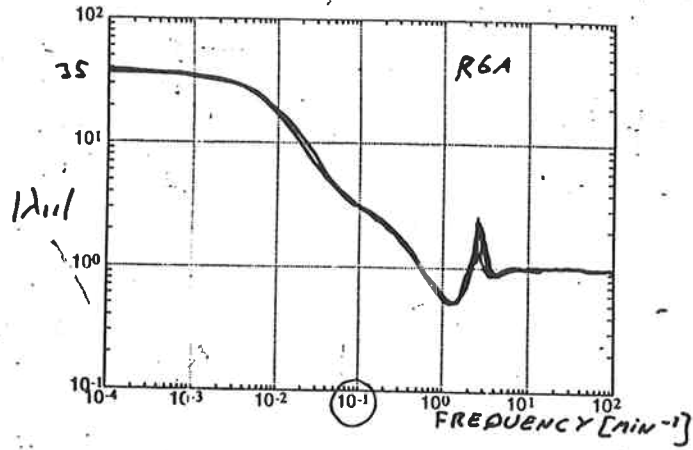


Figure 8.7: Nonlinear response of column A to a 30% increase in feed flow F with and without use of feed preheater for control. Product compositions controlled using single-loop PI-controllers. Lower plot shows response in q_F with feed preheater used for control.

RGA and CLDG for column A (LV-CONTROL)



WITH BUFFER TANK
NEED $\omega_d \approx 0.1$

$\omega_d = 0.3 \text{ rad/min}$
 $\Rightarrow \text{Min. response time} = \frac{1}{0.3} \approx 3.3 \text{ min}$

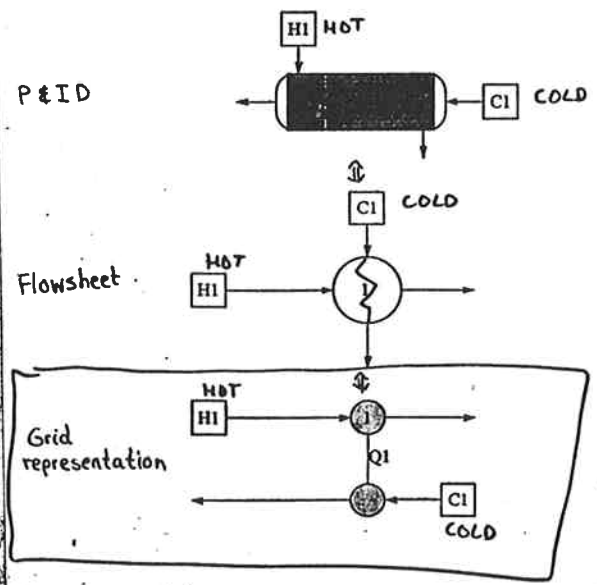
BUT MEAS. DELAY $\Theta = 10 \text{ min} > 3.3 \text{ min}$

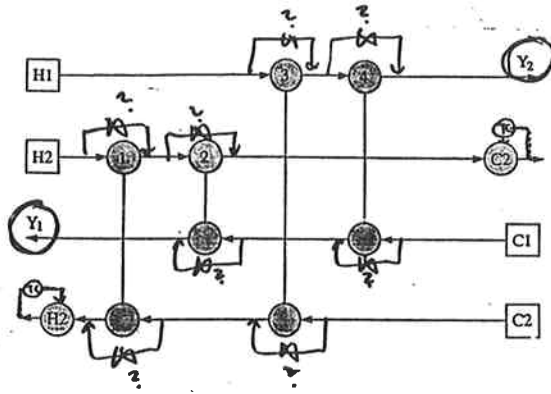
TRY TO ADD BUFFER TANK WITH VARYING LEVEL.
 HOW LARGE TANK?

GET $\tau_{\text{buffer}} = \frac{10}{3.3} \cdot 10 = 30 \text{ min}$ + SLOW LEVEL CONTROL

KNUT W. MATHISEN:
 CONTROL STRUCTURE SELECTION:
 ACTUATORS FOR HEAT EXCHANGER NETWORKS

Heat exchanger representations

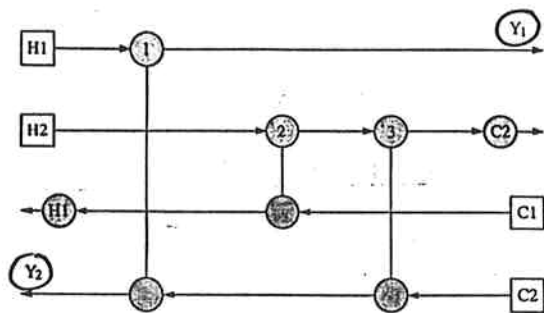




- OBJECTIVE: CONTROL OUTLET TEMPERATURES

- NEED ACTUATORS (BYPASSES)

ISSUE: WHERE PLACE ACTUATORS?



Global optimal design from Gundersen *et al* (1991) of a classic 4 stream problem.

		1H	1C	2H	2C	3H	3C
$G^{all}(0) =$	y_1	1.84	0.91	0.03	0.12	-0.12	-0.20
	y_2	-1.23	-0.61	0.02	0.07	-0.07	-0.11

- Bypass on exchangers 2 and 3 yield a system that is functionally uncontrollable
- Property of the network structure

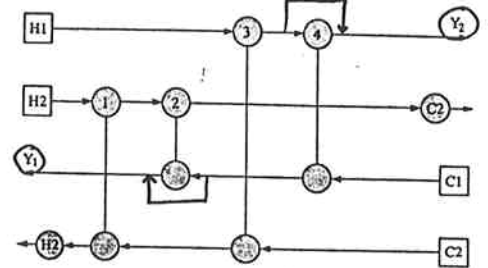
No. of bypass combinations

Consider a general HEN with

- N_{hx} process heat exchangers
- (N_{byp}) single bypasses

The number of bypass combinations:

$$2^{N_{byp}} \frac{N_{hx}!}{N_{byp}!(N_{hx} - N_{byp})!} \quad (3)$$



$N_{hx} = 4, N_{byp} = 2 \Rightarrow 24$ alternative bypass combinations from Eq. 3.

Allow for multi-bypasses and $0 \leq N_{byp} \leq 4$: 2073 different bypass combinations!

A combinatorial problem! \Rightarrow Need insight to simplify and/or effective search algorithms

