Chisa'93

INTERACTIONS BETWEEN PROCESS DESIGN AND CONTROL

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NORWAY

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USE OF CONTROLLABILITY ANALYSIS

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

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OUTLINE

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

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CONTROL HIEARCHY

OBJECTIVE

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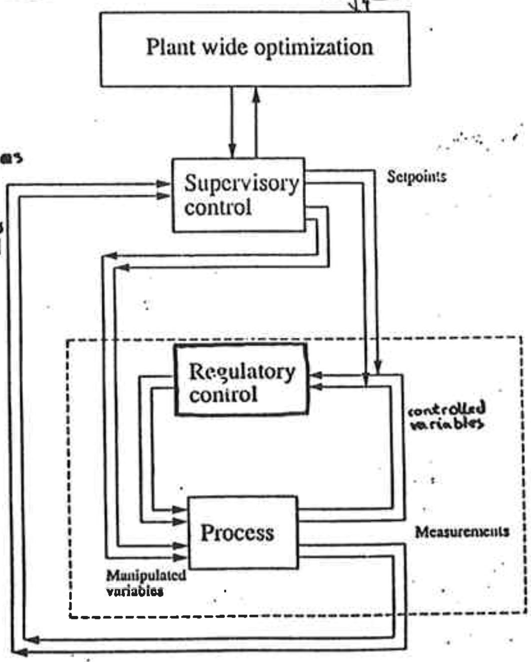
- ($), quality
- Use extra degrees of freedom
- Coordinate
- Avoid constraints in lower level

- Operation



LP

MPC

PID

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

CONTROLLABILITY:

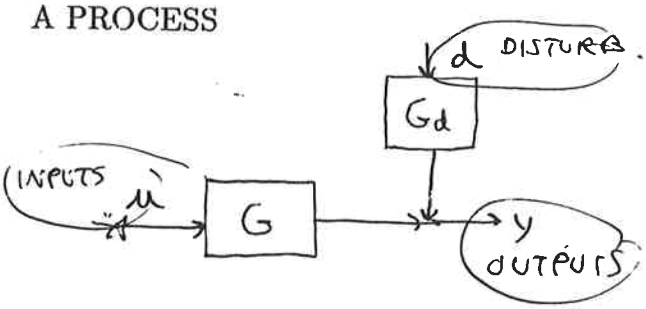
CAN ONLY BE AFFECTED BY "DESIGN" CHANGES:

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

Control structure selection decisions:

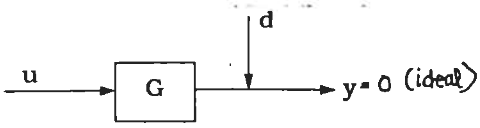
1. Selection of controlled variables. & measurements (y)
2. Selection of manipulated variables. (u)
3. Pairing of controlled and manipulated variables. $(y_1 \leftrightarrow u_1,\ y_2 \leftrightarrow u_2)$

CONTROLLABILITY ANALYSIS OF A PROCESS



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

WHAT LIMITS CONTROLLABILITY?



$$y = Gu + d$$
$$0 \Rightarrow u = -G^{-1}d$$

- "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

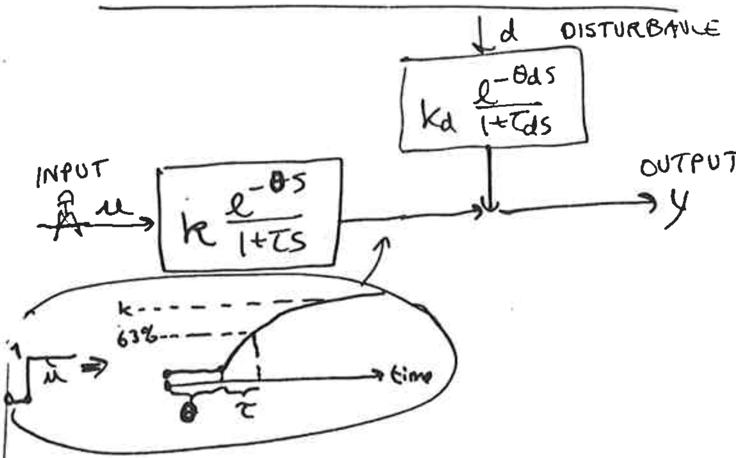
-INTERACTIONS

TOOLS FOR CONTROLLABILITY ANALYSIS

Mostly linear tools, frequency domain :

1. State controllabiity 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

SISO CONTROLLABILITY ANALYSIS

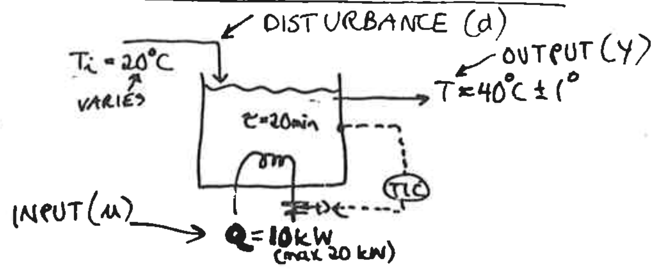


SCALED VARIABLES : $u: \pm 1$, $d: \pm 1$, $y: \pm 1$

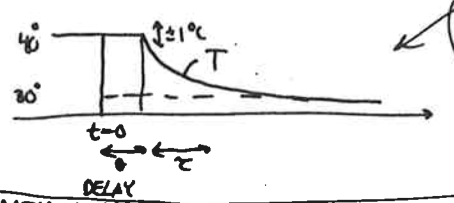
THEN: ACCEPTABLE CONTROLLABILITY :

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 = 10kW):



10 TIMES LARGER THAN ACCEPTED!

MATHEMATICALLY:
SCALED VARIABLES

d = 1: EXPECTED DISTURBANCE $\quad d = \Delta T_i / 10°$

y = 1: ALLOWED DEVIATION $\quad y = \Delta T / 1°$

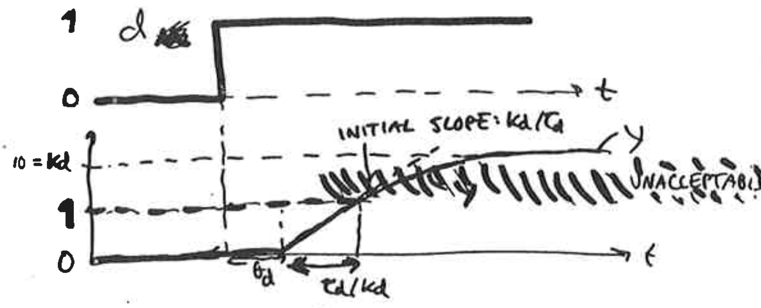
MODEL

$$d \rightarrow \boxed{\frac{k_d \cdot e^{-\theta_d s}}{1 + \tau_d s}} \rightarrow y$$

$k_d = 10$
$\tau_d = 20, \ \theta_d = 5$

$\boxed{\text{CONCLUSION: NEED CONTROL SINCE } k_d > 1}$

GENERALIZATION TO DYNAMICS (STEP RESPONSE)



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

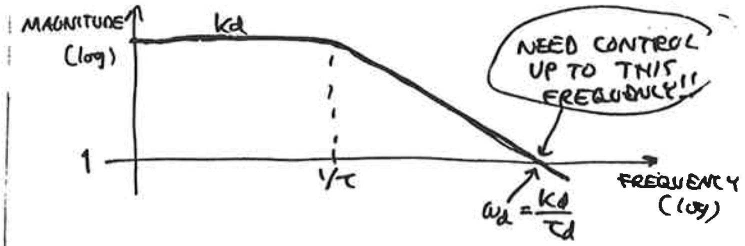
ex: $\tau_d = 20, \ k_d = 10 \Rightarrow \tau_d / k_d = 2$ min

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

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

- 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 FREQUENCY DOMAIN



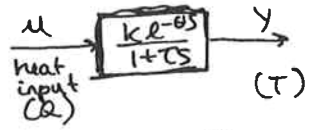
$\frac{k_d}{\tau_d}$ = SLOPE OF INITIAL RESPONSE
- IMPORTANT CONTROLLABILITY PARAMETER (WANT SMALL)

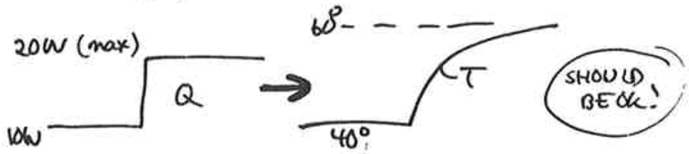
$$\boxed{\text{NEED: } \theta < \frac{1}{\omega_d}}$$

$|g| > |g_d|$

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

INPUT CONSTRAINTS:





SCALED VARIABLES

$u = \Delta Q / 10 \leftarrow$ max change

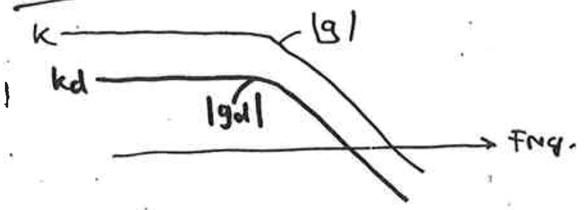
$y = \Delta T / 1$

$\Rightarrow k = \frac{y}{u}\Big|_{\text{steady-state}} = 20$

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).

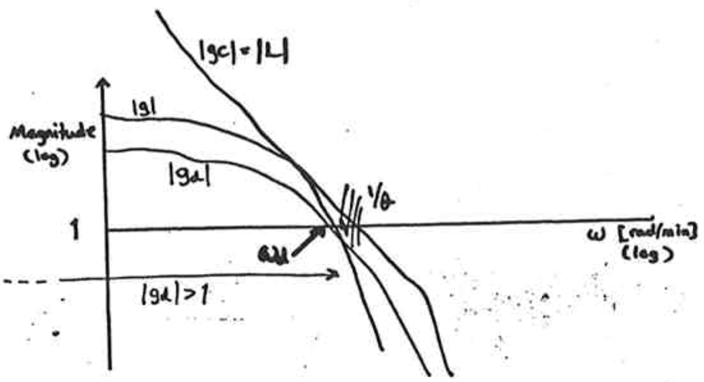
Need:

1. $|g| > |g_d|$: Avoid input constraints ($|u| \leq 1$)

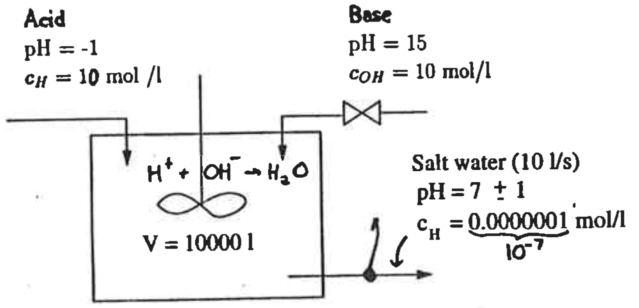
   $\Updownarrow$

   ($|g^{-1}g_d| < 1$)

2. $\underbrace{|gc|}_{L} \gtrsim |g_d|$: Acceptable performance ($|y| \leq 1$).



EXAMPLE. Neutralization process.



Let $y = c_H - \cancel{10^{-7}}\ c_{OH}$ (difference from neutrality)

$u = Flow_{base}$

$d = Flow_{acid}$

Model with appropriate scalings $\quad g = -g_d$

$$y = \frac{k_d}{1+\tau s}(u+d)$$

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

EXTREMELY SENSITIVE TO DISTURBANCES.

Frequency up to which feedback is needed

$$\omega_B > \omega_d = \underbrace{0.25 \cdot 10^7}_{k_d}/\tau = 2500 \text{ rad/s}$$

But delay is $\underline{\theta = 10s}$ so bandwidth must be less than

$$\omega_B < 1/\theta = 0.1 \text{ rad/s}$$

INCOMPATIBLE $\Rightarrow$ POOR CONTROLLABILITY

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

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

pH feed

pH spec

normalized flow change

pH=1 FOR ACID

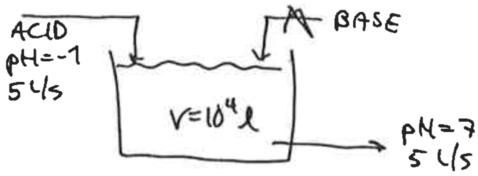
5±2.5 l/s

$$= \frac{10}{10^{-6}} \cdot \frac{2.5}{10} = \underline{0.25 \cdot 10^7}$$

pH: 7±1 IN OUTLET

10 l/s

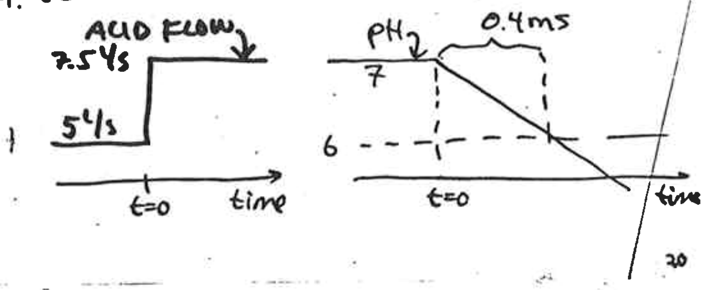
PHYSICAL EXPLANATION:



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

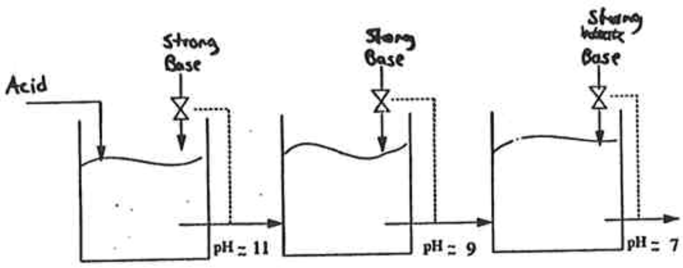
ANSWER:

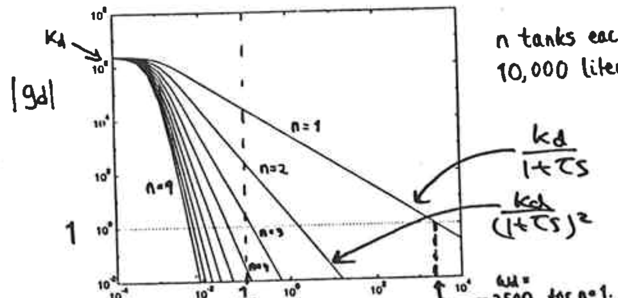
1. MUST ADD H^+ IONS: $\Delta C \cdot V = \underline{10^{-2} \text{ mol } H^+}$ (ΔC: 10^{-6} mol/l; V: 10^4 l)
2. THIS CORRESPONDS TO: 10^{-2} mol / $\underbrace{10^1 \text{ mol/l}}_{pH=-1}$ = $\underline{10^{-3} \text{ l ACID}}$
3. IF THE FEED INCREASES FROM 5 l/s TO 7.5 l/s THIS TAKES: 10^{-3} l / 2.5 l/s = $\underline{0.4 \cdot 10^{-3} s}$
4. GET:



IMPROVE CONTROLLABILITY BY REDESIGN OF PROCESS

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





n tanks each 10,000 liters

$\frac{k_d}{1+\tau s}$

$\frac{k_d}{(1+\tau s)^2}$

With n tanks: $\boxed{g_d(s) = k_d/(1+\tau s)^n}$

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

Assume delay for control is about 10 s in each tank.
Get same controllability with:

$\omega_d = \frac{1}{\theta} = 0.1$

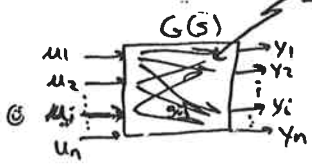
3 tanks of about 13500 l each - 40.5 m^3 total volume
4 tanks of about 4000 l each - 16.0 m^3 total
5 tanks of about 1900 l each 9.5 m^3 total
6 tanks of about 1160 l each - 7.0 m^3 total
7 tanks of about 820 l each - 5.7 m^3 total
8 tanks of about 630 l each - 5.0 m^3 total

Minimum total volume:
16 tanks of about 251 l each - 4.02 m^3 total

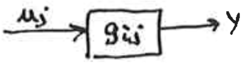
CONCLUSION: USE ~5 tanks.

RELATIVE GAIN ARRAY

MULTIVARIABLE PLANT



"YOUR" LOOP ALONE:



$$g_{ij} = \frac{\Delta y_i}{\Delta u_j}\Big|_{u_k}$$

"YOUR" LOOP WHEN THE OTHERS ARE CLOSED

$$\hat{g}_{ij} = \frac{\Delta y_i}{\Delta u_j}\Big|_{y_k}$$

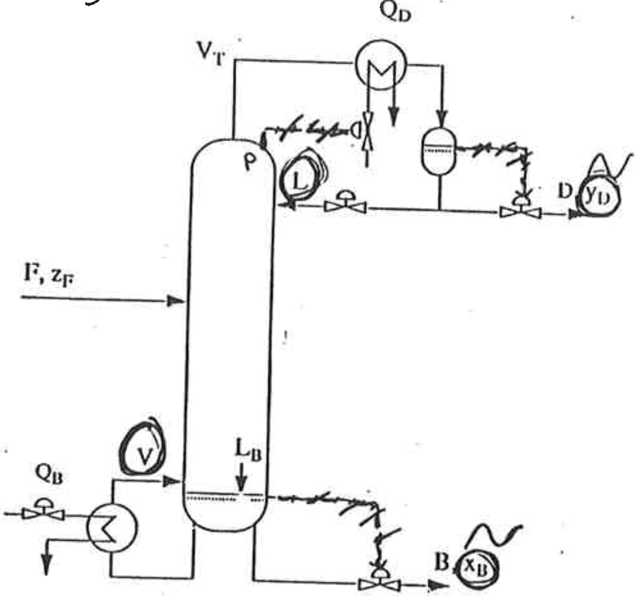
RGA: THE RATIO $= g_{ij}/\hat{g}_{ij} = \dfrac{\text{"OPEN LOOP GAIN"}}{\text{"GAIN WITH OTHER LOOPS CLOSED"}}$

WANT ≈ 1 (NO INTERACTION)

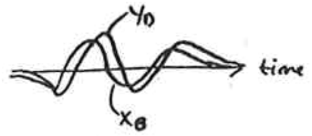
IN GENERAL: $RGA = G \times (G^{-1})^T$

MEASURE OF TWO-WAY INTERACTIONS

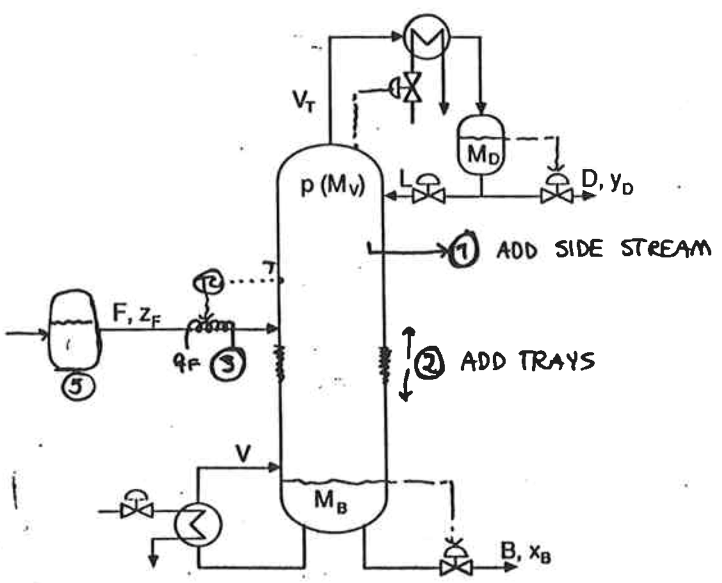
EXAMPLE: DISTILLATION
LV-configuration:



MAIN PROBLEM: strong interactions:



Design changes for improved control

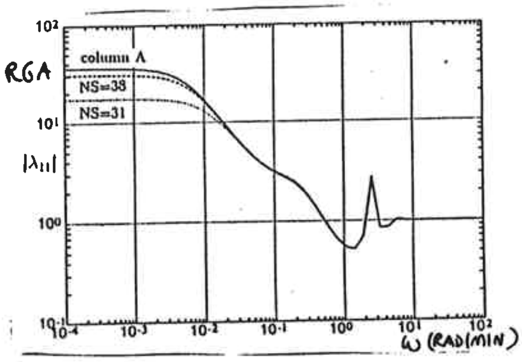


4 INCREASE LIQUID HOLDUP ON TRAYS
5 BUFFER TANK FOR FEED

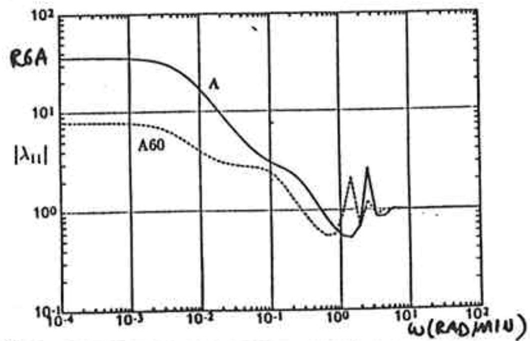
CONSIDER LV-CONFIGURATION

(Jacobsen and Skogestad, 1991c)

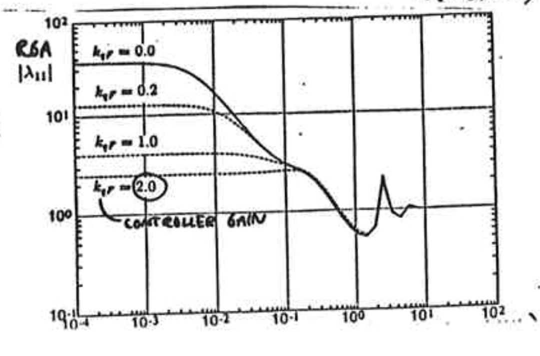
1 SIDE STREAM



2 ADD TRAYS



3 FEED PREHEATER



③ FEED PREHEATER

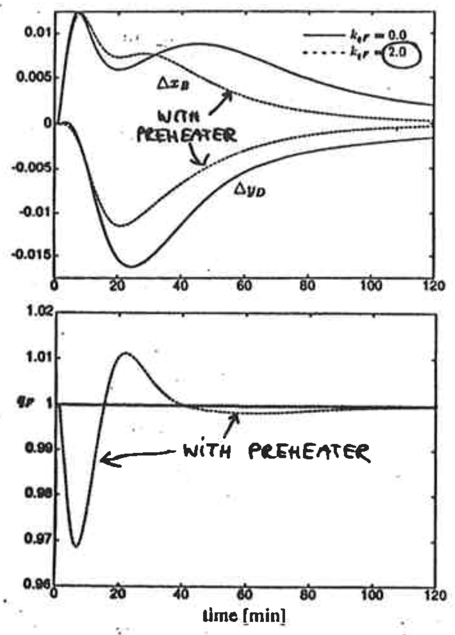
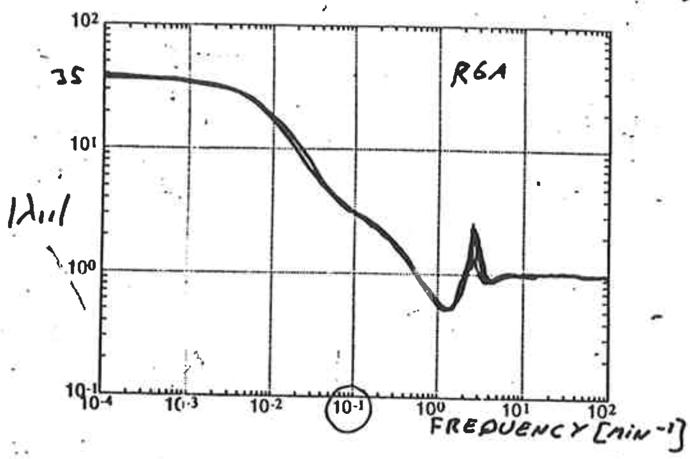
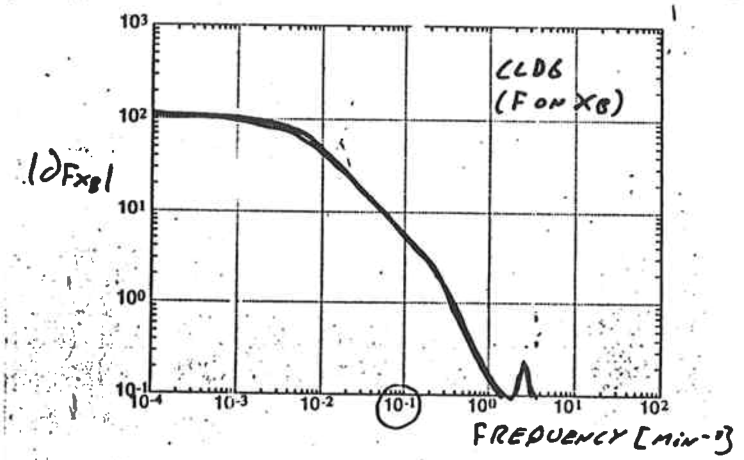
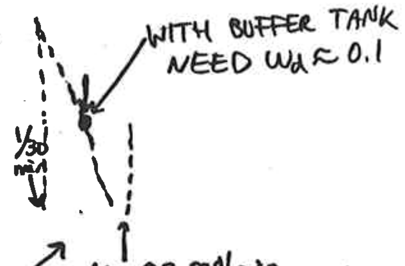


Figure 6.7: Nonlinear response of column A to a 30% increase in feed flow F with and without use of feed preheater for control. Product compositions controlled using single-loop PI-controllers. Lower plot shows response in q_F with feed preheater used for control.

RGA and CLDG for column A (LV-CONFIG.)







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

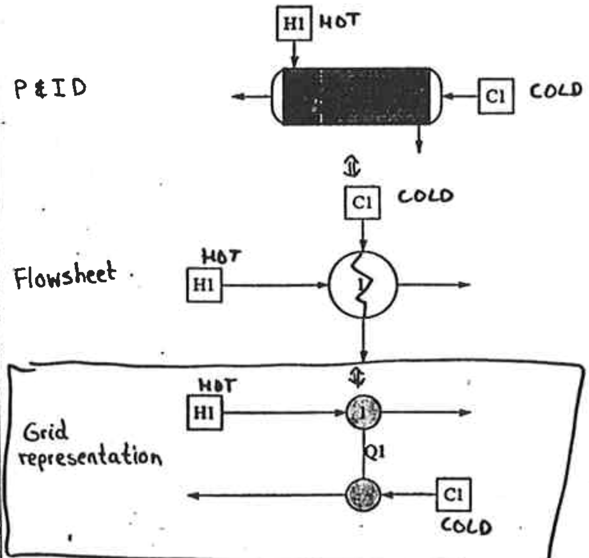
BUT MEAS. DELAY $\theta = 10$ min > 3.3 min

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

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

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

Heat exchanger representations



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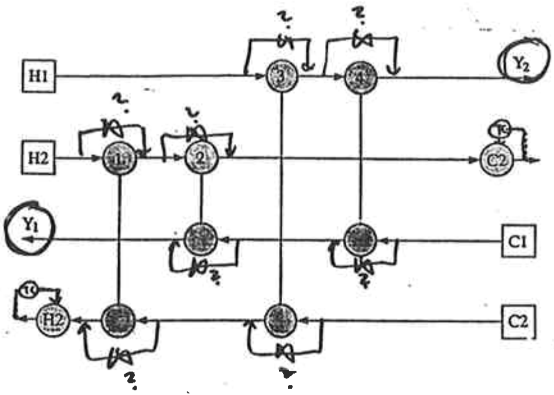

- OBJECTIVE: CONTROL OUTLET TEMPERATURES

- NEED ACTUATORS (BYPASSES)

ISSUE: WHERE PLACE ACTUATORS?

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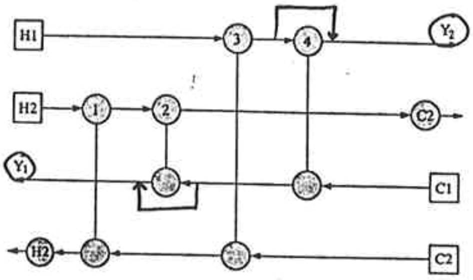
No. of bypass combinations

Consider a general HEN with

- N_{hx} process heat exchangers
- N_{byp} *single* bypasses

The number of bypass combinations:

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



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

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

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

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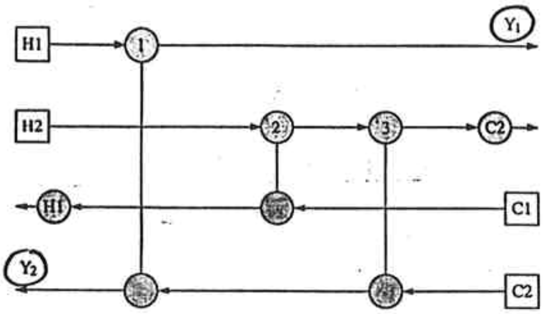

Global optimal design from Gundersen *et al* (1991) of a classic 4 stream problem.

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

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

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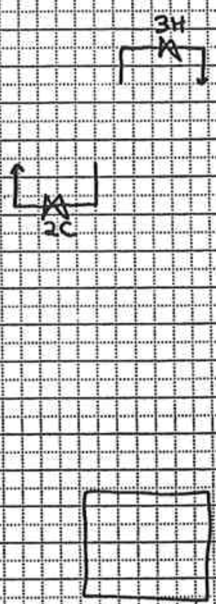
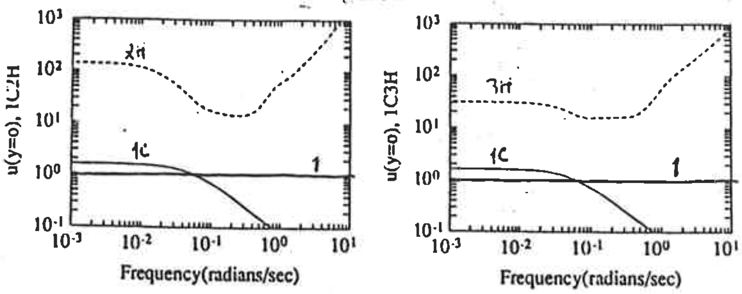


3. Input constraints

- Exchangers 2 and 3 are below pinch, whereas exchanger 1 is mainly above
- The structure of the network forces one to control an output above pinch with a bypass below!



Required manipulation for perfect control $G^{-1}G_d$ for cases $1C2H$ and $1C3H$. ⟵"BEST"

- Multi-bypass does not help for constraints
- This "global optimal solution" must be discarded.

A Procedure for Regulatory Control Structure Selection with Application to the Fluid Catalytic Cracking Process

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AIChE Journal

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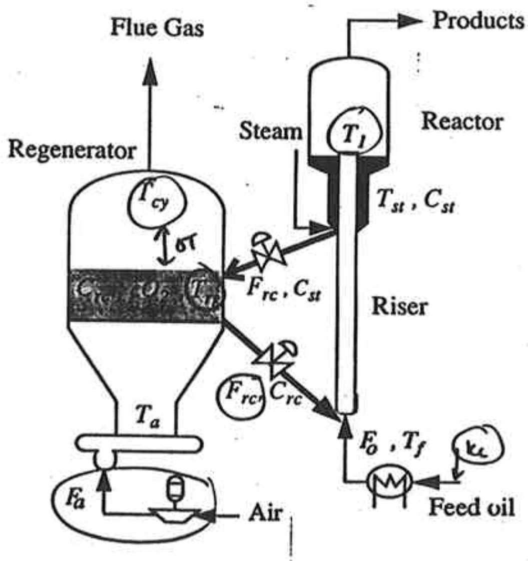
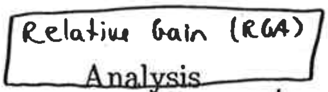
Example: FCC



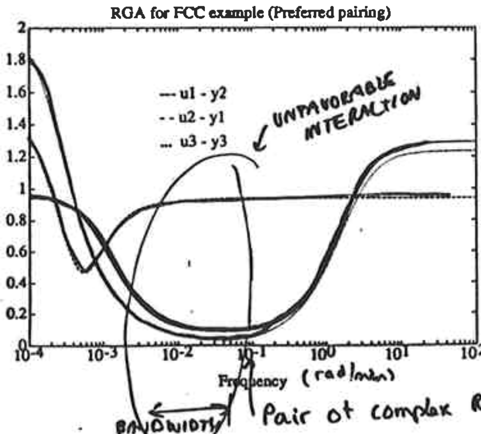
$$u = \begin{pmatrix} F_a \\ F_{rc} \\ k_c \end{pmatrix} \quad y = \begin{pmatrix} T_1 \\ T_{cy} \\ T_{rg} \end{pmatrix} \quad d = \begin{pmatrix} T_f \\ T_a \\ F_f \end{pmatrix}$$

- ISSUE:
  1) IS 3x3 CONTROL A GOOD IDEA?
  2) WHAT SHOULD BE USED FOR 2x2 CONTROL?



Analysis

$$\Lambda(0) = \begin{array}{c} \\ \\ \\ \end{array}\begin{array}{ccc} F_a & F_{rc} & k_c \end{array}$$
$$\Lambda(0) = \begin{pmatrix} 0.99 & 1.50 & -1.47 \\ 0.97 & -0.42 & 0.45 \\ -0.96 & -0.08 & 2.04 \end{pmatrix} \begin{array}{c} T_1 \\ T_{cy} \\ T_{rg} \end{array}$$

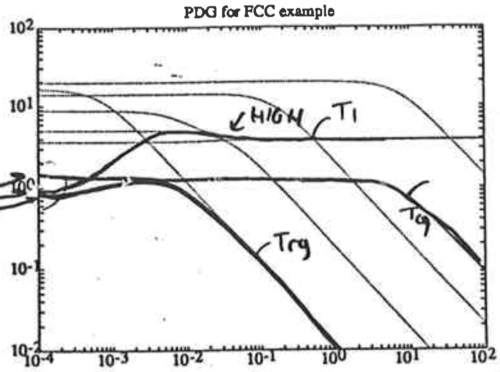


- Significant interaction at intermediate frequency
- Decoupler not advisable
- Only one pairing with $\Lambda(0) > 0$
- PRGA: Not significant triangularity

⟹ 3x3 CONTROL SEEMS DIFFICULT!

Disturbance Analysis for Partial Control (2x2)

PDG for FCC example



NOT USE

UNUSED

$$
G_{PDG}(0) = \begin{array}{c} \\ \end{array}
\begin{array}{cccc}
 & F_a & F_{rc} & k_c \\
 & 9.0 & 3.3 & 0.84 & T_1 \\
 & 15.9 & 20.0 & 1.6 & T_{cy} \\
 & 17.3 & 4.7 & 0.81 & T_{rg}
\end{array}
$$

T_{rg} ← PROBABLY LEAVE UNCONTROLLED

OVERALL EFFECT OF THE THREE DISTURBANCES

- Leaving y_3 uncontrolled and input u_3 in manual gives us a 2×2 system with acceptable disturbance rejection properties.

CONCLUSION: DO NOT USE kc, THEN NEED ONLY CONTROL TWO OUTPUTS!

→ CONSIDER 2x2 !!

2x2 control problem

Controlled variables:

Primary variable: T_{cy} or $\Delta T_{rg} = T_{cy} - T_{rg}$
Secondary variable: T_1 or T_{rg}

Manipulated variables:

- Flowrate of regenerated catalyst F_{rc}
- Flowrate of air to regenerator F_a.

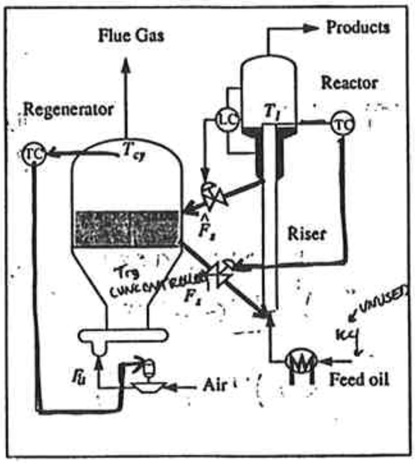
Disturbances:

- Feed oil temperature T_f.
- Air temperature T_a.
- Feed oil flowrate F_f.
- Coke producing tendency of the feed k_c. ← Note! Disturbance for 2x2 problem

Selection of controlled variables.

	Outputs	RHPT zeros [rad/min]
Conventional	$\Delta T_{rg}, T_1$	0.02
Kurihara	$\Delta T_{rg}, T_{rg}$	0.19
Alt. Kurihara	T_{cy}, T_{rg}	0.19
Hicks	T_{cy}, T_1	-
Riser-regenerator	T_1, T_{rg}	-

→ choose Hicks control structure or riser-regenerator control structure.
Will concentrate on the Hicks control structure in the following.



Hicks

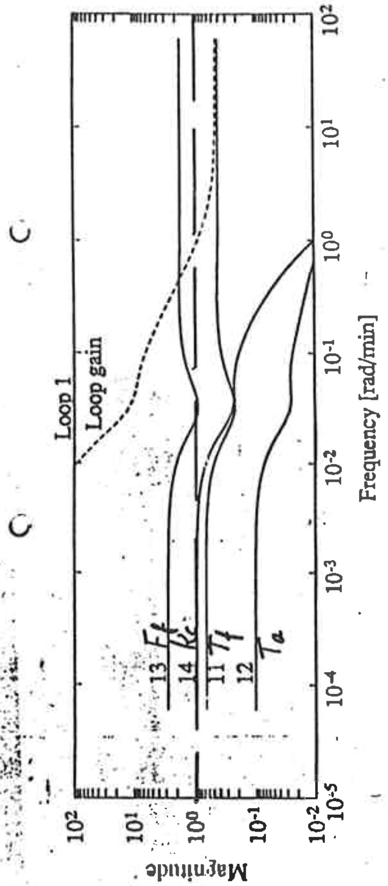
Scaling of outputs and disturbances:

Largest tolerated offset in

- Riser exit temperature: 3K
- Regenerator cyclone temperature: 2K

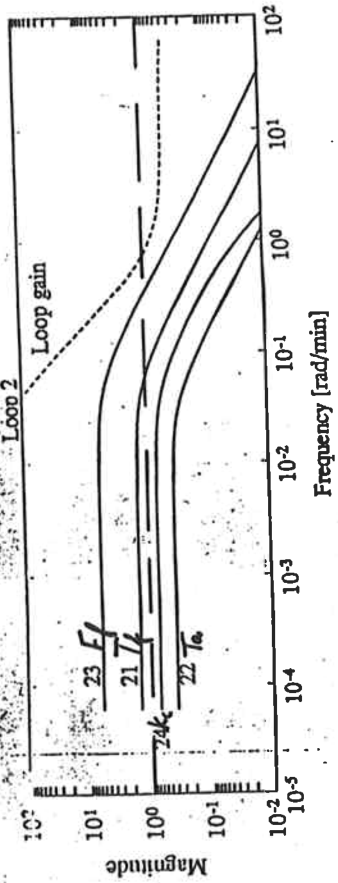
Largest expected size of disturbance:

- Feed oil temperature: 5 K
- Air temperature: 5 K
- Feed oil flowrate: 4 kg/s (ca. 10%)
- Coke producing tendency of feed: 2.5% (relative to original value)

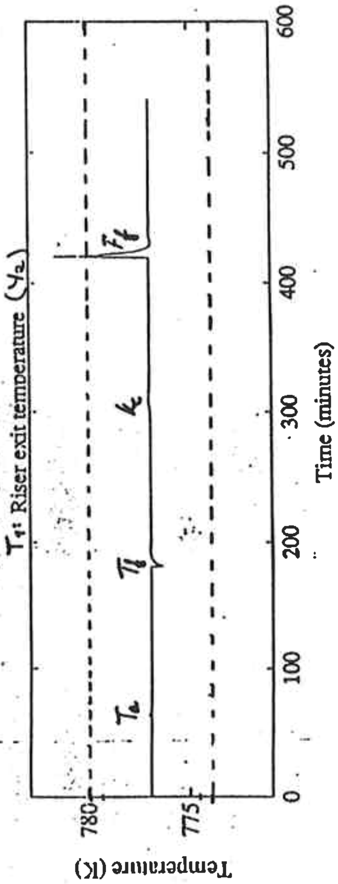


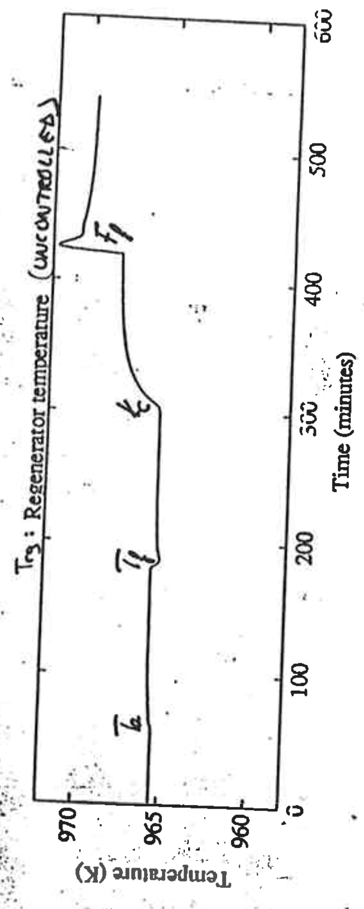
EFFECT OF DISTURBANCES ON RISER EXIT TEMPERATURE

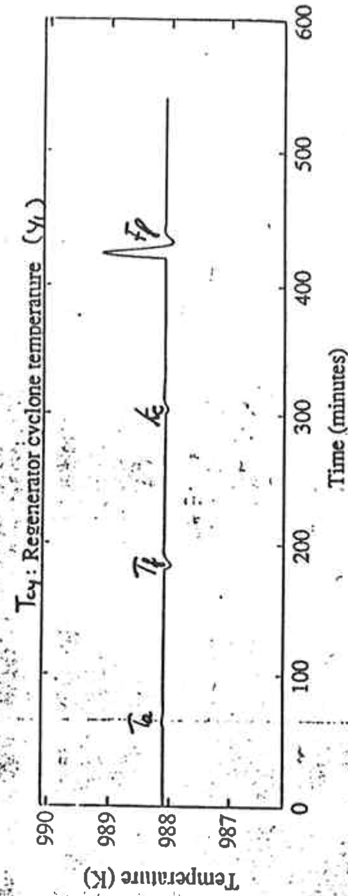
CLDG



EFFECT OF DISTURBANCES ON REGENERATOR CYCLONE TEMPERATURE







Conclusions:

- Regenerator cyclone temperature must be included in model.
- Hicks control structure (or riser-regenerator control structure) preferable.
- Disturbances in F_f will affect T_1.
- Uncertainty w. r. t. model structure can have some effect.

INTERACTIONS BETWEEN PROCESS DESIGN AND CONTROL.

CONCLUSIONS :

- DESIGN CHANGES MAY BE ONLY WAY TO GET ACCEPTABLE CONTRO
- IMPORTANT FOR CONTROLLABILITY
  - DISTURBANCE SENSITIVITY ("SELF-REGULATING PROPERTI
  - TIME DELAYS
  - MULTIVARIABLE: RHP ZEROS
  - INPUT CONSTRAINTS
- BARRIERS
  - ORGANIZATIONAL
  - EDUCATION
  - INTEGRATED TOOLS

Plenary paper for CHISA'93, Prague, 29 Aug - 3 Sep 1993

INTEGRATION BEWTEEN PROCESS DESIGN AND CONTROL.

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A plant should be designed such that it is able to adjust to changes in operation policy, feedstocks, specifications, product loads, and to failures and other disturbances in an economic and safe manner. During normal operation the regulatory (often PID controllers), supervisory (often the operator) and optimizing (often the engineer) control levels are responsible for adapting the process to such changes. In addition there is a control system which deals with startup, shutdown, etc., but this is not considered here. In this talk the focus is on the regulatory control system which deals mainly with counteracting the effect of fast changes (disturbances) to maintain smooth operation.

The link between process design and control is provided by the term "controllability" (of a plant), which has has the meaning of "inherent control characteristics of the plant" or maybe better "achievable performance" (irrespective of the controller). This usage is in agreement with most persons intuitive feeling about the term, and was also how the term was used historically in the control literature. For example, Ziegler and Nichols (1943) define controllability as *"the ability of the process to achieve and maintain the desired equilibrium value"*. Unfortunately, in the 60's Kalman defined the term "controllability" in the very narrow meaning of "state controllability". This concept is of interest for realizations and numerical calculations, but as long as we know that all the unstable modes are both controllable and observable, it has almost no practical significance.

It would be desirable to have a more precise definition of controllability, but on the other hand this is difficult and probably not useful. An exact definition would require selection of a certain norm to measure the control error, and would also require a detailed specification of all external signals such as noise, reference signals and disturbances. Indeed, Ziegler and Nichols (1943) note in their paper that although they *took the area under a recovery curve as one measure of controllability ... this is only one of many possible bases for comparison of control results.* They also stress that it is difficult to narrow controllability down to one single attribute of the plant. They say: *Unfortunately, the authors are not able to give a formula for controllability. It appears that when such a factor is devised it will consist of several factors. One might be called the "recovery factor", the ability of the process to recover from the maximum change in demand or load. Another, a "load factor" must take into account the point in the process at which the disturbance occurs.* Later in the paper they state that the total integrated control error, $\int |e(t)|dt$, is equal to: (Load Factor) $\cdot$ (Recovery Factor). Essentially, the "recovery factor" depends on the process model, $g(s)$, and recovery is poor (and thus the recovery factor is large) if it contains large time delays or if the plant gain is small. The "load factor" expresses the effect of the disturbances and thus depends on the disturbance model, $g_d(s)$, and the load factor is large if the disturbances have a large effect.

The achievable control quality depends strongly on the plant design, for example, a large ship cannot make a sudden turn no matter how sophisticated the control system is. On the other hand, in some cases feedback control can have quite drastic effects on the dynamic response, and it is possible, for example, to achieve fast control (within minutes) of large distillation columns which seemingly are very slow (with an uncontrolled response time of hours or days) provided the measurements are sufficiently fast. Thus, in general a quite careful analysis is required to say how easy the plant is to control. However, we do not want to perform a detailed controller design and simulation for each possible design alternative. Thus, there is a need for relatively simple tools for evaluating controllability, and the main part of the talk is focused on discussing various tools for controllability analysis.

1. Compute the multivariable RHP-poles and RHP-zeros and their associated directions. Test for functional controllability (the rank of G should equal the number of outputs).

2. Perform a frequency-dependent SVD-analysis to understand the multivariable directions.

3. Perform a frequency-dependent RGA-analysis to check for fundamental limitations due to inherently coupled outputs. Compute the plant condition number.

4. Evaluate disturbance sensitivity. For decentralized control the use of the CLDG-matrix, $G_{diag}G^{-1}G_d$, directly generalizes the SISO results. Here G_{diag} is a diagonal matrix consisting of the diagonal elements of G. For the general case it is more complicated, but an SVD-analysis of G_d and $G^{-1}G_d$ yields useful information about which disturbances are difficult, and the bandwidth requirement in certain directions.

The above tools for controllability analysis are simple indicators which are easy to compute, and help the engineer to obtain insight into what the control problems are for the plant in question. In the talk these tools are applies to a distillation column and a reactor example. A number of the tools presented may be applied also to evaluate flexibility (steady-state controllability). Although, there has been good progress during the last few years, the area of controllability analysis is still a very interesting area for future research. Since a detailed controllability analysis at the design stage usually is prohibitive, it is important to focus part of this research on obtaining design rules (rules of thumb) for various classes of processes.