PLANTWIDE CONTROL

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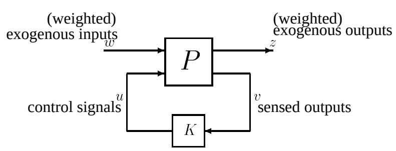
OUTLINE

1. Introduction and Motivation
2. Plantwide control and control structure design
3. Some tools – controllability analysis
4. An outline of a procedure
5. Some examples
   - Tennessee Eastman plantwide challenge problem
   - Reactor with recycle (“snowballing” ??)
   - Buffer tanks for pH-control
   - Distillation column control
   - Optimizing control of Petlyuk distillation
6. Conclusion

CONTROL THEORY

General controller design formulation



- w: Disturbances (d) and setpoints (r)
- v: Measurements (y_m, d_m) and setpoints (r)
- u: Manipulated inputs (u)
- z: Control error, $y - r$
- Find a controller K which based on the information in v, generates a control signal u which counteracts the influence of w on z, thereby minimizing the closed-loop norm from w to z.

PRACTICE
Typical base level control structure

PRACTICE
Typical control hierarchy

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.

Recent developments

- Many case studies (Luyben and others; Tennessee Eastman process)
- Some theoretical tools (e.g. chapter 10 in book by Skogestad and Postlethwaite, Wiley, 1996)
- Several ad-hoc procedures for plantwide control
- BUT: No unified approach (which is the goal of our work)

PLANTWIDE CONTROL

The *control philosphy* for the overall plant with emphasis on the *structural decisions*:

- Which “boxes” (controllers; decision makers) do we have and what information (signals) are send between them

NOT:

- The inside of the boxes (design and tuning of all the controllers)

The most important sub-problem: CONTROL STRUCTURE DESIGN

PLANTWIDE CONTROL
Some issues:

- Where is the production rate set?
- Degrees of freedom - local “tick-off” can be useful
- Configuration for stabilizing control may effect layers above (including easy of model predictive control)

PLANTWIDE CONTROL
Alt.1 ”Cascade of SISO loops” - Control structure design

- Local feedback
- Close loop - same number of DOFs but uses up dynamic range
- Cascades - extra measurements,
- Cascades - extra inputs
- Selectors
- RGA

PLANTWIDE CONTROL
Alt.2 ”Optimization”: Multivariable predictive control

- Model-based
- Mostly feedforward based
- Excellent for extra inputs and changes in active constraint
- Feedback somewhat indirectly through model update.

Alt.3 Usually: A combination of feedback and models.

- How to find the right balance

CONTROL STRUCTURE DESIGN

Tasks:

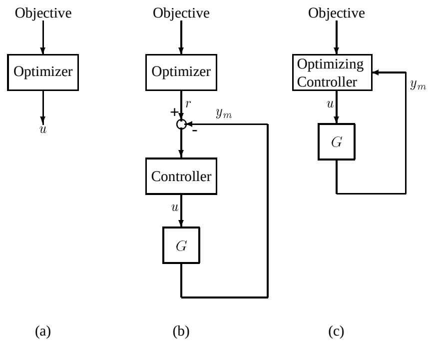
1. *Selection of controlled outputs* (a set of variables which are to be controlled to achieve a set of specific objectives)
2. *Selection of manipulations and measurements* (sets of variables which can be manipulated and measured for control purposes)
3. *Selection of* **control configuration** (a structure interconnecting measurements/commands and manipulated variables)
4. *Selection of controller type* (control law specification, e.g., PID, decoupler, LQG, etc.).

Note distinction between control *structure* (all tasks) and *configuration* (task 3).

Tasks 1 and 2 combined: **input/output selection**
Task 3 (configuration): **input/output pairing**

Shinskey (1967, 1988);Morari (1982); Stephanopoulos (1984); Balchen and Mumme (1988)

Nett (1989); van de Wal and de Jager (1995); Skogestad and Postlethwaite (1996)



(a) Open-loop optimization.
(b) HERE: Closed-loop implementation with separate control layer.
(c) Integrated optimization and control.

Hierarchical structuring:

- *optimization layer* — computes references r
- *control layer* — implements this in practice, $y \approx r$.

Example 1. *Room heating.*
$y =$ room temperature

Other cases: Less obvious.

Example 2. *Cake baking.*
Goal (purpose): well baked inside and nice outside
Manipulated input: $u = Q$ (assume 15 minutes).

(a) Open-loop implementation: Heat input Q

(b) Closed-loop implementation:
$y =$ oven temperature
"Optimizer": Cook book (look-up table)

TASK 1: Selection of controlled outputs

Controlled output y*:* Measured output with reference (r)

Two distinct questions:

1. What should be the controlled variables y?
(includes open-loop by selecting $y = u$)
2. What is their optimal values (y_{opt})?

Second question: A lot of theory.

BUT First question: Almost no theory. Decisions mostly made on experience and intuition.

THEORY: SELECTION OF CONTROLLED OUTPUTS

- Assume we have performed steady-state optimization and have obtained u_{opt}:
$$\min_u J(u) = J(u_{opt})$$
where J is the operating cost ($).
Note: $u_{opt}(d)$ depends on the disturbances (operating point).
- **Obvious:** *The input* u *(possibly generated by feedback to achieve* $y \approx r$*) should be close to the optimal input* $u_{\text{opt}}(d)$.

$$u - u_{\text{opt}} = G^{-1}(0)(y - y_{\text{opt}})$$

where $G(0)$ - effect of small change in u on y, and

$$y - y_{\text{opt}} = \underbrace{y - r}_{\text{Control error}} + \underbrace{r - y_{\text{opt}}(d)}_{\text{Optimization error}}$$

⇒ Select controlled outputs y such that:

1. *Optimization error* $r - y_{\mathrm{opt}}(d)$ *is small;* $y_{\mathrm{opt}}(d)$ *depends only weakly on disturbances.*
2. *Control error* $y - r$ *is small; good measurement and control of* y.
3. $G^{-1}(0)$ is small; the variables y are uncorrelated.

Simple tool for selecting controlled outputs:

- Scale outputs such that $\|y - y_{\mathrm{opt}}(d)\| \approx 1$ (due to measurement errors and disturbances)
- Prefer a set of controlled outputs with large $\underline{\sigma}(G(0))$.

Note: $\bar{\sigma}(G^{-1}(0)) = 1/\underline{\sigma}(G(0))$.

TASK 2: Selection of manipulations and measurements

Dynamics and *controllability* are more important here.

- Manipulations u – usually fixed (the valves), but may not want to use all of them (see task 1) or may change their location.
- Measurements – may want to add secondary measurements y_{2m} to
  1. Compensate for lack of measurements of primary output y
  2. Improve dynamic response – e.g. by adding a measurement of y_2 “close” to the manipulation u the manipulation u

  Can perform **controllability analysis** of alternative combinations.

SUMMARY
Rules for selecting controlled outputs y

Select the controlled outputs y such that:

1. Optimal value $y_{\mathrm{opt}}(d)$ is insensitive to disturbances (changes in the operating point)
2. Result insensitive to expected control error for y.
   (a) “Optimum is flat” and/or
   (b) Can achieve tight control of y (need accurate measurement)
3. The outputs are weakly correlated

This is usually based on a steady-state analysis

PROBLEM: Combinatorial growth

Possibilities with 1 to M inputs and 1 to L outputs (Nett, 1989):

$$\sum_{m=1}^{M}\sum_{l=1}^{L}\binom{L}{l}\binom{M}{m}$$

$M = L = 2$: 4+2+2+1=9 candidates

$M = L = 4$: 225 candidates, etc.

TOOLS THAT AVOID COMBINATORIAL GROWTH DESIRED.

RGA is one such tool.

TASK 3: Selection of control configuration

Controller K connects available measurements/commands (v) and manipulations (u):

$$u = Kv$$

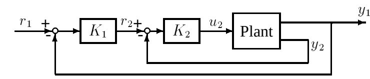
Control configuration: *The restrictions imposed on the structure of the overall controller K by decomposing it into a set of local controllers (subcontrollers, units, elements, blocks) with predetermined links and with a possibly predetermined design sequence.*

Some elements used to build up configuration:

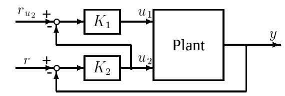
- Decentralized controllers (K diagonal)
- Cascade controllers (with predetermined order for tuning)
- Feedforward elements
- Decoupling elements
- Selectors

$\Rightarrow$ Split the "big" controller into many smaller boxes.

Cascaded controllers



(a) Extra measurements y_2 (conventional cascade control)



(b) Extra inputs u_2 (input resetting)

Why use control configurations?

- Decomposed configurations often quite complex.
- Better performance: Optimization problem – resulting in a centralized multivariable controller.

So why use control configurations?

- Cost associated with obtaining good plant models (needed for centralized control).
- Cascade, decentralized, etc.: Controller is usually tuned **on-line** one at a time with little modelling effort.
- $\Rightarrow$ Rely on feedback rather than on models

Other advantages decentralized/cascade/hierarchical configurations:

- "Stabilize" the plant such that it is can be controlled by operators.
- Simple or even on-line tuning
- Tuning parameters have direct and "localized" effect
- Often easier to understand for operators
- Tend to be insensitive to uncertainty
- Allow simple models when designing higher layers
- Reduce the need for control links
- Allow for decentralized implementation
- Simpler implementation
- Reduced computation load
- Longer sampling intervals for the higher layers

Comment. The terms "stabilize" and "unstable" as used by operating people may not refer to a plant that is unstable in a mathematical sense, but rather to a plant that is *sensitive* to disturbances and which is difficult to control manually.

THEORY FOR CONTROL CONFIGURATIONS

Partial control

Close loop involving u_2 and y_2 using controller K_2:

Figure of partial control from the end goes in here!

IMPORTANT

- Closing a loop does not imply a loss of degrees of freedom (DOFs) (since the setpoint r_2 replaces u_2 as a DOF), **BUT** we usually "use up" some of the dynamic range.

Set $y_2 = r_2 - n_2$

$$y_1 = \underbrace{(G_{11} - G_{12}G_{22}^{-1}G_{21})}_{\triangleq P_u} u_1 + \underbrace{(G_{d1} - G_{12}G_{22}^{-1}G_{d2})}_{\triangleq P_d} d + \underbrace{G_{12}G_{22}^{-1}}_{\triangleq P_r}(r_2 - n_2)$$

Some criteria for selecting u_2 and y_2 in lower-layer:

1. Lower layer must quickly implement the setpoints from higher layers, i.e., controllability of subsystem u_2/ y_2 should be good. (G_{22})
2. Provide for local disturbance rejection. (partial disturbanve gain P_d should be small)
3. Impose no unnecessary control limitations on problem involving u_1 and/or r_2 to control y_1. (P_u or P_r)
   - Avoid negative RGA for pairing u_2/y_2 – otherwise P_u likely has RHP-zero

"Unnecessary": Limitations (RHP-zeros, ill-conditioning, etc.) not in original problem involving u and y

Partial control

	Meas./Control of y_1 ?	Control objective for y_2 ?
Sequential decentralized control	Yes	Yes
Sequential cascade control	Yes	No
"True" partial control	No	Yes
Indirect control	No	No

THEORY FOR CONTROL CONFIGURATIONS

Stabilization

Tool: Pole directions
Example: Tennesse Eastman challenge problem

Summary of procedure for plantwide control

The overall procedure consists of

I. Top-down analysis to identify degrees of freedom and control objectives

II. Bottom-up design of the contol structure

Iteration is required in this overall procedure!

Outline of a Procedure for plantwide control

1. Define overall control objectives
   - Stabilization
   - Product specifications
   - Minimize operating costs (J)
2. Selection of (additional) controlled outputs to minimize operating costs (steady-state analysis)
   - Typically pressures, temperatures, internal compositions etc.
   - No. of steady-state degrees of freedom $= N_u - N_I$
     N_u - number of manipulated inputs (valves)
     N_I - number of inventories with no steady-state effect
3. Identify the manipulated variables
   - Valves - N_u of these
4. Identify the most important disturbances
   - Important to know where **production rate** is set (at inlet or outlet or internally) – since production rate is generally a very important disturbance

5. Design a stabilizing control structure
   - Stabilize inventories with no steady-state effect (buffer tanks; levels in reboilers and condensers)
     - N_I of these
     - Their setpoints can be used dynamically but have no steady-state effect
   - "Stabilize" other parts of the process so that it can be operated manually
     - "True" instabilities, e.g. unstable reactor
     - Variables which are sensitive to disturbances (e.g. average temperature in distillation column) - look out for slowly integrating variables caused by recycle
     - Note: The setpoints of these variables have a steady-state effect (e.g. reactor volume) – so do **not** loose any steady-state degrees of freedom
   - May add extra measurements for dynamic reasons to improve the controllability
6. Design a control system for controlling the remaining variables
   - Product specifications
   - Additional variables from step **??**
7. Can the control structure be simplified? (Eliminate measurements and/or inputs)

CONTROLLABILITY ANALYSIS

Before attempting controller design one should analyze the plant:

- Is it a difficult control problem?
- Does there exist a controller that meets the specs?
- How should the process be changed to improve control?

QUALITATIVE RULES from Seborg et al. (1989)
(chapter on "The art of process control"):

1. *Control outputs that are not self-regulating*
2. *Control outputs that have favorable dynamic and static characteristics, i.e., there should exist an input with a significant, direct and rapid effect.*
3. *Select inputs that have large effects on the outputs.*
4. *Select inputs that rapidly effect the controlled variables*

- We have developed controllability tools which quantify these statements.
- Scale variables. Must then require

  1. Self-regulation: $|G_d| < 1$ at all frequencies
  2. Disturbance rejection: $\left|G_d\left(j\frac{1}{\theta}\right)\right| < 1$
  3. Disturbance rejection: $|G| > |G_d|$ at frequencies where $|G_d| > 1$

"PERFECT CONTROL" and plant inversion. (Morari, 1983)

$$y = G(s)\, u + G_d(s)\, d$$

Ideal feedforward control, $y = r$:

$$u = G^{-1}\, r - G^{-1} G_d\, d \tag{1}$$

Feedback control:

$$u = G^{-1} T\, r - g^{-1} T G_d\, d \tag{2}$$

For frequencies below the bandwidth $(\omega < \omega_B) : T \approx I$: Then **(??)** =**(??)**.

Controllability is limited if G^{-1} cannot be realized:

- Delay (Inverse yields prediction)
- Inverse response = RHP-zero (Inverse yields instability)
- Input constraints (Inverse yields saturation)
- Uncertainty (Inverse not correct)

POOR CONTROLLABILITY CAN BE CAUSED BY:

1. Delay or inverse response in $G(s)$
2. or $G(s)$ is of "high order" (tanks-in-series) so that we have an "apparent delay"
3. Constraints in the plant inputs (a potential problem if the plant gain is small)
4. Large disturbance effects (which require "fast control" and/or large plant inputs to counteract)
5. Instability: Feedback with the active use of plant inputs is required. May be unable to react sufficiently fast if there is an effective delay in the loop. And: May have problems with input saturation if there is measurement noise or disturbances
6. With feedback: Delay/inverse response or infrequent or lacking measurement of y. May try
   (a) Local feedback (cascade) based on another measurement, e.g. temperature
   (b) Estimation of y from other measurements
7. Nonlinearity or large variations in the operating point which make linear control difficult. May try
   (a) Local feedback (inner cascades)
   (b) Nonlinear transformations of the inputs or outputs, e.g. $\ln y$
   (c) Gain scheduling controllers (e.g. batch process)
   (d) Nonlinear controller
8. MIMO RHP-zeros: May have internal couplings resulting in multivariable RHP-zeros $\Rightarrow$ Fundamental problem in controlling some combination of outputs.
9. MIMO plant gain: May not be able to control all outputs independently (if the "worst case" plant gain $\underline{\sigma}(G)$ is small).
10. MIMO interactions: May have large RGA-elements (caused by strong two-way interactions between the outputs) which makes multivariable control difficult.
11. Feedforward control: Should be considered if feedback control is difficult (e.g. due to delays in the feedback loop or MIMO interactions) and an "early" measurement of the disturbance is possible.

There are tools available which quantify this.

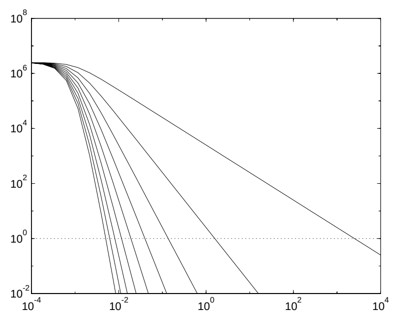
y = concentration of poduct (meas. delay θ=10 s)
$u = \text{Flow}_{base}$
$d = \text{Flow}_{acid}$

Introduce excess of acid $c = c_H - c_{OH}$ [mol/l].
In terms of c the dynamic model is a simple mixing process !!

$$\frac{d}{dt}(Vc) = q_A c_A + q_B c_B - qc$$

EXTREMELY SENSITIVE TO DISTURBANCES.

With n tanks: $G_d(s) = k_d/(1+\tau s)^n$.
τ: residence time in each tank.



To reject disturbance must require

$$|G_d(j\frac{1}{\theta})| < 1$$

where θ is the measurement delay. Gives

$$\tau > \theta\sqrt{(k_d)^{2/n} - 1}$$

Total volume : $V_{tot} = n\tau q$ where $q = 0.01$ m^3/s.

IMPROVE CONTROLLABILITY BY REDESIGN OF PROCESS

- Use several similar tanks in series with gradual adjustment

With $\theta = 10$ s the following designs have the same controllability:

No. of tanks n	Total volume V_{tot} $[m^3]$	Volume each tank $[m^3]$
1	250000	250000
2	316	158
3	40.7	13.6
4	15.9	3.98
5	9.51	1.90
6	6.96	1.16
7	5.70	0.81

Minimum total volume: 3.66 m^3 (18 tanks of 203 l each).
Economic optimum: 3 or 4 tanks.
Agrees with engineering rules.

PLANTWIDE DYNAMICS

- Poles are affected by recycle of energy and mass and by interconnections
- Parallel paths may give zeros - possible control problems
- Recycle yields positive feedback and often large *open-loop* time constants
- This does *not* necessarily mean that *closed-loop* must be slow
- See MYTH on distillation contol where open-loop time constant for compositions is long because of positive feedback from reflux and boilup
- Luyben's "snowball effect" is mostly a steady-state design problem (do not feed more than the system can handle...)

EXAMPLE: Recycle around reactor (snowball efect)

Simple example (Luyben, Yu):

- Reaction $A \rightarrow B$
- Recycle of unreacted A
- Product is pure B

At steady-state

$$\text{Feed of A} = \text{Generation of B in reactor} = \text{Production of B}$$

where Generation of B in reactor is

$$G_B = k(T)x_A V_R$$

$$G_B = k(T)x_A V_R$$

Three ways to increase G_B:

1. Increase reactor temperature T
2. Increase x_A by increasing the recycle ratio RR

$$x_A = \frac{RR}{1+RR}$$

   (the "snowball effect" of Luyben is that $x_A \rightarrow 1$ as $RR \rightarrow \infty$ – occurs when the reactor is too small)

3. Increase the reactor volume V_R
   - BUT: Loose money by not operating at maximum volume (Possible trade-off between operating costs and controllability)
   - Gas phase reactor: Increasing the pressure has the same effect (larger inventory in reactor).

DISTILLATION EXAMPLE

$$u = \begin{bmatrix} L \\ V \end{bmatrix}; \quad y = \begin{bmatrix} y_D \\ x_B \end{bmatrix} [\text{mol} - \%\ \text{light}]$$

Steady-state gains $y = Gu$ (LV-configuration)

$$G(0) = \begin{bmatrix} 87.8 & -86.4 \\ 108.2 & -109.6 \end{bmatrix}$$

RGA-value about 35 at steady-state $\Rightarrow$ Strong two-way interaction

OVERALL DISTILLATION PROBLEM

Typically, overall control problem has 5 inputs

$$u = [\, L \quad V \quad D \quad B \quad V_T \,]$$

(flows: reflux L, boilup V, distillate D, bottom flow B, overhead vapour V_T)
and 5 outputs

$$y = [\, y_D \quad x_B \quad M_D \quad M_B \quad p \,]$$

(compositions and inventories: top composition y_D, bottom composition x_B, condenser holdup M_D, reboiler holdup M_B, pressure p)

Without any control we have a 5×5 model

$$y = Gu + G_d d$$

(which generally has some large RGA-elements at steady-state)

Analyze G^{u_1} and $G_d^{u_1}$ with respect to

1. No composition control
   - Consider disturbance gain $G_d^{u_1}$ (e.g. effect of feedrate on compositions)
2. Close one composition loop ("one-point control")
   - Consider partial disturbance gain (e.g. effect of feedrate on y_D with constant x_B)
3. Close two composition loops ("two-point control")
   - Consider interactions in terms of RGA
   - Consider "closed-loop disturbance gains" (CLDG) for single-loop control

Problem:

- No single best configuration
- Generally, get different conclusion on each of the three cases
- $\Rightarrow$ Stabilizing control is not necessarily a trivial issue

DISTILLATION CONFIGURATIONS

There are usually three "unstable" outputs with no or little steady-state effect

$$y_2 = [\, M_D \quad M_B \quad p \,]$$

Remaining outputs

$$y_1 = [\, y_D \quad x_B \,]$$

Many possible choices for the three inputs for stabilization. For example, with

$$u_2 = [\, D \quad B \quad V_T \,]$$

we get the LV-configuration where

$$u_1 = [\, L \quad V \,]$$

are left for composition control.
Another configuration is the DV-configuration (has small RGA-elements) where

$$u_1 = [\, D \quad V \,]$$

After closing the stabilizing loops ($u_2 \leftrightarrow y_2$) we get a 2×2 model for the remaining "partially controlled" system

$$y_1 = G^{u_1} u_1 + G_d^{u_1} d$$

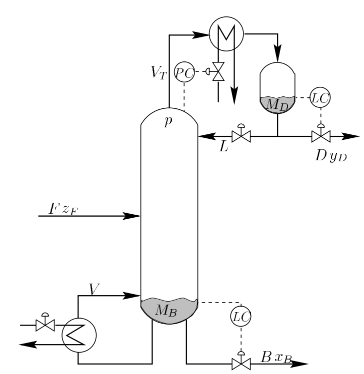
Which configurations is the best?

A PARADOX

Distillation columns have large RGA-elements
$\Rightarrow$ Fundamental control problems (cannot have decoupling control)

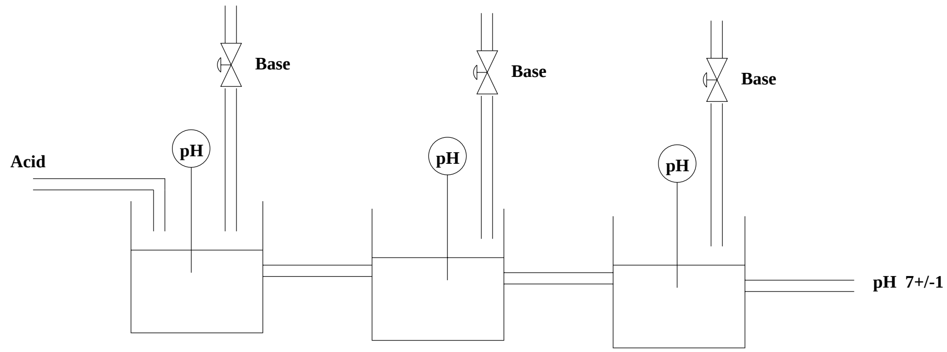
BUT: DV configuration has small RGA-elements and we can decouple the compositions loops
How is this possible?

Solution to paradox: DV configuration has coupling between composition and level loops
(whereas LV has decoupling between level and composition)

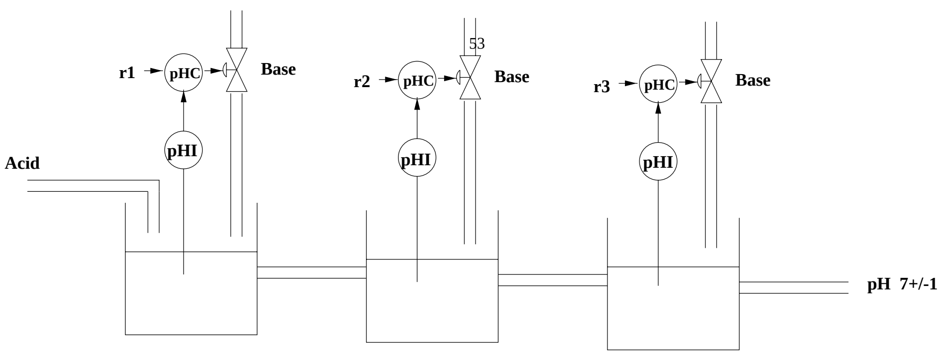


V_T
PC
p
LC
M_D
L
$D\,y_D$
$F\,z_F$
V
M_B
LC
$B\,x_B$

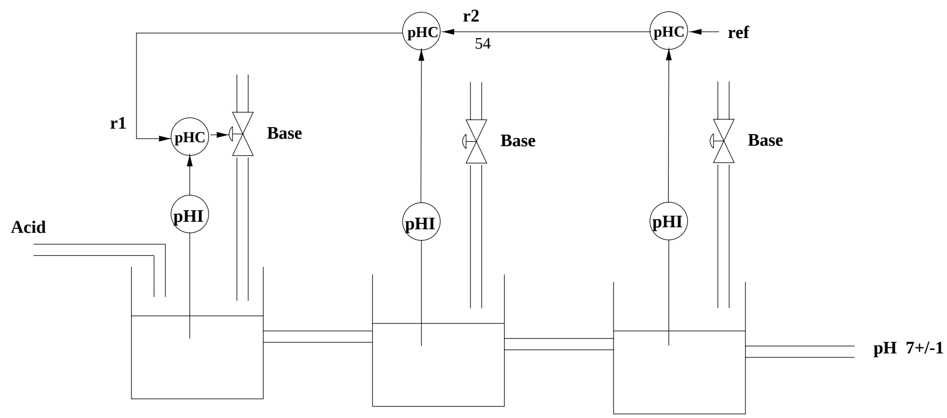
Inputs	Measurements	No of combinations	No of combinations pH in last tank used
3	3	1	1
3	2	3	2
3	1	3	1
2	3	3	3
2	2	9	6
2	1	9	3
1	3	3	3
1	2	9	6
1	1	9	3
Total		**49**	**28**



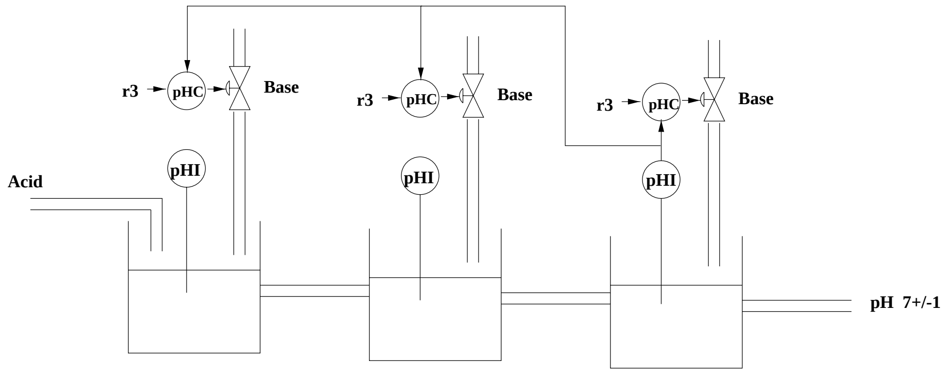
Local control



Cascade control

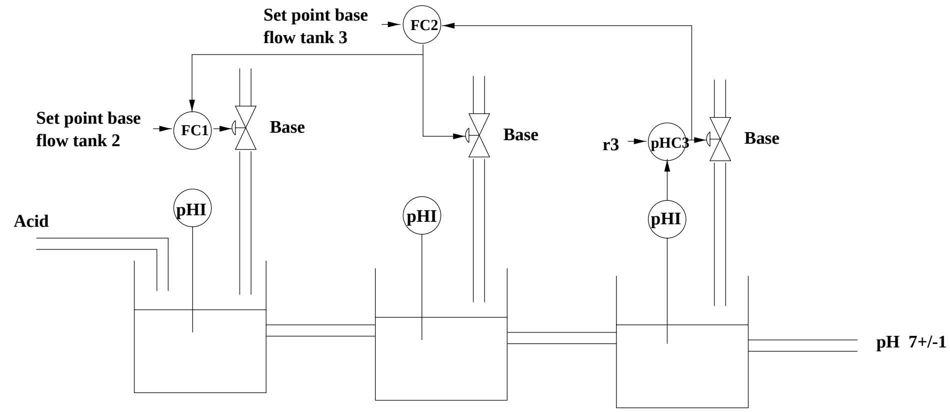


Extra inputs, parallel control



EXAMPLE: Petlyuk distillation (extra degrees of freedom)

Extra inputs, cascade control



CONCLUSIONS / FUTURE WORK

1. Want to make a procedure which applies generally (not only process control)
2. Many theoretical tools are already there – still some effort left to get a unifying approach
3. Want to avoid "case study approach" (but the case studies are useful for understanding the issues)
4. Hope to make good progress in near future