# Chapter 10

# Tuning of PID controllers

## 10.1 Introduction

No exercises here.

## 10.2 The Good Gain method

### Exercise 10.1

Figure 10.1 shows the response in the temperature of a simulated temperature control system with P controller with the following “good gain” value:

$$K_c = 4.0 \tag{10.1}$$

Tune a PI controller for this process using the Good Gain method.

What can you do with the controller tuning if turns out that the stability of the control system is too bad with this value of $K_p$?

## 10.3 Skogestad’s PID tuning method

### Exercise 10.2

Given a process where the relation between the control signal and the process measurement can be well represented as “time-constant with time-delay”:

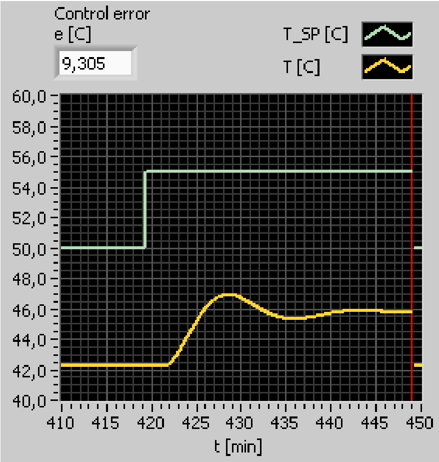


Figure 10.1:

Gain:

$$K = 0.5 \tag{10.2}$$

Time-constant:

$$T = 5 \text{ s} \tag{10.3}$$

Time-delay:

$$\tau = 1 \text{ s} \tag{10.4}$$

Tune a PI(D) controller for this process using Skogestad's method.

**Exercise 10.3**

Figure 10.2 shows a level control system for a wood-chip tank with feed screw and conveyor belt which runs with constant speed. (This system is described in Section 7.2 of the text-book, however the measurement filter is omitted in the present example, assuming it has negligible effect on the dynamics of the control loop.)Based on mass balance of the wood-chip of the tank, a mathematical model is

$$\rho A\dot{y}(t) = K_s u(t-\tau) - w_{out}(t) \tag{10.5}$$

where $\rho$ [kg/m$^3$] is chip density, $A$ [m$^2$] is cross-sectional area, $K_s$ [(kg/min)/%] is feed screw gain, $\tau$ [min] is time-delay of the conveyor belt. The level $y$ is measured with a level sensor, and the measurement is

$$y_m(t) = K_m y(t) \tag{10.6}$$

where $K_m$ [%/m] is the sensor gain.

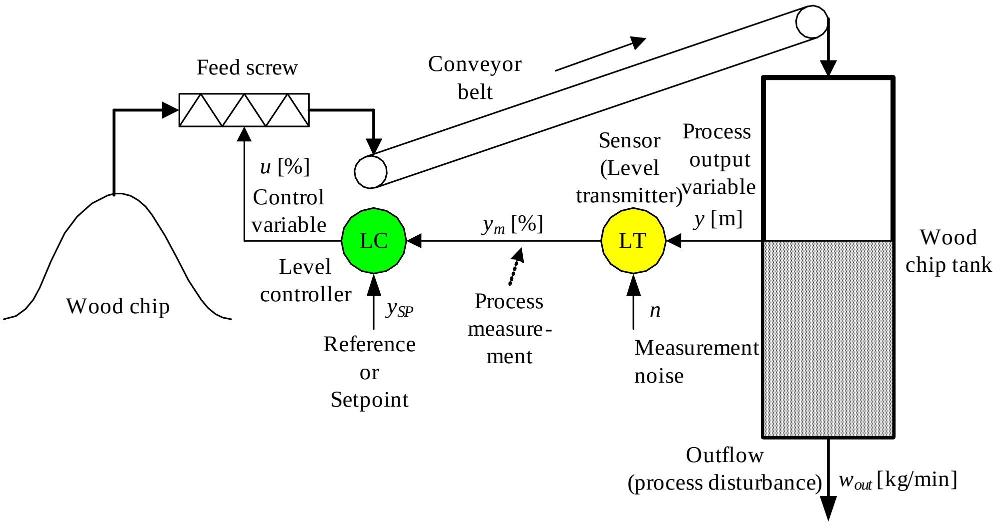


Figure 10.2:

1. Calculate the transfer function $H(s)$ from control variable $u$ to process measurement $y_m$.
2. Calculate the PI(D) parameters for the process using Skogestad's method.

### Exercise 10.4

Figure 10.3 shows the response in the filtered process measurement $y_{m_f}$ due to a step of amplitude $U = 2$ in the control signal $u$ (the step comes at time $t = 0$). Calculate PID settings for this process.

## 10.4 Auto-tuning

### Exercise 10.5

1. In general, if you in a given application are to select between open loop excitation and closed loop excitation, which one would you select?
2. Are there any processes for which open loop excitation should not be used?

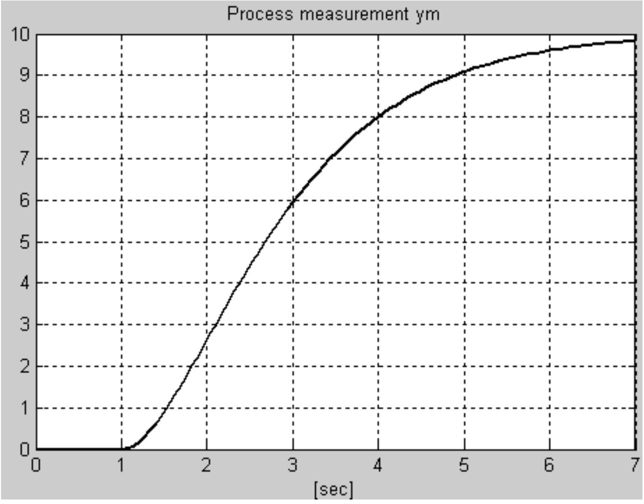


Figure 10.3:

## 10.5 PID tuning when process dynamics varies

**Exercise 10.6**

Assume that the process to be controlled has varying process dynamics, which may cause stability problems or sluggish control. Both of the solutions A and B below are possible. Which is the best one with respect to control performance, and which is the simplest one?

A: The controller is tunes the the most critical operating point, and the controller parameters are then kept constant.

B: The controller parameters are adjusted continually so that they fit to the dynamic properties of the process at any operating point.

**Exercise 10.7**

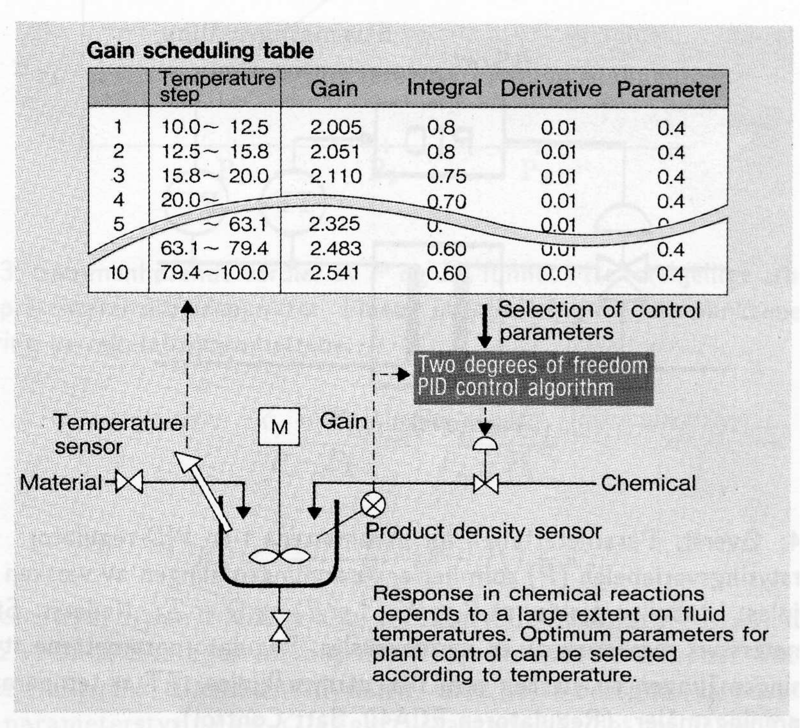
Assume that you in a given control system for a “time-constant with time-delay” process have found proper PI parameters in one specific operating point. Assume that the process gain increases.

1. How would this process gain increase influence the stability of the control system?

2. Derive formulas for the new controller parameters. You can indicate the initial values of the controller parameters and the process parameters (before the change) with index 0, and new values (after the changes) with index 1.

**Exercise 10.8**

Figure 10.4 shows a chemical reactor and a PID parameter table which is the basis of a PID controller with gain scheduling. Assume that gain scheduling is not to be used, but fixed PID settings in stead. Should the controller be tuned at high temperature or at low temperature, given that it is crucial that the stability of the control system is satisfactory at any temperature?



**Gain scheduling table**

|  | Temperature step | Gain | Integral | Derivative | Parameter |
|---|---|---|---|---|---|
| 1 | 10.0~ 12.5 | 2.005 | 0.8 | 0.01 | 0.4 |
| 2 | 12.5~ 15.8 | 2.051 | 0.8 | 0.01 | 0.4 |
| 3 | 15.8~ 20.0 | 2.110 | 0.75 | 0.01 | 0.4 |
| 4 | 20.0~ |  | 0.70 | 0.01 | 0.4 |
| 5 | ~ 63.1 | 2.325 | 0. | 0.01 |  |
|  | 63.1 ~ 79.4 | 2.483 | 0.60 | 0.01 | 0.4 |
| 10 | 79.4~100.0 | 2.541 | 0.60 | 0.01 | 0.4 |

Response in chemical reactions depends to a large extent on fluid temperatures. Optimum parameters for plant control can be selected according to temperature.

Figure 10.4:

**Exercise 10.9**

Table 10.1 shows parts of a gain scheduling based PID controller.

| $GS$ | $K_p$ | $T_i$ | $T_d$ |
|---|---|---|---|
| $\vdots$ | $\vdots$ | $\vdots$ | $\vdots$ |
| 20% | 0.4 | 5.2 | 1.3 |
| 30% | 0.5 | 4.5 | 1.6 |
| $\vdots$ | $\vdots$ | $\vdots$ | $\vdots$ |

Table 10.1: PID Gain Schedule

Find $K_p$ as a function of the gain scheduling variable $GS$ between the operating points shown in the table. The function should be based on linear interpolation.

**Exercise 10.10**

Figure 10.5 shows a process with a PID control system where the actuator is represented with a nonlinear relation between the control signal $u$ and the internal process variable $z$:

$$z = f(u) \tag{10.7}$$

For example, the actuator can be a control valve with some nonlinear relation between the control signal ($u$) and the flow ($z$). In most cases it is benefical if the PID controller sees a *linear* process – not a nonlinear process, because this makes the controller tuning easier, and the dynamic properties of the control system may be independent of the operating point. This can be achieved by including the *inverse* of the nonlinear function in the controller:

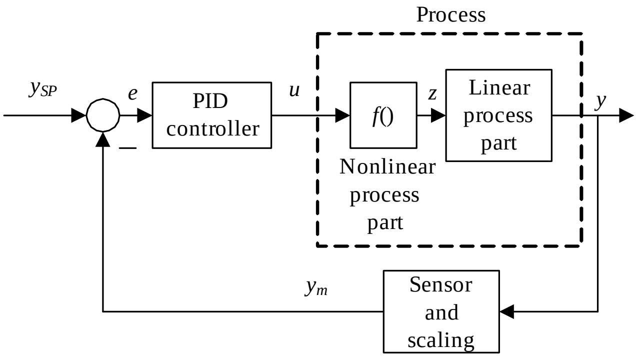


Figure 10.5:

$$u = f^{-1}(z) \tag{10.8}$$

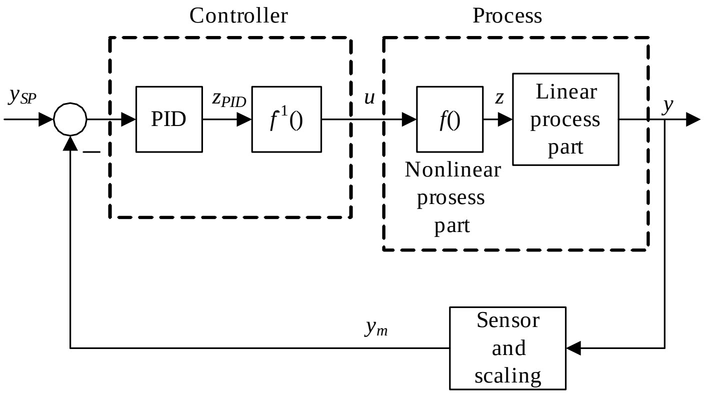


Figure 10.6:

The $z$-value that the PID controller demands can be denoted $z_{PID}$. See Figure 10.6.

Assume that the nonlinear function $z = f(u)$ can be represented with $n$ tabular data points (which can stem from a data sheet or from experiments):

| $z$ | $u$ |
|---|---|
| $z_1$ | $u_1$ |
| $z_2$ | $u_2$ |
| $\vdots$ | $\vdots$ |
| $z_n$ | $u_n$ |

Explain how you can implement the inverse function using *table-lookup*. Table-lookup functions implements linear interpolation between the data points in the table.[1]

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[1] Table lookup functions are available in computer tools as MATLAB and LabVIEW.

# Chapter 11

# Various control methods and control structures

## 11.1 Cascade control

**Exercise 11.1**

In the neutralization section of a fertilizer production plant, intermediate mother liquor flows into and out of a tank. In the tank the pH value of the liquid is controlled by adjusting the inflow of ammonia gas to the tank. The ammonia flow is flow controlled using a control valve.

Draw an instrumentation diagram of this process section. (You can use Q (for quality) as symbol for pH.) What can be the purpose of the ammonia flow control loop?

**Exercise 11.2**

Figure 11.1 shows a control valve being used to manipulate the flow of a heating medium (liquid) into a heat exchanger where the temperature is to be controlled. The output of the temperature controller is flow command signal (flow setpoint) to the valve, and the output of the flow controller is a valve stem position command (position setpoint) to the stem moving mechanism.

Draw an instrumentation diagram of the total control system. You can use symbol G for position of the valve stem. (Hint: There are three control loops.)

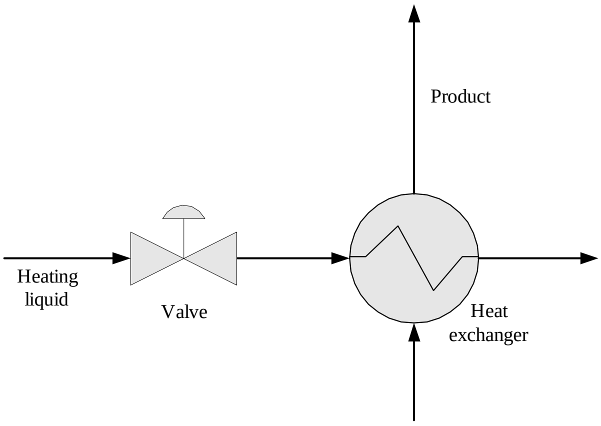

Figure 11.1:

### Exercise 11.3

Figure 11.2 shows a ship. The position of the ship is controlled. Assume that it is benefical for the positional control system that the rotational speed of the propeller is controlled.

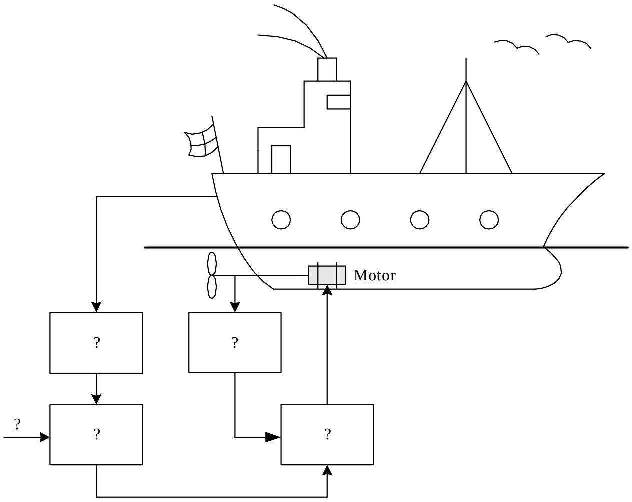

Figure 11.2:

Based on the given information, substitute the question marks with proper functions (text). What are the purposes of the control loops?

## 11.2 Ratio control

### Exercise 11.4

Figure 11.3 shows a tank with two inlet flows. The liquid level of the tank is to be controlled by manipulating (controlling) flow A. It is assumed that flow A is much larger than flow B. The ratio between flow B and A is specified as

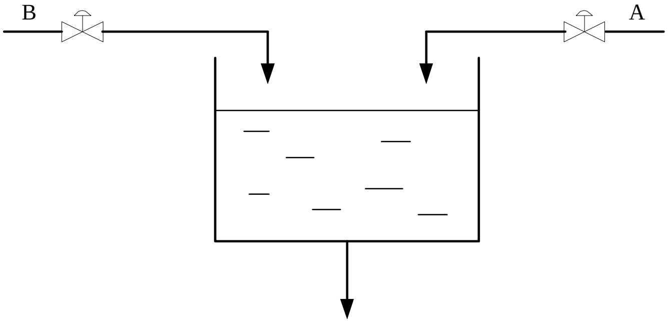

Figure 11.3:

$$\frac{F_B}{F_A} = k \tag{11.1}$$

where $k$ is a given ratio. Assume that it is necessary to have local flow control loops around each valve.

Draw a Process & Instrumentation diagram of a control system for this process.

## 11.3 Split-range control

### Exercise 11.5

Figure 11.4 shows a liquid tank where the pH value of the liquid is to be controlled with split-range control where acid flow and base flow are adjusted. Both the acid flow and the base flow are controlled with (local) flow control loops. Draw an instrumentation diagram of the tank with control system.

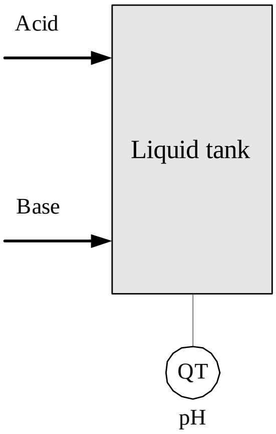

Figure 11.4:

## 11.4 Flow smoothing with sluggish level control

### Exercise 11.6

See Figure 11.13 in the text-book. Assume that the process model is given by Eq. (11.4) in the text-book. The model is repeated here:

$$\rho A\dot{h} = F_{in} - F_{out} \tag{11.2}$$

Assume that the level is controller with a P controller:

$$F_{out} = K_p(h_{SP} - h) \tag{11.3}$$

where $K_p$ is the controller gain. (The value of $K_p$ will be negative because the controller must have direct action mode.)

1. Express the level $h$ as a function of the setpoint $h_{SP}$ and the inflow $F_{in}$ under static conditions. (Hint: At static conditions the time-derivative is zero.) You can use index s for "static".
2. Assume that for a given flow, $F_{in_0}$, the level are allowed to depart from the setpoint by $\Delta h_s$. (Thus, $\Delta h_s$ is the level control error, $e_s$.) Calculate the corresponding controller gain.

   Is $\Delta h_s$ reduced or increased if $K_p$ is increased?
3. What is the drawback, regarding the static control error, of using a P controller for level control of the tank?

## 11.5 Plantwide control

### Exercise 11.7

Figure 11.5 shows a wood-chip tank, which is in the beginning of the pulp & paper production line. Spruce, pine and eucalyptus are used as feeds into the tank, via a conveyor belt. The percentages of each of these flows are indicated in Figure 11.5. There is a flow disturbance before the belt which is due to sieving the chip flow to remove large parts of chip.

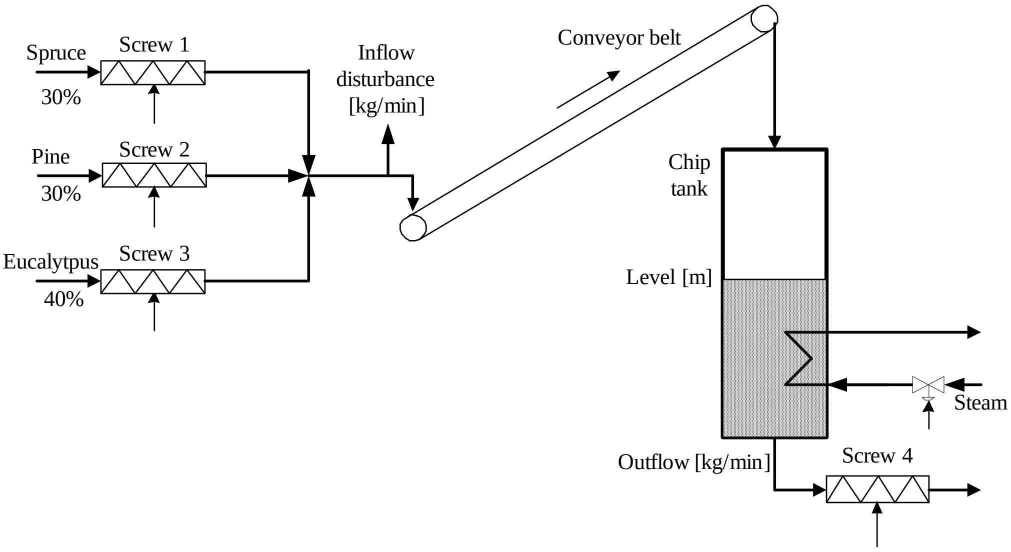

Figure 11.5:

Draw a P&I (Process & Instrumentation) diagram of a control system for this part of the production line according to the following specifications:

- The production rate is controlled to a setpoint with flow control of Screw 4.
- The level of chip in the tank is controlled to a setpoint by manipulating the total inflow to the conveyor belt.
- The total chip flow into the belt is splitted into percentage flows shown in Figure 11.5. The splitting can be represented with a block with total control signal (100%) as input and three flow value outputs (30%, 30% and 40% respectively). The flows out of the three inflow screws are flow controlled.

- A flow control loop is used to compensate for the flow disturbance due to the sieveing. This flow loop is based on the measurement of the flow with a flow sensor at the beginning of the belt (cf. Figure 11.7 in the text-book).
- The temperature of the chip in the tank is controlled to a setpoint using the steam valve.

### Exercise 11.8

Figure 11.6 shows an incomplete P&I (Process & Instrumentation) diagram of a controlled distillation column.

If you need it, here is some basic information about distillations columns: A distillation column contains a number of trays from where liquid can pour downwards (to the next tray) and vapour can rise upwards (to the next tray). The purpose of the distillation column is to separate the "light" component and the "heavy" component by exploiting their different boiling points of temperature. Heat is supplied to the boiler at the bottom of the column. Vapour leaving the column is condensed in the condenser. The liquid leaving the condenser is accumulated or stored in the accumulator. Part of the liquid leaving the accumulator is directed back to the column, and the rest – the distillation product – is directed to e.g. a storage tank. Ideally, the concentration of "heavy" component in the top product is zero, and the concentration of the "light" component in the bottoms product is zero. In principle this can be achieved by one quality control loop for the top product and one quality control loop for the bottoms product, but due to the dynamic properties of distrillation columns such "two-point" control is difficult to realize. Therefore, there is typically either quality control of the top product or quality control of the bottoms product.

Make the diagram shown in Figure 11.6 complete by entering letter codes in the instrumentation symbols according to these specifications: The quality of the distillate product is controlled, and there is mass balance control of various parts of the column. (The heating medium supplied to the boiler is manually controlled, so it is not adjusted by an automatic controller.)