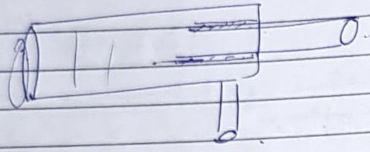
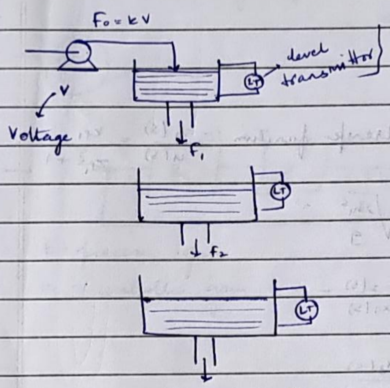


Madhuri

PDC



$$\frac{dh_1}{dt} = \frac{F_0}{A_1} - \frac{a_1}{A_1}\sqrt{2gh_1}$$

$$\frac{dh_2}{dt} = \frac{a_1}{A_1}\sqrt{2gh_1} - \frac{a_2}{A_2}\sqrt{2gh_2}$$

$$\frac{dh_3}{dt} = \frac{a_2}{A_2}\sqrt{2gh_2} - \frac{a_3}{A_3}\sqrt{2gh_3}$$

input = V

Output = level of tank 3.

a_1, a_2, a_3 → area of outflow pipe

A_1, A_2, A_3 → area of tank

→ PID controller $\quad c = c_b + k_c e + \frac{k_c}{\tau_I}\int e\,dt + k_c \tau_D \frac{de}{dt}$

$e = 0$ → at ss → can be used to obtain c_b ($c = c_b$) → controller bias

$e = SP - PV$ (SP → set point, PV → process variable)

→ Tuning of PID controller → obtain k_c, τ_I, τ_D → using transfer function model find k_c, τ_I, τ_D

Dynamic model → Transfer function model → find k_c, τ_I, τ_D

→ current to voltage → use resistor

→ laplace transform of non-linear equation is not possible

Transfer function model

Define: $x_1 = h_1 - h_1^s$; $x_2 = h_2 - h_2^s$, $x_3 = h_3 - h_3^s$, $u = V - V^s$

(deviation variable) ⇒ ss values are desired operating conditions

→ linearize using taylor series approx. about ss value :-

$$\frac{dh_1}{dt} = \frac{k}{A_1}V - \frac{a_1}{A_1}\sqrt{2g}\left[\sqrt{h_1^s} + \frac{1}{2\sqrt{h_1^s}}(h_1 - h_1^s)\right] \quad (1)$$

SS :- $\frac{k}{A_1} v^s - \frac{a_1\sqrt{2gh_1^s}}{A_1} = 0 \to ②$

$= \frac{dh_1^s}{dt}$

① − ②

$\frac{dx_1}{dt} = \frac{k}{A_1} u - \frac{a_1\sqrt{g}}{A_1\sqrt{2h_1^s}} x_1 \Rightarrow$ transfer function :- $\frac{x_1(s)}{u(s)} = \frac{k_{p_1}}{\tau_{p_1} s + 1}$

$k_{p_1} = \frac{k}{a_1}\sqrt{\frac{2h_1^s}{g}}$ $\tau_{p_1} = \frac{A_1}{a_1}\sqrt{\frac{2h_1^s}{g}}$

$\frac{dx_2}{dt} = ---$ $\frac{x_2(s)}{x_1(s)} =$

$\frac{dx_3}{dt} = ---$ $\frac{x_3(s)}{x_2(s)} =$

Process transfer function $= G(s) = \frac{x_3(s)}{u(s)} = \frac{k_{p_1} k_{p_2} k_{p_3}}{(\tau_{p_1}s+1)(\tau_{p_2}s+1)(\tau_{p_3}s+1)}$

Servo, regulatory → ≡

DOF ⇒ ✓

→ Quadruple tank system

$\frac{dh_1}{dt}$

$\frac{dh_2}{dt}$

$\frac{dh_3}{dt}$

$\frac{dh_4}{dt}$

No: of inputs = No: of outputs
↳ DOF = 0

Jacketed CSTR
output = T
MV = F_j

CSTR
A → B → C
A → 2D
Van-de-Vusse reaction

C_B = output
F = Input
plot to know what is the desired value C_B v/s F

desired product = B

Input = $\frac{F}{V} = \tau$ → residence time (this will change the conc)

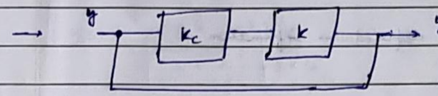
{Donot require C_C, C_D}

eqn ③ & ④ are not required (can be neglected)

Depending on the control objective, we can reduce the no. of equations

Input multiplicity system → same output can be achieved for multiple different inputs

⇒ gain



If k = +ve, k_c = ?
If k changes from +ve to −ve, k_c should change accordingly

k = process gain
k_c = controller gain

→ MATLAB

sfunctions [S-function takes the differential equation in the form of state space model]

>>[edit sfuntmpl]

$\dot{x} = f(x,u)$ → state equation

$y = h(x,u)$ → output map

→ 3 tank system

$\frac{d}{dt}\begin{bmatrix} h_1 \\ h_2 \\ h_3 \end{bmatrix} =$

$u = F_0$ | $x = \begin{bmatrix} h_1 \\ h_2 \\ h_3 \end{bmatrix}$

$\dot{x}$

→ Control relevant state-space model

$\dot{x} = f(x) + g(x)\cdot u$ → non-linear S-S model

$\dot{x} = AX + BU$ → linear S-S model

$Y = CX + DU$.

→ No: of continuous states = ③ → no: of continuous equations.
h_1, h_2, h_3

No: of discrete states = 0

No: of inputs = 1

No: of outputs = 1 (h_3)
↳ we want to measure the level of 3rd tank only.
or 3 (h_1, h_2, h_3) → to get the info about all 3.

Directfeedthrough → while calculating the output whether input is involved or not
If input is involved → then enter 1 else 0

⇒ x_0 = [ ]; → initial conditions for the states

classmate Date Page

→ Code

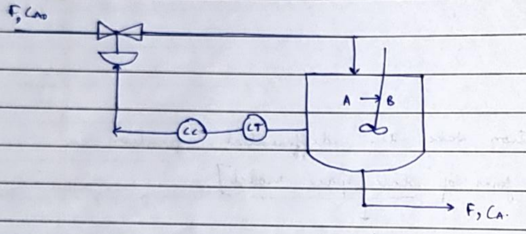
sys = [der1; der2]; → ⊗

→ Simulink

↓

library browser → simulink → user defined → s-function

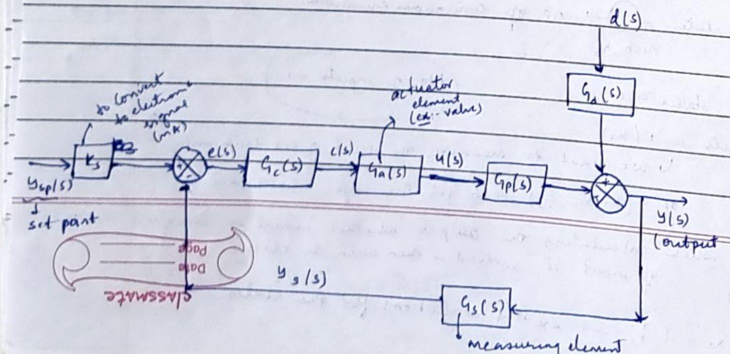
Reactor control



Designing control system

1. select control architecture
   - Feedback (initial)
2. find transfer function model of process
   - (a) using model equation
   - (b) Using process data
3. select controller algorithm
   - (a) PID controller
   - (b) Other types.

Feedback control loop



Reactor: $\frac{dC_A}{dt} = \frac{F}{V}(C_{A0} - C_A) - kC_A^2$

V → const   F → manipulative variable (u(s))

linearize and take laplace.   left over input → disturbance (d(s))

$$y(s) = \frac{(C_{A0}^s - C_A^s)^1/_V}{s + \frac{F^s}{V} + 2kC_A^s} u(s) + \frac{F^s/V}{s + \frac{F^s}{V} + 2kC_A^s} d(s)$$

$y = C_A - C_A^s$ $u = F - F^s$ $d = C_{A0} - C_{A0}^s$

Control valve :- (actuator) $u(s) = \frac{K_a}{\tau_a s + 1} c(s)$

[Generally pneumatic control valve]

→ air to close control valve

→ air to open

- Normally first order (specified by vendor)
- Time const specified by vendor
- Gain (calculated from installed characteristics)

Types of Globe valves

Quick opening

Equal percentage

linear

⇒ valve eqn :- $Q = C_v \underbrace{f(x)}_{} \sqrt{\frac{\Delta P}{s.g.}}$

this will decide which type of valve it is

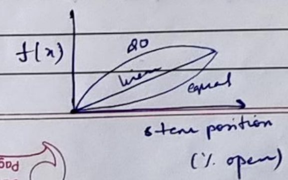
$f(x) = x$ — linear

$f(x) = \sqrt{x}$ — Quick opening

$f(x) = R^{x-1}$ — equal %

R → rangibility of valve

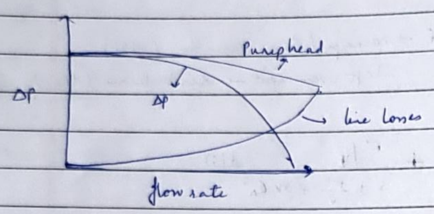
⇒ Inherent control valve.



(pressure drop across the valve is const)

But if flow changes ΔP changes.

ΔP v/s Q

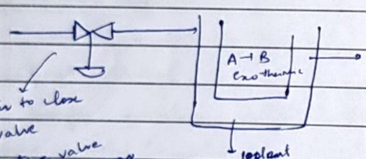


if ΔP changes, valve characteristics change

→ Installed flow characteristics

→ When to use air to open valve or air to close valve?

- depends on the safety of the process



air to close valve

∴ If the valve fails, the flow will still continue

But if we use air to open valve, No flow will occur & explosion might take place (exothermic)

Sensor :- $y_s(s) = \dfrac{k_s e^{-\alpha s}}{\tau_s s + 1}$

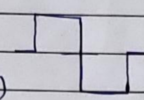
→ Gain is calculated from range setting

→ Time const & measurement delay is specified by vendor

→ Obtain process data using specified signal

- step function
- Pulse function
- Sinusoidal function (rarely used)
- Pseudo random Binary sequence (PRBS)

Doublet
(2 pulses opp)



→ Process data includes the actuator and sensor response

→ Evaluate transfer function model from process.

$y_s(s) = g_p(s)\, c(s) + g_d(s)\, d(s)$

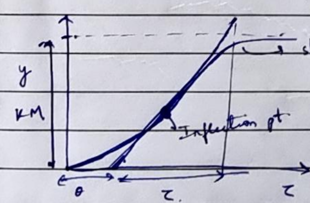
→ 1st order ~~process~~ with dead time (FOPDT)

$$g(s) = \frac{k\, e^{-\theta s}}{\tau s + 1}$$

→ 4 classical methods (using step response data)

1. Zeigler Nichols method
2. Smith's
3. Sundaresan & Krishnaswamy
4. Nishikawa's

→ Z - N method



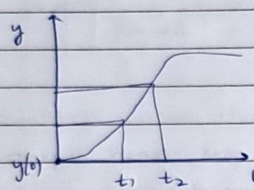
Draw a tangent at the inflection pt

→ Difficult to find inflection point & to draw tangent

Smith's method

$K = \dfrac{\Delta \text{output}}{\Delta \text{input}}$ $\quad \tau = \dfrac{3}{2}(t_2 - t_1)$

$\theta = t_2 - \tau$



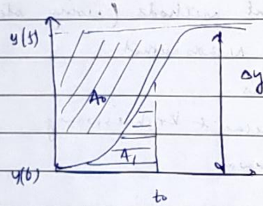
Sundaresan & krishnaswamy

$k = \dfrac{\Delta \text{output}}{\Delta \text{input}}$ $\quad \tau = \dfrac{2}{3}(t_2 - t_1)$ $\quad \theta = 1.3t_1 - 0.29t_2$

Nishikawa's

$A_0 = \int_0^\infty (\Delta y(\infty) - \Delta y(t))\,dt$

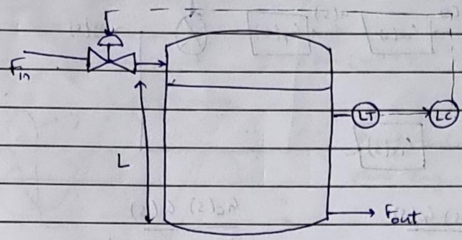
$A_1 = \int_0^{t_0} \Delta y(t)\,dt$, $\quad t_0 = \dfrac{A_0}{\Delta y(\infty)}$



$\tau = \dfrac{A_1}{0.368\,\Delta y(\infty)}$ $\quad \theta = t_0 - \tau$ $\quad K = \dfrac{\Delta y(\infty)}{\Delta c(\infty)}$

PID Controller

Direct / Reverse acting controller



Input ↑, output ↑ → sign of gain = +
level

sign of gain of air to open valve = + ve

product of these 2 = + ve

∴ use reverse control element

Process variable ↑, controller output ↑ → direct acting controller → controller gain = K_c = -ve

PV ↑, controller output ↓ → reverse acting → sign of controller gain = +ve.

$e = SP - PV$

set point

$$\left[\frac{dC_{out}}{dt} = K_c \frac{de}{dt}\right]$$

for direct acting PV ↑, e = -ve.

but $\dfrac{dC_{out}}{dt}$ = +ve $\quad \therefore K_c$ = -ve

→ fail close / fail open valve

→ air to open / air to close

C_{out} ↑ valve will open more

direct acting final control element

(~~[illegible]~~ / ~~acting~~)

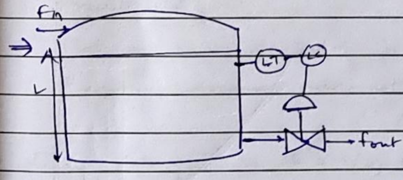
So whatever action controller is taking, valve will be in same direction.

(if the level ↑ the Cout ↓ (close the valve)

(gain = +ve)

(controller ↓ reverse acting controller

Guidelines of selecting a controller



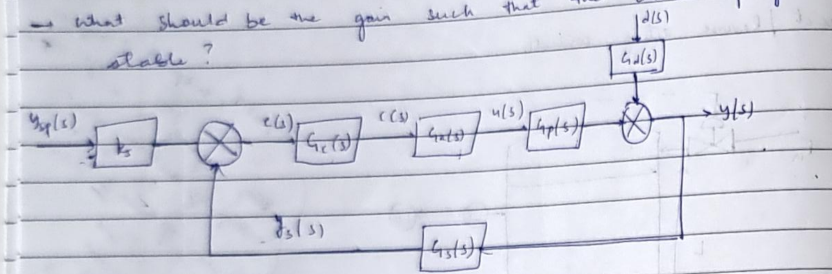
Process gain = -ve → Fout ↑ h ↓.

If we use air to open valve → gain = +ve

product = -ve

∴ use direct acting controller

→ What should be the gain such that the closed loop system is stable?



$$\frac{y(s)}{y_{sp}(s)} = \frac{k_s\, G_p(s)\, G_a(s)\, G_c(s)}{1 + G_p(s)\, G_a(s) * G_c(s)\, G_s(s)} = \frac{G_c(s)\, G(s)}{1 + G_c(s)G(s)}$$

→ characteristic equation

$1 + G_c(s)G(s) = 0$. find the poles.

controller gain $= k_c$ (proportional controller)

Proportional controller

Integral control

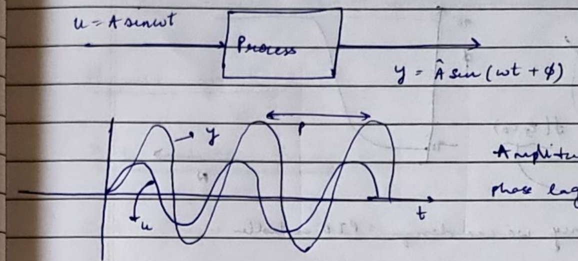
PI control

closed loop stability

BIBO → bounded input bounded output

→ Stability analysis

Frequency Response



(AR)
Amplitude ratio $= \hat{A}/A$.
phase lag $= -\Delta t/p \times 360$

$s = j\omega$

$G(s) \to s = j\omega$

$G(j\omega) = Re(\omega) + j(Im(\omega)) = e(\omega)\, e^{j\phi(\omega)}$

$e(\omega) = \sqrt{(Re(\omega))^2 + Im^2(\omega)} = |G(\omega)| = AR$

$\phi =$ phase difference $= \arctan\left(\frac{Im(\omega)}{Re(\omega)}\right) = \angle G(j\omega)$

→ Two ways of representing the frequency response.

- Bode plot

  Log Log AR v/s ω, semi-log φ v/s ω

- Nyquist plot

  Plot of $Re(\omega)$ v/s $Im(\omega)$ for various values of ω.

  Frequency response does not appear explicitly in this plot

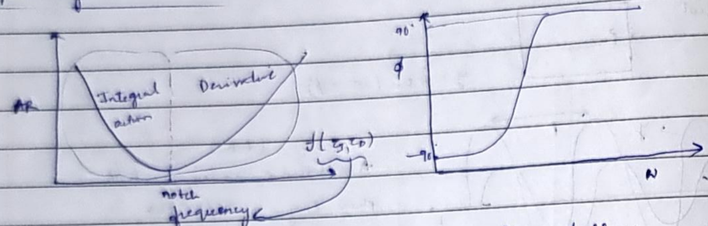
→ steady state gain ⇒ $k = e(0)\, e^{j\phi(0)}$ → Bode plot

wrt stabilization of process & is faster

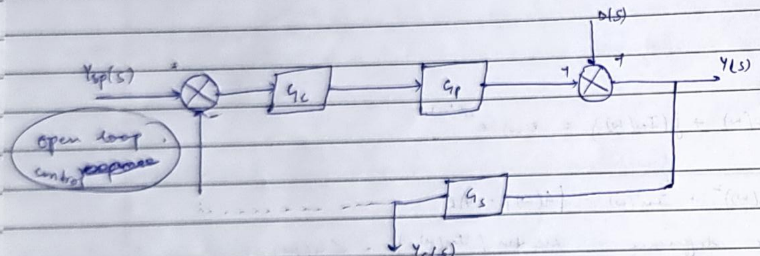
→ Nyquist Diagram

Bode stability

Bode plot of PID controller



by placing the notch frequency we can design PID controller.



(open loop response to $Y_{sp}(s)$).

→ AR < 1 → at some k_c value

→ if $Y_{sp}(s) = 0$ and complete the loop.

Input = $A \sin \omega t$ $\quad$ AR = 1 ← $k_c \uparrow$

Output = $A \sin(\omega t + \phi)$ $\quad$ Signal will not decay.

($\phi = -180$ → Signal will remain same in the loop) ✓ $A \sin \omega t$

→ ↑ k_c further, AR > 1 → amplitude will go on increasing
↳ the system will become unstable.

→ ↓ k_c → AR < 1 → amplitude ↓ → oscillations will die → System is stable

→ System is on the verge of instability when $\phi = -180°$, AR = 1

Bode stability criteria

k_{cu} = ultimate gain → ~~proportional controller~~

Nyquist stability

PID controller tuning

$\tau_I \uparrow$ → high Integral action → means more oscillations

What value of k_c & τ_I should be selected for good performance?

→ Optimization methods use Integral error.

Direct synthesis method

$G_c = ?$ → unknown

$Y/Y_{sp} = ?$

$\frac{Y}{Y_{sp}} = 1$ → ideally → $G_c \to \infty$.

$$\Rightarrow G_c = \frac{Y/Y_{sp}}{G(1 - Y/Y_{sp})} \qquad \left(\therefore \frac{Y}{Y_{sp}} \neq 1 \to \text{never}\right)$$

(1) ⇒ $\frac{Y}{Y_{sp}}$ can be considered a 1st order process → $\left(\frac{Y}{Y_{sp}}\right)_d = \frac{1}{\tau_c s + 1}$
↳ valid for any process without time delay

$\tilde{G}$ → process transfer function

$\tilde{G} = k \qquad \left(\frac{Y}{Y_{sp}}\right)_d = \frac{1}{\tau_c s + 1}$

$$G_c = \frac{\frac{1}{\tau_c s + 1}}{k\left(1 - \frac{1}{\tau_c s + 1}\right)} \qquad G_c = \frac{1}{k \tau_c s}$$

$G_c = \frac{K_I}{s}$ → Integral Controller

$$\tilde{g} = \frac{k}{(\tau_1 s+1)(\tau_2 s+1)} \longrightarrow G_c = \longrightarrow \text{PID}$$

$$G_c = \frac{\tau_1+\tau_2}{k\tau_c}\left(\frac{\tau_1\tau_2 s^2 + (\tau_1+\tau_2)s + 1}{(\tau_1+\tau_2)s}\right)$$

τ_c = tuning parameter

$\tau_c > \tau_T$

$\tau_c > \theta$

$$\Rightarrow \tilde{G}(s) = \frac{k_p e^{-\theta_p s}}{\tau_p s+1}$$

$$\left(1+\frac{1}{\tau_I s}+\tau_D s\right) \rightarrow \text{PID}$$

$$\frac{y(s)}{y_{sp}(s)} = \frac{1}{\tau_c s+1}$$

$$G_c = \frac{1/\tau_c s+1}{\frac{k_p e^{-\theta_p s}}{\tau_p s+1} \times \frac{\tau_c s}{\tau_c s+1}} = \frac{\tau_p s+1}{k_p e^{-\theta_p s}\tau_c s}$$

$$\frac{c(s)}{e(s)} = G_c = \frac{\tau_p s+1}{k_p \tau_c s}\,e^{\theta_p s}$$

✗ cannot use this

Not physically realisable

$$e = y_{sp} - y$$

$\Rightarrow$ use $e^{-\theta s} = 1-\theta s$

$$G_c = k_c\left(1+\frac{1}{\tau_I s}\right)$$

Q) $g = \frac{2e^{-s}}{(10s+1)(5s+1)}$

$e^{-s} = 1-s$

$$\frac{Y}{Y_{sp}} = \frac{e^{-s}}{\tau_c s+1}$$

$$\frac{1-e^{-s}}{\tau_c s+}$$

$$G = \frac{2(1-s)}{(10s+1)(5s+1)}$$

$$\frac{Y}{Y_{sp}} = \frac{e^{-s}}{\tau_c s+1}$$

$$G_c(s) = \frac{1}{\tilde{g}}\frac{e^{-s}}{(\tau_c+1)s} = \frac{(10s+1)(5s+1)}{2(\tau_c+1)s} \longrightarrow \text{PID}$$

$\longrightarrow$ $k_c = ?$

$\tau_I = ?$

$\tau_D = ?$

Open loop control

$$y_{sp} \longrightarrow \boxed{q} \longrightarrow \boxed{g} \longrightarrow y$$

for $y = y_{sp}$, ideally $q(s) = \frac{1}{g(s)}$

$$y = q(s)\,g(s)\,y_{sp}$$

Generally, $y \neq y_{sp}$

$\therefore$ we consider $\frac{y(s)}{y_{sp}(s)} = \frac{1}{(\lambda s+1)^n}$

$$\therefore q(s) = \frac{1}{g(s)}\frac{1}{(\lambda s+1)^n} = \frac{f(s)}{g(s)}$$

$g(s)$ → actual transfer function of the process.

→ $g(s)$ is estimated as $g_m(s)$

$$q(s) = \frac{f(s)}{g_m(s)}$$

$f(s)$ → filter. It should be chosen in such a way that

- $q(s)$ should not becomes $\boxed{\text{improper}}$ (no: of zeros > no: of poles)
- process dynamic characteristics are not violated

Non-invertible → TF containing right hand plane zeros & dead time

$$g_m(s) = \frac{k e^{-\theta s}}{\tau s+1}$$

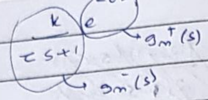
$$\frac{1}{g_m(s)} = \frac{(\tau s+1)e^{\theta s}}{k} \rightarrow \text{non-invertable}$$

because of dead time.

$$\frac{1}{g_m(s)} = \frac{(\tau_1 s+1)(\tau_2 s+1)}{-\eta s+1}$$

$s = 1/\eta$

$g_m(s) = g_m^-(s)\, g_m^+(s)$



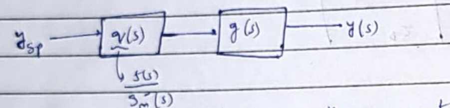
~~g_m~~ $g_m^-(s)$ is invertible part

$g_m^+(s)$ is non-invertible

$$\underbrace{q(s)}_{\text{open}} = \frac{f(s)}{g_m^-(s)}$$

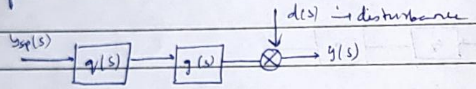
loop control law. → only for servo control (no disturbance)

Open loop control is always stable if the process is stable.

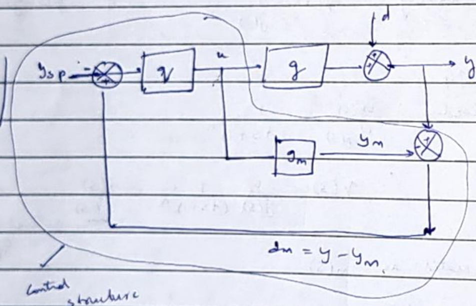


if we have disturbance, the open control law fails.

Internal model control



solves both regulatory problem & servo problem



internal model control architecture

control structure

$$y(s) = \frac{g(s)\,q(s)}{1 + q(s)(g(s) - g_m(s))}\, y_{sp}(s) + \frac{1 - g_m(s)q(s)}{1 + q(s)(g(s) - g_m(s))}\, d(s)$$

ideal: $q(s) = \frac{1}{g_m(s)}$ → $y = y_{sp}$ → It can take care of servo control

→ if d(s) changes also there is no change in the process variable → takes care of regulatory control

---

→ $g_m(s) = g_m^-(s)\, g_m^+(s)$

$$q_I(s) = \frac{1}{g_m^-(s)}$$

$$f(s) = \frac{1}{(\lambda s+1)^n}$$

$$q(s) = q_I(s)\, f(s) = \frac{f(s)}{g_m^-(s)}$$

(λ, n are the parameters)

Q) $g(s) = \frac{1}{10s+1} e^{-3s}$

No process/model mismatch → $g_m(s) = g(s)$

$g_m^-(s) = \frac{1}{10s+1}$ $\quad g_m^+(s) = e^{-3s}$

$q_I(s) = 10s+1$

$q(s) = q_I(s)\, f(s) = \frac{10s+1}{\lambda s+1}$ → lead-lag TF. → lead lag element is available.

$y(s) = g_m^+(s)\, f(s)\, y_{sp}(s) = \frac{e^{-3s}}{\lambda s+1} y_{sp}(s)$

$\frac{10s+1}{(\lambda s+1)^n}$ $\quad n \geq 1$

(n = 1 has advantages)

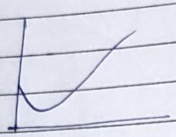
Q) $g(s) = \frac{-9s+1}{(15s+1)(3s+1)}$

$g_m^-(s) = \frac{9s+1}{(15s+1)(3s+1)}$ $\quad g_m^+(s) = \frac{-9s+1}{9s+1}$

$q_I(s) = q_I(s)\, f(s) = \frac{(15s+1)(3s+1)}{9s+1} \frac{1}{\lambda s+1}$

$y(s) = g_m^+(s)\, f(s)\, y_{sp}(s) = \frac{-9s+1}{(9s+1)(\lambda s+1)} y_{sp}(s)$

– If we have Right hand plane zero, we will have inverse response



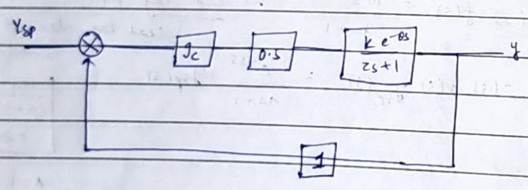
PID controller

$$g_c(s) = \frac{q(s)}{1 - g_m(s)\, q(s)}$$

Q) FODT system using PI controller
Control valve gain = 0.5
measurement device gain = 1

Method	k_c	τ_I
Z-N	0.3426	14.2546
IMC-PI	0.2	10.0

ZN PI relation :- $k_c = 0.45\, k_u$, $\tau_I = P_u/1.2$



Find k, τ, θ. $\qquad g_c^{(Z-N)}$ = known

g_c^{IMC} = known.

$$\frac{y}{y_{sp}} = \frac{\dfrac{g_c(0.5)\, k e^{-\theta s}}{\tau s+1}}{1 + \dfrac{g_c(0.5) k e^{-\theta s}}{\tau s+1}}$$

$$G_c = k_c\left(1 + \frac{1}{\tau_I s}\right) =$$

$$k_c = \frac{1}{k}\left(\frac{\tau}{\theta + \tau_c}\right) \qquad \underline{\tau_I = \tau}$$

Z-N method

→ $k_c = 0.3426$ $\qquad k_u = \frac{0.3426}{0.45}$ $\qquad P_u = 1.2 \times 14.2546$

$P_u = \frac{2\pi}{w_{co}}$ → find w_{co}. $\qquad w_{co}$ – cross over frequency. ($\phi = -180°$) $\quad \phi = -\pi$

$\boxed{w_{co} = 0.3673}$ $\qquad \phi = \tan^{-1}(-w_{co}\tau) - w_{co}\theta$. get θ

$k_u = \frac{1}{AR_c}$ $\qquad AR_c = \frac{k\, 0.5}{\sqrt{1 + \tau^2 w_{co}^2}}$

for IMC PI :- $\boxed{\tau_I = \tau}$

$k_c = \frac{\tau}{k(\lambda + \theta)}$

find τ from here. $\quad \tau = 10 = \tau_I$

$\boxed{\tau = 10}$

$-\pi = \tan^{-1}(-3.673) - w_{co} \times \theta$.

$+\pi = +1.304 + 0.367 \times \theta$.

$\theta = \frac{5.006 \times 180}{\pi} = 286.83°$
rad

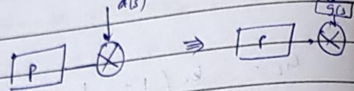
$k = 10$.

→ PID for servo → $f(s) = \frac{1}{(\lambda s+1)^n}$

→ $f(s) = \frac{\gamma s + 1}{(\lambda s + 1)^n}$

IMC eqn :- $y(s) = \frac{(1 - q(s)\, g(s))}{1 + q(s)(g(s) - g_m(s))}\, d(s)$

$= \frac{1 - q(s)\, g(s)}{1 + q(s)(g(s) - g_m(s))}\; g(s)\, L(s)$ → for disturbance.



$$y(s) = \frac{1 - q(s)\,g(s)}{1 + q(s)\left(g(s) - g_m(s)\right)}\, g(s)\, d(s)$$

$$y(s) = \left(1 - q(s)\,g(s)\right) g(s)\, d(s)$$

$\gamma = f(\tau)$

should be calculated using this

Q2) $g(s) = \dfrac{2}{(5s+1)(2s-1)}$

$$q(s) = \frac{(5s+1)(2s-1)}{2} \times \frac{\gamma s+1}{(\lambda s+1)^2}$$

$P_u = 1/2$

$$f(s) = \frac{\gamma s+1}{(\lambda s+1)^2}$$

$$1 = \frac{\gamma(1/2)+1}{(\lambda(1/2)+1)^2}$$

$$\gamma = \left[\left(\lambda\left(\tfrac{1}{2}\right)+1\right)^2 - 1\right]2$$

$$\gamma = \left(\frac{\lambda^2}{4} + \lambda\right)2$$

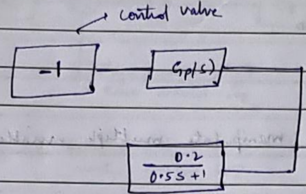
$$\gamma = \frac{\lambda^2}{2} + 2\lambda$$

$$g_c(s) = \frac{1}{2}\left[\frac{(5s+1)(2s-1)(\gamma s+1)}{\lambda^2 s^2 + (2\lambda-\gamma)s}\right]$$

$$g_c(s) = \frac{1}{2}\left[\frac{(5s+1)(\gamma s+1)}{\frac{\lambda^2}{2} s}\right] \quad \text{PID}$$

($\cancel{5s+1}$)

Q3) $g_p(s) = \dfrac{-0.7}{(5s+1)(1.5s+1)}$



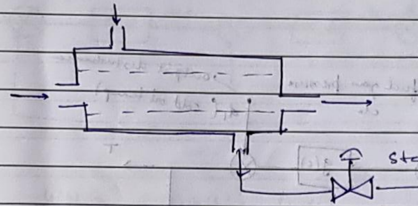
30sec lag

for temp measurement device gain

$$\frac{20-4}{100-20}$$

$$\frac{\Delta\,\text{output}}{\Delta\,\text{input}}$$

Q4)



$$\text{gain} = \frac{4^\circ C}{1\ \text{kg/min}}$$

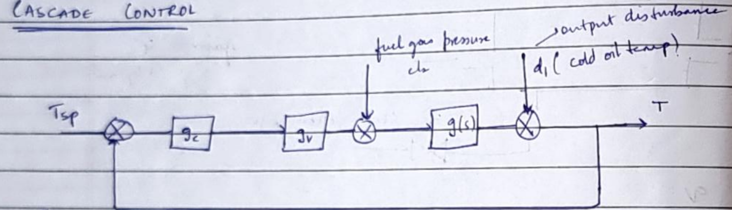
$$g_c(s) = \frac{q}{1 - q\,g_m(s)}$$

Traditional advanced control strategies

① High level application of PID
- Time delay compensator
- Cascade control
- split range control — manipulate multiple variables using 1 controller

③ Feedforward and ratio control.

CASCADE CONTROL

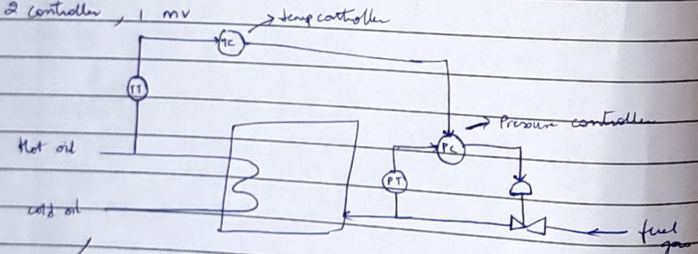


furnace (feedback control) (normal loop)

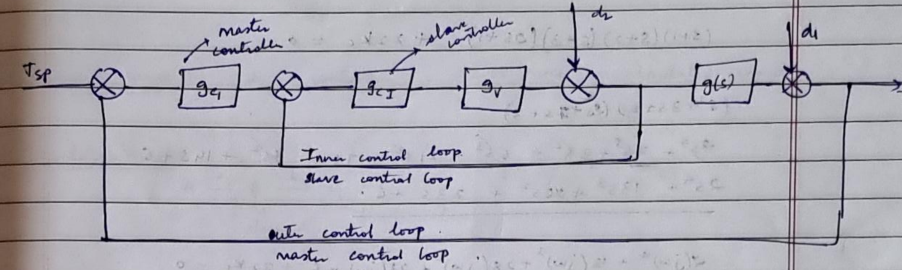
→ regulatory control strategy cannot take care of input disturbances. (fuel gas pressure)

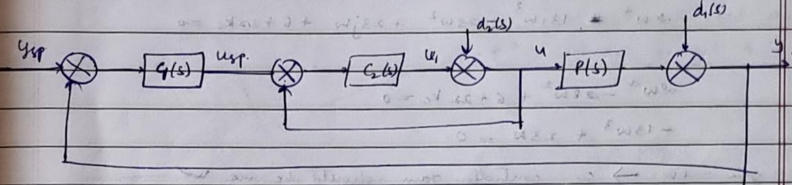
This can be controlled using another controller at the input

→ 2 controller, 1 mv



cascading temp controller using pressure controller.
(TC-PC cascade controller)

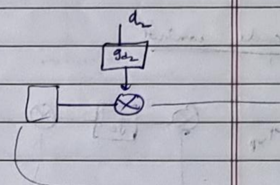




$$u(s) = \frac{C_2(s)V(s)}{1 + C_2(s)V(s)}\, u_{sp}(s) + \frac{1}{1 + C_2(s)V(s)}\, d_2(s)$$

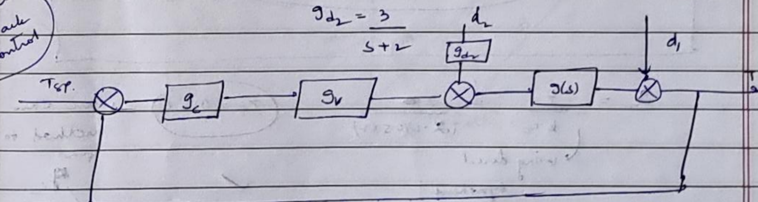
Q) $y(s) = \frac{10}{(s+1)(s+2)(s+3)}\, u(s)$

$g(s) = \frac{10}{(s+1)(s+2)(s+3)}$ $g_v = \frac{2.2}{2s+1}$



Normal feedback control

$g_{d_2} = \frac{3}{s+2}$



$1 + k_c g_v g = 0$

$1 + k_c\left(\frac{2.2}{2s+1}\right)\left(\frac{10}{(s+1)(s+2)(s+3)}\right) = 0.$ $K_{cu} = ?$

$s = j\omega.$

$(2s+1)(s+1)(s+2)(s+3) + k_c(22) = 0.$

$(s+1)(s+2)(s+3)(2s+1) + 22k_c = 0$

$(s^2+3s+2)(2s^2+7s+3)$

$2s^4 + 7s^3 + 3s^2 + 6s^3 + 21s^2 + 9s + 4s^2 + 14s + 6$

$2s^4 + 13s^3 + 28s^2 + 23s + 6.$

$2(j\omega)^4 + 13(j\omega)^3 + 28(j\omega)^2 + 23(j\omega) + 6 + 22k_c = 0.$

$2\omega^4 \; 13j\omega^3 - 28\omega^2 + 23j\omega + 6 + 22k_c = 0.$

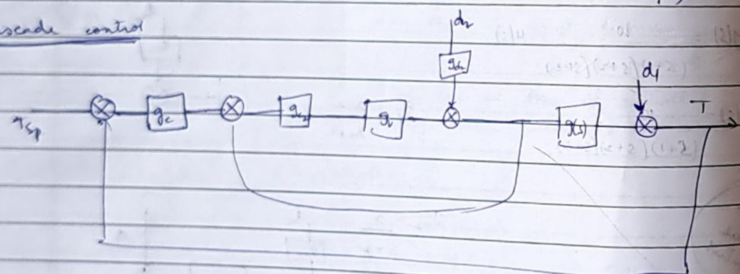
$2\omega^4 - 28\omega^2 + 6 + 22k_c = 0.$

$-13\omega^3 + 23\omega = 0.$

→ Process gain = +ve → so controller gain should be +ve ✓
overall (10×22)
stable process

→ If the process is unstable, controller gain = −ve.
for unstable process Process gain & controller gain can be of diff signs)

Cascade control



$g_{c_2} = \frac{2}{k\tau_c} = \frac{2}{(2.2)(0.5\times2)}$ $\tau_c = 0.5\tau$ → use this method to find g_{c_2}
using direct synthesis approach.
Proportional controller

Cascade controller
Controller
2 or more

Manipulative Input
1

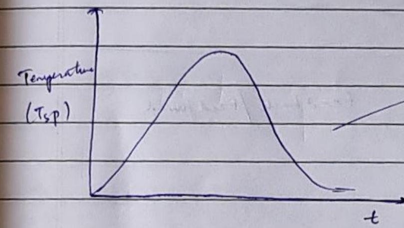
SPLIT RANGE CONTROL

split range :- 1 controller, 2 or more manipulated input

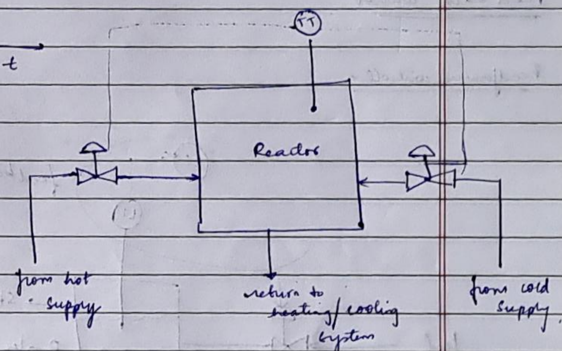
→ Batch reactor

$A \xrightarrow{k_1} B \xrightarrow{k_2} C$
B: desired
C: undesired

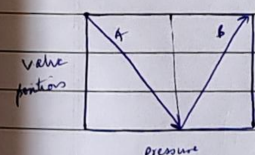
$k_i = A_i e^{-\frac{\Delta H_i}{RT}}$



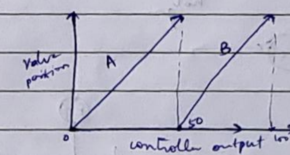
hot stream & cold stream are blended to get this temp profile



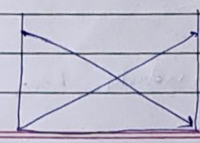
split range



Sequence operation (split range)



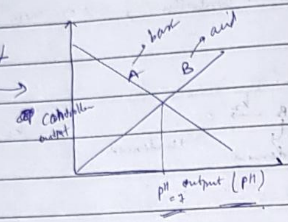
opposition action split range control



Reflux drum

→ pH neutralization reactor

pH ↓, controller output ↓

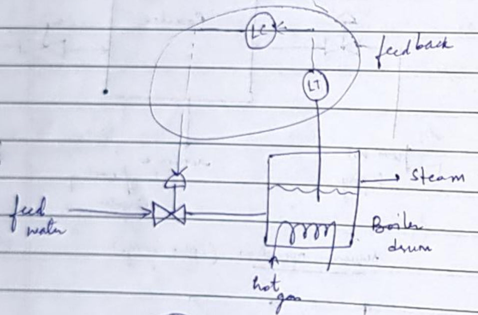


FEEDFORWARD AND RATIO CONTROL

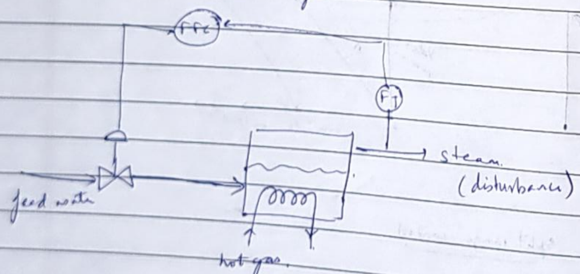
feedback / Feed forward

Q2 Feed back control

Feedforward controller.

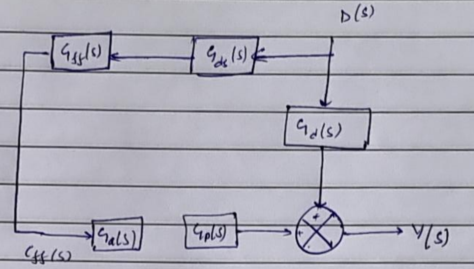


liq level is output



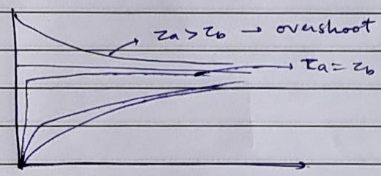
Only FFC strategy is never applied in an industry becoz it cannot solve servo problem.

→



~~$Y(s) = D(s)\, G_{ds}(s)\, G_{ff}(s)\, G_a(s)\, G_p(s) + D(s)\, G_d(s) = 0$~~

→ $\frac{Y(s)}{U(s)} = G_c(s) = \frac{\tau_a s + 1}{\tau_b s + s}$



→ Combined FF & FB control

If we use a pure FFC strategy, there is no stability issue

If we have combination of FF & FB → stability will come into picture

Adding FFC will not hamper the stability of the overall control loop.

Ratio control

$R = \frac{u \to mv}{d \to \text{disturbance}}$