DCS Controller Tuning

- Process Dynamics
- Feedback Control
- Controller Tuning Procedure
- Initial Tuning from Design Data

Process Dynamics

Two Main Types of Processes

- Stable Processes
- Unstable Processes
  - Ramp Processes
  - Exothermic Chemical Reactions

Stable Processes

- First Order
  - Process Heaters
  - Flash Drums
  - Flow Controllers
- Second Order / Higher Order
  - Distillation Columns
  - Feed / Effluent Exchangers

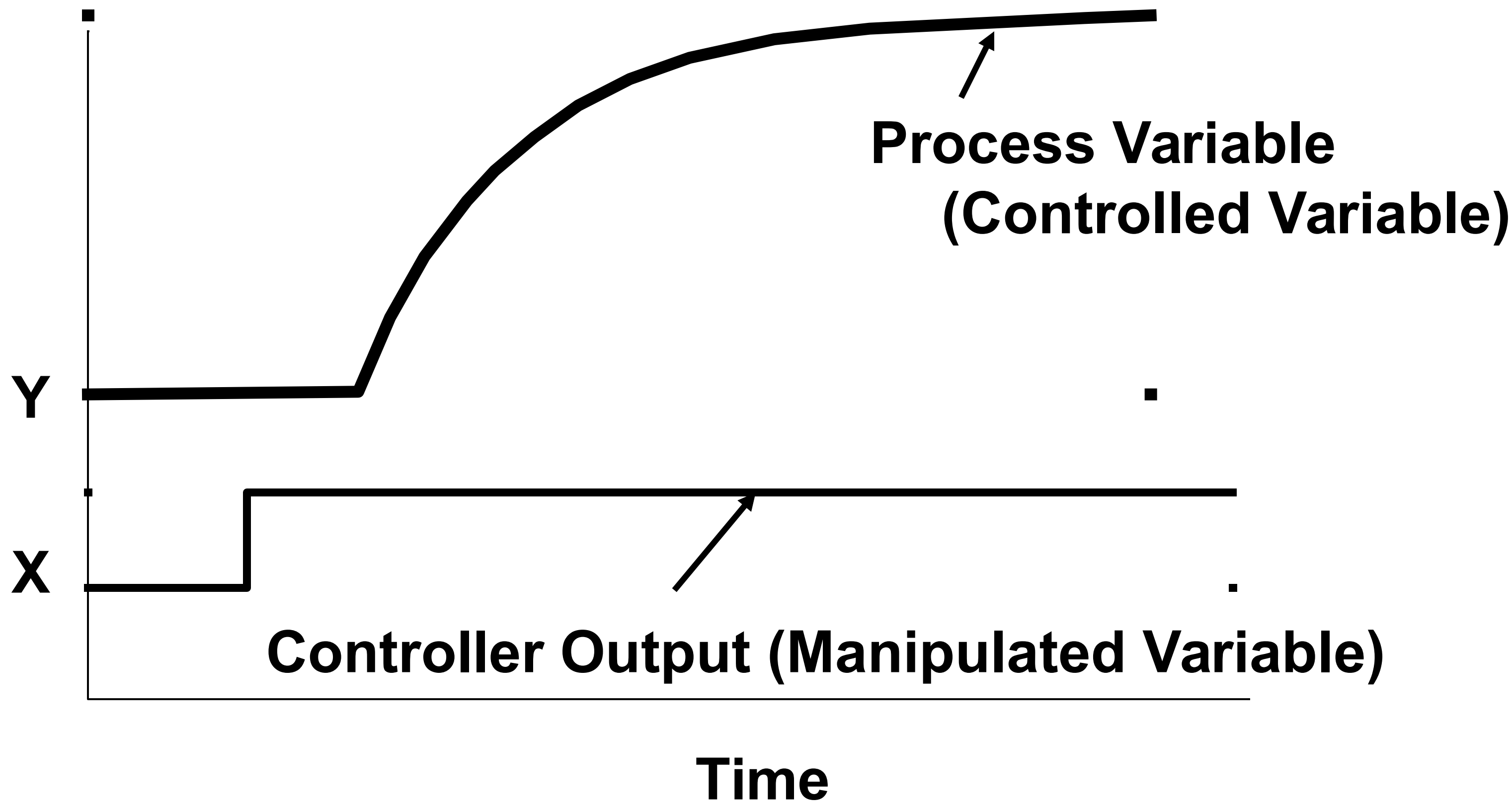
First Order Dynamics

Defined by 3 Parameters

- **Process Gain - Kp**
  - **Change in Process Variable (% of Scale) Steady State per 1% Change in Controller Output**
- **Dead Time - θ (Theta)**
  - **Elapsed Time Before Process Reacts to Change in Controller Output**
- **First Order Time Constant - τ (Tau)**
  - **Time for Process to Reach 63.2% of its Steady State Process Gain**

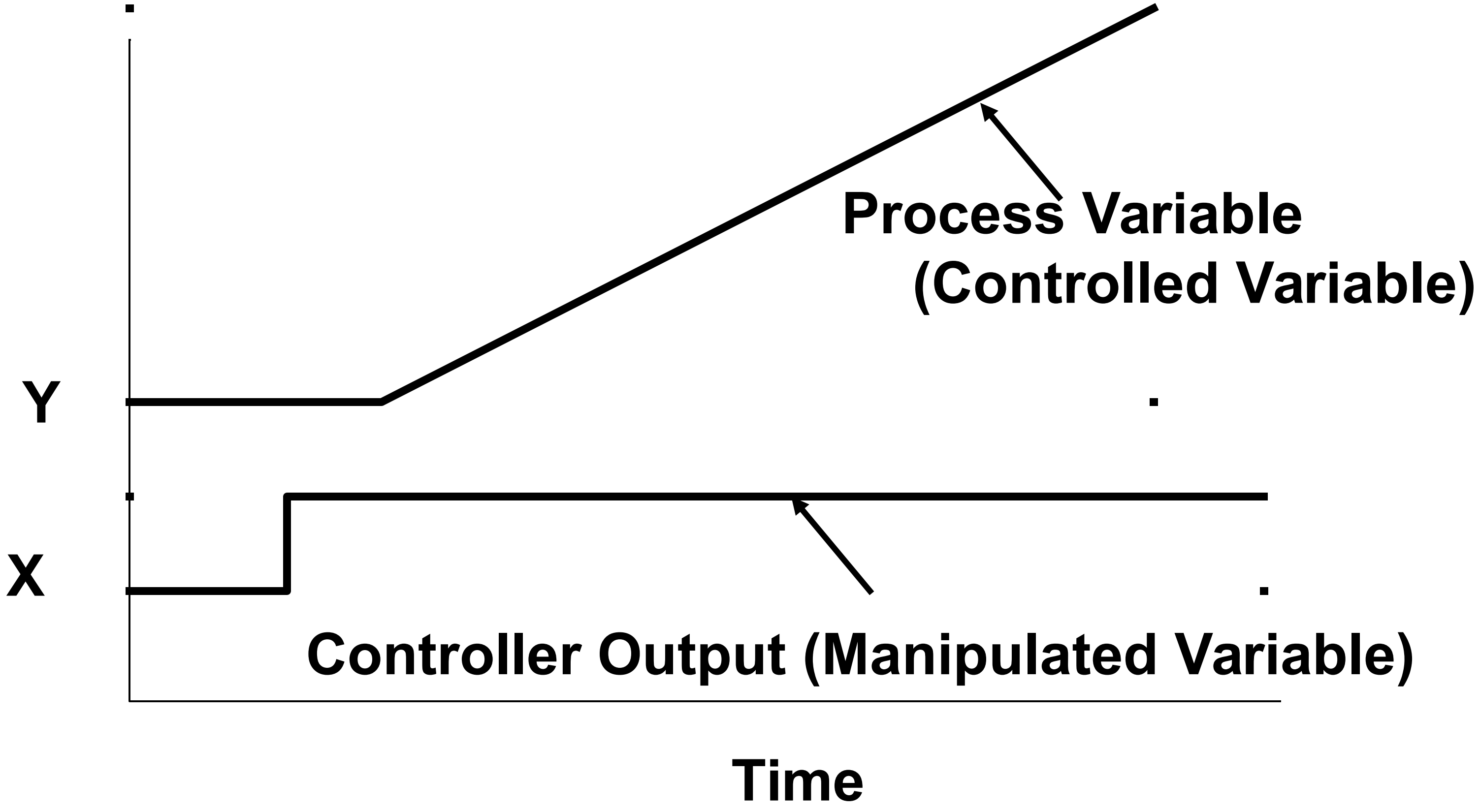
Stable Process

- Temperatures
- Flows

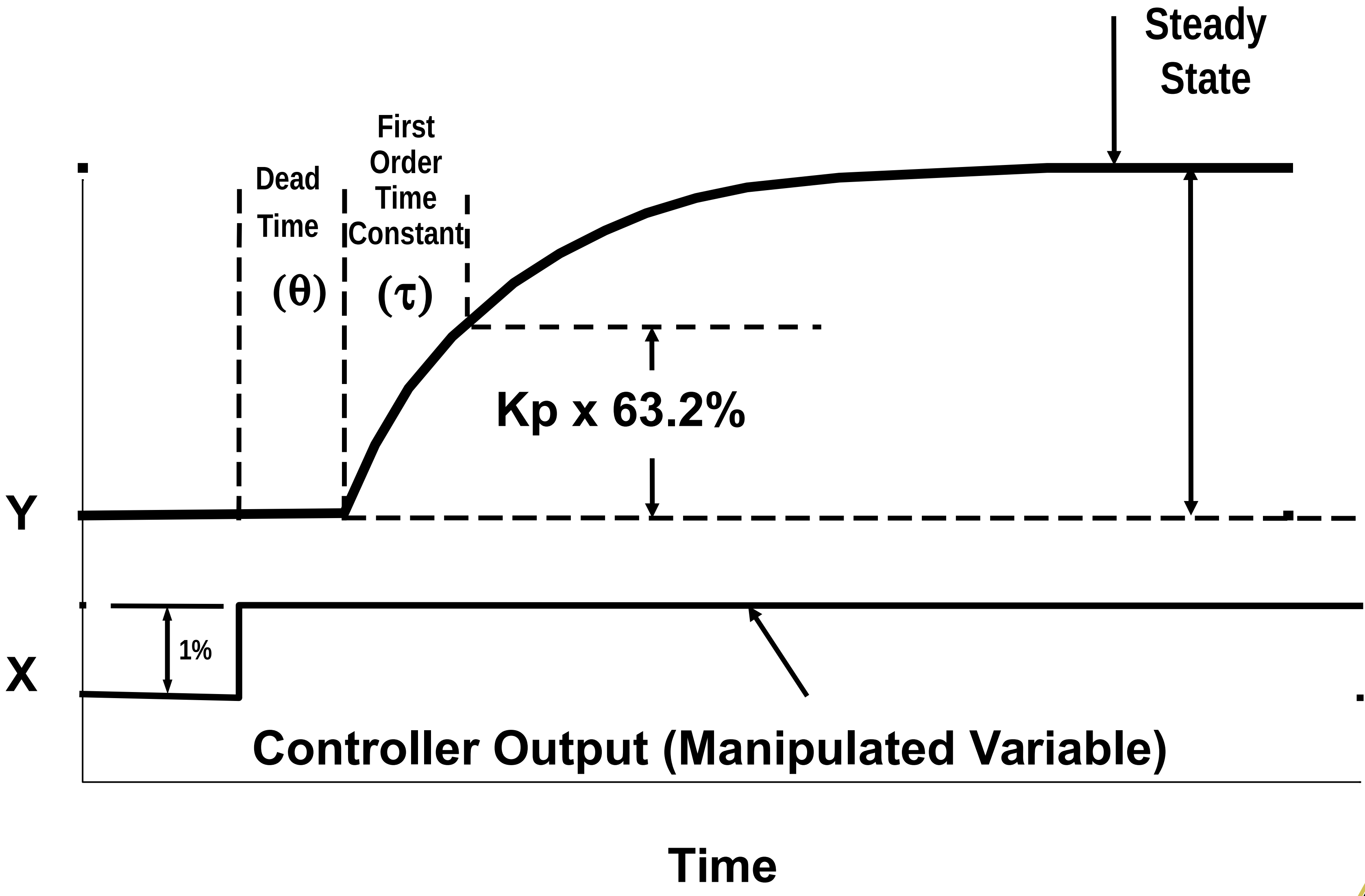


Unstable (Ramp) Processes

- Levels
- Pressures



First Order Dynamics



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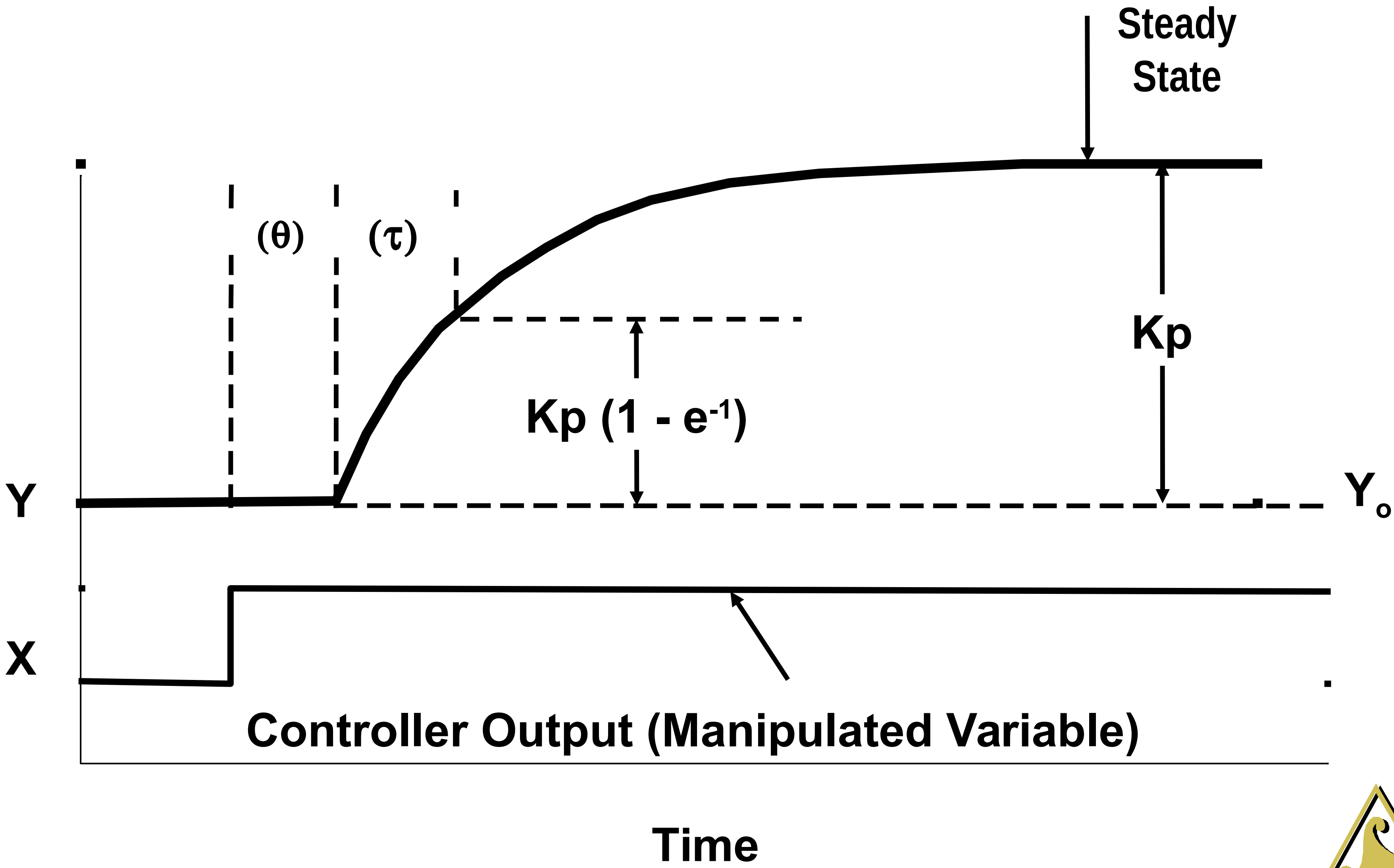
First Order Dynamics



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First Order Dynamics

$$Ay + \frac{dy}{dt} + C = 0 \qquad (1)$$

$$A,\ C = \text{Constant}$$

$$y = y_0 + Kp \cdot \Delta x \left[1 - e^{-\left\{\frac{t - \theta}{\tau}\right\}}\right] \qquad (2)$$

$\text{IF, } Kp = 1,\ \Delta x = 1, \text{ then}$

a) $[\text{for } t < \theta]\ y = y_0$

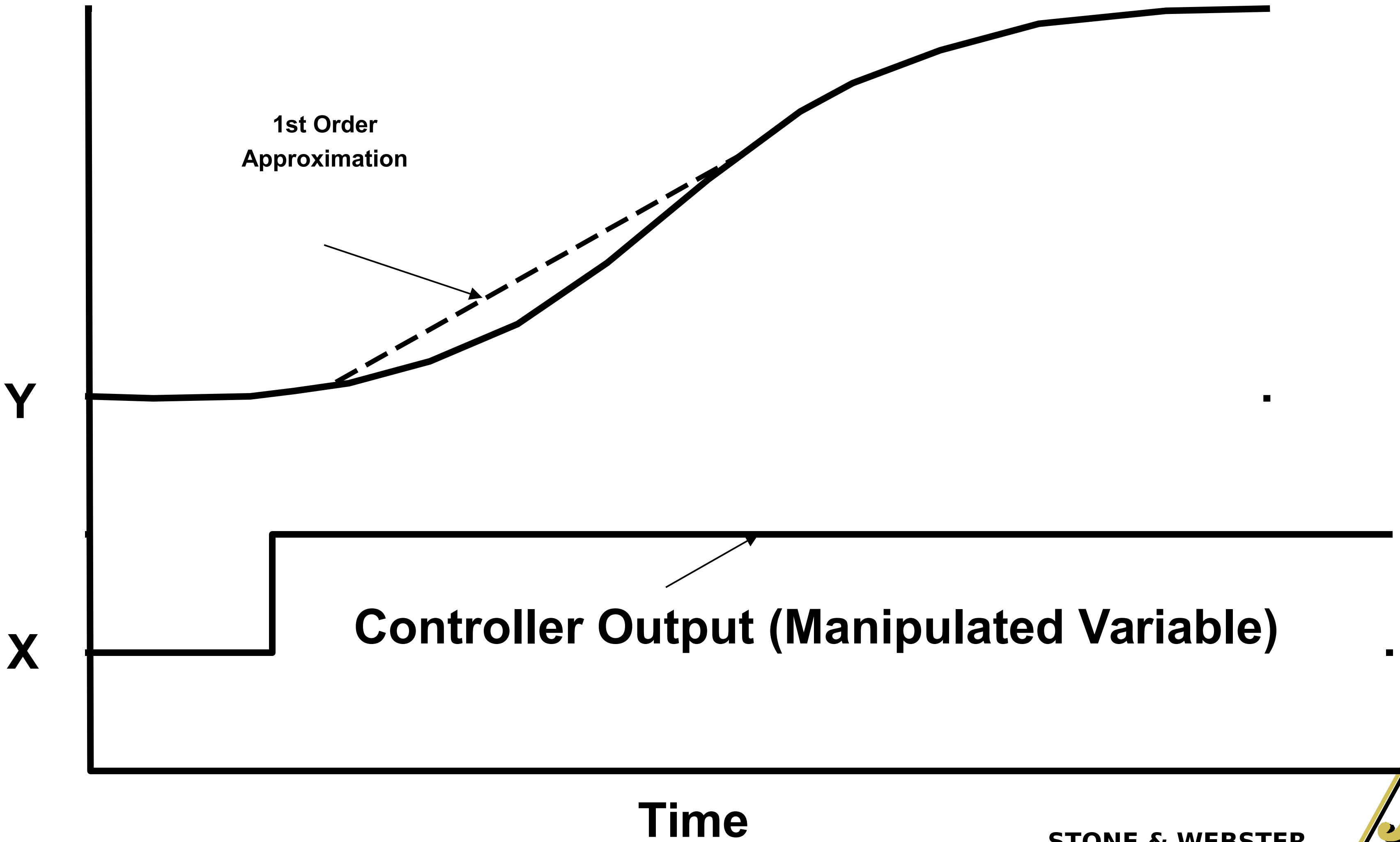
b) $[\text{for } t = \theta + \tau]\ y = y_0 + 1 - e1 = y_0 + 0.632$

c) $[\text{for } t = \infty]\ y = y_0 + 1$

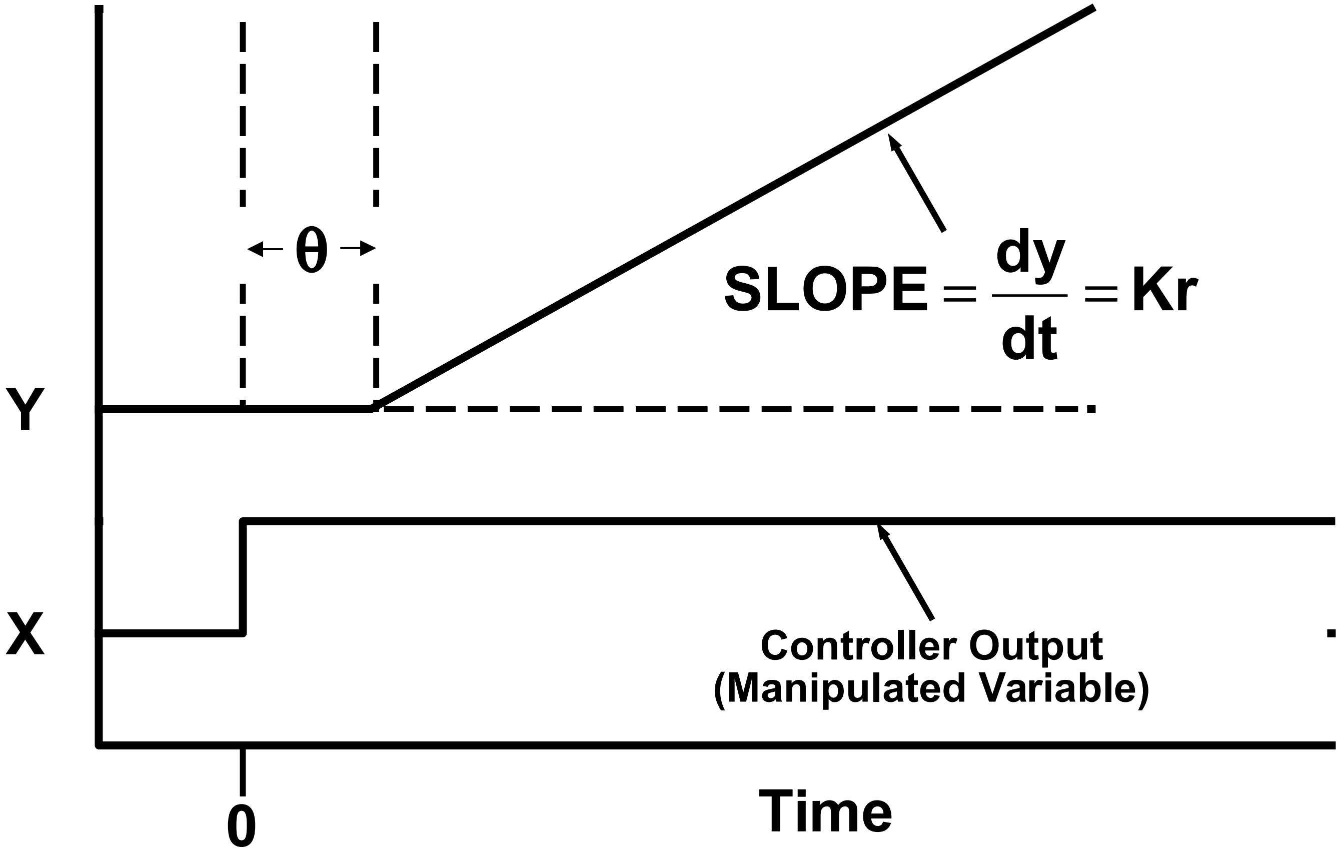
Second Order and Higher Order Dynamics

- **Summation of Two or More First Order Processes**
- **Can be Approximated by First Order Dynamics for Purposes of Controller Tuning**

Second Order Dynamics



Ramp Process Mathematics



$$\frac{dy}{dt} + C = 0 \qquad (1)$$

$C = \text{Constant}$

$$y = y_0 + Kr \Big| \quad \theta \qquad (2)$$

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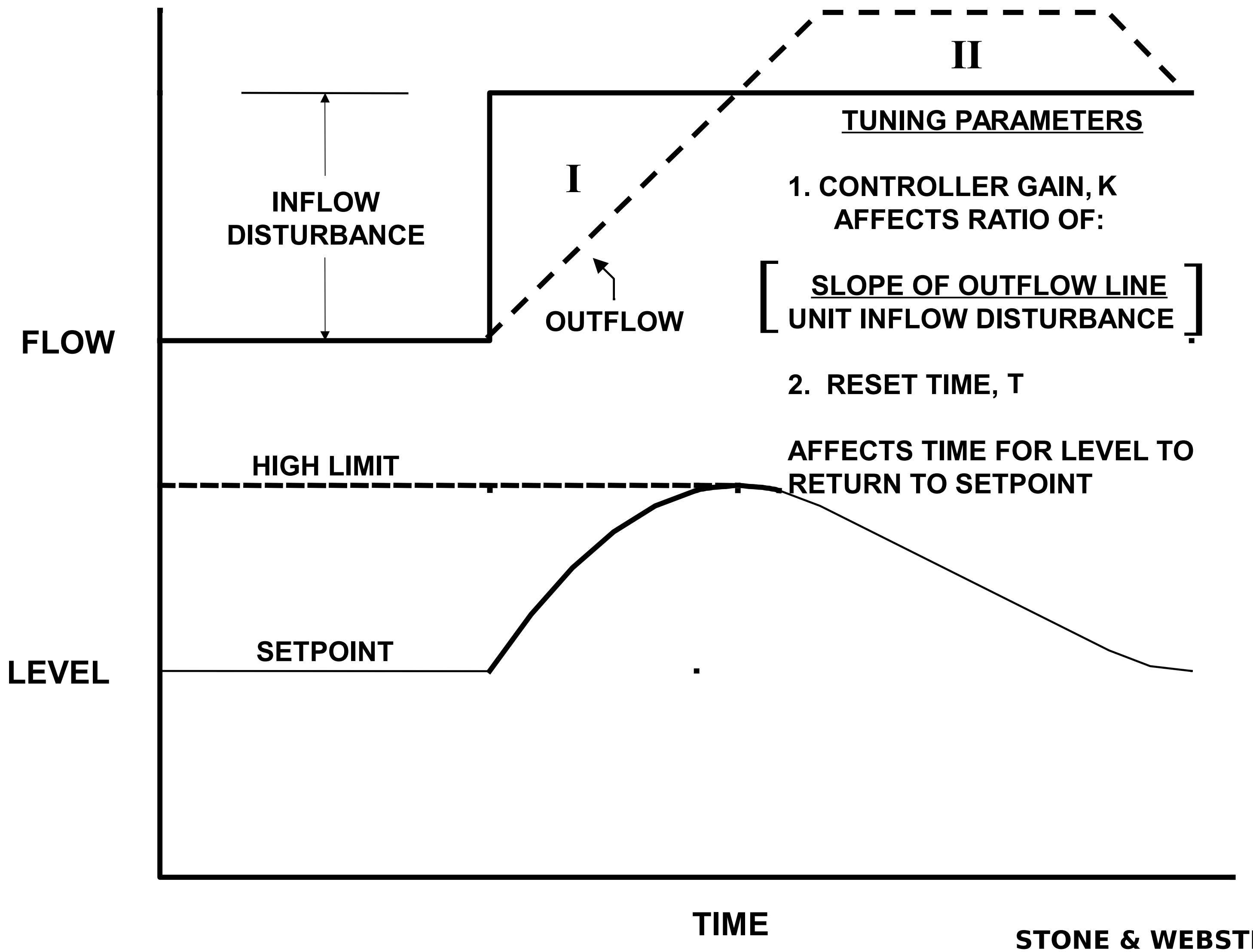
Ramp Process Dynamics

<u>Defined by 2 Parameters</u>

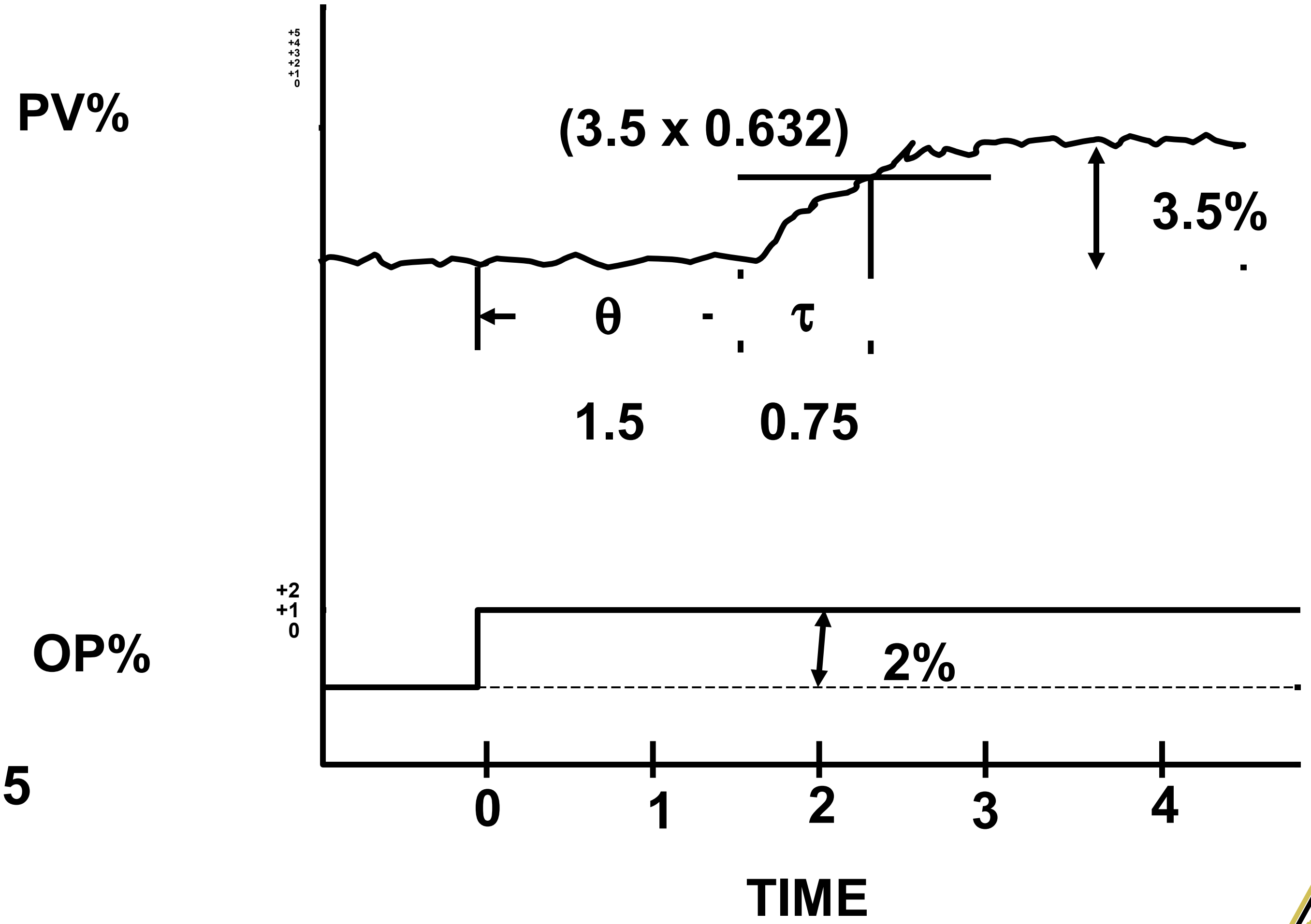
- **Process Gain - Kr**
  - **Change of Slope of Process Variable (% of Scale/Minute) (per 1% Change in Controller Output)**
- **Dead Time - θ**

Inflow, Outflow, and Accumulation



Calculation of Model Parameters

A) Stable Process



$$Kp = \frac{3.5}{2.0} = 1.75$$

$$\theta = 1.5 \text{min}$$

$$\tau = 0.75 \text{min}$$

Identifying Process Dynamics

1) Determine Whether Process is Stable or Unstable
2) Conduct Response Test
    a) Put Controller in Manual
    b) Make Small Change in Controller Output
    c) Use Trend Recorder to Plot Change in Process Variable
    d) Move Controller Output Back to Original Position
3) Calculate Model Parameters
    a) Kp, θ, τ for Stable Process
    b) Kr, θ, for Ramp Processes

DCS Controller Tuning

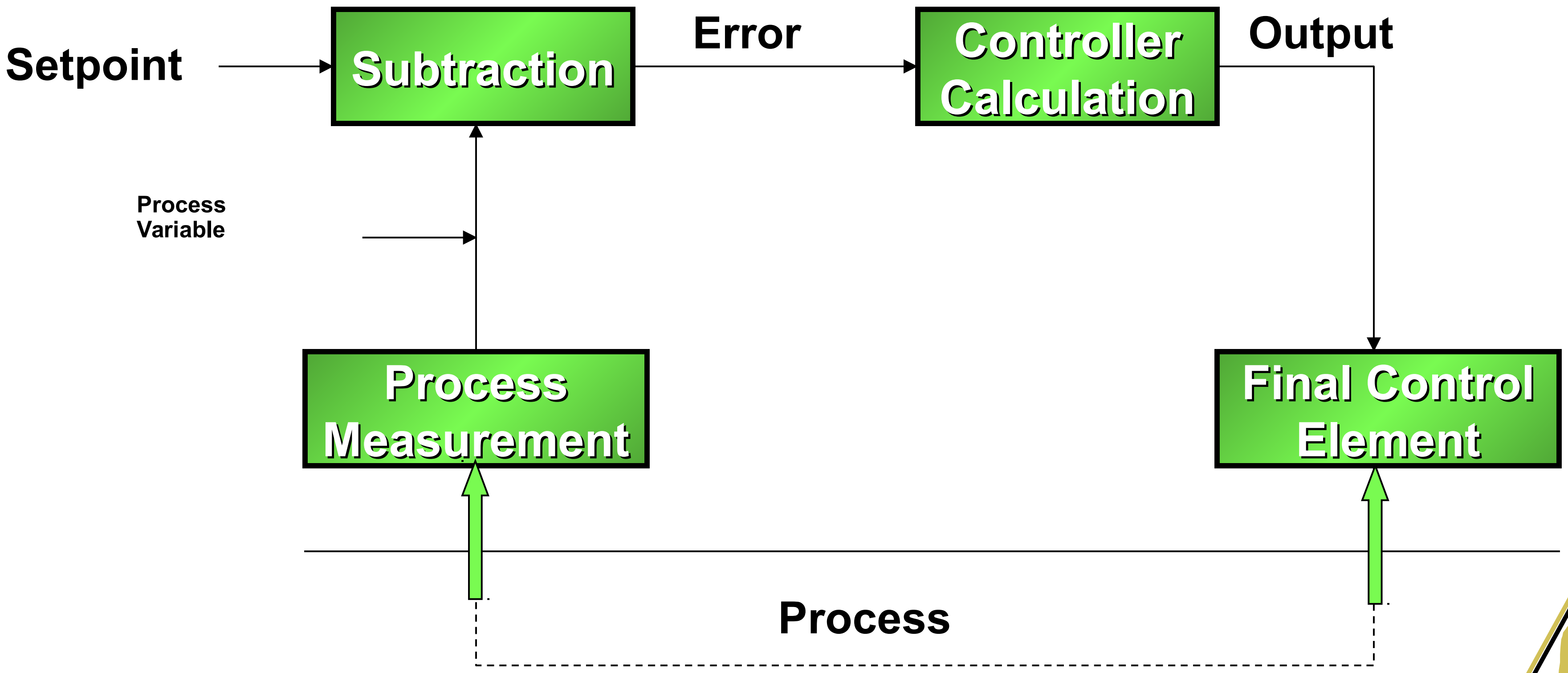
- **Feedback Control**

Types of Feedback Control

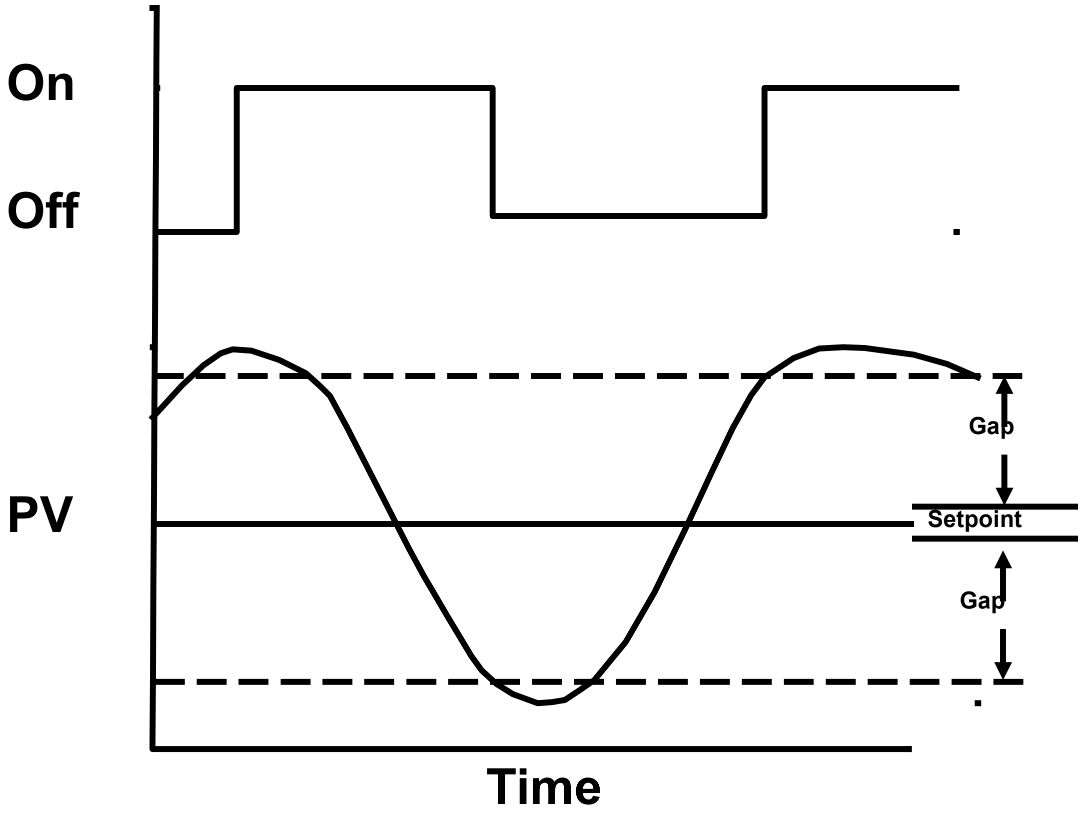
- **Manual Operation (Human Controller)**
- **On/Off Control (e.g. Thermostat)**
- **Proportional**
- **Integral**
- **Derivative**
- **PID**

Feedback Control Elements

- Measurement
- Setpoint Mechanism
- Controller
- Final Control Elements



On/Off Control



- Size of gap is only tuning parameter.
- Small adjustments impossible without frequently switching on/off.
- Air cooler is a common refinery example of on/off control.
- Sump pump another example.

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Objectives of Proportional Control

- **React to disturbances (regulatory).**
- **Move process toward new setpoint entered by operator.**

Proportional Control Parameters

PID Tuning Parameters are Defined Differently by Different Manufacturers

Proportional Control Parameters

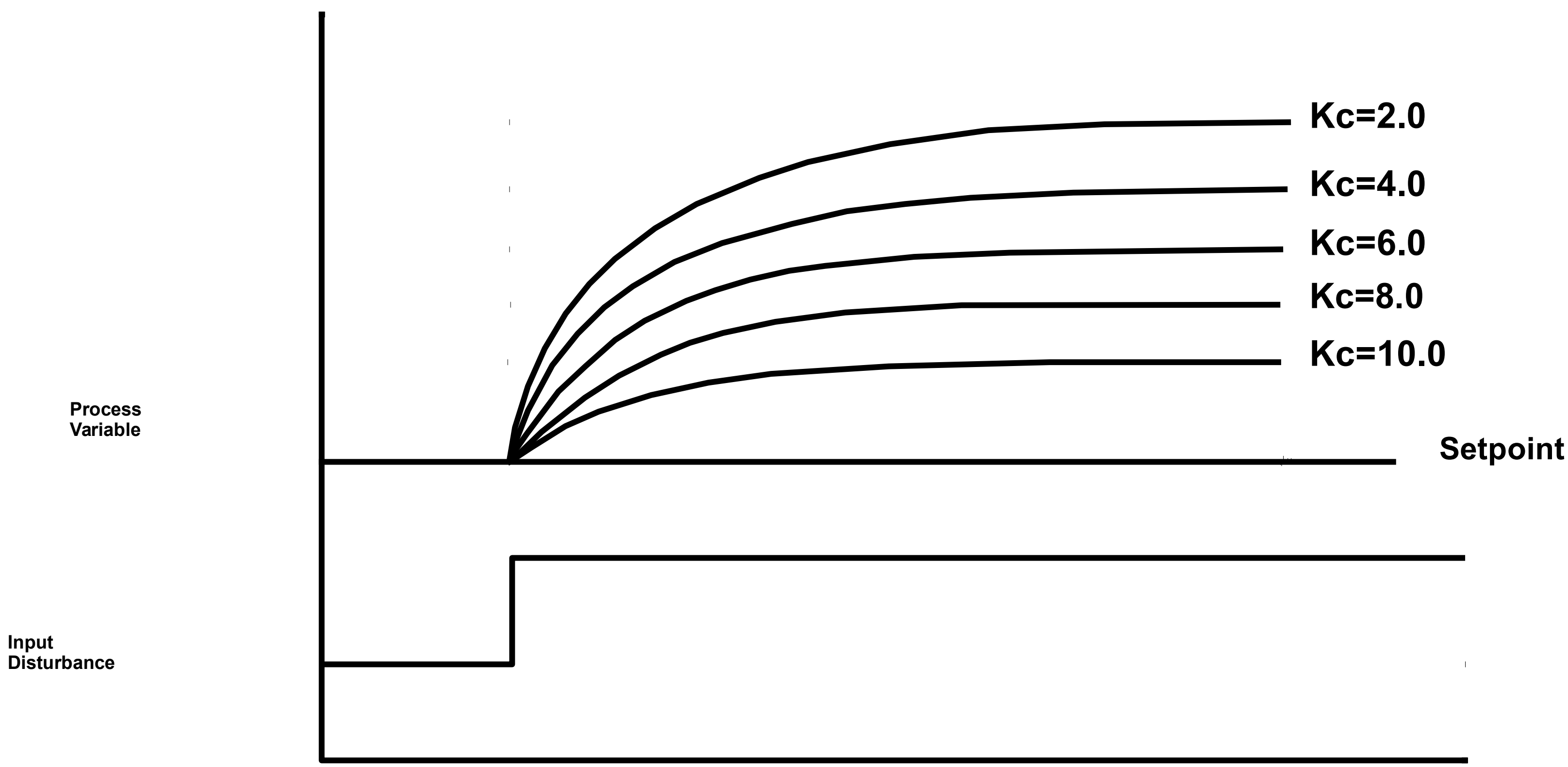
Kc = Controller Gain
Increase Kc To Speed Up Action

PB = Proportional Band
Decrease PB to Speed Up Action

$$\boxed{Kc = \frac{100}{PB}}$$

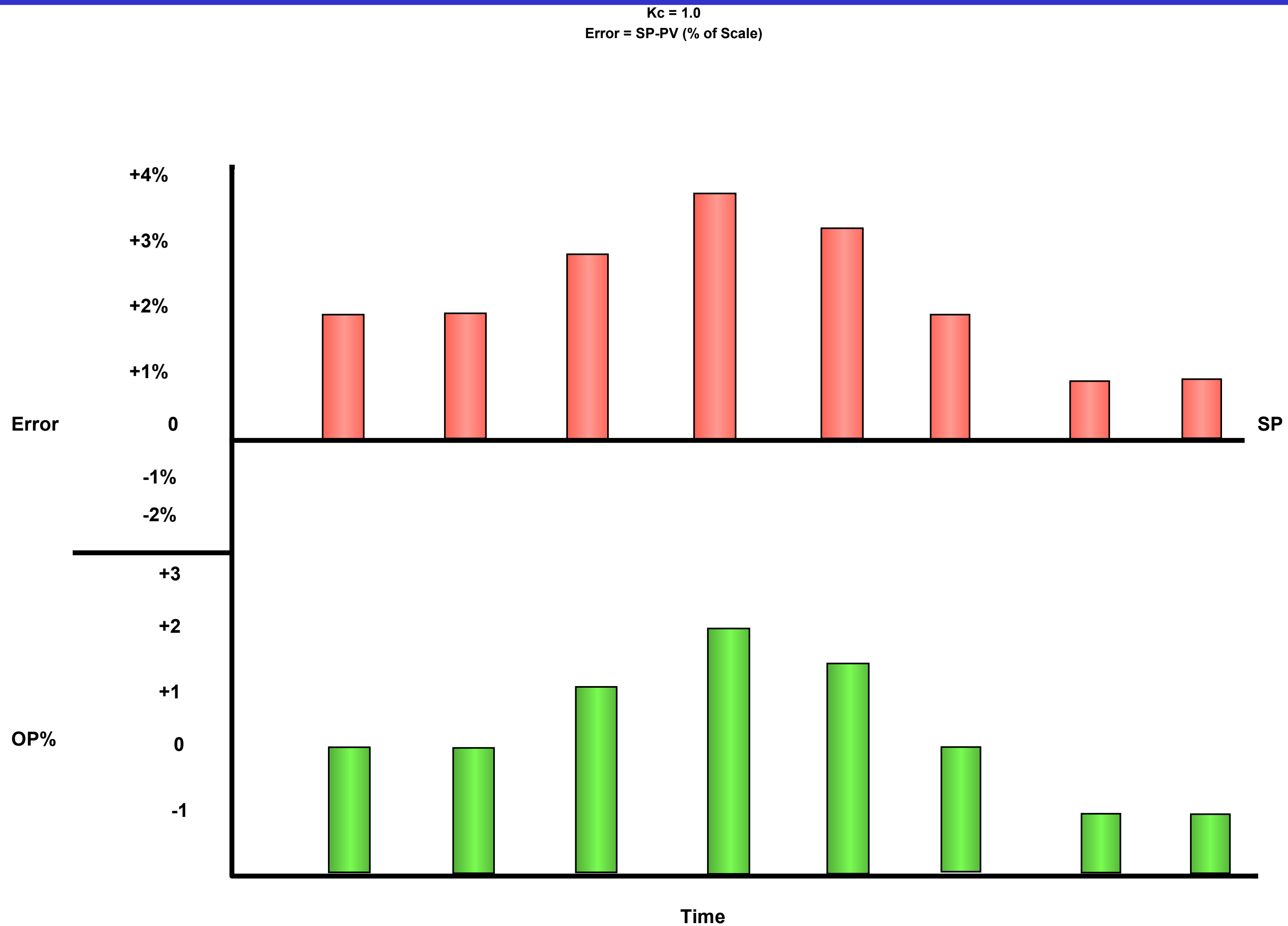
Proportional Control

Effect of Tuning Constant on Control Action



Question: Why does the process variable not return to setpoint?

Proportional Control Example



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Proportional Control Example

a) T = 0: E = +2

b) T = 1: E = +2, ΔE = (2-2) = 0, ΔOP = (1.0) (0) = 0

c) T = 2: E = +3, ΔE = (3-2) = 1, ΔOP = (1.0) (1) = +1.0

d) T = 3: E = +4, ΔE = (4-3) = 1, ΔOP = (1.0) (1) = +1.0

e) T = 4: E = +3.5, ΔE = (3.5-4) = -0.5, ΔOP = (1.0) (-0.5) = -0.5

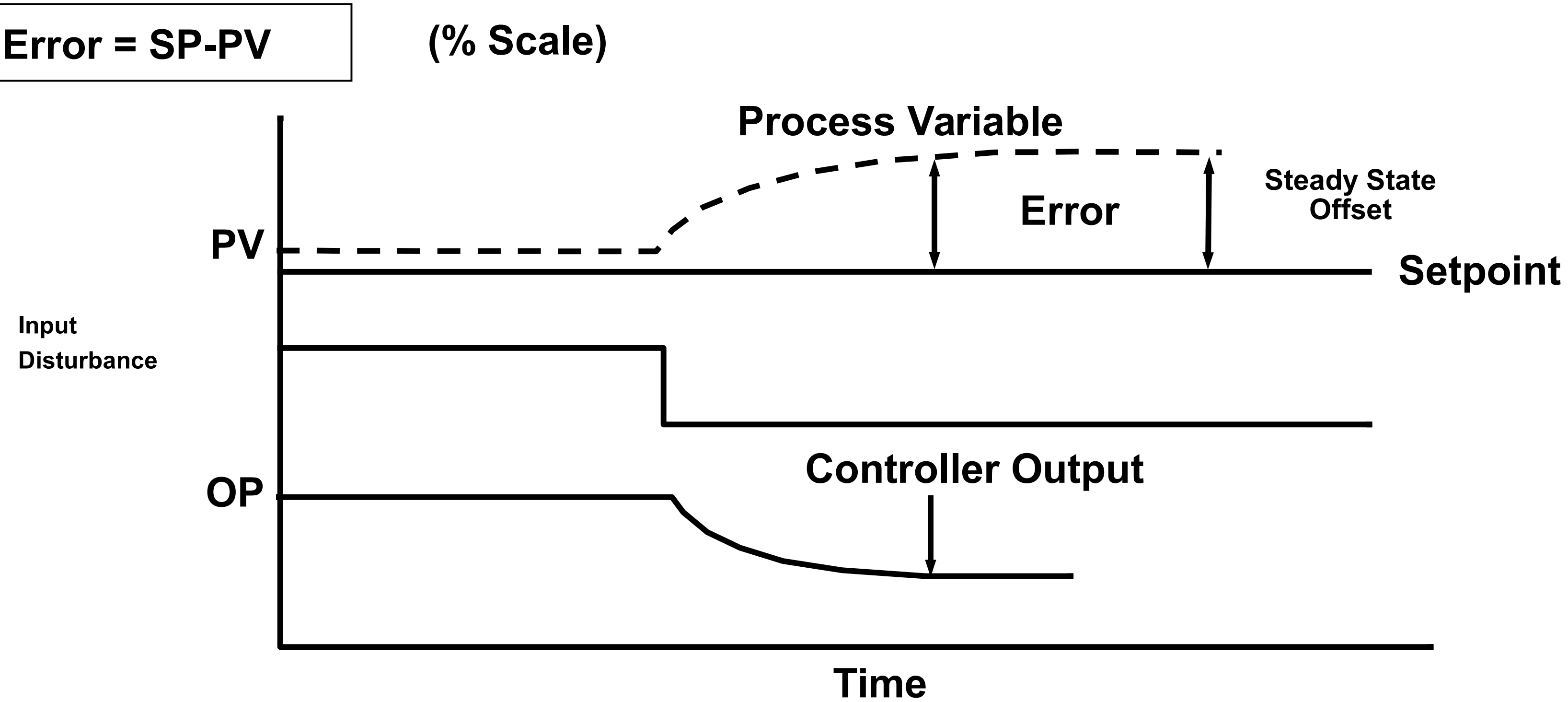
f) T = 5: E = +2.0, ΔE = (2.0-3.5) = -1.5, ΔOP = (1.0) (-1.5) = -1.5

g) T = 6: E = +1.0, ΔE = (1.0-2.0) = -1.0, ΔOP = (1.0) (-1.0) = -1.0

h) T = 7: E = +1.0, ΔE = (1.0-1.0) = 0.0, ΔOP = (1.0) (0) = -0.0

Proportional Control

Error = SP-PV



- Change in controller output proportional to change in error.
- Proportional-only controller stops moving when error is constant.
- Proportional-only controllers have steady-state offset.

Proportional Control

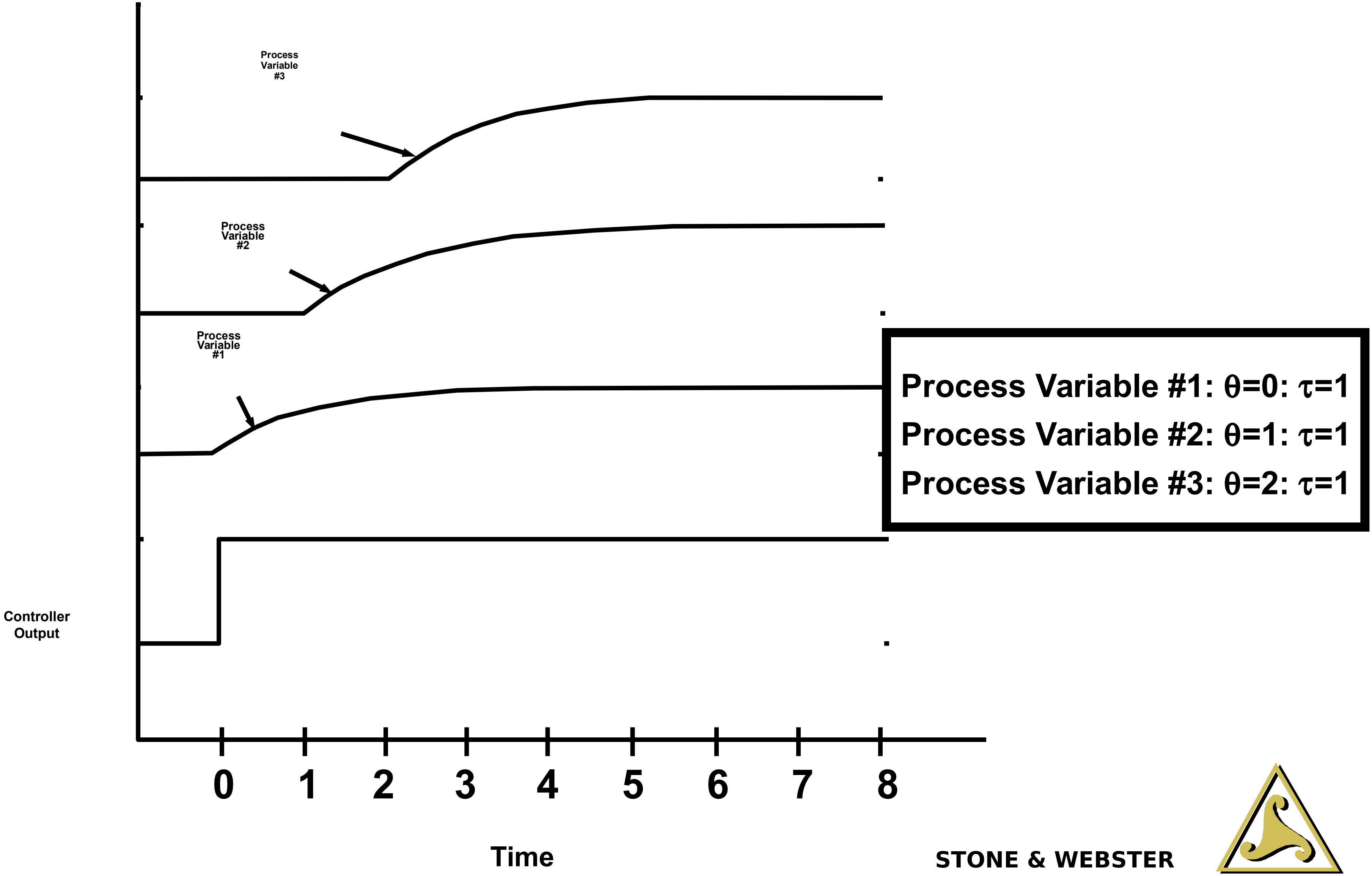
Proportional - Only Calculation

Output Change = Kc x [Error - Old Error]

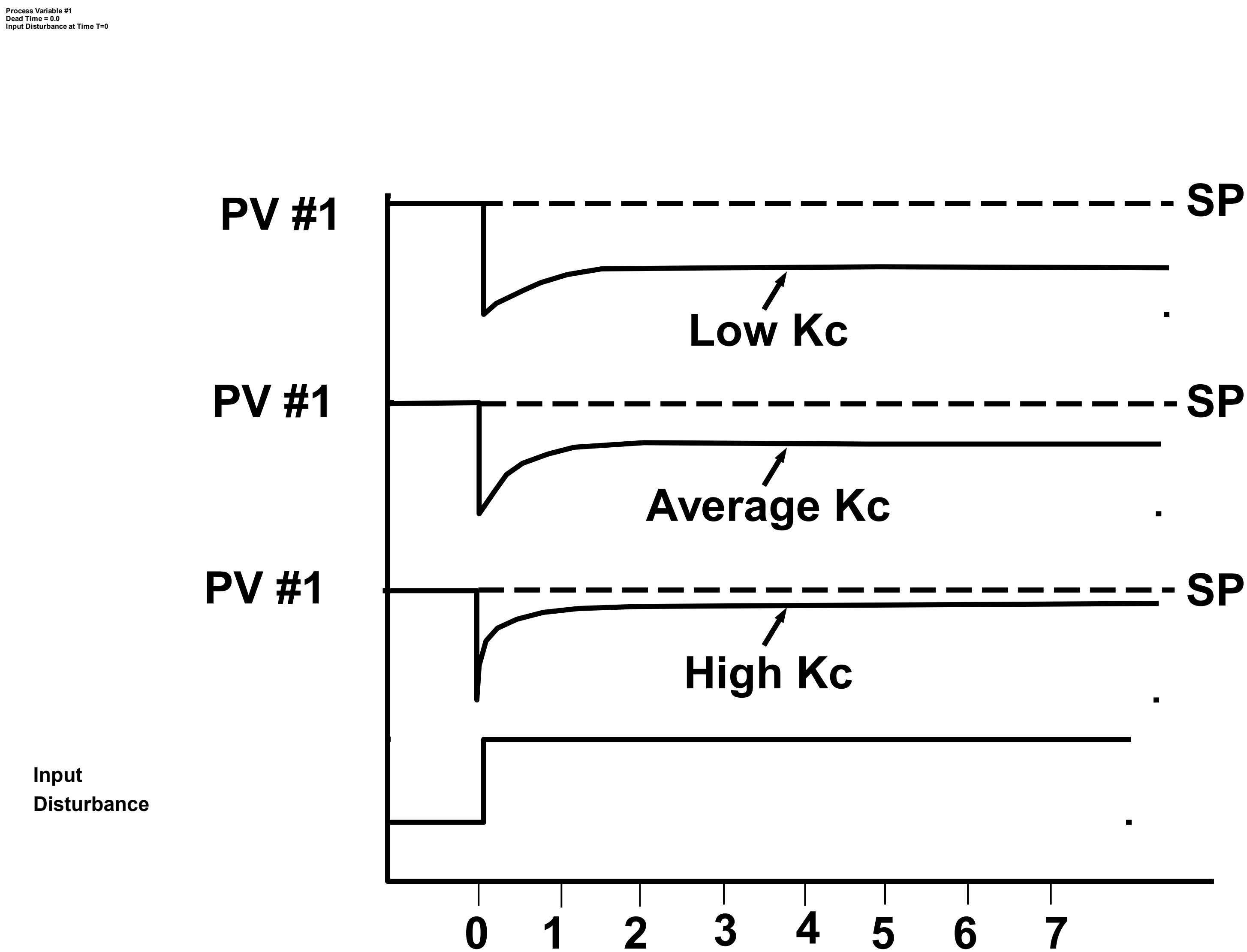
ΔOP = Kc x ΔE

Kc = Controller Gain (Tuning Parameter)

Proportional Control - Dead Time



Proportional Control - Dead Time



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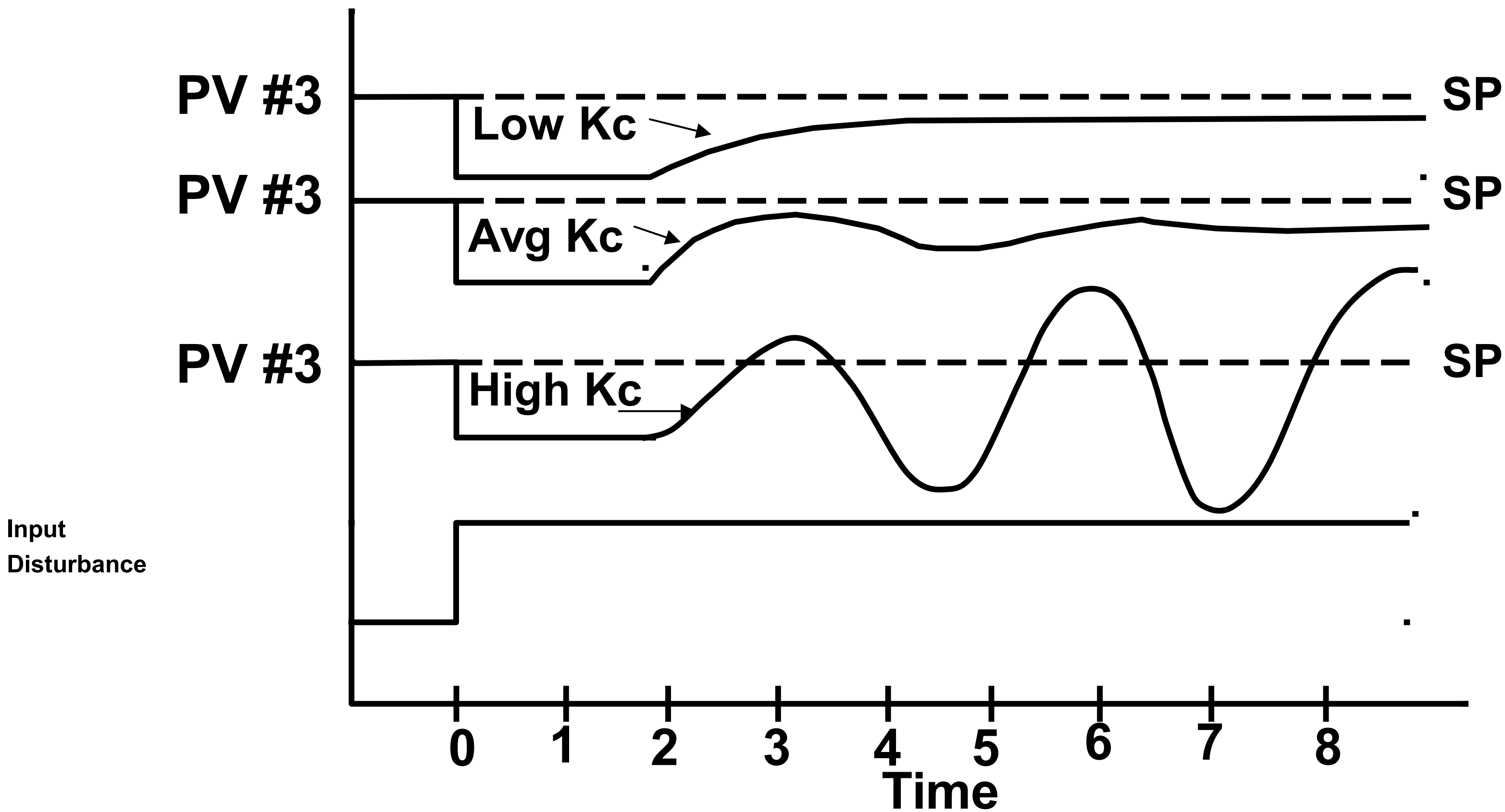
Proportional Control - Dead Time

Process Variable #3
Dead Time = 2.0
Input Disturbance at Time T=0



Question: Why does PV #3 tend to oscillate as the controller gain is increased?

Objectives of Integral Control

- **Eliminate Controller Offset**
- **Bring Controlled Variable to New Setpoint**

Integral Control Parameters

T_I = Reset Time or Integral Time (Minutes)

Decrease Reset Time to Speed Up Action

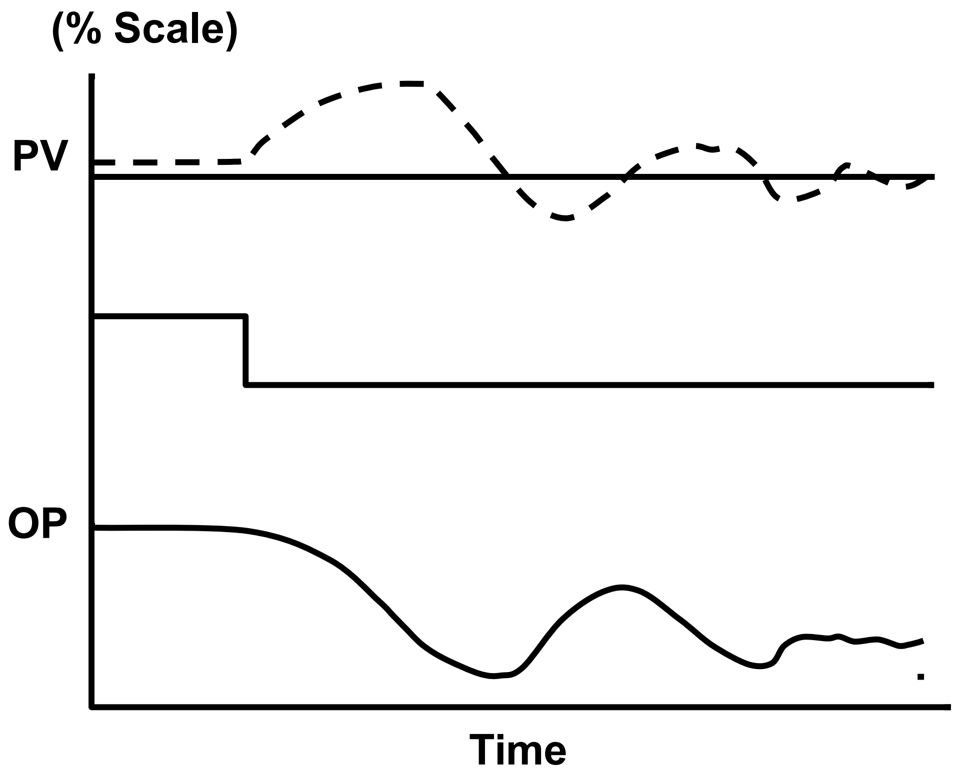
R = Reset Rate (Repeats per Minute)

Increase Reset Rate to Speed Up Action

$$\boxed{T_I = \frac{1}{R}}$$

Integral Control

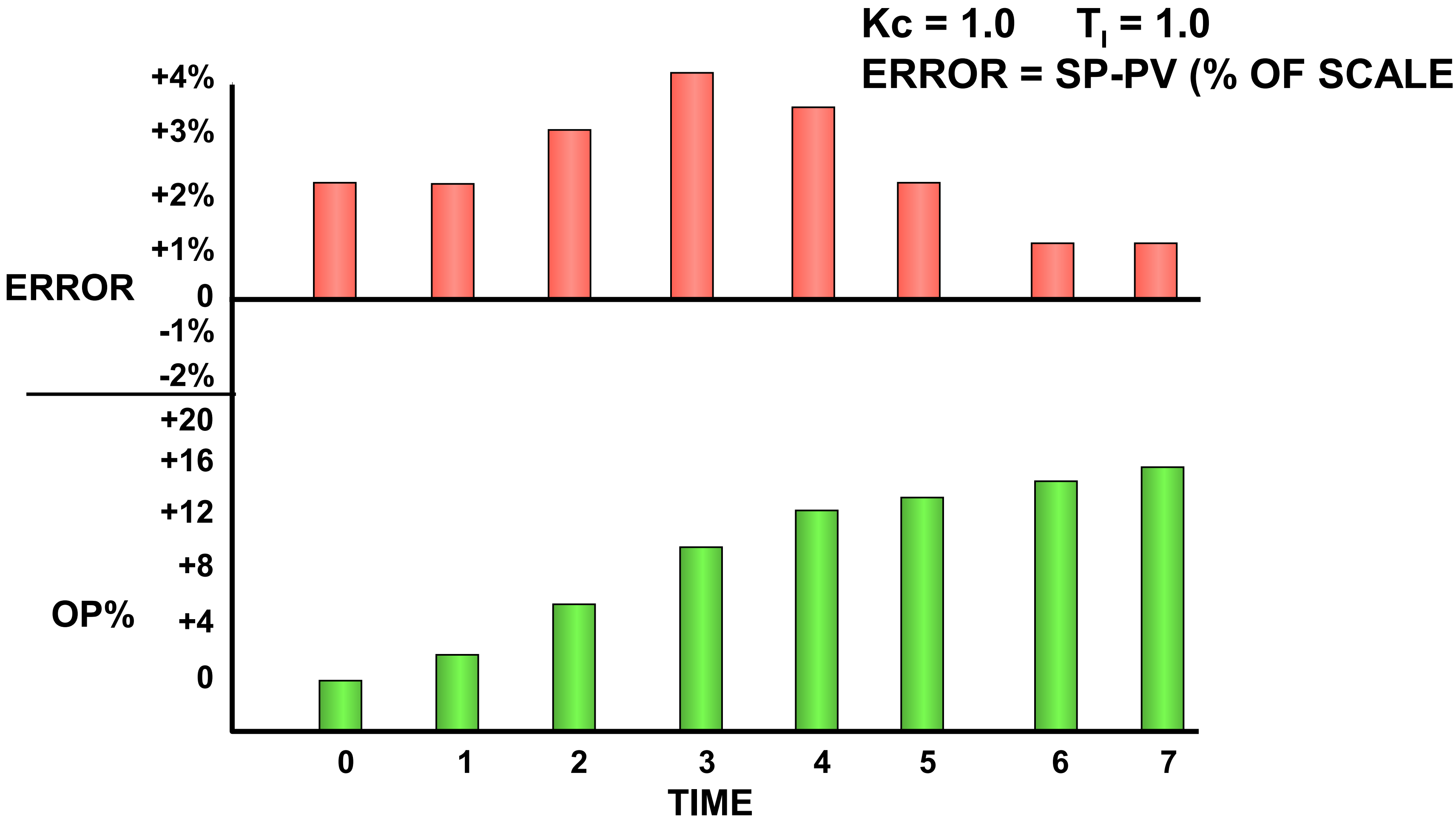
Error = SP-PV



- **Change in controller output proportional to current error.**
- **Integral control is used to eliminate steady state offset.**

Integral - Only Calculation
Output Change = [Kc/T_I] x [ERROR]
ΔOP = [Kc/T_I] x E
T_I = Integral (Reset) Time

Integral Control Example



Question: The process variable changes direction at T=3 and begins to return to setpoint. Why does the controller output continue to increase?

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Derivative Control

Error = SP-PV

Δ Error = Error - Old Error

Δ (Δ Error) = Δ Error - Old Δ Error

- **Change in controller output proportional to change of slope in error.**
- **Derivative control is used to give the controller a slight kick when the process variable begins to change, or when it changes direction.**
- **Implementation of derivative control requires in-depth knowledge. Noisy signals can cause derivative controllers to become oscillatory.**
- **Details of derivative control are beyond the scope of this course.**

Proportional - Integral - Derivative Control (PID)

PI CONTROL CALCULATION

$$\Delta OP = K_c \left[\Delta ERROR + \frac{1}{T_I}\left(ERROR\right) \right]$$

$$K_c = \text{Controller Gain}$$

$$T_I = \text{Integral (Reset) Time}$$

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Proportional - Integral - Derivative Control (PID)

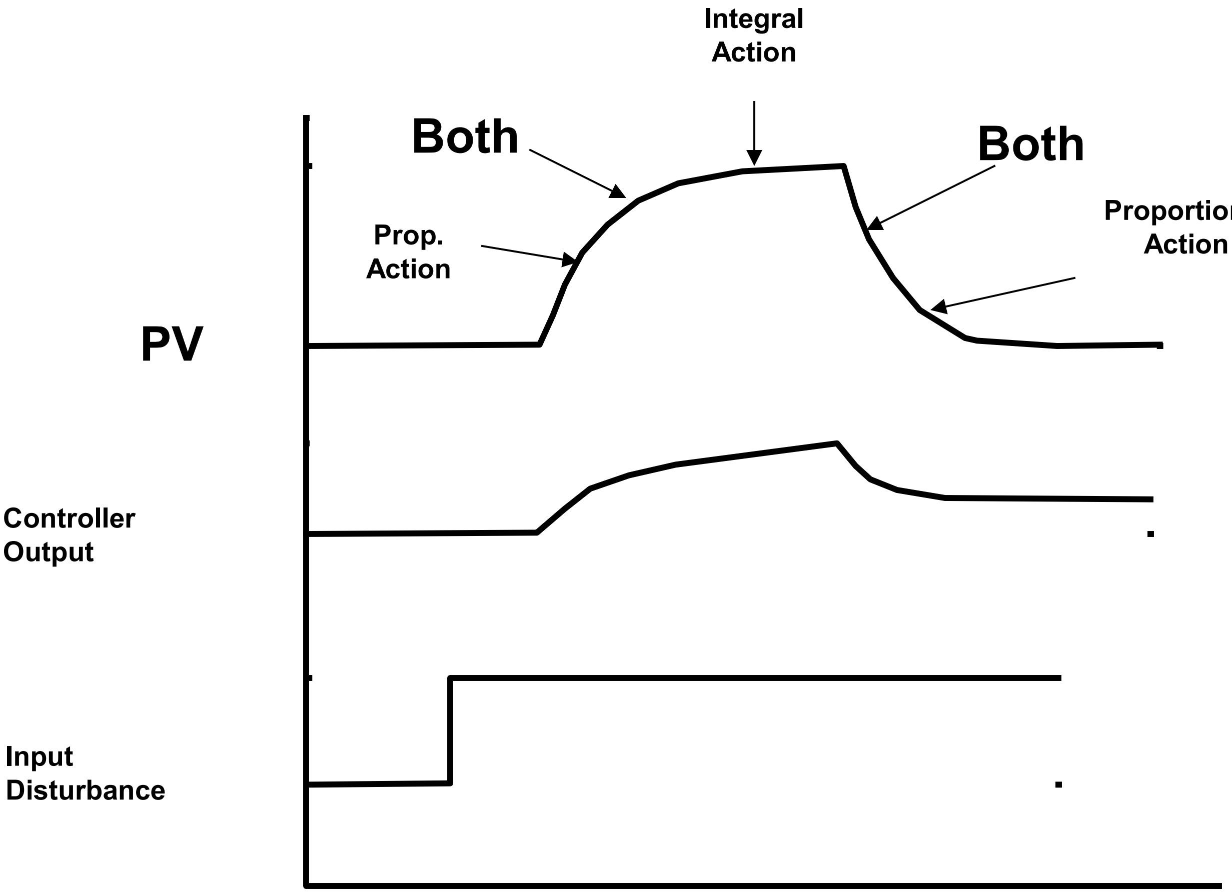
- **PID is the Most Commonly Used Control Algorithm in the Refining / Petrochemical Industry**

- **PID Control is the Sum of the Individual Elements, Proportional + Integral + Derivative Control**
- **PID Control Reduces to PI**

  (Proportional + Integral) Control,

  if Derivative Control is not Used

PID Control

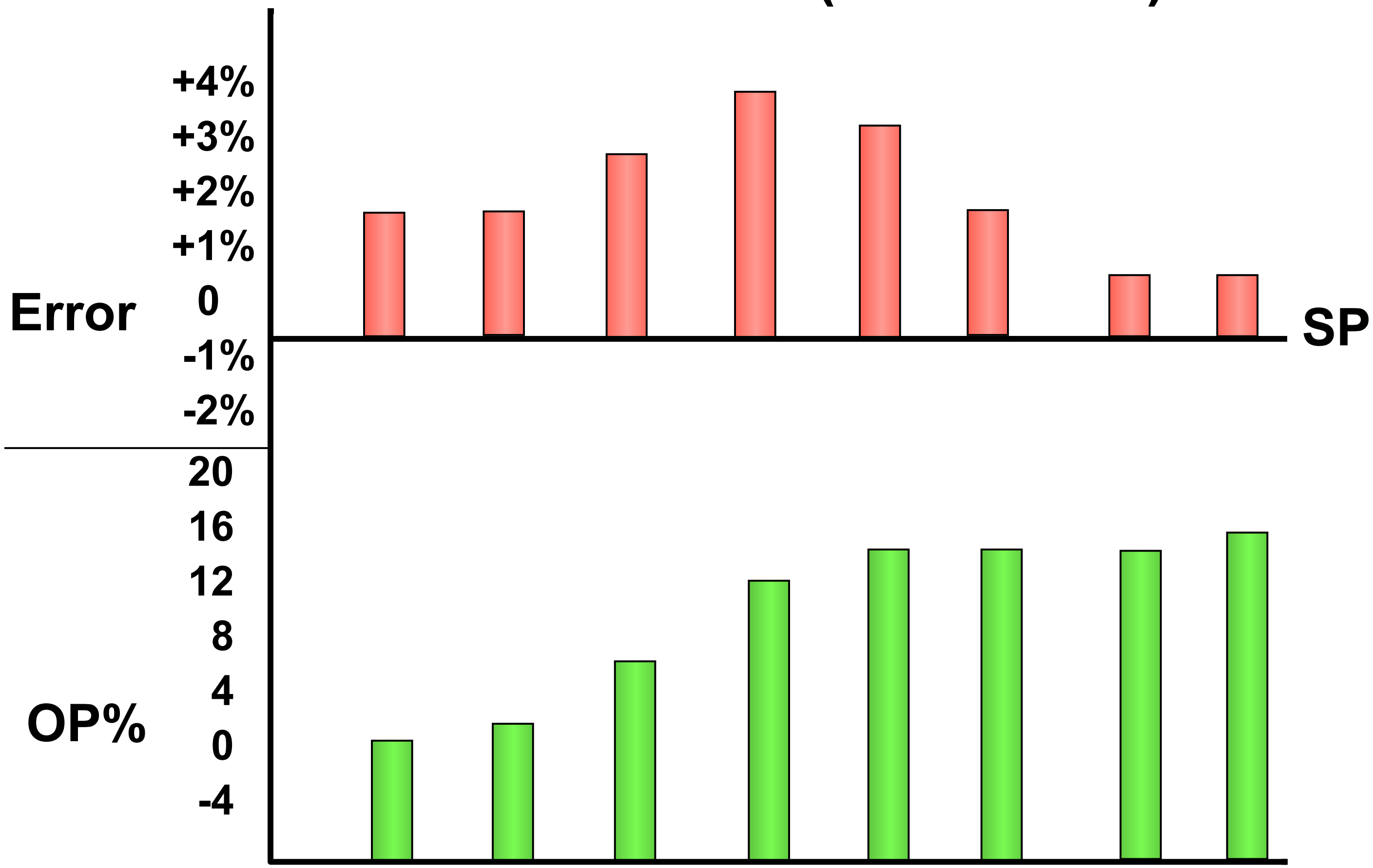


- Proportional action dominates when the PV is changing rapidly, but is close to SP.
- Integral action dominates when the PV is flat, but far away from SP.

PID Control Example

Kc = 1.0 T_I = 1.0

Error = SP-PV (% of Scale)



DCS Controller Tuning

- **Controller Tuning Procedure**

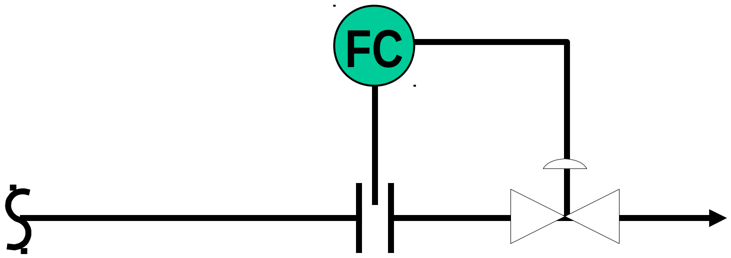
Controller Tuning
Define Process Objectives

- **Minimize Error Between PV and SP**
- **Minimize Movement of Controller Output**
- **Eliminate Swinging**
- **Respond to Upstream Disturbances**
- **Objectives must be Prioritized for each Controller; no Single Rule Applies for all Situations**

Controller Tuning Procedure

- Define Process Control Objectives
- Determine Type of Process: Pressure, Flow, Temperature, or Level
- Identify Process Dynamics with Response Test
- Use Tuning Formulas for First Guess
- Evaluate Performance
- Adjust as Necessary

Controller Tuning - Flow Controllers

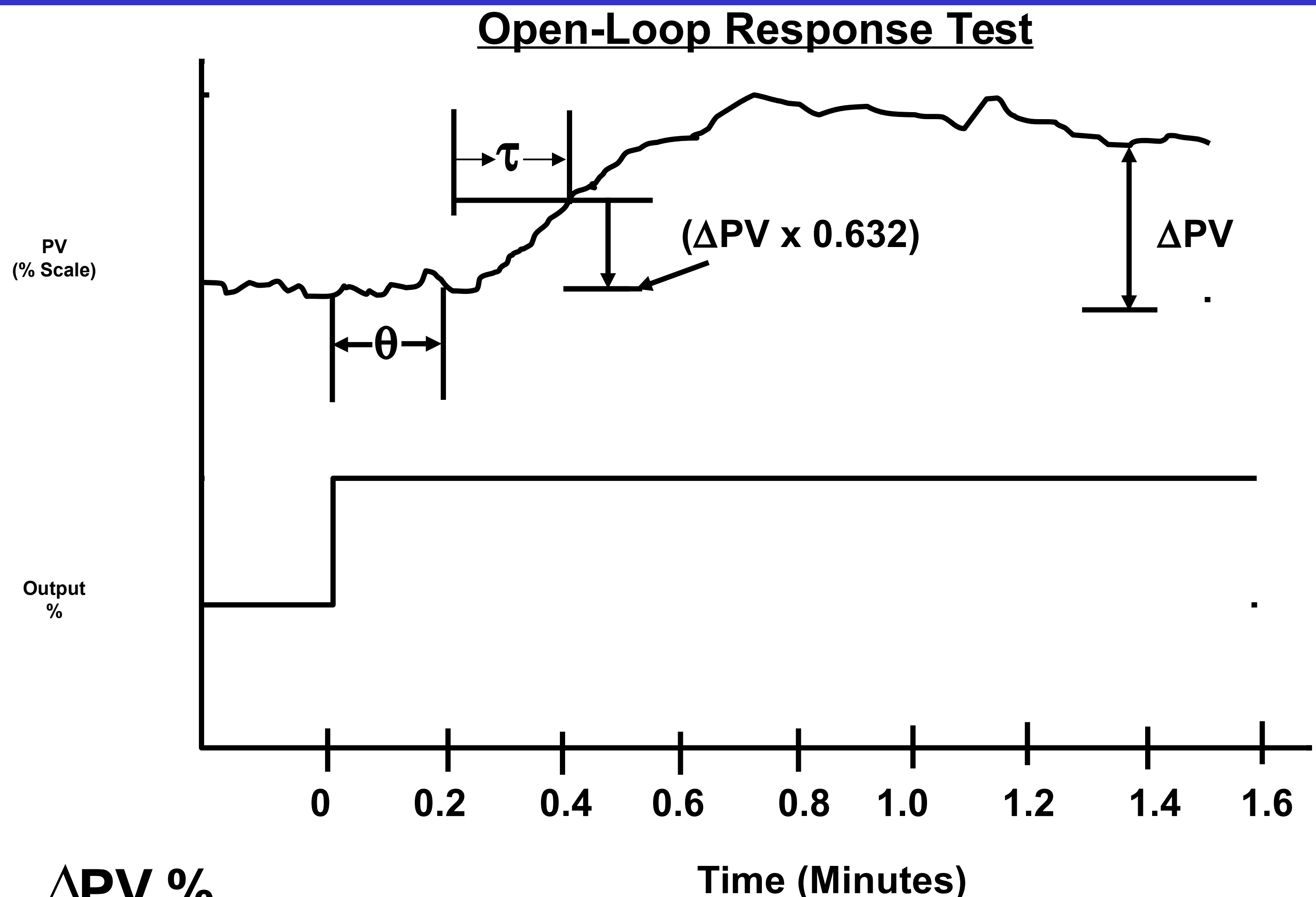


- First Order Dynamics
- Very Small Dead Time
- Generally Tune as Fast as Possible Without Swinging

<u>Outside Disturbances Affecting Flow Controllers</u>

- Change in Upstream/Downstream Pressure
- Change in Phase (Vapor/Liquid) Due to Composition or Temperature Change

Controller Tuning - Flow Controller



$$Kp = \frac{\Delta PV\ \%}{\Delta OP\ \%}$$ (Process Gain)

Controller Tuning - Flow Controller

Tuning Formula

$$Kc = \frac{1}{Kp} \times \left[\frac{\theta + 2\tau}{2\theta + 3\tau}\right] \text{ Controller Gain}$$

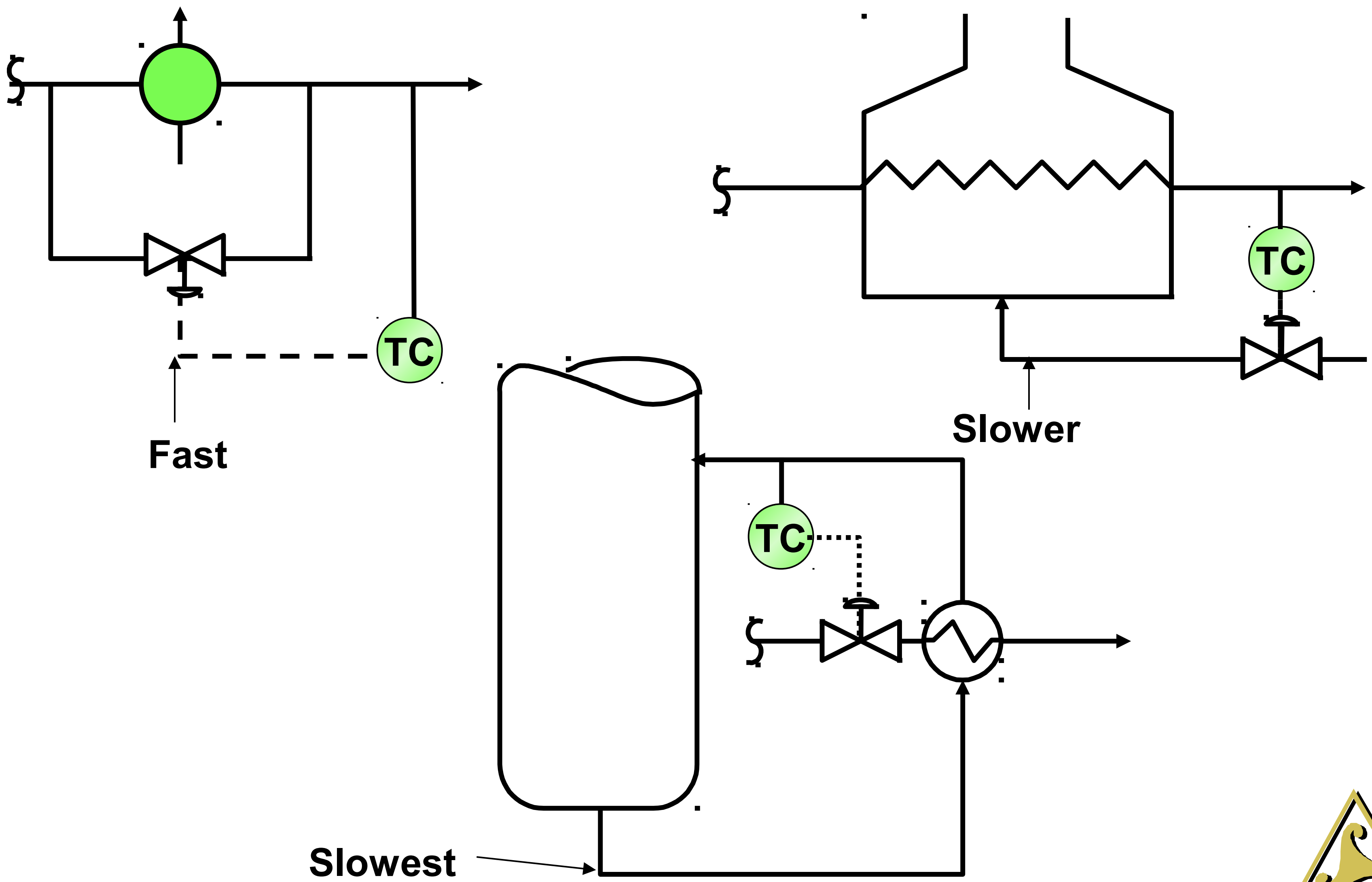
[PB = Proportional Band = 100/Kc]

$T_I = \theta + (0.5)\,\tau$

Reset Time (Minutes)

$$R = \frac{1}{T_I} = \text{ Repeats/ Min}$$

Controller Tuning - Temperature Controllers



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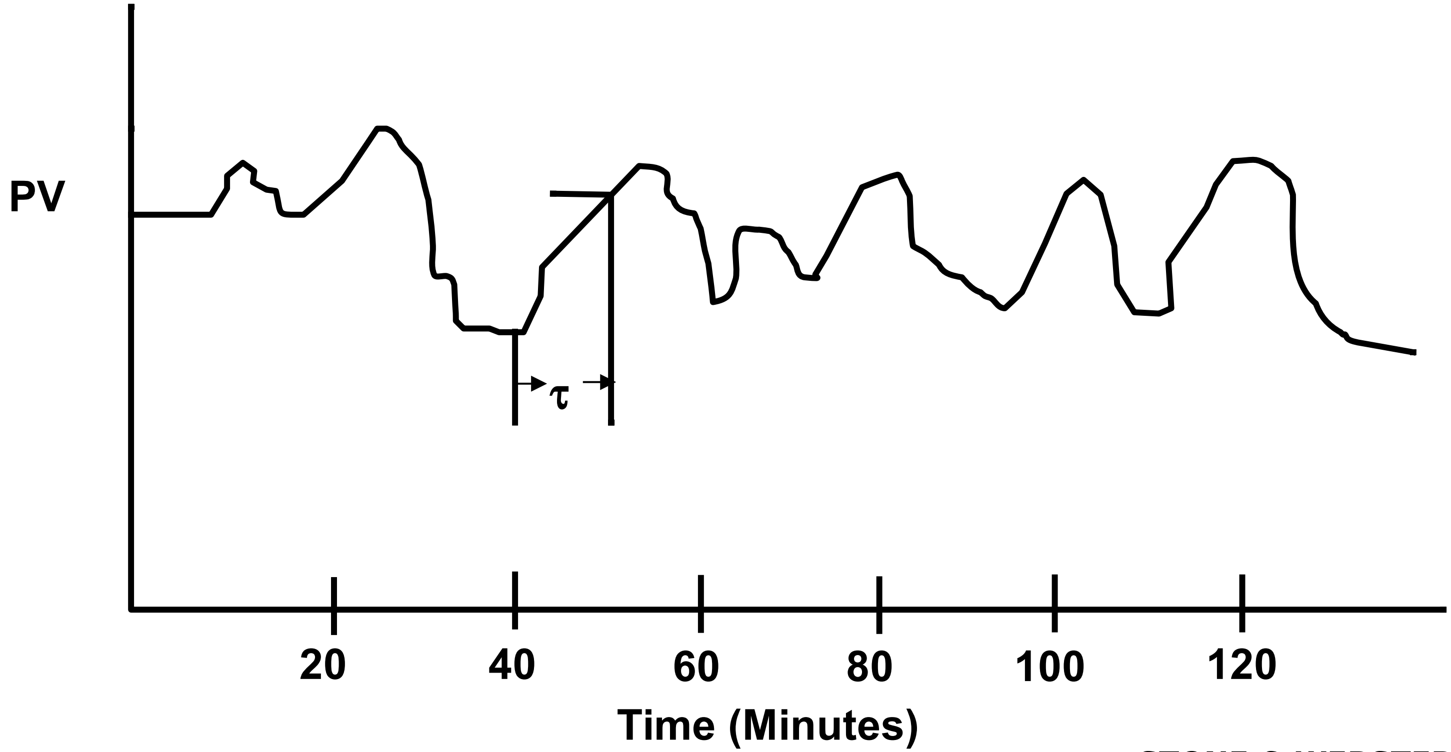
Controller Tuning - Temperature Controllers

- Use Same Response Test and Tuning Formula as Flow Controller
- Allow Response Test to Run to Completion (Sometimes more than 1/2 Hour for Columns)
- Use Trend System for Additional Data



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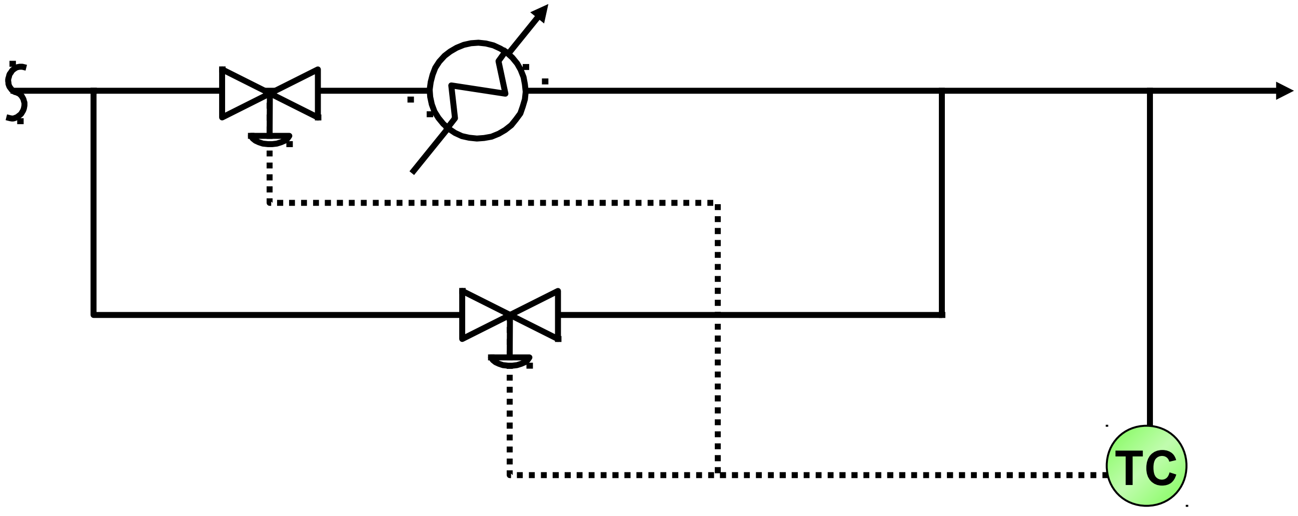
Controller Tuning - Temperature Controllers

Bypass Controllers

- Fastest of all Temperature Controllers
- Can be One or Two-Valve System
- Rangeability and Windup can be Problems

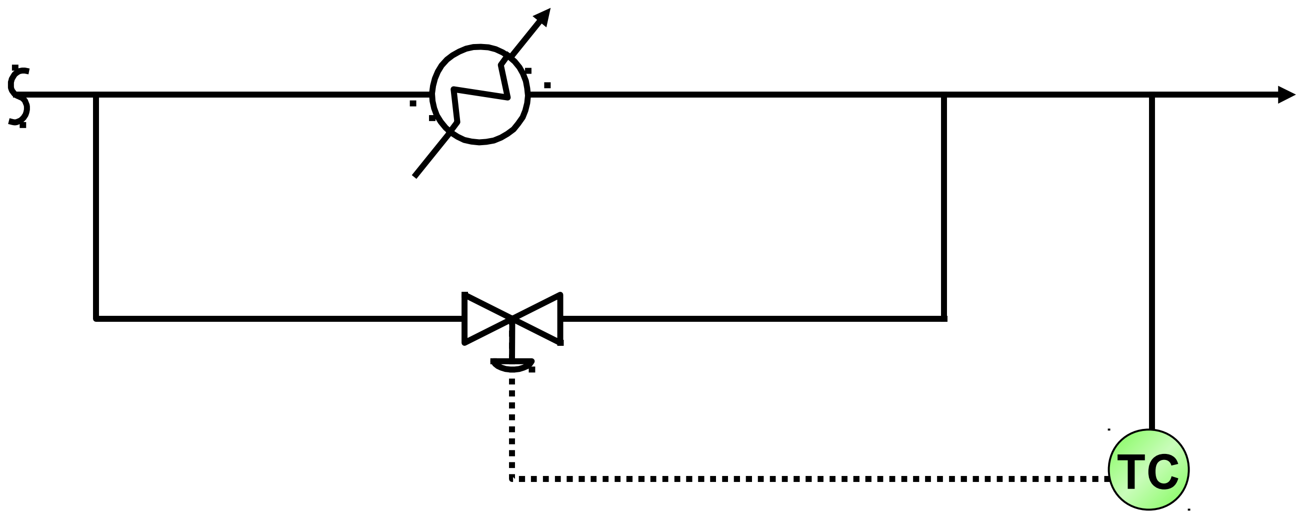
1)



Controller Tuning - Temperature Controllers

- Which Valve are we Operating On?
- Are we Having Problems when one of the Valves Begins to Open or Close?

2)

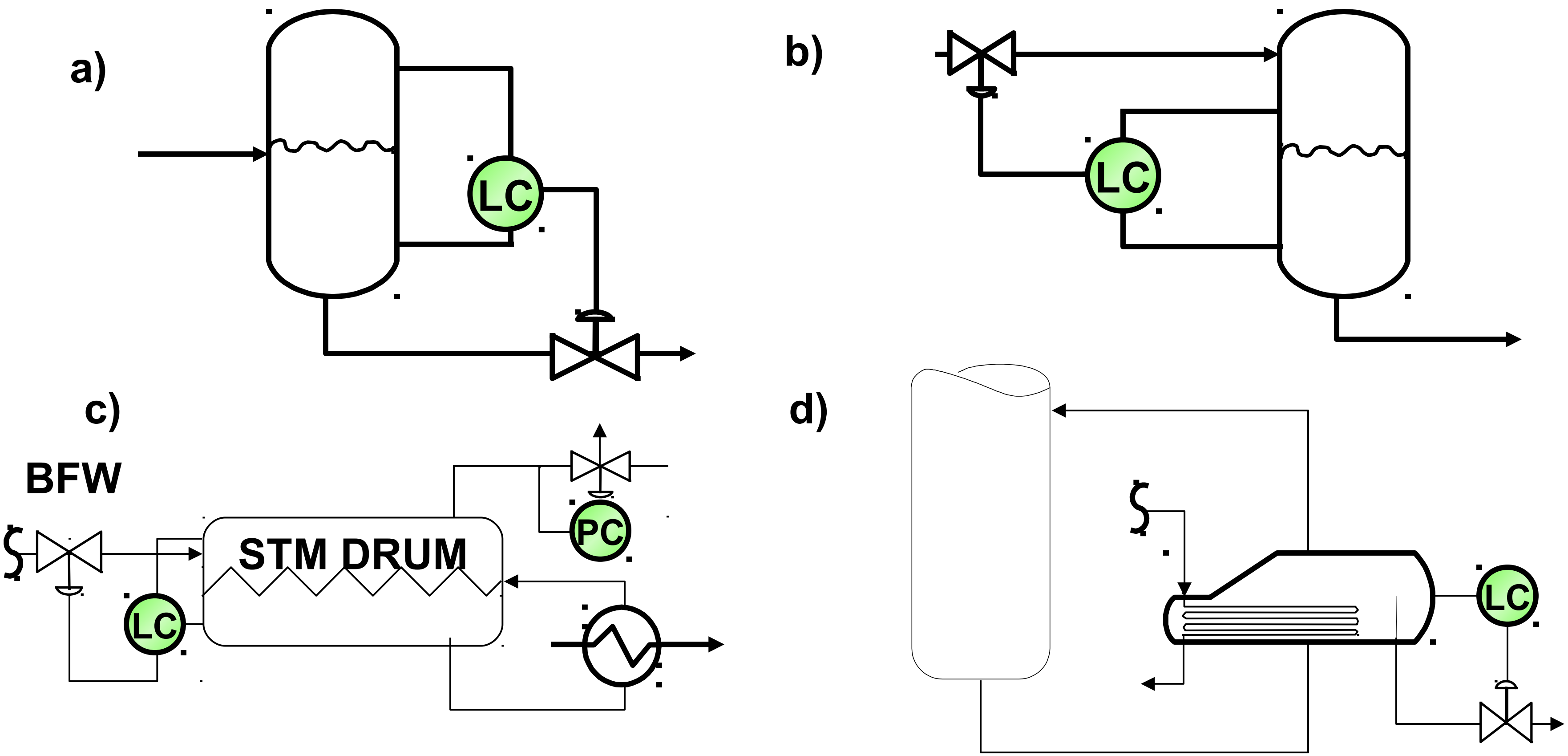


- **Does this controller wind up?**

Objectives of Level Control

- **Reject Normal Feed Rate Disturbances**
- **Prevent Level From Exceeding High or Low Limit**
- **Use Vessel Volume to Minimize Flow Change to Downstream Units**
- **Restore Level to Setpoint in Controlled Fashion After Disturbance Has Passed**

Controller Tuning - Level Controllers

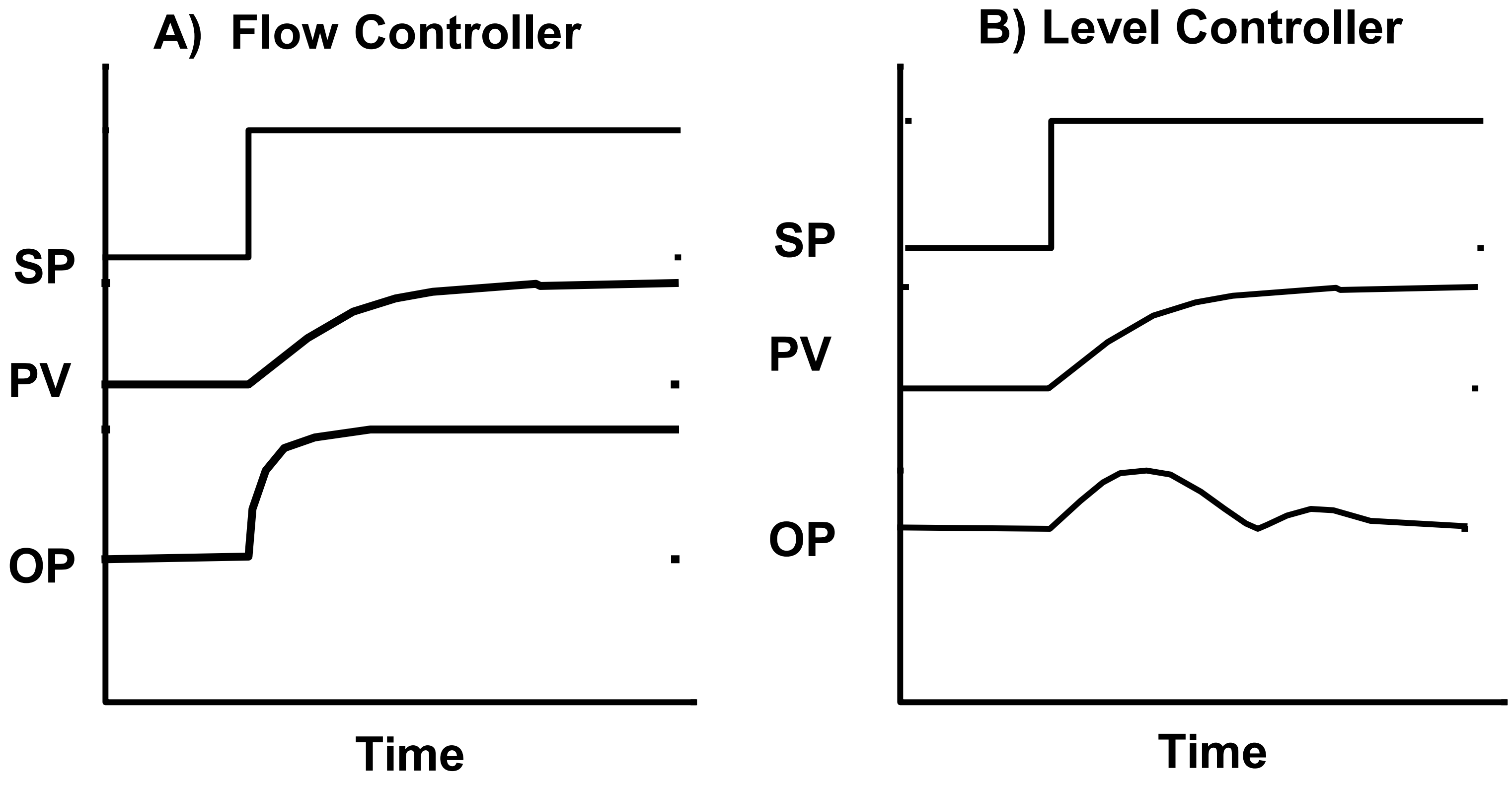


- Objectives different in each case.
- Outside disturbances different in each case.
- Know the process.

Controller Tuning - Level Controllers

Understanding How Ramp Processes Differ from Stable Processes

<u>**Setpoint Change**</u>



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Controller Tuning Level Controller

Level Control Consists of Two Distinct Events:

- **Turning the Level When a Disturbance Causes It to Change (Proportional)**
- **Bringing the Level Back to Setpoint After It Has Been Turned (Integral)**

Controller Tuning Level Controller

In Order to Tune a Level Controller, We Must Decide (Process Decision) How Quickly We Want These Events to Occur

<u>Limits to Consider</u>

- **Dead Time/oscillations**
- **Rate of Change of Controller Output**
- **High/low Level Limit**

Simple Method Controller Tuning - Level Controller

1) Turn the Level

$$Kc = \frac{1}{Kr} \cdot \left(\frac{\theta + 1}{2\theta + 1} \right) \cdot [\text{Speed Factor}]$$

Speed Factor <0.5

2) Bring Level Back to Setpoint

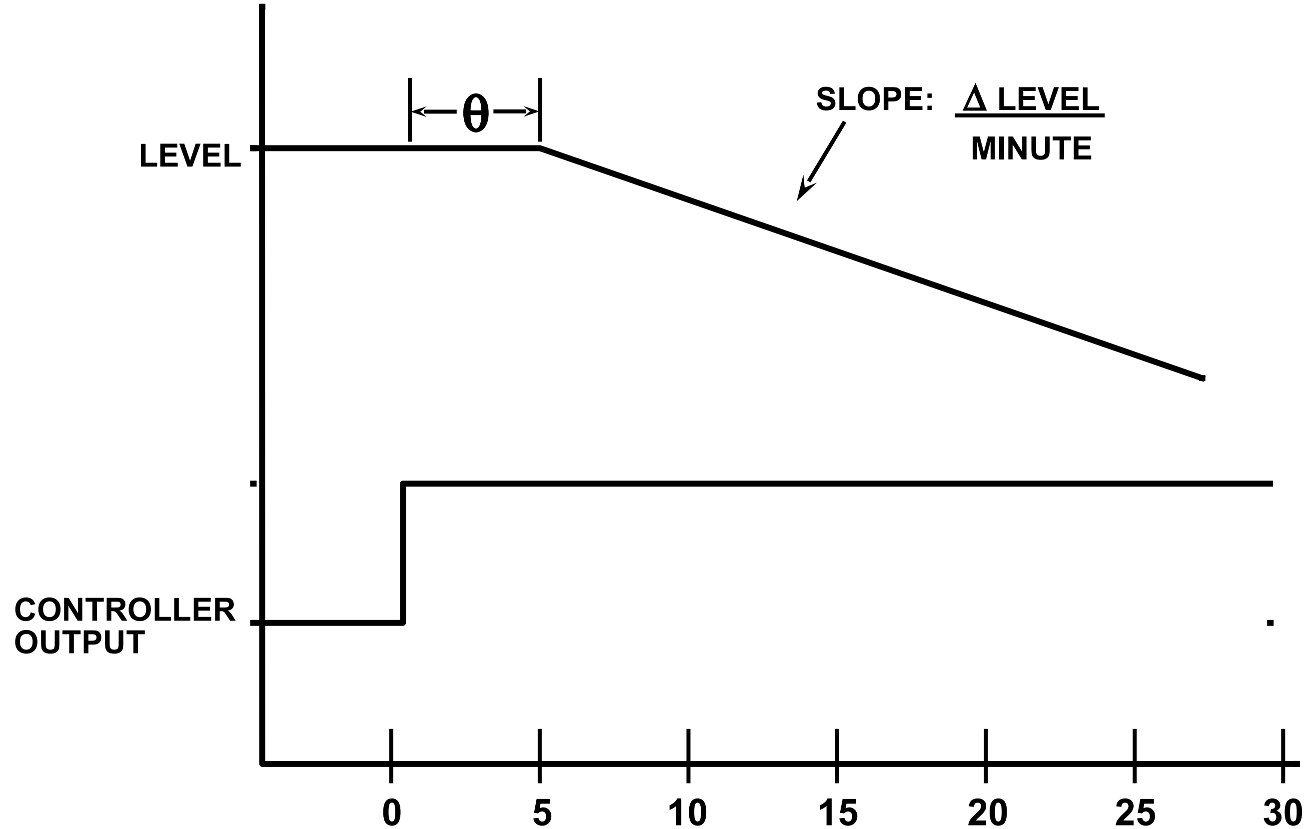
T_I = 20 min (Average)

T_I = 10 min (Fast)

T_I = 40 min (Slow)

Controller Tuning: Level Controllers (Simple Method)

RESPONSE TEST



$$\text{PROCESS GAIN: } K_R = \frac{\frac{\Delta\ \text{LEVEL}}{\text{MINUTE}}}{\Delta\ \text{CTR. OUTPUT}}$$

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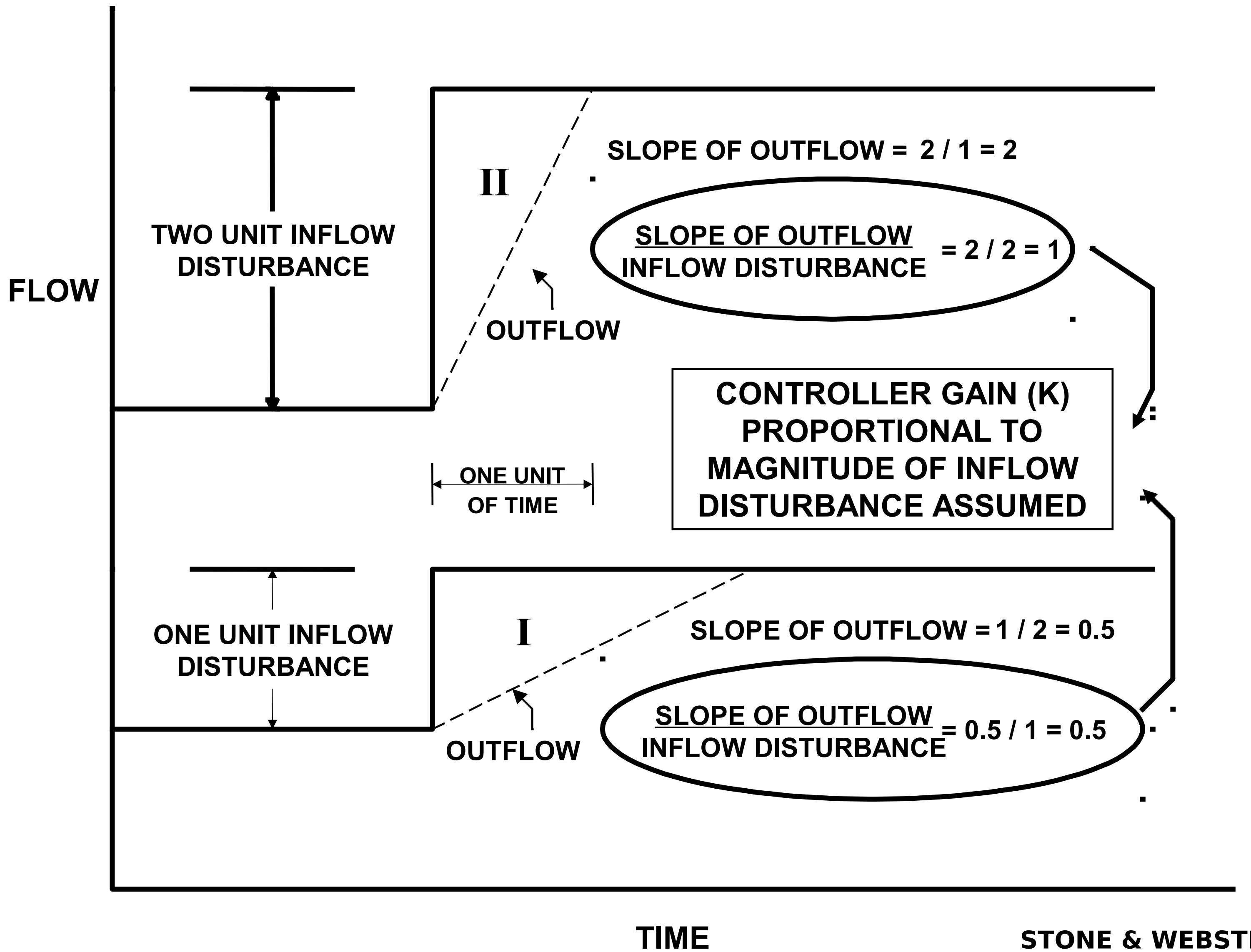
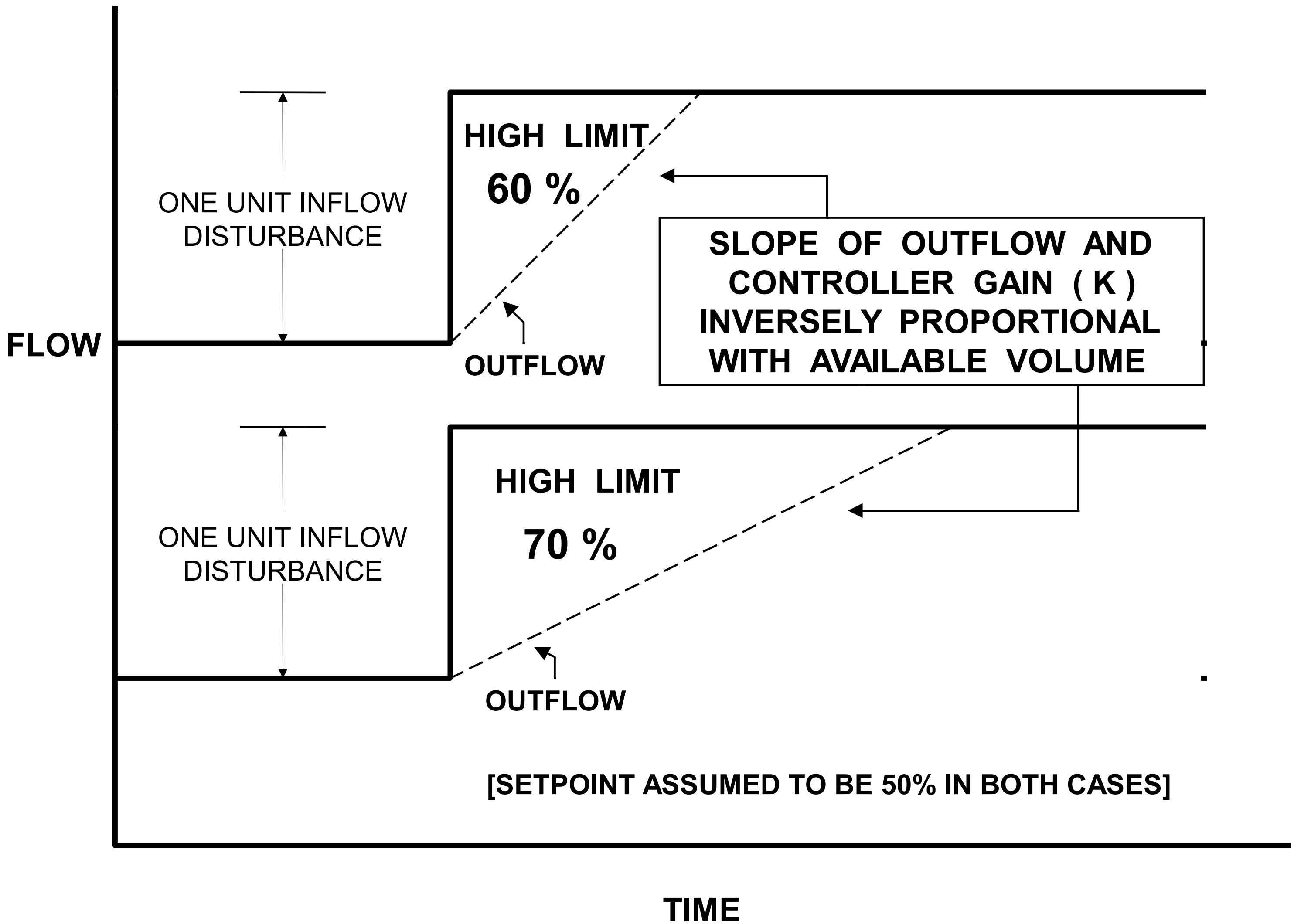
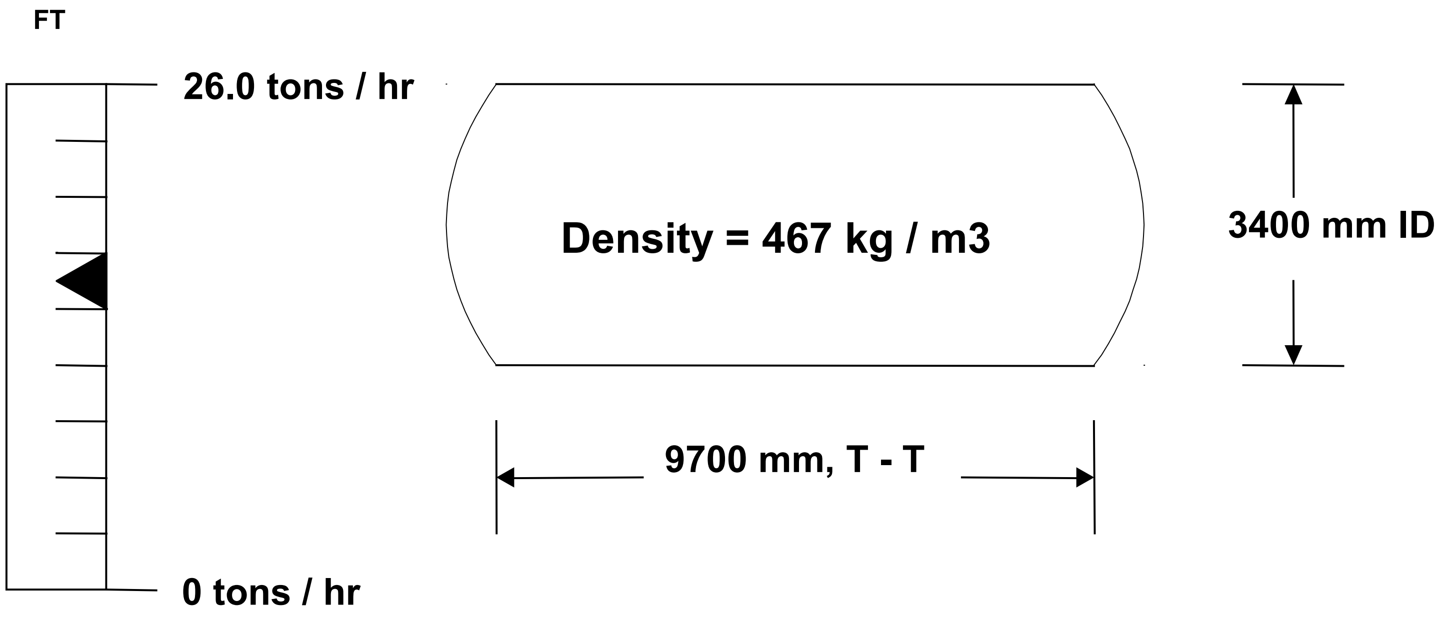
Effect of Magnitude of Inflow Disturbance



Effect of Desired High or Low Level Limit



Level Control Tuning, Example Problem



Determination of Controller Gain, K, for Level Control Applications

1. Assume 'D%', magnitude of inflow disturbance, % level/min.

   (Largest disturbance that occurs in normal operations without failure of process equipment).
2. Calculate 'M', mass of liquid contained in 1% of level.
3. Calculate mass magnitude of inflow disturbance = D% x M.
4. Define approximate 'upper level limit'. Calculate maximum accumulation, 'A', mass of liquid contained between upper limit and setpoint.

   A = M x (Upper limit-setpoint)
5. Calculate average flow imbalance = (D% x M) / 2.

Determination of Controller Gain, K, for Level Control Applications

6. Calculate time available to equalize inflow with outflow,

   T, min = (2 x A) / (D % x M) = 2 x (% lim - % SP) / D (%/min)

7. Calc. Req'd mass flow change per minute, F=(D% x M) /T.

8. Convert required mass flow change to percent of scale,

   F % per minute = (D x M) x 100 / Flowmeter range.

9. Estimate controller gain K = F% / D%.

10. Set reset time as follows: 8 min, rapid return to setpoint

    12 min, avg return to setpoint

    20 min, slow return to setpoint

Level Control Tuning, Example Problem

1. CROSS SECTION AREA = 3.4 m X (9.7 m + 1.7 m) = 38.9 m2
2. M = MASS OF LIQUID IN 1% LEVEL = 38.9 m2 X 0.031 m X 467 kg / m3
   = 563 kg / % level
3. PREDICTED DISTURBANCE = D % = 0.1 % level / min
4. MASS FLOW DISTURBANCE = 0.1 % X 563 kg / % = 56.3 kg / min
5. AVERAGE FLOW IMBALANCE = 56.3 kg / min X 60 min / hr / 2 = 1689 kg / hr
6. UPPER LEVEL LIMIT = 55% thus A = 563 kg / % level X 5% = 2815 kg
7. TIME AVAILABLE = 2815 kg / 1689 kg / hr = 1.67 hr = 100 min.
8. AVG FLOW CHANGE = 56.3 kg / min * 60 min / hr / 100 min = 33.8 kg/hr / min
9. % SCALE FLOW CHANGE = 33.8 kg / hr X 100 / 26000 kg/hr = 0.13 % / min
10. CONTROLLER GAIN K = 0.13 % / 0.1 % = 1.3

Controller Tuning
Pressure Controllers

- Pressure Control is a Ramp Process Similar to Level Control
- Pressure Measures Vapor Inventory; Level Measures Liquid Inventory
- In General, Pressure Needs to be Controlled Tightly, Because Vapor/Liquid Equilibrium is Affected
- This is Accomplished by Increasing Controller Gain

$$Kc = \frac{1}{Kr} \cdot \left(\frac{\theta + 1}{2\theta + 1} \right) \cdot [\text{Speed Factor}]$$

0.5< Speed Factor <0.8

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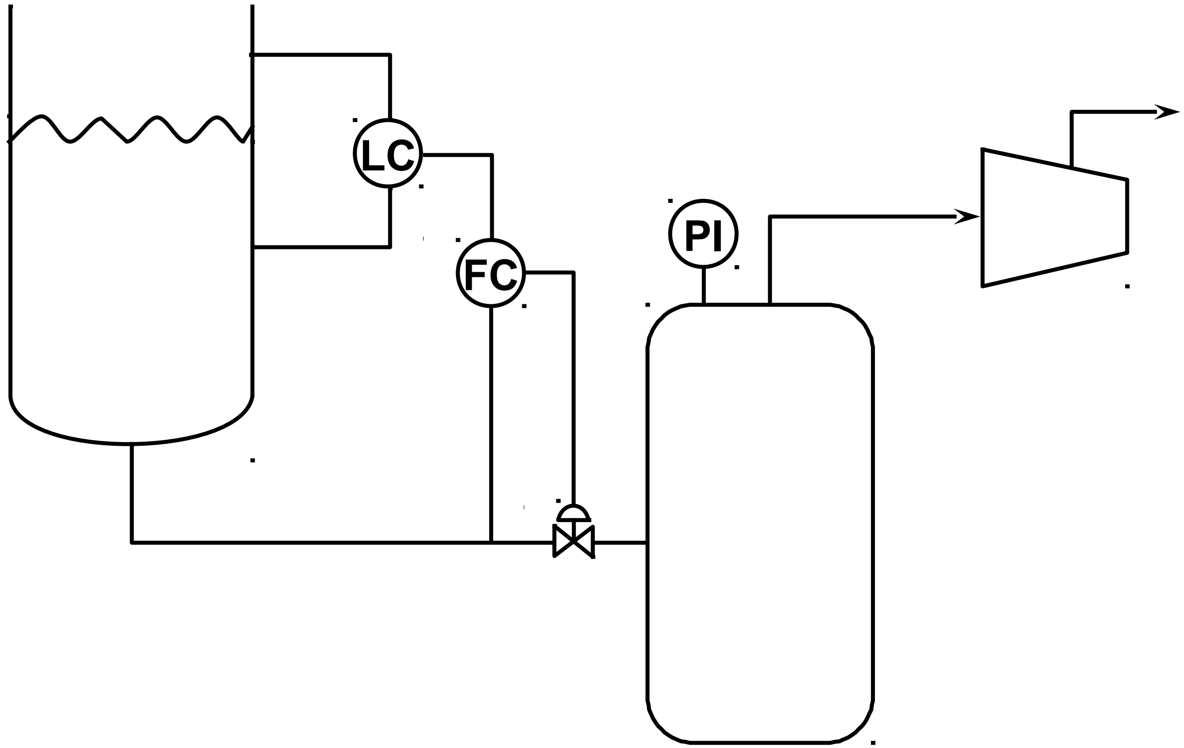
Cascade Control

- Why Use Cascade Control?
- To Reject Measurable Disturbances That Have Faster Dynamics Than The Primary Controller

EXAMPLE

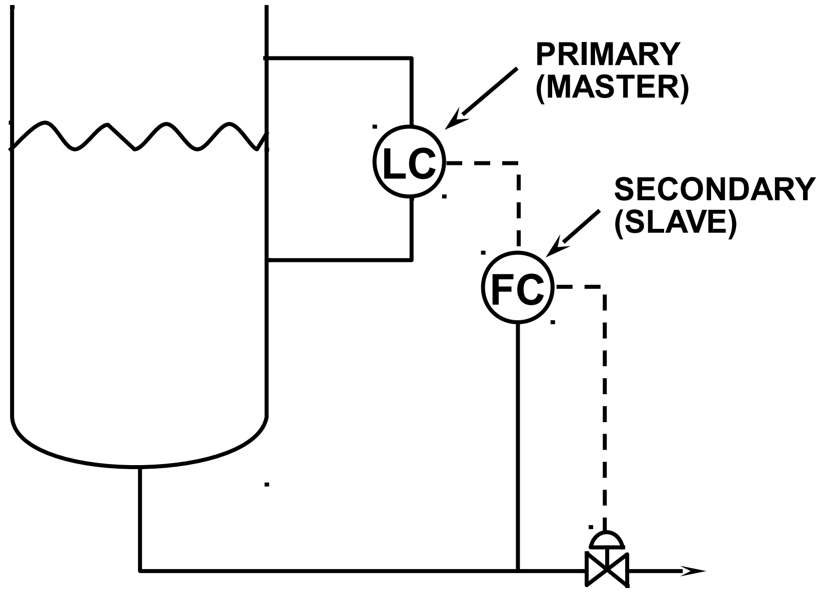


- The FC responds to a change in downstream pressure sooner than the LC.
- The FC can be tuned fast because flow control has faster dynamics than level control.

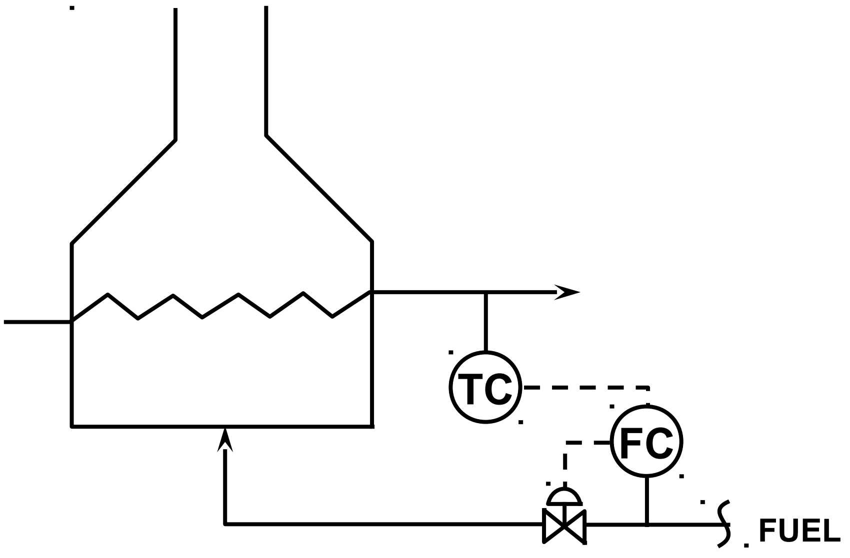
Cascade Control

CONTROL SYSTEM WHERE OUTPUT OF PRIMARY CONTROLLER BECOMES SETPOINT OF SECONDARY CONTROLLER

A.)



B.)



Cascade Control

<u>**Requirements**</u>

- **Secondary (inner) loop dynamics must be at least 3-5 times faster than primary loop.**
- **Sensors and control hardware for the secondary loop must be installed/maintained.**
- **The PV scale of the secondary loop must have wide enough range so that the primary loop does not wind up prematurely.**

Tuning Cascade Controllers

- Tune from the inside out
- Put primary (outer) loop in manual
- Tune secondary (inner) loop
- Tune primary (outer) loop

Note: reset time of primary loop (T), must be at least 3-5 times as great as secondary loop, otherwise, primary loop may ‘chase’ secondary

DCS Controller Tuning

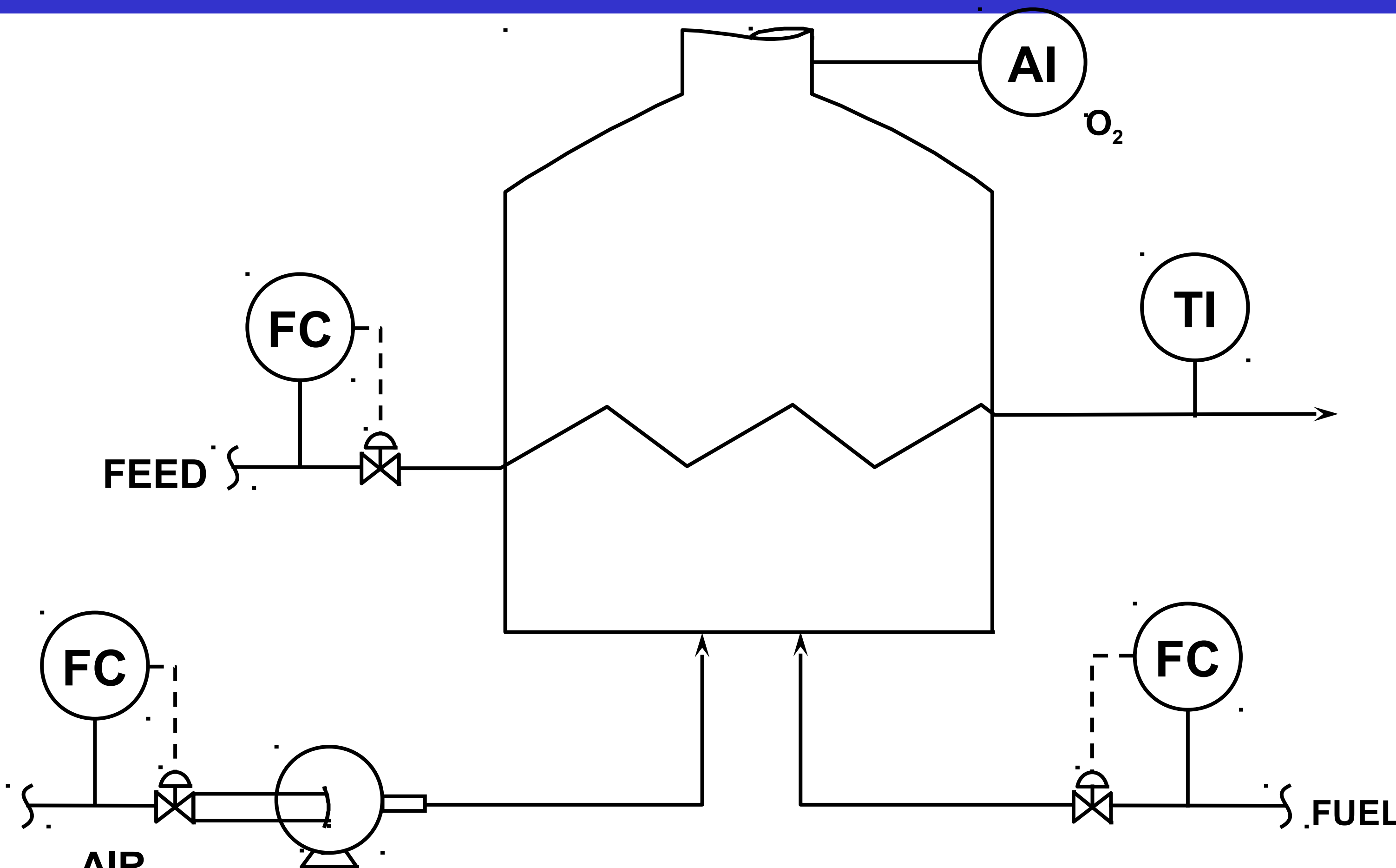
- **Introduction to Multivariable Control**

Multivariable Control

- Simultaneous manipulation of multiple independent variables in order to control multiple dependent variables.

- The primary advantage of multivariable control is that interactions between variables are accounted for in the simultaneous move calculation.

Multivariable Control Example



(INDEPENDENT)
MANIPULATED
VARIABLES

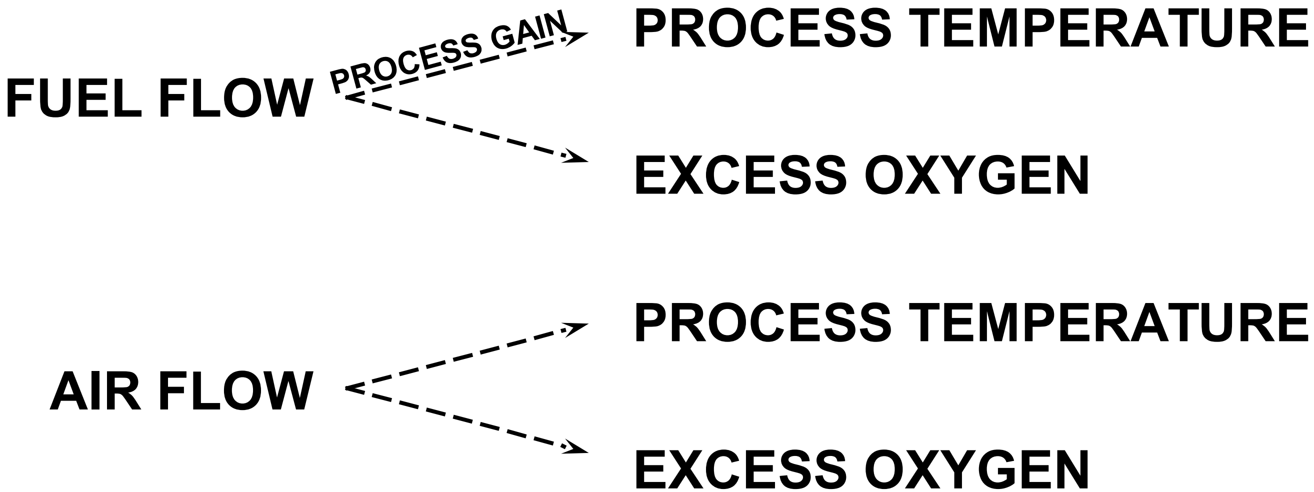
FUEL FLOW
AIR FLOW

(DEPENDENT)
CONTROLED
VARIABLES

PROCESS OUTLET TEMPERATURE
FLUE GAS EXCESS O_2

STONE & WEBSTER

Variable Interaction



Variable Interaction

MULTIVARIABLE GAIN MATRIX

	PROCESS TEMP.	EXCESS OXYGEN
FUEL FLOW	G11	G12
AIR FLOW	G21	G22

Multivariable Control Strategy Degrees of Freedom

Definition

Degrees of freedom = # independent manipulated variables
- # controlled variables

If D.F. = 0, then there is a unique steady state solution

If D.F. < 0, then the problem is constrained

If D.F. > 0, then the problem has infinite steady state solutions and optimization is possible

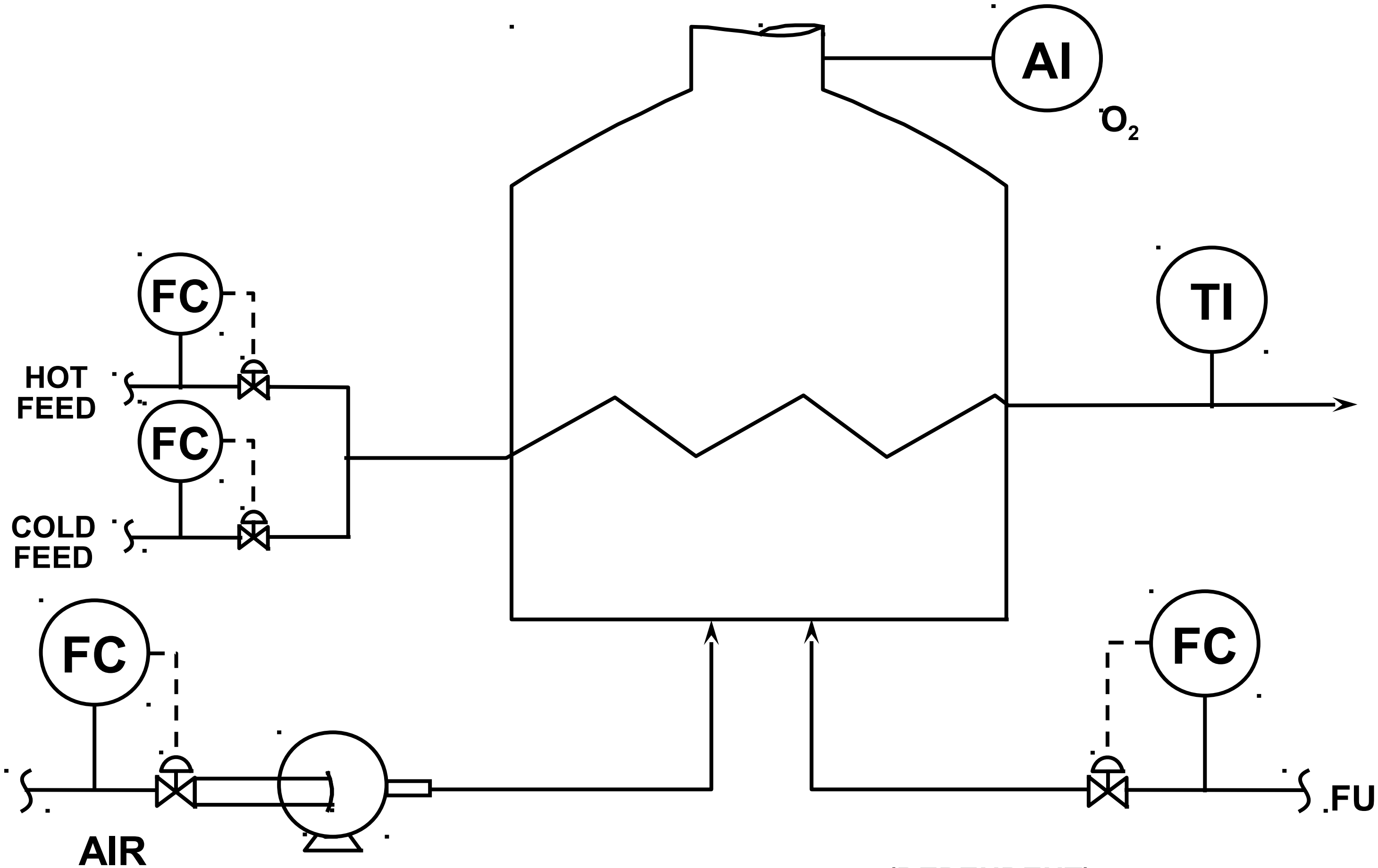
Using the Multivariable Controller To Control Constrained Systems

- Controlled variables must be prioritized by some method
- Prioritization of variables is the primary method of enforcing multivariable control strategy
- APC engineer must consider all constraint scenarios when defining priorities
- Degrees of freedom can change as manipulated variables reach their maximum or minimum limits
- This means that a given APC problem can change among the 3 types
  - Optimizing
  - Square
  - Constrained

Using the Multivariable Controller to Optimize Unit Operation

- Optimization is used to drive manipulated variables towards an economic or operational optimum.
- The optimum point is often at either the maximum or minimum end of the operating range.
- Optimization strategies must be prioritized relative to constraint priorities.
- More complicated optimization strategies require a more rigorous, model-based optimization system.

Optimizing System



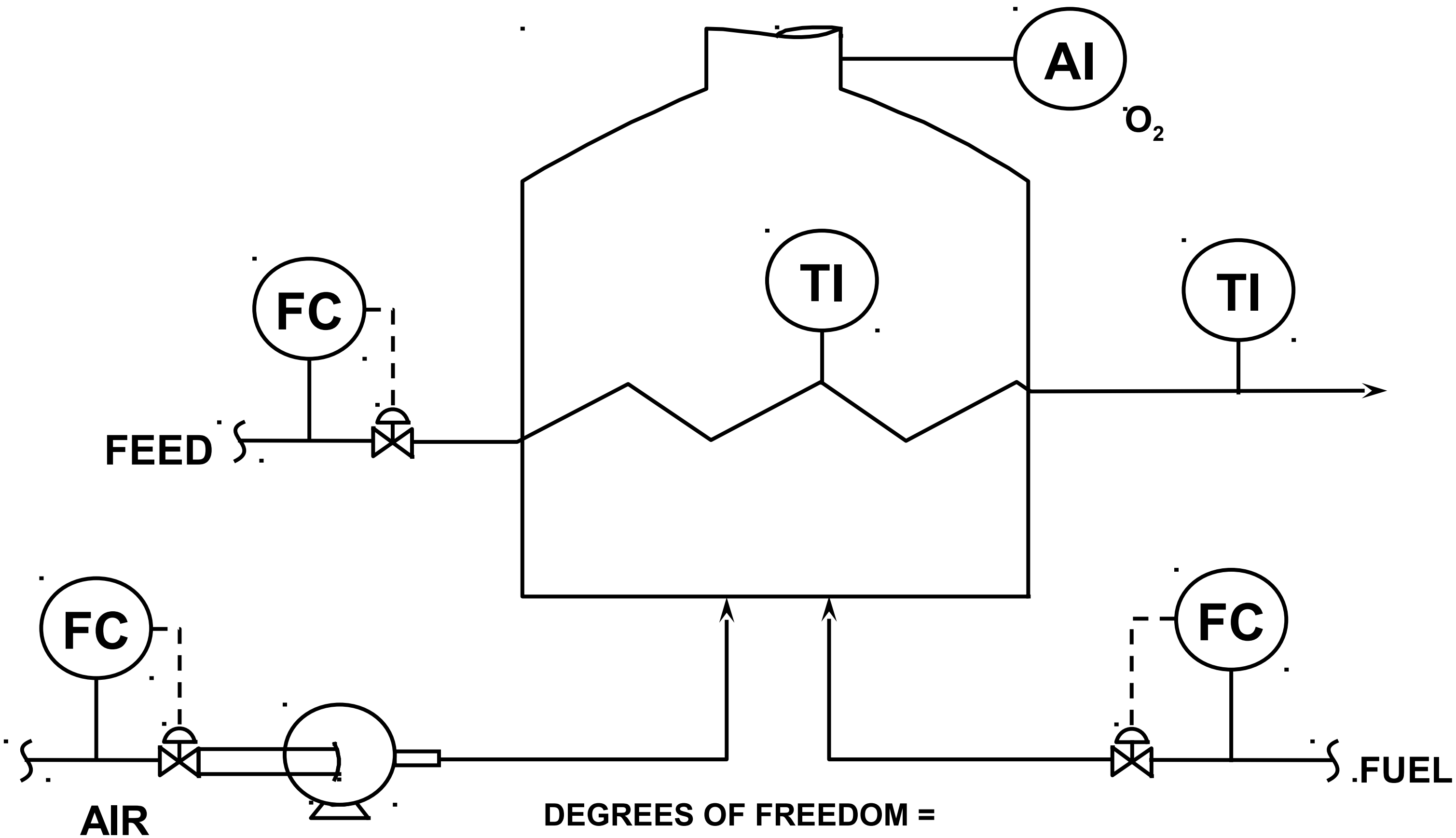
(INDEPENDENT)
MANIPULATED
VARIABLES

FUEL FLOW
AIR FLOW
HOT FEED FLOW
COLD FEED FLOW

(DEPENDENT)
CONTROLED
VARIABLES

OUTLET TEMPERATURE
FLUE GAS EXCESS O_2
TOTAL FEED

Constrained System



DEGREES OF FREEDOM =

(INDEPENDENT)
MANIPULATED
VARIABLES

FUEL FLOW
AIR FLOW

(DEPENDENT)
CONTROLED
VARIABLES

PROCESS OUTLET TEMPERATURE
FLUE GAS EXCESS O_2
TUBE METAL TEMPERATURE

Multivariable Controller Calculation

- Calculation of Current Error

  Ej = (SP-PV) j for Each Controlled Variable (j)
- Calculation of Predicted Future Error

  Ej (Future)

$$= Ej(current) + \sum_{1=1}^{MV} (G_{ij})(X_i)$$

  - where G_{ij} = Process Gain
  - X_i = Moves of # Independent
- Calculation of Multivariable Solution

  $G_{11} X_1 + G_{12} X_2 + \dots G_{1n}X_n - E_1 = 0$

  $G_{21} X_1 + G_{22} X_2 + \dots G_{2n}X_n - E_2 = 0$

  $G_{n1} X_1 + G_{n2} X_2 + \dots G_{nn}X_n - E_n = 0$

Solve for Xi by Mathematical Technique

Controller Tuning - Benefits

- Simplifies reconciliation of test data
- Increases credibility of conclusions
- Permits operation at higher capacity
- Allows more precise identification of limits for expansion planning
- Generates good will with client

Use of Controller Tuning Skills

- **Generate “first guess” tuning constants from design data only**
- **Make adjustments during operation**
- **Solve operating problems**

Initial Tuning from Design Data

- Flow Controllers
- Temperature Controllers
- Level Controllers
- Pressure Controllers

Initial Tuning - Flow Controller

- Estimate process gain Kp
  - Kp ~ 1.0 % flow / % valve for flow controllers
- Controller gain Kc ~ 0.6 / Kp
- Reset time ~ 0.5 min for flow control

Initial Tuning - Temp Controller

- **Use design data to estimate incremental energy for 1 ° change**
- **Convert 1 ° to % scale**
- **Use design data to calculate change in heat medium flow required**
- **Assume 1% valve = 1% design flow**
- **Calculate Kp % / %**

Initial Tuning - Temp Controller

- Kc ~ 0.5 / Kp
- Reset time depends on system inventory
  - About 2 - 3 min for once through exchanger
  - About 5 - 7 min for tower bottoms
  - About 10 min for large recirculated system

Controller Tuning Etiquette

- Get permission from Client engineer
- Share tuning strategy with operator
- Make sure operator has time to monitor performance
- Make sure you have time to monitor performance
- Keep records of changes

Controller Tuning Efficiency

- Minimize impact on schedule
- Be prepared to take advantage of opportunity windows
- Plan other tasks that can be performed while monitoring tuning
- Establish partnership with operator

Controller Tuning as Deliverable

- Identify controller tuning as project deliverable
- Add hours to budget as necessary
- Include in lump sum estimates
- Submit change order if not in original scope of work
- Don’t give away what can be sold