PROCESS CONTROL FUNDAMENTALS

For The Pulp & Paper Industry

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TAPPI Process Control Textbook
Task Force of the TAPPI PCE & I Division

But main brain behind is W.L. Bialkowski (see Ch. 17)

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LAMBDA-tuning

19 jan. 2024

To: Jose Luis, Tore, Krister.

From: Sigurd Skogestad

Hello

The saga on the history of lambda tuning continues.
I found two book chapters from a book from Tappi Press from 1995. Both chapter discuss lambda tuning.

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@inbook{Thomasson95,
 author = {Frederick Y. Thomasson},
 title = {Chapter 6: Controller tuning methods},
 booktitle = {Process control fundamentals – For the pulp & paper industry},
 publisher = {Tappi Press},
 year = {1995},
 pages = {202-274}
}

@inbook{Bialkowski95,
 author = {W.L. Bialkowski},
 title = {Chapter 7: Control objectives for uniformity in pulp and paper manufacturing},
 booktitle = {Process control fundamentals – For the pulp & paper industry},
 publisher = {Tappi Press},
 year = {1995},
 pages = {275-325}
}
```

Comments from Sigurd Skogestad:

1.In Chapter 6, Thomasson refers to Chien and Fruehauf (1990) who uses the symbol tau_cl. In Chapter 7, I didn't find any reference.
However, I note that Bialkowski in the chapter 8 on IMC refers to the 1989-book by Morari
so this is probably where he got the symbol lambda (for tau_cl) from.

2. In eq (79), Thomposon recommends lambda > tau + theta (the SIMC-rule is lambda > tau).
His justification is to that "there will be no overshoot in response to setpoint changes"
(I was thinking earlier that this must be the reason, but I never saw it written before).
It seems from his examples that he is mostly concerned with overshoot of the output y,
but actually this rule also avoids overshoot in the input u, because then there will be no "speed-up" by control.
Proof of no overshoot input: Let k be the steady-state gain and assume no process delay (theta=0). Then with tauc=lambda=tau, we get Kc=1/k.
Steady state: u = (1/k) ys
Initial (before I-action has time to respond): u = (1/Kc) ys = (1/k) ys
QED
Sigurds comment: I think this emphasis on setpoints is strange. One can just filter the setpoint and avoid the overshoots.

Chapter 6

Controller Tuning Methods

Frederick Y. Thomasson

6.1 Introduction

The overwhelming majority (>99%) of controllers used in the pulp and paper industry are PID controllers. There are many reasons for this. PID has been the process control standard for many years. As controller hardware evolved from pneumatic to analog electronic to digital, the PID algorithms have remained virtually unchanged. The PID controller is understood by operational, technical, and maintenance personnel. For most control applications, a well-designed and properly tuned PID controller is all that is needed to achieve the control objectives.

The majority of PID controllers are tuned by the trial error method. However, there are several controller tuning methods that are mentioned in practically every process control book. The Zeigler-Nichols method is perhaps the best known. Actually, Zeigler and Nichols proposed two methods. One was a closed loop method, while the other was an open-loop method. Both of these methods result in a damped oscillation response to a stop change in the setpoint. The decay ratio is 1:4. The Cohen-Coon Method is a variation of the Zeigler-Nichols open-loop method. It also uses the 1:4 decay ratio criterion. Most technicians and engineers in the pulp and paper industry refrain from using these methods because of the oscillations that can result and the poor stability margin of the controller.

In recent years a more applicable tuning method, called Lambda tuning, has been described in the literature. This method is based on simple process models that can be obtained by simple "bump" tests. It results in a closed-loop response that is similar to the open-loop response for most non-integrating processes. This method is the one that is recommended for use in the pulp and paper industry.

6.1.1 Types of PID Algorithms

There are two types of PID algorithms used extensively in process control equipment. Nearly all of the distributed control systems (DCS), analog controllers, digital controllers, and PLCs utilize one or both of these algorithms. However, most of the vendors offer a variety of options so the user can

It can be shown that:

$$\tau = \frac{K_P}{R} \tag{6}$$

Using these definitions the controller tuning settings can be obtained using the Cohen-Coon tuning equations shown in Table 6.3.

Table 6.3. Cohen-Coon tuning equations

TYPE	K_C	K_I	T_D
P	$\frac{1}{K_p}\left(\frac{1}{\alpha} + 0.333\right)$	-	-
PI	$\frac{1}{K_p}\left(\frac{0.9}{\alpha} + 0.082\right)$	$\frac{1}{\tau}\left(\frac{1 + 2.2\alpha}{3.33\alpha + 0.33\alpha^2}\right)$	-
PID	$\frac{1}{K_p}\left(\frac{1.35}{\alpha} + 0.270\right)$	$\frac{1}{\tau}\left(\frac{1 + 0.62}{25\alpha + 0.5\alpha^2}\right)$	$\tau\left(\frac{0.37\alpha}{1 + 0.2\alpha}\right)$

6.1.3 Modern Tuning Methods (Lambda Tuning)

The Lambda tuning method, based on the use of simple dynamic models, is presented in the next three sections. The parameters for the simple models are obtained by open-loop tests on the actual process. Simple step changes (bumps) are made with the controller in manual, and the response of the process variable is recorded. The process gain, time constant(s), and deadtime are estimated from this response.

Most processes in a pulp and paper mill can be represented by one of the following simple models:

1) First-order (e.g., many flows):

$$G_P(s) = \frac{K_P}{1 + \tau s} \tag{7}$$

2) First-order plus deadtime (e.g., consistencies, some flows and some pressures):

$$G_P = \frac{K_P e^{-\theta_D s}}{1 + \tau s} \tag{8}$$

3) Second-order (e.g., temperature):

$$G_P(s) = \frac{K_P}{(1 + \tau_1 s)\ (1 + \tau_2 s)} \tag{9}$$

4) Integrating processes (e.g., levels and some pressures):

$$G_P(s) = \frac{K_P}{s} \tag{10}$$

The first step in the Lambda tuning method is to identify the process dynamics. This is done by placing the controller in manual and letting the process stabilize. The output of the controller is changed in step fashion. The process response is recorded and the process model parameters are estimated from the response curve. The main parameters are the process gain (K_P), the process time constant/s (τ or τ_1, τ_2), and the deadtime (θ_D). The parameters do not need to be highly accurate, only reasonable approximations.

A typical feedback control loop is shown in Fig. 6.4.

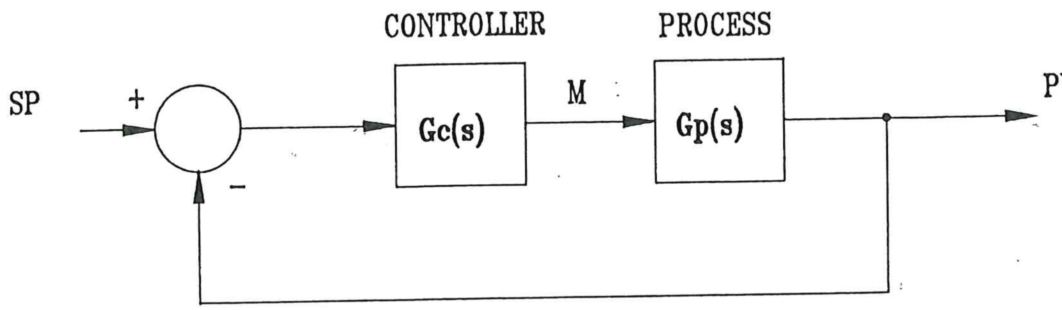

Fig. 6.4. Typical control loop

In this control loop let:

$$G_C(s) = \frac{1}{G_P(s)} \cdot \frac{1}{\lambda s} \tag{11}$$

Where:

$\lambda = \tau_{CL}$ = Desired closed loop time constant (seconds)

and the open loop transfer function is:

$$G(s) = G_C(s) \cdot G_P(s) = \frac{1}{\lambda s} \qquad (12)$$

The closed-loop transfer function can be written as:

$$H(s) = \frac{PV(s)}{SP(s)} = \frac{G(s)}{1 + G(s)} = \frac{1}{1 + \lambda s} \qquad (13)$$

The closed-loop response is first order with a gain of 1.0. The transfer function of the classic PID algorithm can be written as:

$$G_C(s) = K_C\left(\frac{s + K_I/60}{s}\right)\left(\frac{1 + T_D s}{1 + \alpha T_D s}\right) \qquad (14)$$

The derivative mode is a lead-lag transfer function with the lag set at about 1/10th of the lead (α = 0.1). When T_D = 0, or no derivative action, the controller reduces to a PI controller.

Let us apply Eq. 11 to a typical flow loop to see how the tuning equations are developed. Most flow loops can be represented by a first-order lag and gain with the transfer function given by:

$$G_P(s) = \frac{K_P}{1 + \tau s} \qquad (15)$$

Substitution of Eq. 15 into Eq. 11 results in:

$$G_C(s) = \frac{1 + \tau s}{K_P}\,\frac{1}{\lambda s} \qquad (16)$$

With $T_D = 0$, Eq. 14 becomes:

$$G_C(s) = \frac{K_C\ K_I}{60}\left[\frac{1 + \frac{60}{K_I}s}{s}\right] \qquad (17)$$

Comparison of coefficients in Eq. 16 and Eq. 17 for the PI control results in the following tuning equations:

$$\tau = \frac{60}{K_I} \quad (seconds) \qquad (18)$$

Or:

$$K_I = \frac{60}{\tau} \quad (repeats/minute) \qquad (19)$$

The proportional gain is:

$$K_C = \frac{\tau}{K_p \lambda} \qquad (20)$$

Define:

$$\lambda = \tau_{CL} \quad (\ seconds\ ) \qquad (21)$$

Using this expression, the proportional gain can be written as:

$$K_C = \frac{\tau}{K_p \lambda} \qquad (22)$$

Where:

τ = Process time constant (seconds)

τ_{CL} = Desired closed-loop time constant (seconds).

$$K_P = \frac{\Delta PV(\%)}{\Delta CO(\%)} = \frac{\text{Change in process variable}}{\text{Change in controller output}} \quad (23)$$

In this approach to tuning, the PID parameters are selected such that the zeroes created in the transfer function of the controller cancel out the poles or the open-loop process transfer function. For a first-order process, such as a flow loop, the derivative rate should be set to zero. The location of controller zero can be determined by the proportional and integral gains. This is not the case for a second-order process, such as a temperature control loop. The derivative rate should be used to cancel out the second pole of the process transfer function.

A similar approach is used for a second-order process described in Eq. 9. The tuning equations for the PID controller are obtained as follows:

$$K_I = \frac{60}{\tau_1} \qquad T_D = \frac{\tau_2}{60} \quad (24)$$

And:

$$K_C = \frac{\tau_1}{K_P \lambda}, \quad \lambda = \tau_{CL}, \quad \tau_1 > \tau_2 \ (\text{seconds}) \quad (25)$$

This says to set the reset time equal to the longest time constant, the derivative rate equal to the shorter time constant, and adjust the proportional gain to obtain the desired closed-loop speed of response.

A PI controller can be used to control processes includes a pure transport lag in the transfer function (see Eq. 8). If the transport lag is not large compared to the process time constant, the following tuning equations can be used for a first-order plus deadtime process:

$$K_I = \frac{60}{\tau} \quad (26)$$

$$K_C = \frac{\tau}{K_P(\lambda + \theta_D)} \quad (27)$$

For the same first-order plus deadtime process and using a PI controller, the tuning equations are:

$$K_I = \frac{60}{\tau} \quad (repeats\ per\ minute)$$

$$T_D = \frac{\theta_D}{120} \quad (minutes) \tag{28}$$

$$K_C = \frac{\tau}{K_P\left(\lambda + \dfrac{\theta_D}{2}\right)}$$

Where:

τ = Process time constant in seconds (assuming process can be represented by first-order lag plus delay)

θ_D = Process delay time (seconds)

K_P = Process steady state gain (%/%).

If the time delay (the deadtime) is equal to or greater than the process time constant, a Smith Predictor controller should be used instead of a PID controller. Most modern distributed control systems (DCS) have a controller that compensates for deadtime (e.g., Smith Predictor). This is discussed further in Section 6.2.3.

6.1.4 PID Tuning Procedures

There are some well-defined goals to keep in mind as you start to tune basic control loops. Some of the more important ones are:

1. Obtain satisfactory response speeds to changes in the process setpoint.
2. Hold deviations from setpoint to a minimum (minimum variance).
3. Provide adequate control over the entire operating range.
4. Handle process disturbances satisfactorily.
5. Obtain minimum sensitivity to noise.

The mere tuning of the controller alone will not guarantee that these goals are met. The structure of the control loop, the appropriate use of feedforward, the appropriate filtering of the process variable, and gain scheduling also must be considered.

6.2.1 Obtaining Simple Models

There are a number of processes that can be represented by first-order or first-order plus delay systems. The transfer function for these systems is written as:

$$\frac{Y(s)}{M(s)} = \frac{K_P e^{-\theta_D s}}{1 + \tau s} \tag{29}$$

Where:

K_P = The steady state process gain
θ_D = The system delay (deadtime)
τ = The system time constant
M(s) = The manipulated variable.

And:

Y(s) = The process variable.

Estimates of the three parameters (K_P, θ_D, and τ) are needed to utilize the modern tuning procedure described in this section. These parameters can be obtained from open-loop "bump" tests on the process.

The following procedure can be used to obtain these parameters.

STEP 1: Set up a trend to record process variable, setpoint, and controller output.

STEP 2: Place the loop in manual. Make sure the process is in a normal operating region. Move the controller output in a step fashion a small amount (say, 5%). Discuss the size of the move with the operator. Let the process settle out. The trend should resemble the one shown in Fig. 6.5. The process parameters can be estimated from this response.

STEP 3: Determine θ_D and τ directly from the response. The units will be in seconds. The process gain is calculated as follows:

$$K_P = \frac{\Delta Y}{\Delta M} \cdot K \tag{30}$$

Where:

- ΔY = Incremental change in process variable in engineering units (e.g., GPM).
- ΔM = Incremental change in controller output (%).
- K = 100/SPAN. This changes ΔY to percent (%) units.

Example

Assume that:

- ΔM = 5%
- ΔY = 520 GPM

And flow measurement was ranged 0–10,000 GPM.

Thus:

$$K = \frac{100}{10{,}000} = 0.01 = \frac{(520)(.01)(\%)}{5(\%)} = 1.04 \tag{31}$$

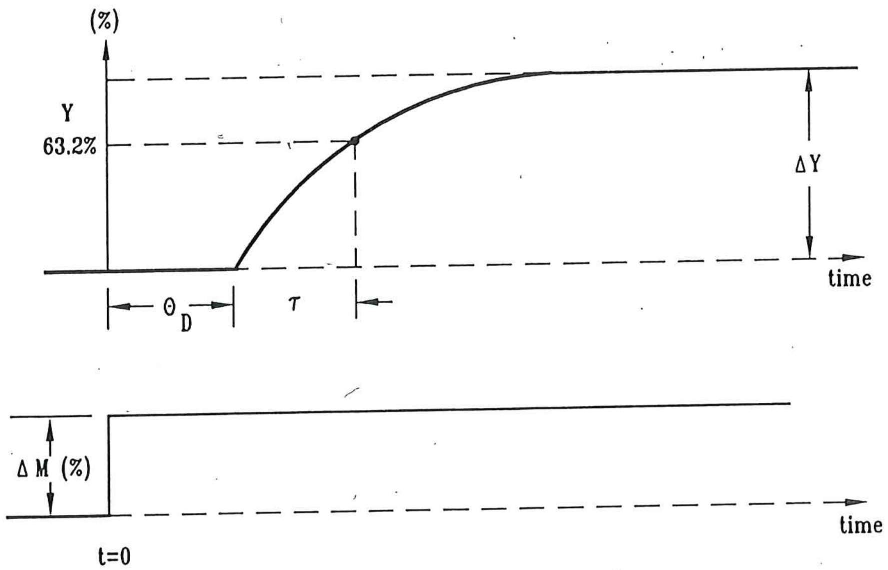

Fig. 6.5. First-order plus deadtime response to step change

The open-loop transfer functions for many of the temperature loops can be represented as second-order systems. The transfer function can be expressed as:

$$\frac{Y(s)}{M(s)} = \frac{K_P}{(1 + \tau_1)(1 + \tau_2)} \tag{32}$$

This system has two time constants and a process gain. An open-loop "bump" test is used to obtain a process response such as the one shown in Fig. 6.6.

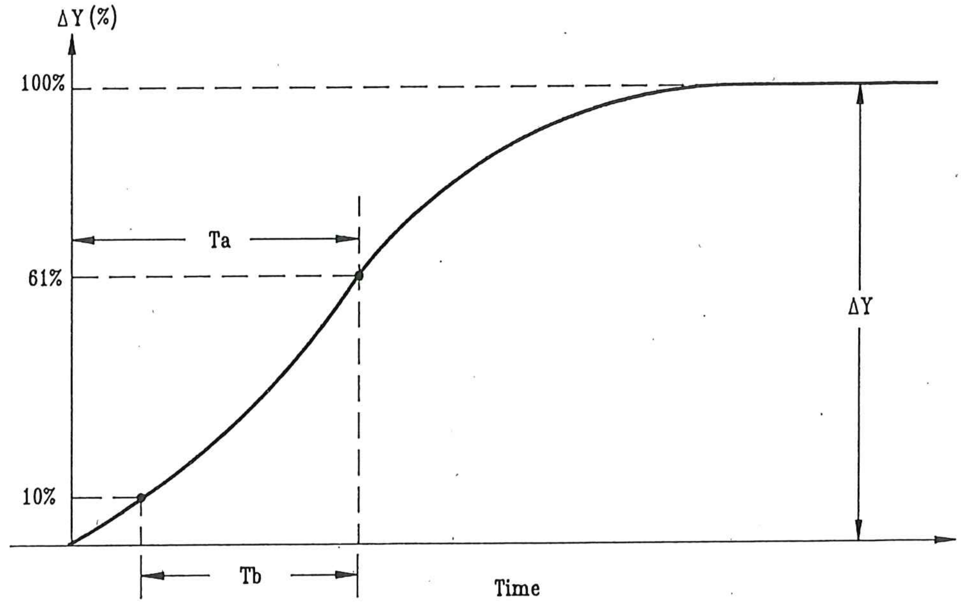

Fig. 6.6. Second-order response to step change

The process gain is calculated the same way as it was for the first order plus deadtime system. The difficulty arises in estimating the two time constants τ_1 and τ_2. The following method will be offered without any mathematical proof or justification. It provides estimates that are adequate for tuning PID controllers.

Define:

T_A = Time it takes process to reach 61% of its final value.
T_B = Time it takes process to go from 10% to 61% of its final value.

Let:

$\tau_1 = T_A$ (in seconds)
$\tau_2 = T_A - T_B$ (in seconds).

There are numerous tanks, chests, or vessels in pulp and paper mills that are operated on level control. They are called integrating processes. The transfer function for these processes is simply a gain plus integral and can be written:

$$\frac{Y(s)}{M(s)} = \frac{K_P}{s} \tag{33}$$

Where:

- Y = Tank level in percent (%)
- M = Controller output in percent (%)
- K_p = The process gain.

Perform another simple "bump" test on the process. Actually, a series of two or three "bumps" will be needed to obtain the process gain, K_P. A typical series of "bump" tests and the process response is shown in Fig. 6.7.

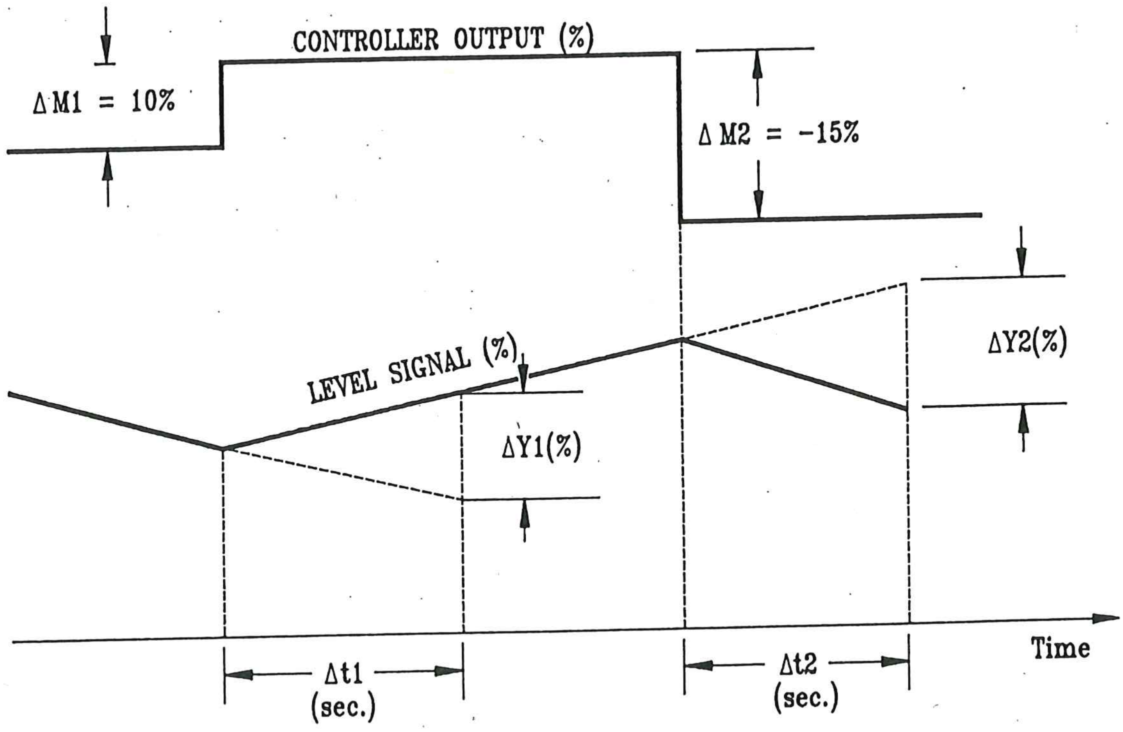

Fig. 6.7. Integrating process response to step change

The process gain, as determined from the first bump, is given by:

$$K_P = \frac{(\Delta Y_1/\Delta t_1)}{\Delta M_1} \tag{34}$$

And from the second bump:

$$K_P = \frac{(\Delta Y_2/\Delta t_2)}{\Delta M_2} \tag{35}$$

The two gains should be the same. If they are not, a third "bump" test should be run.

6.2.2 Determining P, I, and D Parameters

It is a simple matter to select the tuning parameters once the process model has been obtained. The PID tuning equations for classical (interactive) PI/PID controllers is shown in Table 6.4.

Table 6.4. PID tuning equations for interactive algorithms

PROCESS MODEL	$K_C K$	K_I	T_D
$\frac{K_p}{1 + \tau s}$	$\frac{\tau}{K_p \cdot \lambda}$	$\frac{60}{\tau}$	-
$\frac{K_p e^{-\theta_D s}}{1 + \tau s}$	$\frac{\tau}{K_p(\lambda + \theta_D)}$	$\frac{60}{\tau}$	-
$\frac{K_p e^{-\theta_D s}}{1 + \tau s}$	$\frac{\tau}{K_p\left(\lambda + \frac{\theta_D}{2}\right)}$	$\frac{60}{\tau}$	$\frac{\theta_D}{120}$
$\frac{K_p e^{-\theta_D s}}{(1 + \tau_1 s)(1 + \tau_2)}$	$\frac{\tau_1}{K_p(\lambda + \theta_D)}$	$\frac{60}{\tau_1}$	$\frac{\tau_2}{60}$
$\frac{K_p}{s}$	$\frac{2}{K_p\lambda} = \frac{100}{ALV}$	$\frac{30}{\lambda}$	-
$\frac{K_p e^{-\theta_D s}}{s}$	$\frac{2\lambda + \theta_D}{K_p(\lambda + \theta_D)^2}$	$\frac{60}{2\lambda + \theta_D}$	-

The λ term is a user specified value that determines the closed -loop time constant. A higher value provides a slower response. A value of $\lambda = \tau$ sets the closed-loop speed of response equal to the open-loop response. Typically the closed-loop speed of response is set at half the speed of the open-loop response ($\lambda = 2 \cdot \tau$). This prevents excessive control action. In the case of a flow control loop, it would prevent the valve from making large, rapid moves.

Tuning level controllers is a separate problem. More will be said about this later in this section. The development of the tuning equations for various process models is similar to Chien.

Where:

K_C = Proportional gain
K_I = Integral gain (repeats / minute)
T_D = Derivative rate (minutes).

All of the time constants in the model are expressed in seconds. Several other definitions are as follows:

$$K = \frac{100}{Span}$$

$\tau_1 > \tau_2$ for a second-order system

And:

ALV = Allowable level variation (%).

The tuning values calculated using the simple process models (obtained by the "bump" tests) should be entered into the controller. The controller should be placed in automatic and the setpoint should be stepped several times to see if the desired closed-loop response is obtained. Do not be surprised if a slight adjustment of one or more of the tuning parameters is required in order to obtain the desired response. This usually means that the model is not quite matched to the actual process. In most situations, it does not pay to spend a lot of time obtaining a highly accurate model. A rough approximation of the process is all that is needed to get a decent starting point for your tuning parameters.

6.2.3 Other Tuning Considerations

It was assumed that the process in the previous section was linear. This is never the case in the real world. Most valve characteristics are not linear, which leads to a varying process gain. Often the process gain varies as a function of load. In some situations, the process time constant will also vary with load. What all this means is that the correct tuning values for one operating point may not give acceptable performance at other operating conditions. The user should strive to tune the loop at the nominal operating point. The controller performance should then be checked at other operating conditions. If the performance is not satisfactory for certain operating regions, the user has several options available. He can tune the loop at several operating conditions and use a gain-scheduling feature that many controllers possess. Another option may be to linearize the process gain by sending the output of the controller through a characterizer (function generator). Yet another option may be to find a compromise set of tuning parameters that gives adequate control performance over the entire operating range, but this sacrifices performance that could be obtained at the nominal operating point.

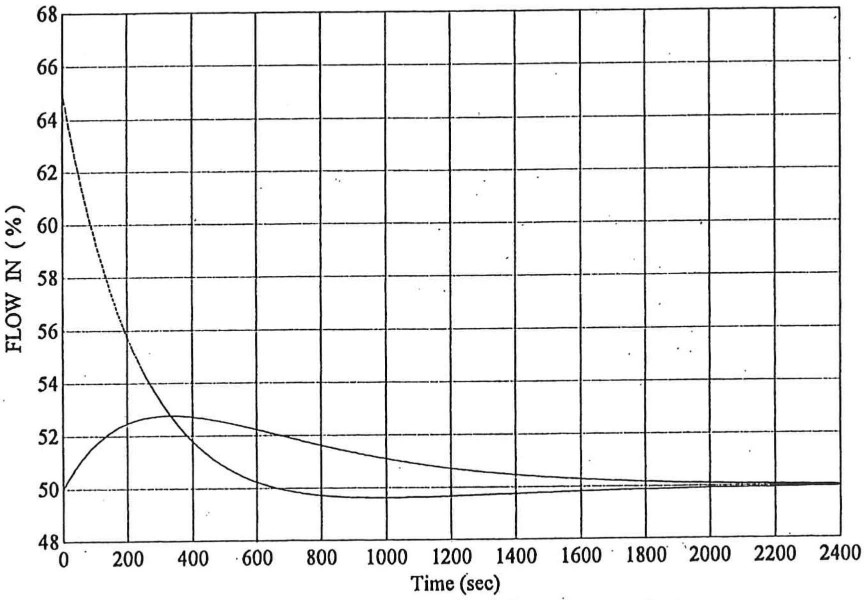

Fig. 6.33. Two methods for changing level setpoint

There will be overshoot in level when a level setpoint change is made with the controller in automatic mode. Overshoot can be avoided if the operator places the controller in manual mode, makes the setpoint change, and then places the controller back in automatic. The flow and level responses for both methods are shown in Figs. 6.32 and 6.33. The flow out of the tank remained constant and $K_C = 3.0$ and $K_I = 0.09$. In both cases, the setpoint was increased from 50% to 55%.

Note that the flow jumped from 50% to 65% when the controller was left in automatic while the setpoint was changed. However, the flow never exceeded 53% when the controller was in manual when the setpoint was changed. Also, there was no overshoot for the latter. Which method do you suppose causes the biggest upset to the upstream process? The sudden change comes from the proportional action of the controller and is called the proportional "kick."

Another way (probably the best way) to avoid proportional "kick" is to send the setpoint through a first-order lag before it enters the PID control block. The time constant (seconds) should be set to $60/K_I$. The responses will be the same whether the controller is in automatic or manual mode when the setpoint change is made.

6.3.4 Tuning Control Loops for First Order Plus Deadtime Processes (Consistency Control)

A first-order plus deadtime transfer function adequately represents many pulp and paper processes for tuning purposes. A PID controller can be used to control this type of process if the deadtime is less than three times the process time constant:

$$\theta_D \leq 3\tau \qquad (74)$$

The first-order plus deadtime transfer function is as follows:

$$G_p(s) = \frac{K_p e^{-\theta_D s}}{1 + \tau s} \qquad (75)$$

The regulatory control performance of the loop (disturbance rejection) will deteriorate rapidly when θ_D exceeds τ, even though the response to setpoint changes remains acceptable. An algorithm such as a Smith Predictor should be used (if available) if the deadtime exceeds the process time constant. Most DCS vendors have an algorithm (such as a Smith Predictor) to handle loops with deadtime. It is beyond the scope of this chapter to cover tuning of deadtime compensation algorithms, but this author encourages readers to examine the DCS application guides for instruction on tuning their deadtime compensation algorithm.

Follow the procedure outlined in Section 6.2.1 to determine the process gain (K_P), the process time constant (τ), and the deadtime (θ_D) from the recorded process response to an open-loop "bump" test. The tuning equations are obtained from Table 6.3. For the interactive PI algorithm, the tuning equations are as follows:

$$K_C = \frac{\tau}{K_p K(\lambda + \theta_D)} \quad (Proportional Gain) \qquad (76)$$

$$K = \frac{100}{Span} \qquad (77)$$

And:

$$K_I = \frac{60}{\tau} \quad [repeats/\min.] \qquad (78)$$

Where:

τ = Process time constant (seconds)
K_P = Process gain
θ_D = Deadtime (seconds)
λ = Desired closed loop time constant (seconds)
K = Normalizing value.

The tuning equations are valid when the deadtime is less than the process time constant. Notice that the integral gain remains constant as deadtime increases, but the proportional gain decreases. The smallest value of λ that the author recommends using is:

$$\lambda \min = \tau + \theta_D \tag{79}$$

For this value, there will be no overshoot in response to setpoint changes. The values obtained from the tuning equations should be considered the starting points. After these values are entered into the controller slight adjustments may be required. The setpoint should be changed (in a step change) and the response noted. Continue to make slight adjustments until the response is satisfactory.

Often the primary control objective of the loop is to hold the process variable near the setpoint as disturbances push it away from the setpoint. The setpoint may only be changed infrequently. When tuning this type of loop, actually "bump" the disturbance variable if you can. Often the main disturbance is the load, and it can usually be changed. If the disturbance rejection is not satisfactory, increase the proportional gain, but not the integral gain. After you are satisfied with the response to disturbances, go back and change the setpoint and watch the response. If there is significant overshoot, you should consider decreasing the proportional gain.

If a process is discovered where the deadtime exceeds the time constant, look for the cause of the delay. Try to get the process modified if possible to decrease the deadtime. Sometimes this can be as simple as moving the measurement sensor. Deadtime is an enemy to good regulatory control.

Example 5 — Tuning a Consistency Control Loop

In this example, the process simulated is a consistency loop. It has a transfer function which is first order with deadtime and represented as:

$$G_P(s) = \frac{0.03\, e^{-\theta_D s}}{1 + 10s} \tag{80}$$

Where:

Span	= 4%, (2% - 6%)
Process gain (K_P)	= 0.03
Time constant (τ)	= 10 seconds
Deadtime (θ_D)	= 5, 10, 20, and 30 seconds

STEP 1: Perform open-loop "bump" tests to obtain process response and then calculate the process parameters. The controller was placed in manual mode and the controller output was stepped (-16.67%). A family of consistency responses (for θ_D = 5, 10, 20, and 30 seconds) is shown in Fig. 6.34.

STEP 2: A. [For θ_D = 5 seconds, τ = 10 seconds]
Let λ = 5, 10, 15, and 20 seconds
Using the tuning equations for interactive PI algorithm:

$$K_C = \frac{\tau}{K_P K(\lambda + \theta_D)}, \quad K_I = \frac{60}{\tau}, \quad K = \frac{100}{Span} \tag{81}$$

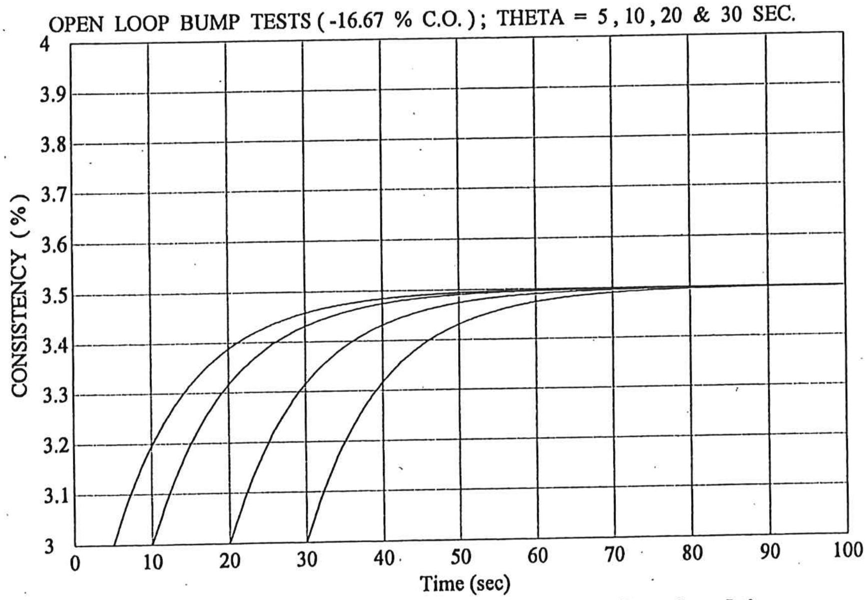

Fig. 6.34. Consistency responses (first-order plus deadtime example)

Table 6.11. Several combinations of possible λ and K_C values

λ	K_C	K_I
5	1.33	6
10	0.89	6
15*	0.67	6
20	0.53	6

*Recommended value: $\lambda = \tau + \theta_D$ = 10+5 = 15 seconds

A family of consistency responses is shown in Fig. 6.35. Notice that overshoot occurs for λ = 5. A value of λ = 15 seconds is recommended.

B. [For θ_D = 10 seconds, τ = 10 seconds]
λ = 5, 10 and 20 seconds

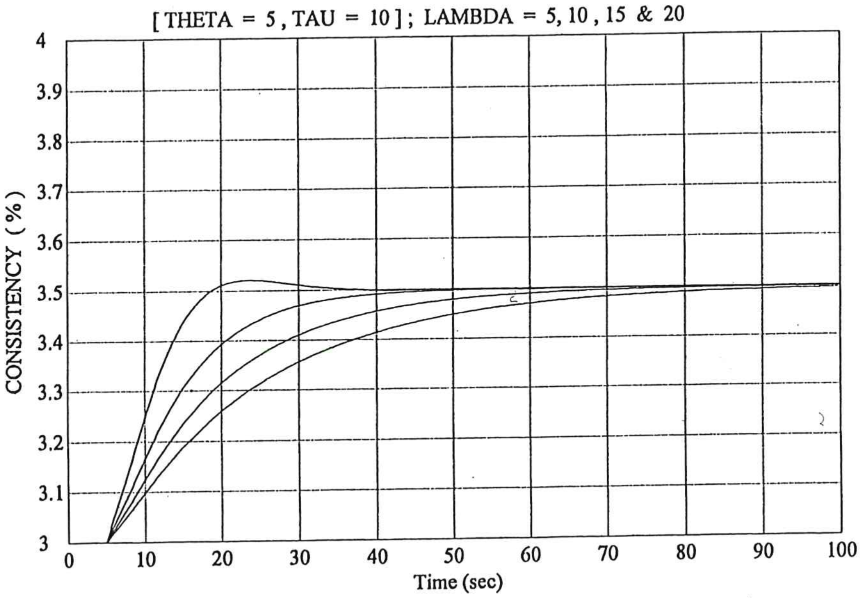

Fig. 6.35. Consistency responses (first-order plus deadtime example)

Table 6.12. Several combinations of possible λ and K_C values

λ	K_C	K_I
5	0.89	6
10	0.67	6
20*	0.44	6

*Recommended value: $\lambda = \tau + \theta_D$ = 10 + 10 = 20 seconds

A family of consistency responses is shown in Fig. 6.36. Notice that overshoot occurs for λ = 5 and 10 seconds. A value of λ = 20 seconds is recommended.

C. [For $\theta_D = 20$ seconds, $\tau = 10$ seconds]
$\lambda = 5, 10, 20$ and 30 seconds

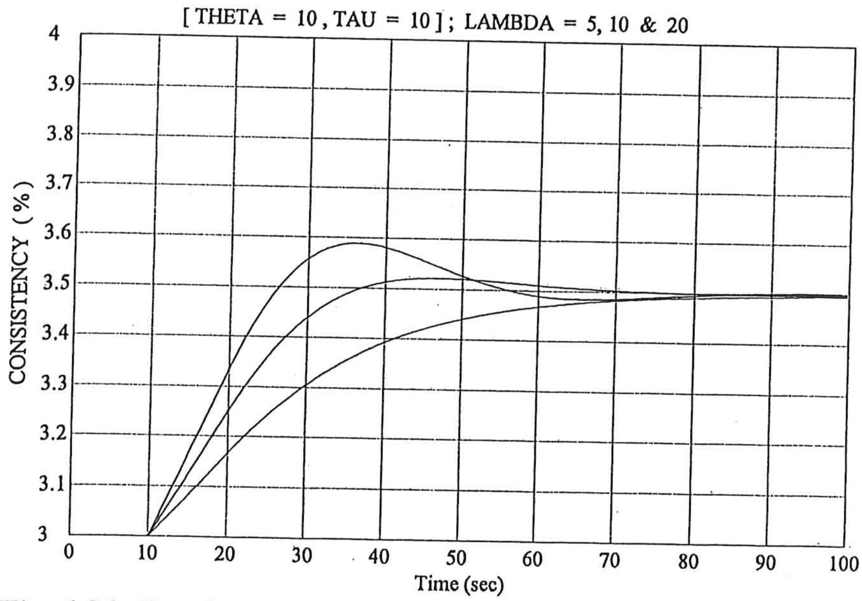

Fig. 6.36. Consistency responses (first-order plus deadtime example)

Table 6.13. Several combinations of possible λ and K_C values

λ	K_C	K_I
5	0.533	6
10	0.444	6
20	0.333	6
30*	0.267	6

*Recommended value: $\lambda = \tau + \theta_D = 20+10 = 30$ seconds

A family of consistency responses is shown in Fig. 6.37. Notice that overshoot occurs for $\lambda = 5$, 10, and 20. A value of $\lambda = 30$ seconds is recommended.

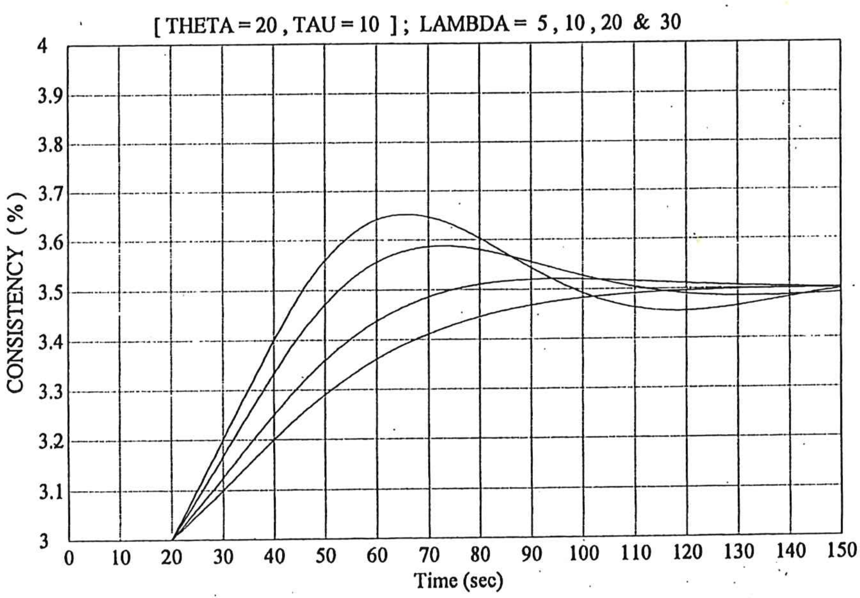

Fig. 6.37. Consistency responses (first-order plus deadtime example)

For the $\lambda = 20$ tuning, K_C was kept at 0.333 but K_I was varied: 6.0, 5.5, and 5.0. The three responses are shown in Fig. 6.38. All of the overshoot is eliminated when K_I is reduced to less than 5.5.

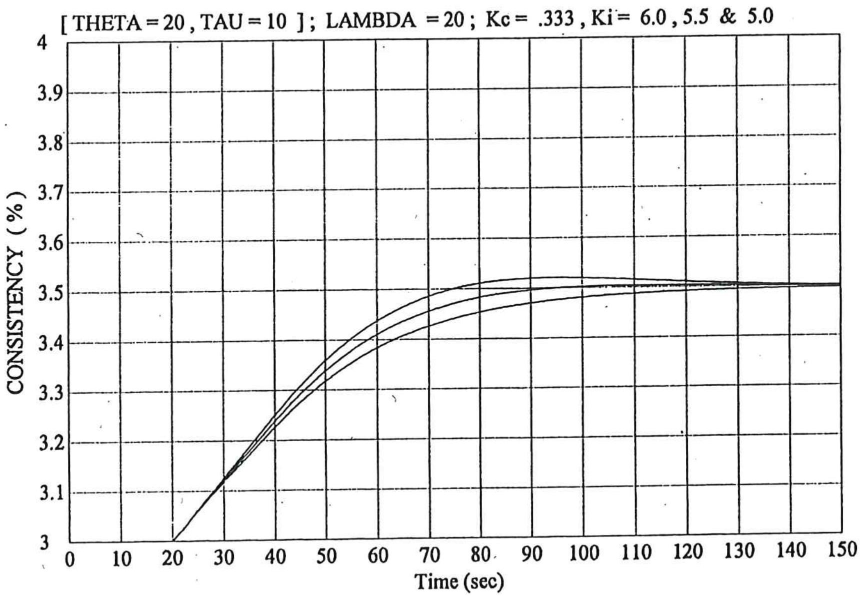

Fig. 6.38. Consistency responses (first-order plus deadtime example)

D. [For θ_D = 30 seconds, τ = 10 seconds]
λ = 30 and 40 seconds

Table 6.14. Possible λ and K_C values

λ	K_C	K_I
30	0.222	6
40*	0.19	6

*Recommended value $\lambda = \tau + \theta_D$ = 10+30 = 40 seconds

The consistency responses are shown in Fig. 6.39. Overshoot occurs when λ = 30 seconds. The recommended value is λ = 40 seconds.

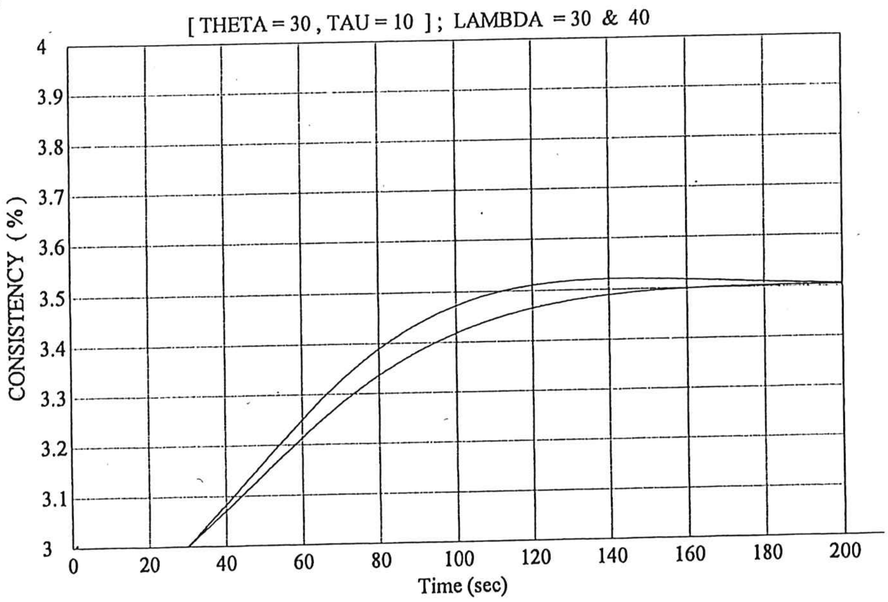

Fig. 6.39. Consistency responses (first-order plus deadtime example)

As the deadtime increased, the closed-loop time constant (λ) had to be increased also. The following table shows how K_C is decreased based on setting $\lambda = \tau + \theta_D$.

Table 6.15. Several combinations of possible λ and K_C values

θ_D	λ	K_C	K_I
5	15	0.667	6
10	20	0.44	6
20	30	0.267	6
30	40	0.19	6

The proportional gain (K_C) decreases from 0.667 to 0.190. The main impact is that the loop's disturbance rejection capability is diminished. This is demonstrated in Fig. 6.40. A step disturbance of +0.5% is applied at t = 0 seconds. The response for θ_D = 30 seconds shows poor disturbance rejection and would not provide acceptable consistency control. In fact, the author considers only the response for θ_D = 5 seconds (λ = 15 seconds) as acceptable.

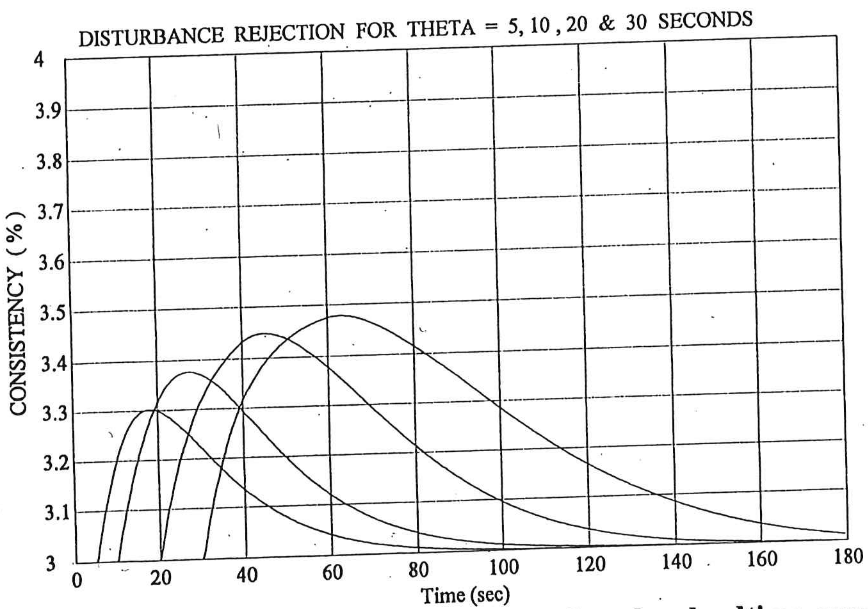

Fig. 6.40. Consistency responses (first-order plus deadtime example)

6.3.5 Tuning Cascade Control Loops

Often a process can be considered as two processes connected in series, as shown in Fig. 6.41. Each process may be subject to one or more process disturbances. Control performance can usually be improved if an additional control loop is used. This results in a control configuration known as cascade control (see Fig. 6.42). The output of the primary controller (outer loop) becomes the setpoint for the secondary controller (inner loop). Disturbances to the secondary process are attenuated (rejected) by the inner loop and, therefore, have little impact on the primary process. The secondary process often has a nonlinear process gain. This is the case for most flow processes because of valve characteristics. By using a feedback control loop on flow, the closed-loop gain for the primary loop is linearized to some degree. There are a number of processes in pulp and paper industries where cascaded loops are commonly used. A partial list is given in Table 6.16.

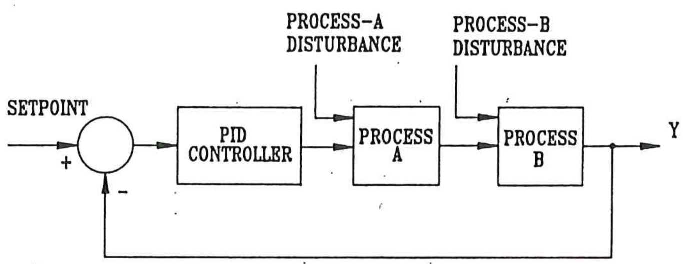

Fig. 6.41. Cascaded processed with single loop control

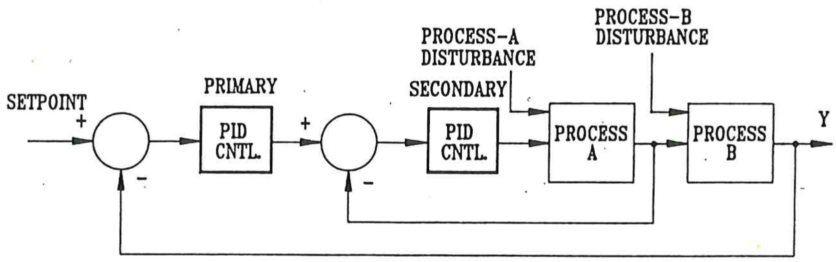

Fig. 6.42. Cascaded process with cascaded control loops

Table 6.16. Typical cascaded loops

PROCESS	PRIMARY LOOP PV	SECONDARY LOOP PV
MD Moisture	MD Moisture	Dryer Pressure
MD Basis Weight	MD Basis Weight	Dry Stock Flow
Stock Consistency	Consistency	Dilution Flow
Stock Refining	HP Days/Ton	Power (KW or HP)
Chest Level	Level	Flow
Hot Water	Temperature	Steam Flow
Drum Level	Level	Feedwater Flow

Example 6 — Tuning Cascade Control Loops

A typical cascade control is shown in Fig. 6.43. This is a hot water heater where low pressure steam is used to heat water to a desired temperature. The secondary loop (inner loop) is steam flow. The process disturbance is the steam header pressure. The primary loop (outer loop) is hot water temperature. The disturbance to this process is the water flow through the heater.

STEP 1: Place flow loop and temperature loop in manual mode.

STEP 2: Perform "bump" test on flow loop and tune as previously described. The control should be tuned so that it responses promptly but without any overshoot.

STEP 3: Leave the flow loop in automatic mode with the temperature loop in manual. Bump the output of the temperature controller and record the temperature response. Note that the inner loop (flow) is now part of the process. The temperature loop (outer loop or primary loop) is tuned using the procedures described in previous sections. The closed loop response time (λ) should be about two to three times the process time constant. This insures that the outer loop will be two or three times slower than the inner loop. It is necessary for the inner loop to be faster than the outer loop for good control performance.

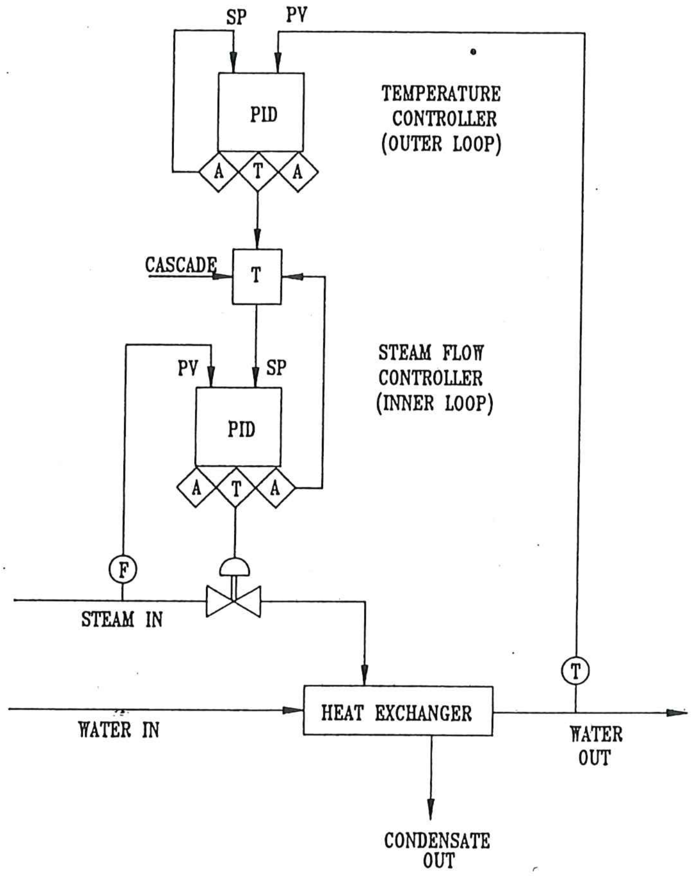

Fig. 6.43. Hot water heater control

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Chapter 7

Control Objectives for Uniformity in Pulp and Paper Manufacturing

W. L. Bialkowski

7.1 Introduction

This chapter attempts to put into perspective the problems and challenges of pulp and paper manufacturing and its associated control engineering concepts. The material presented owes its origin primarily to two main activities — mill variability audits and control engineering training. The many pulp and paper mill variability and control audits, together with succeeding optimization programs, have produced a large array of plant data, some of which is presented here and in Chapter 9 for illustrative purposes. The experience brought to light many practical pulp and paper manufacturing problems which had not been carefully examined before from the perspective of control engineering. The second activity, that of teaching professional control engineering short courses to pulp and paper engineers, has forced the reconciliation and proving of new control techniques, such as the internal model control concept (IMC) and Lambda Tuning, with actual pulp and paper practice. This has generated new tuning rules for specific pulp and paper uses.

7.1.1 Historical Perspective

The pulp and paper industry is steeped in history and is still in transition from "art" to "science." Terms like "couch" and "deckle" precede even the invention of the paper machine by the Fourdrinier brothers in 1801, yet they are still with us today. Although the application of science has been extensive in recent years in both pulping and papermaking, this has been largely focused on developing a fundamental understanding of the behavior of pulp and paper chemistry and processes. The science of dynamics and control engineering has not been as extensively applied in our industry to date. This is not surprising, since the chemists and chemical engineers have focused primarily on the steady-state behavior of pulping and papermaking reactions and processes.

Rule 1) Choose an Actuator Which Behaves Linearly

The final actuator should be near linear and as free as possible of both backlash (hysteresis) and stiction, thus allowing repeatable, small magnitude changes in the process variable to be made. Assuming that the desired process variability can be defined for the loop in question, then the maximum allowed backlash/stiction specification can be stated as:

$$\Delta m_{max} = \frac{(2\sigma_y)}{(K_p\alpha)} \tag{1}$$

Where:

Δm_{max} = Maximum allowed backlash and/or stiction
$2\sigma_y$ = Desired process variable "2- sigma" under normal controlled conditions
α = Safety factor (typically five).

The intent is to ensure that reliable changes in the process variable can be made, which are significantly smaller (say, five times) than the desired process variability on control. Dynamic specifications are essential to establish guidelines for both equipment purchase and on-going maintenance practice (EnTech, Control Valve Dynamic).

Rule 2) Select a Closed-Loop Time Constant Lambda (λ)

The selection of the closed loop time constant Lambda (λ) is of paramount importance in control loop tuning in the pulp and paper industry. The ability to select Lambda allows loops to be tuned in a coordinated manner in order to meet the manufacturing objectives. For instance, the closed-loop time constants of all of the flow loops supplying furnish ingredients to a blend chest should be the same, so that the blend ratio is not disturbed dynamically. Hence, the control algorithm should be chosen so that the gain and tuning parameters can be pre-calculated in order to achieve a desired closed loop time constant, Lambda (λ). For a first-order process this will provide the closed-loop setpoint response of:

$$\frac{1}{\lambda s + 1} \tag{2}$$

It will also give a regulation response of:

$$\frac{\lambda s}{\lambda s + 1} \tag{3}$$

The period at the cut-off frequency will be two. In the presence of deadtime, the period at the cut-off for the setpoint response is still two, while the period at the cut-off frequency for the regulation response is slowed down by the deadtime and is $2\pi(\lambda+\theta_D)$.

Rule 3) Derive the Control Algorithm Using Pole-Zero Cancellation

The control algorithm basic structure should be derived using Internal Model Control (IMC) structuring rules to be discussed in Chapter 8, or pole-zero cancellation. This approach leads to robust control structures for linear systems, and results in controller zeros (integral times, derivative times) which are set to cancel out the dominant process poles. In so doing, the natural dynamic behavior of the process will be modified and replaced by the desired dynamic behavior of the control loop. Naturally, this approach requires the process dynamics to be near linear. When nonlinearities are present, even after best efforts have been made to remove these effects through equipment selection and maintenance, then loop tuning procedures need to be modified to accommodate the presence of backlash and stiction. Avoidance of limit cycling is important. This requires higher gains and slower integral times to be used.

Rule 4) Minimize Oscillatory Behavior

The control loop should be designed so it does not inadvertently become oscillatory as process conditions change. Robustness can now be defined as the desire to ensure that the dominant closed-loop poles are on the negative real axis of the s-plane (or very near to it). This statement can be formalized by specifying that the lowest allowable value of damping coefficient should be approximately 0.8.

In the presence of deadtime, it is often not possible to avoid some resonance in the control loop, regardless of the algorithm. The resonance occurs in the regulation response because of the deadtime-induced phase lag. Consider a first-order plus deadtime process (such as consistency) controlled by a Lambda Tuned PI controller. The process transfer function is:

$$G_P(s) = \frac{K_P e^{-\theta_D s}}{\tau s + 1} \tag{4}$$

The controller transfer function is:

$$G_C(s) = \frac{K_C(\tau_I s + 1)}{\tau_I s} \tag{5}$$

The loop transfer function is:

$$G_L(s) = G_C G_P = \frac{K_C K_P(\tau_I s + 1)s^{-\theta_D s}}{\tau_I s} \quad (6)$$

Lambda Tuning this controller involves the following settings:

$$\tau_I = \tau \quad \textit{and} \quad K_C = \frac{\tau_I}{K_P(\lambda + \theta_D)} \quad (7)$$

This results in the loop transfer function:

$$G_L(s) = \frac{e^{-\theta_D s}}{(\lambda + \theta_D)s} \quad (8)$$

The setpoint response transfer function is:

$$G_{SP}(s) = \frac{G_L}{1 + G_L} = \frac{e^{-\theta_D s}}{(\lambda + \theta_D)s + e^{-\theta_D s}} \quad (9)$$

Using the first-order Taylor series approximation for $e^{-\theta s} = 1 - \theta_D s$ we get:

$$G_{SP}(s) \cong \frac{e^{-\theta_D s}}{(\lambda s + 1)} \quad (10)$$

This is a first-order response with deadtime. Given Eq. 9, the load response now is:

$$G_{LD}(s) = 1 - G_{SP}(s) = \frac{(\lambda + \theta_D)s}{(\lambda + \theta_D)s + e^{-\theta_D s}} \cong \frac{(\lambda + \theta_D)s}{(\lambda s + 1)} \qquad (11)$$

Figure 7.26 is a plot of the Bode Plot (Thompson) for this example, for Lambda set equal to twice the deadtime (θ_D = five seconds; λ = ten seconds).

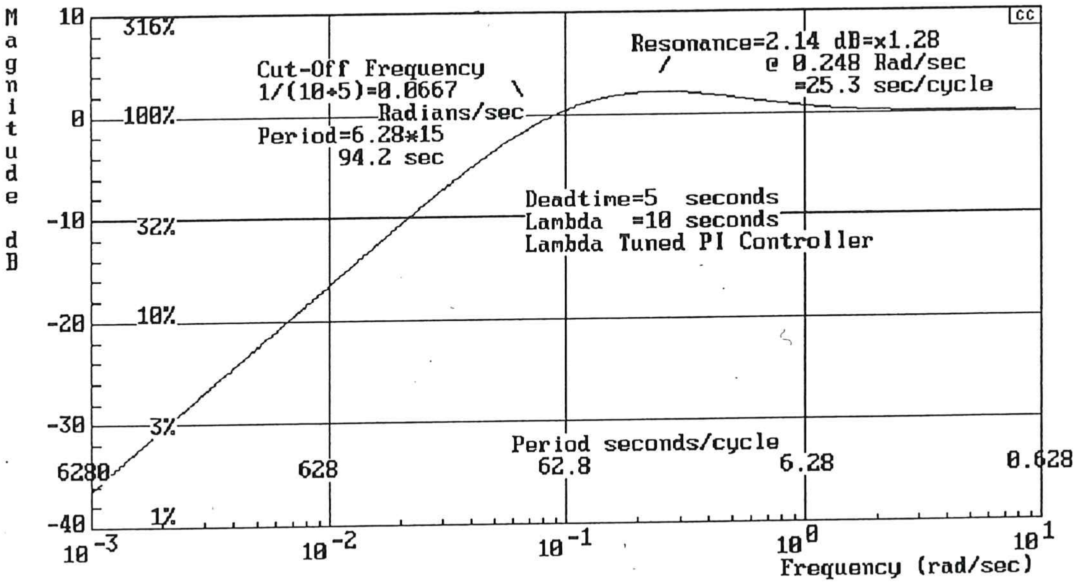

Fig. 7.26. Consistency control load Bode plot

Figure 7.26 illustrates the properties of a regulator in which the cut-off frequency determines the low-frequency attenuation, starting at a period of $2\pi(\lambda + \theta_D)$, which in this case works out to be 94.2 seconds/cycle. It also illustrates that in the presence of deadtime the loop resonates, in this case by +2.14 dB, equivalent to an amplification ratio of 1.28. Hence, this consistency loop will attenuate low frequency variability of 0.00667 radians/second (or one decade slower than the cut-off period of 942 seconds/cycle or 15.7 minutes/cycle) by -20 dB or down to ten percent of its amplitude. At the same time, it will amplify variability at a frequency of 0.248 radians/second or 25.3 seconds/cycle by +1.28 dB or +28%. Figure 7.27 illustrates what happens when the closed-loop time constant Lambda is varied over a range from very fast to very slow.

As expected, the Lambda value determines the cut-off frequency and the degree of resonance — the faster the loop, the more resonance. From this information it would be dangerous to recommend tuning settings faster than a Lambda value of ten seconds, or twice the deadtime, as in so doing the loop would be amplifying high-frequency noise by as much as 28%. On this basis, it reasonable to say that non-integrating deadtime processes should not be tuned with Lambda values faster than two deadtimes in order to limit the degree of resonance.

Rule 5) Choose the Closed-Loop Time Constant (λ) for Robustness

The closed loop-time constant should be selected carefully to ensure that the demand for fast response does not overtax the worst-case model mismatch and sacrifice robustness. A simple rule of thumb is that as long as λ is three times slower than the dominant open-loop time constant, there is little danger of model mismatch causing serious problems. The faster values of λ increasingly endanger the robustness of the loop.

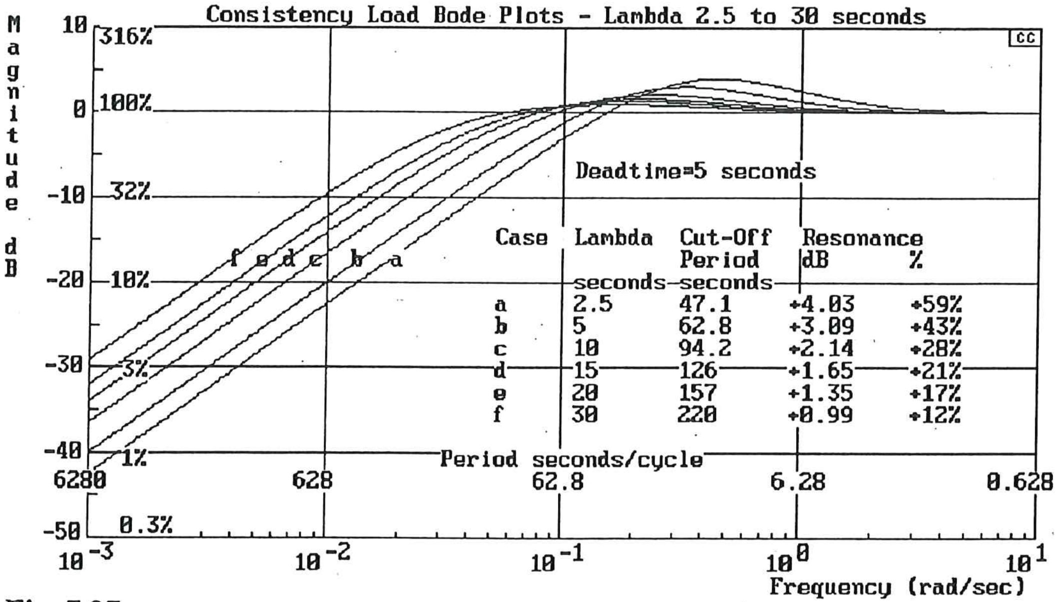

Fig. 7.27.

Rule 6) Choose the Closed-Loop Time Constant to Minimize Loop Interaction

The closed-loop time constant should be compatible with the needs of adjacent and related loops. Outer cascade loops must be slower than inner cascade loops. A rule of thumb is to plan for outer cascade loops to be five to ten times slower than inner loops. There are occasions when faster tuning is possible, and there are also occasions when it is difficult to even achieve this goal of ten times slower. Naturally, once the inner cascade loop has been tuned to a given Lambda value, the outer loop Lambda value will naturally be established through the normal bump test and tuning process. Adjacent interacting loops should also be five to ten times faster or slower than loops with which they interact. The choice of which loop is tuned to be the faster of a given pair must be made on the basis of process knowledge. The alternative is to design decoupling control logic to effectively decouple the loops. This requires knowledge of decoupler design techniques.

Rule 7) Choose the Closed-Loop Time Constant Based on Regulator Bandwidth

The closed-loop time constant should be chosen so that the loop bandwidth ($1/2\pi(\lambda + \theta_D)$) is matched to the bandwidth of the disturbance energy affecting the loop (minimum variance control concept).

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Chapter 8

Internal Model Control: A General Unifying Concept

W. L. Bialkowski

8.1 Introduction

The internal model control concept (IMC) provides the control engineer with a powerful methodology for deriving the control structure necessary to control a process of arbitrary dynamics. Given a process transfer function, IMC provides a methodology for arriving at a control algorithm which includes an internal model of the process and reduces the control problem to the selection of a desired closed-loop time constant (often called Lambda: λ). The Lambda tuning method is a subset of this general framework. IMC controllers can be easily transformed into conventional feedback controllers, and for simple low-order processes the resulting control structure is often the common PID algorithm, for which the tuning parameters can be calculated exactly once the closed- loop time constant λ has been selected. Depending on the process dynamics, the IMC controller can be transformed into a P, I, PI, PD, PID, PID form, or sometimes even a filter. For deadtime processes, the IMC structure yields identical results to the Smith Predictor and Dahlin algorithm. IMC hence encompasses the majority of common algorithms that have been in use in process control for some time, providing both a tuning methodology and a unifying framework of thought.

Beyond IMC is the amplifying concept of robustness. To design and tune a control loop, we need a model of the process. We usually get this by testing the process dynamics on-line through a series of "bump tests." Once we have the dynamics, we tune the loop. However, the dynamics are not constant with either time or operating point. How well will the control loop work with the old tuning parameters after the process has changed? A robust control loop will continue to deliver acceptable performance as long as the process dynamics are bounded within a given region. The concept of robustness is of great importance. The reason that many process control loops have not worked well in the past is because they were not robust enough.

Chapter 11

Beyond the Loop

Henry Wells, Jr.

11.1 Introduction

Assuming all the technical material contained in the first ten chapters of this text had been known and mastered at the time, a 1960s engineer was fairly well equipped to do process control work in a paper mill. In 1995, the same engineer may find that that body of knowledge, while still crucial, makes up only a small part of what he/she needs to know to do the job.

The process control world of today resembles the stereotypical Middle Eastern bazaar of a thousand years ago. It is a place where many people and cultures come together uneasily. Strange sounding languages and customs mix, and chaos can easily gain the upper hand. More than a few people get stabbed in the back in the process. To survive, control engineers must not only have a command of their own field, but they must be "street-smart" in the overall environment and have enough knowledge of the more important foreign cultures to do business with their adherents comfortably. The last seven chapters of this book can be viewed as the cultural diversity liberal arts requirement for the well-rounded control engineer.

Linear process control is a mature and powerful technology, but as traditionally practiced, it is a very localized technology in a corporate sense. Within a corporate mission statement and business strategy, there is very little that constrains how an engineer goes about tuning a loop. In the era of analog controllers, only a small group of people, perhaps 10 or 20 in operations, maintenance, engineering, and technology had any need to know that the loop even existed. This made process control an invisible technology to corporate management, and thus it rarely appeared in corporate strategic plans.

The advent of the microprocessor has changed this situation drastically. Economics dictated that control hardware would evolve in the direction of large numbers of microprocessor-based controllers connected by local and wide-area networks. For better or worse, this evolution has destroyed forever the invisibility of process control. A corporation's computer/communications network is its central nervous system and determines what sort of organization a company can have and what sort of business strategy it can pursue. Conversely, all of the various functions within the company now have a vested interest in "the network" and decisions that affect it. Network architecture issues may selection, loop tuning, process and instrumentation trouble shooting with time series analysis, spotting process variability, and minimizing variability through control. Finally, the control engineer should be a good communicator, teacher, trainer and coach who is equally at home with a process operator, technician, or mill manager.

Impact on Industry Thinking

Not only has the industry failed to develop the technical skills of control engineers, but it has experienced a void in leadership that these engineers should have been exerting. What has happened to establishing effective programs for loop tuning, E/I technician training (still being taught the highly oscillatory (Ziegler and Nichols) quarter amplitude damping loop tuning method), and effective management decision making? In short, with the absence (in effect) of the control engineer from the ranks of the pulp and paper industry, industry thinking has been allowed to "slide" to the point where the concepts of process variability and dynamics are effectively absent. This absence occurs at a time when variability, uniformity and quality are increasingly defining competitive position and company survival.

Advances in Process Control

In the last decade, modern model-based methods such as Lambda Tuning (Dahlin), Internal Model Control (IMC), Robust Control (Morari and Zafiriou), and Stochastic Control (Astrom) have been used extensively in pulp and paper mills to reduce variability. These concepts are well established in the technical literature and collectively capture the essence of the variability and control problems in the pulp and paper industry. However, to be able to take advantage of these ideas, the right academic training, practical experience, organizational mandate, and personal motivation are required — in short, to be a qualified, competent control engineer.

17.3 A View from the Chemical Industry

It is informative to see how other industries have handled similar control issues. An excellent discussion of process control in the chemical industry was presented by J. J. Downs and J. E. Doss of Tennessee Eastman Company, Kingsport, TN, at the Fourth International Conference on Chemical Process Control, Padre Island, Texas, in 1991. The main points from their paper, "Present status and future needs — A view from North American industry," are summarized below.

Chapter 17

The Marketplace and the Future

W. L. Bialkowski

17.1 Introduction

This textbook, *Process Control Fundamentals for the Pulp & Paper Industry*, is intended to help correct a serious and complex problem — that process control as practiced in the pulp and paper industry to date has failed in its primary role of helping paper producers become more efficient manufacturers of uniform product. Mill variability audit experience (Bialkowski, 1993) has shown conclusively that process control frequently serves to increase both process and product variability in the short term — the very opposite of the intended goal! The practice of process control should be an engineering discipline based on an established theoretical foundation (the subject of this text). Instead, it often appears as a haphazard process based on guesswork, old habits, and folklore. The causes of this situation are complex, and they extend far beyond mathematics into areas relating to human resources, culture, organization, staffing, education, and training. Yet the need for effective process control has never been stronger.

Throughout the industrialized world, the trend in manufacturing is toward customer driven quality consciousness. Quality is becoming synonymous with product uniformity, and is it increasingly expected by the customer as the "cost of admission." This makes reducing process and product variability the chief manufacturing objectives for on-going success and improved competitiveness. Although these are things which process control should be able to achieve, it has so far not been able to. The pulp and paper industry, however, is not alone. Similar events have occurred in the chemical industry, where recently the following prediction has been made:

> *In ten years, the ability to produce highly consistent product will not be an issue because those companies that fail in this effort will be out of the market or possibly out of business. Among the remaining companies, the central issue ten years from now will be the efficient manufacture of products that conform to customer variability expectations* (Downs and Doss).

7) What is the quality of the control tuning relative to being able to produce uniform product efficiently?

8) How effectively do the operating staff understand the design of the control strategy and its intended use over the grade structure? How well do they use it?

9) How well does the maintenance staff understand the control strategy, and are they able to maintain the health of the control loops?

10) Is there a variability maintenance and optimization program in place? Does it work effectively?

Naturally, each mill will react to competitive pressures and their available resources differently.

17.6 Conclusion

This chapter has attempted to tie the concepts of variability and control to the "big-picture," in which a pulp and paper mill is seen as a supplier of product in a competitive marketplace — a marketplace which is increasingly demanding product uniformity. Indeed, the issues are complex and seemingly very far removed from the transfer function, the core subject of this textbook. Yet *Process Control Fundamentals for the Pulp and Paper Industry* sets out to help correct a serious problem in a complex setting. Process control as practiced in the pulp and paper industry to date has failed in its primary role — helping paper producers become more efficient manufacturers of uniform product. Why did this failure occur? The heart of the problem, which caused the original "slide" in competence, is that control engineers allowed themselves to forget the fundamental principles on which dynamics and control are based — mere "math!" Once this happened, control design became guesswork, tuning became "tweaking," and control engineering no longer deserved to be called engineering. Engineering is a discipline of thought and design. It is based on an understanding of the principles which govern the physical world. In the case of control engineering, these are the principles set out by Newton, Nyquist, Bode, Nichols, Evans, Kalman, Astrom, Dahlin, Morari, and others. The essence of these principles is captured in the transfer function. Once we embrace these concepts and build our practical expertise around them, we will become "real control engineers." We can then go forward and tackle the other problems in our complex world. We will know we have at least the "basics" in place and are building on a solid foundation.