PROCESS DYNAMICS FOR ENGINEERS I

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Foreword:

Process Dynamics for Engineers I is a five day seminar intended for Process Engineers, Control Engineers and Process Design Engineers working in the process industries. The aim of the seminar is to illustrate how the dynamic behaviour of all elements of a process operation (the process, the controller, the measurement device and the final control element) can influence process variability. In this way the participant can better understand the sources of process variability, how process variability propagates through a process operation and how process variability ultimately affects both product uniformity and process efficiency. The course promotes the idea that proper process design and control strategy design are necessary prerequisites for reducing the variability in a process operation. There is often very little appreciation at the process and control strategy design stage for the dynamics of the process system. This failure to account for system dynamics often needlessly limits control performance, machine efficiency and process variability.

To enable a full discussion of dynamic behaviour, useful mathematical tools, such as linear transfer functions, time and frequency response behaviour and time series analysis are reviewed. The seminar is heavily lab-based, in the belief that that true learning occurs through practical application of theoretical concepts. The laboratory employs a comprehensive dynamic simulation of a paper machine together with its associated process and control equipment dynamics. In the lab the participant is able to explore the many sources of process variability in a process operation. In addition, participants are introduced to Program CC for the analysis of linear transfer functions and their behaviour.

The participants also work with real plant data which they analyze using some of the software tools which EnTech engineers use operationally during mill variability audits and optimization programs. These include software for time series analysis and controller tuning. The concluding session deals with process variability and the role of integrated process design as well as on-going process control management.

The industrial student will find the course challenging and stimulating. The overall intent is to provide the student with a useful toolkit to assist in diagnosing and correcting process variability problems.
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1. INTRODUCTION TO PROCESS VARIABILITY AND AUDITING
Variability and How to Handle it

Sources of Variability

Variability can be introduced into any process via:
- composition of incoming raw materials
- composition of incoming additives and chemicals
- process equipment, process and control design
- air entrainment in stock
- inadequate mixing/agitation
- cycling loops
- ineffective tuning of control loops
Techniques for Dealing with Product Variability

1. Eliminate Variability at its source
2. Use Modern Control Engineering Principles
   - Process Design
   - Control Strategy Design
   - Loop Tuning
3. Use Time Series Analysis Tools to monitor and evaluate
4. Process Control Training focused on Pulp and Paper
   - Process Control Engineers
   - Process and Production Engineers
   - E & I Staff
   - Operators
5. Variability Management programs
Identifying the Source(s) of Variability

The process variability generated upstream at a given frequency will be felt at the reel at the same frequency and at an amplitude depending on:

- coupling gain
- effect of downstream control (high pass)
- effect of downstream mixing (low pass)
Time Series Analysis

Examples
Coupling Gain Example:

Basis Weight Response to a Thick Stock Flow Bump
Coupling Gain Example:
Basis Weight Response to an Additive Bump
The Function of Control

To help manufacture a uniform product in the face of nonuniform, randomly varying raw materials, chemicals, additives, sources of energy, together with uneven process channeling, mixing and other disturbances

However, control loops can cycle for different reasons, such as:
- poor valve/actuator/positioner: backlash, stiction, positioner overshoot
- tuning method: quarter amplitude damping, aggressive tuning by "feel"
- cascade loops: inner/outer tuning mismatch
- difficult dynamics: integrating processes (levels), variable process gain, variable
- dynamics, interacting loops, deadtime and variable deadtime.
State of Process Control in Industry

Loops which:
- reduce variability (good/adequate tuning) 20 %
- cycle and increase variability due to tuning 20 %
- cycle and increase variability due to instrument maintenance issues (valve backlash, etc.) 25 %
- require control strategy redesign to work properly 15 %
- require process redesign in order to reduce variability 20%  

Total Number of Loops 100%
The Need for Action

Process variability is excessive and causing loss of productivity.

Product quality is suffering as a result of process variability.

Process control elements may be contributing to poor performance.

Action required: Process Variability and Control Audit
Process Variability and Control Audits

Audit Goals and Objectives

- Determine product variability and compare to market needs
- Determine process variability by process variable and link to product variability
- Identify causes of process variability
- Identify requirements for: maintenance, modifications, in order to improve variability
- Raise awareness of variability/process control/competitiveness
- Remove variability by tuning loops where possible
Elements of the Audit

1. Preparation - P&ID Review, Planning
2. Field Data Collection and Preliminary Analysis
3. Data Analysis and Interpretation
4. Tune where Possible
5. Report Preparation and Presentation
6. Follow Up
Audit Examples

Closed-loop example of backlash
Open-loop bump test to determine amount of backlash. How much backlash is in this valve assembly?
Closed-loop example of stiction.

Open-loop bump test to determine amount of stiction. How much stiction is in the valve assembly?
Effect of Sampling Rate

Basis Weight Scan Average, sampling every 30 seconds
Basis Weight Single Point, sampling every 1 second

Basis Weight Single Point, sampling every 0.1 seconds
APPENDIX 1A
AUDIT LITERATURE
PROCESS CONTROL AUDITS HAVE MAJOR IMPACT ON UNIFORMITY

By William L. Biakowski

Paper companies are spending millions of dollars on the latest control systems in an attempt to stay competitive. And a large portion of the budget is going toward machine-direction (MD) and cross-direction (CD) control functions for basis weight, moisture, and several other dry-end variables.

What is most surprising about these efforts is that they ignore or give little attention to the other 300 or so control loops in the stock preparation, lean stock, whitewater, steam, and condensate areas. Moreover, little is known about the influence of these control loops on final product uniformity.

This article examines a linerboard machine process control audit conducted by EnTech Control Engineering Inc., Toronto, Ont. It will illustrate some key issues on paper machine variability, what papermakers think about this subject, and the potential for improving paper uniformity to meet customer needs.

LINERBOARD MACHINE

The linerboard machine in question manufactures 42-lb and 69-lb linerboard, with occasional heavyweight runs at 90 lb. The production rate of 69-lb linerboard, which represents 50% of the machine's overall production, is typically about 1,800 tpd at an average speed of 1,500 fpm with a 226-in. trim width.

The linerboard machine was built in the early 1970s and has a conventional fourdriner with primary and secondary headboxes. The machine has a conventional press section, four dryer sections, and two calender stacks.

The primary headbox runs pine stock cooked to 90 Kappa, while the secondary headbox runs pine stock cooked to 65 Kappa. Both the primary and secondary stocks have variable-speed fan pumps and screens, and the secondary fan pump also has a stock cleaning system.

The paper machine has a conventional basis weight and moisture MD/CD control system at the reel. In 1988, the mill installed a DCS to control all other variables.

CUSTOMER COMPLAINTS

About 40% of the linerboard produced on this machine is sold to one large corrugated box manufacturer. In 1989, this manufacturer installed two new high-speed corrugators at its largest box plant.

As soon as the new corrugators began operating, the corrugated box manufacturer began complaining about the quality of 69-lb linerboard. Soon, nearly 5% of the machine's production was in question and the working situation was becoming intolerable.

The mill undertook pulsation and vibration studies on the machine but did not identify any operating problems except for "random low frequency variability."

The paper machine control system...
The EnTech audit team used high-speed data collection equipment to measure the voltage signal directly off the beta gauge.

Figure 3: Fixed point basis weight variability based on rapid measurement of beta gauge voltage readings for 69-lb linerboard.

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<th>P</th>
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Mean = 69.876  2-Sigma = 5.832  < 8.346%

Figure 4: Dilution header pressure measurement.

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</table>

Mean = 47.325  2-Sigma = .850  < 1.795%

Vendor verified that MD variability was a bit over 1%, but added that the control system was doing a fine job in spite of the fast basis-weight variability coming from the wet end of the machine.

The mill's electronics and instrumentation department checked all instrumentation and confirmed that all was fine.

At this juncture, the mill called in EnTech to perform a process control audit.

Process Control Audit

The audit team first confirmed that the variability observed by mill personnel was correct. This was done by collecting scan-average, basis weight data and analyzing the results using EnTech's time series analysis software.

Fig. 1 shows about 90 minutes of basis-weight scan-average data on 69-lb linerboard. The machine direction "2-Sigma" was found to be 0.87 lb or 1.25% off target, which confirmed the mill staff's observation that the variability was a bit over 1%, based on reel turn-up 2-Sigma reports generated about every 30 minutes.

The left graph of Fig. 1 shows the basis weight scan average variability as a function of time. The right hand graph of Fig 1 shows a power spectrum, which decomposes the basis weight variability into its frequency components.

Analysis of the graph data in Fig 1 showed how much the variability at a given cyclic period contributed to the total 2-Sigma. It was obvious from this data that there was significant variability present at about 14 minutes per cycle due to a clearly visible tendency of basis weight to cycle at this time period.

The audit team next examined basis weight variability in single point with the scanner positioned at the center of the sheet for about 18 minutes. The data consisted of 5 second average samples collected by the computer. This was done because experience has shown that a 45-second scan average tends to hide a great deal of fast variability (see Fig 2).

Analysis of the data in Fig. 2 showed a 2-sigma variability of about 3.2%, which was much higher than the 1.25% measured previously. Other cyclic disturbances became visible with the shorter time periods—the biggest of which occurred at about 65 seconds—but there were others at 3 minutes and 6 minutes. These data have tended to confirm that the wet end of the machine was not very stable and that there were very clear cyclic disturbances.

Because the five-second averaging process was also known to hide some very fast variability, the EnTech audit team used high-speed data collection equipment to measure the voltage signal directly off the beta gauge, taking care first to scale the data to pounds of basis weight in the analysis computer.

Fig. 3 showed the results of this analysis over a 14-minute period. The variability was now much higher with a '2-sigma' of more than 8%. The 65 second cycle was...
During the audit, all consistency and stock flow controls were evaluated.

The audit team carried out a thorough survey of the wet end control loops and process variables. During this investigation, the audit team measured whitewater dilution header pressure as shown in Fig. 4. The header pressure cycles at both 65 seconds and 14 minutes per cycle.

After a thorough review of the whitewater users that were fed off this dilution header, the secondary headbox rejects tank level controller, LC-203, was found to cycle with a period of 65 seconds. This in turn was caused by an inappropriate tuning of the level controller, the reset time of which was at 0.1 minute—a high oscillatory setting.

This cycle, which was caused by the controller, had the effect of destabilizing the dilution header, which in turn destabilized the consistency controllers in both machine chests, the primary blend chest, and the secondary LD chest. It also destabilized the operation of the secondary headbox.

Fig 5, which plots the 5-second average reel moisture in single point, showed the deleterious impact of the controller. The left data on Fig. 5 was with LC-203 in the manual mode. It showed no evidence of a 65-second cycle. The right hand data showed moisture with LC-203 in automatic with the original tuning values. Here, the 65-second cycle was clearly visible.

**CORRECTIONS**

LC-203 was re-tuned using a critically damped tuning method to ensure that level oscillations did not occur.

The elimination of the 65-second cycle had a major impact on the operation of the paper machine. MD stability and CD performance improved significantly. Moreover, CD control was stabilized.

The wet end investigation also revealed that with its level controller on automatic, the whitewater chest cycle with a period of 14 minutes (see Fig. 6). Because this result was caused by both process and loop design problems that could not be overcome by tuning, a new whitewater chest level control arrangement design was necessary.

Analysis of the four interconnected whitewater tanks (see Fig. 7) disclosed that the system was inherently unstable and had to cycle. As a result, EnTech recommended a redesign of the level control to eliminate this problem.

**AUDIT ANALYSIS**

The 65-second and 14-minute cycles had a significant impact onlinerboard uniformity. However, it should be noted that they were only 2 of 60 process variables that were closely investigated during the audit.

During the audit, all consistency and stock flow controls were evaluated. All stock chest and whitewater level controls were inspected. All loops that had a major impact on transporting or processing fiber or affecting the sheet were reviewed.

The results of this thorough investigation revealed that of the 60 loops, five had no significant problems, seven required minor tuning, and 32, including LC-203, required extensive retuning to minimize variability. Twenty-three required instrument repair or maintenance before the loop could be tuned effectively, nine required control strategy redesign before they could perform properly, and five required changes to make the loop function effectively.

**Figure 5:** Fixed point moisture variability with LC-203 on Manual (left) and Automatic (right).
The process audit also had a major impact on mill management and staff.

Figure 6: Whitewater chest level variability.

![Graph showing whitewater chest level variability with curves and data points]

- Mean = 51.653
- 2-Sigma = 4.214 (8.158%)

Figure 7: Whitewater chest level control.

![Diagram of whitewater chest level control system]

RESULTS
The audit had an immediate impact on variability, machine efficiency, and customer satisfaction, even before all the recommended improvements and repairs were carried out.

The major finding of the audit was that the 65-second cycle was a major cause of machine instability. Its effect on sheet moisture was such that the corrugator operation could not compensate. Elimination of the instability meant that most of the complaints and rejects, which comprised nearly 5% of the mill's production, disappeared.

The process audit also had a major impact on mill management and staff. The audit served to open the eyes of mill personnel to the operation of their machine. Who would have thought that a little level controller as seemingly insignificant as LC-203 could have done so much damage?

The thorough examination of paper machine control raised a number of issues for mill personnel such as loop tuning skills, valve maintenance standards, the mill's ability to check performance in a proactive manner, and the tools needed to measure variability.

The audit exposed what could be a major chink in any mill's armor. Closer attention to the performance of process control loops is necessary if mills are going to meet their customer's product uniformity requirements.

William L Baikowska is president of EnTech Control Engineering Inc., Toronto, Ont., a company that specializes in the application of process control for the pulp and paper industry.
Quarter Amplitude Damping Method Is No Longer The Industry Standard

By W.L. Bialkowsk and Brian Haggman

Times have changed and there now exists within the industry new modern control tuning methods...

This year marks the 50th anniversary of a pioneer paper written by J.G. Ziegler and N.B. Nichols on how to tune PID controllers to obtain quarter amplitude response dammed to setpoint changes. Chances are good that engineers and technicians attending loop tuning training classes today would be taught some variation of this method, which was first written about in 1942.

Although the Q.A.D method espoused by Ziegler and Nichols was a good starting point, times have changed and there now exists within the industry new modern control tuning methods that are better at minimizing variability. Indeed, that Q.A.D method that is used so commonly throughout the pulp and paper industry is not an effective way to insure product uniformity.

**METHODOLOGY**
What is the quarter amplitude damping tuning method? Figure 1 shows a simulated consistency control loop tuned for a quarter amplitude damped response to a setpoint change. The loop has a natural process response consisting of 5 seconds of deadtime due to a transport delay to the consistency transmitter and 2 second time constant. The quarter amplitude damping loop tuning rule requires that the loop overshoots the new value of the setpoint by 50% and then oscillates in a damped sinusoid in such a way that the next overshoot is a quarter of the first.

People in the process industries, including pulp and paper, have been using this tuning method for 50 years. Nearly all of these people have applied this tuning method based on the assumption that it produced loops that gave “the best possible response.”

Analysis of the quarter amplitude damped response to setpoint change will show that this tuning method is no longer the best possible method and that response of the process variable (PV) to setpoint changes, the loop will tend to resonate and amplify setpoint changes that were made at a frequency of 0.3 radians/sec or about 20 seconds. The magnitude of the resonant peak is an amplitude ratio of 2.4. This means that if the setpoint were cycled at 20 seconds per cycle, then the PV would cycle with an amplitude more than twice as large as that of the setpoint.

Figure 2 shows the load response Bode plot for the same loop, which relates the effect of external disturbances, such as dilution pressure header changes, on the consistency PV over a range of frequencies. This means that the loop is very effective at reducing the impact of slow disturbances such as at the frequency of 0.01 radians/sec or a period of about 10 minutes/cycle, at which point the effect of the disturbance would be reduced to 0.05, i.e., that 95% of the disturbance would be eliminated. However, at a frequency of 0.03 radians/sec (a period of 20 seconds/cycle),
the loop would amplify any variation present in the PV by a factor of more than three times.

Figure 3 shows the performance of the loop under normal control in the typical mill environment, where most signals are noisy and tend to drift due to process disturbances (n.b., the integrating moving average type noise structure is typical of many consistency problems.) The left hand graph illustrates a time trace of the consistency on control. The right hand graph shows the power spectrum. This figure shows the total variability broken down by frequency harmonics plotted on a log-log scale that is similar to a Bode plot. Note that at a frequency of about 0.04 Hz, or a period of 20 sec./cycle, variability is being amplified by more than a factor of 10.

The performance of the same loop under the same conditions, after it has been tuned using the Lambda tuning method with a close loop time constant (Lambda) chosen to be 12 seconds in order to minimize the impact of the specific random variability typically present in this loop, shows that the variability in consistency has been reduced by 60% based on the 2-Sigma value.

MORAL OF STORY

For the last 50 years, the only tuning method that has been commonly available to loop tuners has been the quarter amplitude damped method. The only alternative to Q.A.D. was to tune "by the seat of the pants."

In looking at the figures presented in the article, it is probable that the Q.A.D method alone has probably been responsible for destabilizing paper uniformity by as much as a factor of 2 in a noisy environment. With a cycle time of about 20 seconds, the consistency loop described above would contribute too much fiber for 10 seconds and too little fiber for 10 seconds and so on forever. At 3,000 fpm the typical paper machine would produce a heavy weight for 1,500 feet followed by a light weight for the next 1,500 feet. This is not an effective way to ensure product uniformity and maintain market share.

The moral of this story is that today's control engineers have available to them modern tuning methods that minimize variability much better than the Q.A.D method. These new tools include Lambda tuning, internal mode control, minimum variance control and robust control loop tuning concepts.

The improved results shown in Figure 3 were derived from a pulp and paper control audit and optimization program that was undertaken by EnTech Control Engineering. A number of courses are readily available to teach these tuning methods to pulp and paper mill engineers and technicians.
PROCESS CONTROL RELATED VARIABILITY AND THE LINK TO END-USE PERFORMANCE

W.L. BiaŁkowski
EnTech Control Engineering Inc.

The link between end-use performance and product uniformity has not been clearly established in the pulp and paper industry to date. Recent audits of paper machines and Kraft mill operations have provided considerable insight into process and product variability in the process control frequency spectrum from one cycle per second down to one cycle per hour and longer. It is common to find process uniformity destabilized by control loops which cycle due to poor loop tuning or instrument state of repair. The paper presents the process variability results from typical audits of digesters, bleach plants, stock preparation areas and paper machines and discusses the implications of such variability in the Kraft pulp, newsprint, fine paper and linerboard sectors of the industry. A new variability index which classifies variability by frequency decades is presented. The index is used to discuss the possibility of predicting end-use problems as pulp bales and paper rolls are used in paper operations, pressrooms, converting operations and box plants. The ability of a pulp or paper manufacturer to improve competitive position by improving product uniformity through a more disciplined application of process control techniques is emphasized.

1.0 INTRODUCTION

The competitive position of pulp and paper manufacturers is increasingly being tested as global trends towards tighter competition continue to strengthen. As printers and paper converters strive to meet tighter print quality specifications they install new technology equipment and escalate the demands for paper quality. Chief among these demands is for product uniformity and consistency. To meet these demands for uniformity the paper manufacturer must improve process equipment, improve process control practice and ultimately demand more uniform pulp supply. When these competitive pressures reach the pulp manufacturer, his options are essentially the same: improve the process, improve process control and demand more uniform wood chip supply. This paper builds on an earlier work which dealt specifically with newsprint [1] and reviews the process variability typically present in pulping and papermaking operations especially as it relates to operation of control systems. The impact that the
present in pulping and papermaking operations especially as it relates to operation of control systems. The impact that the resulting product variability may have on the customer’s operation will be explored.

2.0 OLD CONCEPTS OF VARIABILITY

The pulp and paper industry has historically gauged product quality simply by reporting averaged values of pulp or paper test properties such as K.No., viscosity, dirt count, basis weight, tear, opacity, brightness etc. once per shift hour or reel. The concept of variability has been that of calculating the standard deviation or ‘2-sigma’ of these tests about their average values. The industry also uses the variance analysis reported in the typical paper machine computer reel report, which analyses basis weight and moisture variability in terms of total, machine direction (MD), cross direction (CD) and short term (ST) or residual variability expressed as a ‘2-sigma spread’ or two standard deviations. All of these statistics report the variability spread but give little indication of rates of change. Unfortunately these combined statistics have given the pulp and paper maker little practical insight into the variability in the product which relates to runnability and end use performance at his customer’s plant. Recent articles on end-use properties stress the need to quantify and understand paper variability [2] especially as it relates to the dynamics of the process used in the next stage of conversion. The rates of change of product uniformity are critically important issues in this context.

2.1 Process Variability Spectrum - As Seen at the Paper Machine

Consider the spectrum of process variability on a 1000 m/min paper machine. Variability can be considered from the very rapid surface topography of the sheet as determined by fibre dimensions (3 mm x 30 μm) all the way through to shift, daily and seasonal variation. Causes of variability can often be traced all the way back to the pulp mill in an integrated mill or bale-to-bale variation when pulp is purchased on the market. Table I tabulates typical causes of MD variability from very high frequency to very low frequency for a paper machine operating at 1000 m/min (3300 ft/min). Figure 1 presents similar information in graphical form. The frequency and oscillatory period are plotted on a logarithmic scale in which the major divisions are decades of frequency. The scale shows a range over 13 decades from 0.1 μHz or once per 4 months representing seasonal variation and operating problems, all the way through to 0.5 MHz which is equivalent to the width of a softwood fibre at 1000 m/min.
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<td></td>
<td>9-28µHz</td>
</tr>
<tr>
<td><strong>Sampling/Testing Frequencies</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Set of Paper - period/set</td>
<td>15km (9mi)</td>
<td>15min</td>
</tr>
<tr>
<td>Jumbo Reel - period/reel</td>
<td>45km (28mi)</td>
<td>45min</td>
</tr>
<tr>
<td>SPC Sample Frequencies</td>
<td></td>
<td></td>
</tr>
<tr>
<td>per reel - period c/2 reels</td>
<td>90Km (56mi)</td>
<td>90min</td>
</tr>
<tr>
<td>per hour - period c/2 hours</td>
<td>120Km (75mi)</td>
<td>2hr</td>
</tr>
<tr>
<td>per shift - period c/2 shifts</td>
<td>960Km (600mi)</td>
<td>16hr</td>
</tr>
<tr>
<td>per day - period c/2 days</td>
<td>2880Km (1800mi)</td>
<td>48hr</td>
</tr>
<tr>
<td>per month - period c/2 months</td>
<td>87000Km (55000mi)</td>
<td>2months</td>
</tr>
<tr>
<td></td>
<td></td>
<td>0.2µHz</td>
</tr>
</tbody>
</table>
2.2 Very High Frequency - Furnish Mix

The variability in sheet surface uniformity caused by adjacent fibres can be decomposed into wavelengths as short as 30 μm. At 1000 m/min (3300 ft/min) this translates into periods of 1.8 μs or frequencies of 0.56 MHz. This represents the high frequency end of the paper variability spectrum. Similarly, fibre length, wire mark and floc size are in the 2 mm to 2 cm range (about 0.2 to 13 KHz). All of the above relate to the sheet micro-structure and are largely determined by furnish mix and fibre structure. To this extent the wood species choice and pulping process have great influence. Consider the differences between softwood and hardwood fibre or the differences in the final fibre structure for black spruce after processing via either the stone groundwood, TMP or Kraft processes. Furnish mix is, in turn, determined by the blend chest fibre and additive flow ratio controls and may well continuously vary if these controls cycle.

2.3 High Frequency - Process Design/Maintenance Problems

In a similar vein, process design and maintenance consideration include headbox vibration [3], hydraulic pulsations caused by fan pumps, eddies, turbulence and clothing rotation. These have their impact on variability in roughly the 0.2 to 200 Hz or 5 ms to 6 sec range. (Paper length of between 8 cm (3 in) to 100 m (300 ft)).

2.4 Low Frequency - Process Control Considerations and Problems

Of special interest is the variability related to process control from seconds to hours per cycle. This is probably the most ignored and least understood range of the frequency spectrum. The process control spectrum starts at about 1 Hz or 1 second/cycle and extends to slower and slower periods. It can be measured reliably with beta gauges and other on-line sensors. The characteristics of control loops will be discussed in detail later, however the influence of fast loops such as fan pump speed regulators start at about 1.0 Hz. Stock flows start at a few seconds/cycle while the influence of slower control loops such as basis weight start at about 20 minutes/cycle and extend down scale to very long periods of a day and slower. K.No. control based on hourly pulp samples is even slower and has potential oscillatory periods as slow as 10 to 30 hours per cycle. Level control represents a significant source of both pulp and paper mill variability due to the tendency of most level controls to cycle. Often this is due to the way that most people tend to tune these. Level controls represent about 30% of all the control loops in a mill or at least 1000 loops in a typical mill. Level controls will cycle at once per minute for a stand pipe to once per
hour for a stock chest. Another source of significant variability in a non-integrated paper mill is caused by pulper operation. Continuous pulpers operate at perhaps 1 bale a minute, while batch pulpers make up batches at typically 20 minutes per batch. These operations often significantly destabilize consistency controls in a stock preparation department at frequencies which cannot be easily controlled out.

2.5 Very Low Frequency - Pulp/Paper Tests

Pulp mill tests are typically taken once per shift, every 2 hours or every hour. SPC done on shift samples is only sensitive to cycles every 16 hours and slower. Samples taken every hour can only detect variability with a period of 2 hours. Typically a jumbo reel turn-up on a paper machine occurs every 45 minutes on a 1000 m/min paper machine. This represents 45 km (28 miles) of paper and contains variability down to 0.5 MHz. Samples taken from the reel and used for testing, reporting and SPC contain cyclic information at periods of double the sampling interval or 90 km of paper. Similarly, hourly, shift and daily samples represent even longer cyclic periods. Clearly, doing SPC based on once per reel based on grab samples makes for extremely ineffective control, especially when each grab sample has very low statistical significance on, say, an 8 m (315 inch) wide paper machine running at 1000 m/min (3300 ft/min). (A 0.3 m x 0.3 m (1 ft x 1 ft) sample represents 4 ppm of the area of the reel (1 sq. ft. in 85 acres).)

3.0 SENSORS, SIGNAL FILTERING AND INTERPRETATION

Let us briefly review sensor measurement and signal processing principles as they apply to most sensors in order to understand exactly how to correctly interpret on-line data. Sensor electronics normally respond through low pass measurement dynamics. The time constant of a beta gauge, for instance, typically varies from about 20 ms to 100 ms. The time constant is the time necessary to register 63% of a step change in the true signal. It takes 4 time constants to register 98% of the full step change. The cut-off frequency is the point at which faster variability will be increasingly filtered out. This starts at a period of 6.3 time constants (2π). Hence for a beta gauge with time constants of 20 - 100 ms, the output voltage measurement ‘footprint’ is 1 - 2 cm wide in the cross direction and exponentially fades in the machine direction with a time constant of between 0.3 m and 1.7 m (1 to 6 ft) at 1000 m/min (3300 ft/min).

All sensors have some degree of signal filtering, as a result of both the measurement principle and filtering, which has been added intentionally to provide noise smoothing. In many cases additional filtering has also been added in the
mill by pneumatic snubbers, 'RC' filters and software filtering. This all serves to alter how mill staff see and 'think' of their process variability.

Table II illustrates the effect of filtering on attenuating high frequency variability for various choices of time constant. In many cases filtering is added to 'make things look smooth' - to get 'straight lines'. Clearly this hides the true variability which is really there and prevents the mill from being conscious of it. This variability will undoubtedly affect the operation of the customer's plant. For instance, a bleach plant brightness sensor with a filter time constant of 30 seconds is able to measure variability only as fast as 3 minutes per cycle reliably. Faster variability is attenuated. Since the lay-boy drop rate is likely faster than 3 minutes, then there can easily be bale-to-bale variability in brightness actually present which is too fast for the chlorination stage sensors to have measured properly when the pulp in question was entering the chlorination stage.

<table>
<thead>
<tr>
<th>Typical Sensor</th>
<th>Time Constant $\tau$</th>
<th>Cut-off Period $2\pi \tau$</th>
<th>Period at 90% Attenuation (-20 db)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fast Beta Gauge</td>
<td>20 ms</td>
<td>120 ms/cycle</td>
<td>12 ms/cycle</td>
</tr>
<tr>
<td>Fast Pressure</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Slow Beta Gauge</td>
<td>100 ms</td>
<td>0.63 s/cycle</td>
<td>0.063 s/cycle</td>
</tr>
<tr>
<td>Pressure</td>
<td>1 sec</td>
<td>6 s/cycle</td>
<td>0.6 s/cycle</td>
</tr>
<tr>
<td>Flow</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Fast Consistency</td>
<td>3 sec</td>
<td>20 s/cycle</td>
<td>2 s/cycle</td>
</tr>
<tr>
<td>Slow Consistency</td>
<td>10 sec</td>
<td>60 s/cycle</td>
<td>6 s/cycle</td>
</tr>
<tr>
<td>Slow Flow</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Fast Pulp Brightness Sensor</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Heavy Snubber or Filter</td>
<td>1 min</td>
<td>6 min/cycle</td>
<td>40 s/cycle</td>
</tr>
<tr>
<td>Extreme Snubber or Filter</td>
<td>10 min</td>
<td>60 min/cycle</td>
<td>6 min/cycle</td>
</tr>
</tbody>
</table>
4.0 CONTROL LOOP CHARACTERISTICS

A modern pulp and paper mill has some 3000 to 5000 control loops and cannot be operated without most of these being in automatic control. Control loops provide two functions: to allow the process to be operated at chosen targets; and also to reduce the impact of raw material variability or process upsets and hence to maximize product uniformity. Thus the operation of control loops is of critical importance to market acceptance of the pulp and paper products. Well designed, tuned and maintained loops will reduce variability. Badly maintained and tuned loops may well increase variability. Control loops can be implemented in pneumatic controllers, electronic controllers, the latest DCS systems and computer systems. Analog controllers use mainly proportional and integral control (PI). Modern digital systems also primarily use PI control. Occasionally more advanced algorithms are used in a special application. Control loops have a number of generic characteristics which are independent of the control system equipment used.

4.1 Regulatory Control Loops - Generic Characteristics

Control loops such as headbox total head, stock consistency, basis weight and moisture are intended to regulate out the effects of process upsets and hence to maximize product uniformity. In all cases a control loop will be more effective at regulating out slower disturbances as opposed to fast ones. This generic property holds for all disturbances at frequencies slower than the loop’s cut-off frequency.

4.2 Loop Cut-off Frequency

The cut-off frequency of a regulatory loop is a function of: process dynamics (time constant, deadtime); type of algorithm used; and tuning employed - the faster the tuning, the faster the cut-off frequency. Unfortunately, as the loop tuning becomes more aggressive, the more likely it is that the loop will oscillate near its cut-off frequency and hence degrade product uniformity. The actual speed of response of the loop is determined by how the loop has been tuned and is measured by its closed loop time constant. For stability and robustness this is often about three times slower than the natural process time constant, although the actual value depends on many tuning issues. In turn, the period at the cut-off frequency is $2\pi \tau$ (about 6.3) times the closed loop time constant. Examples of typical values for various control loops are listed in Table III.

Figure 2 shows how effectively four of the examples listed in Table III are able to regulate out disturbances at various frequencies. Shown are both the ideal cases where the
### TABLE III

Typical Control Loop Dynamics

<table>
<thead>
<tr>
<th>Process</th>
<th>Process Time Constant</th>
<th>Closed-Loop Time Constant</th>
<th>Cut-Off Period</th>
<th>Period at 90% Attenuation (-20 db)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fan Pump Speed</td>
<td>0.1 sec</td>
<td>0.3 sec</td>
<td>2 s/cycle</td>
<td>20 s/cycle</td>
</tr>
<tr>
<td>Total Head</td>
<td>1 sec</td>
<td>3 sec</td>
<td>20 s/cycle</td>
<td>3 min/cycle</td>
</tr>
<tr>
<td>Consistency</td>
<td>5 sec</td>
<td>15 sec</td>
<td>1.5 min/cycle</td>
<td>15 min/cycle</td>
</tr>
<tr>
<td>C-Stage Brightness</td>
<td>0.5 min</td>
<td>1.5 min</td>
<td>9 min/cycle</td>
<td>1.5 hr/cycle</td>
</tr>
<tr>
<td>Basis Weight</td>
<td>1 min</td>
<td>3 min</td>
<td>20 min/cycle</td>
<td>3 hr/cycle</td>
</tr>
<tr>
<td>Moisture</td>
<td>2 min</td>
<td>6 min</td>
<td>40 min/cycle</td>
<td>7 hr/cycle</td>
</tr>
<tr>
<td>PM Colour Control @ Fan Pump</td>
<td>5 min</td>
<td>15 min</td>
<td>1.5 hr/cycle</td>
<td>15 hr/cycle</td>
</tr>
<tr>
<td>O₂ Delignitation K.No. F.B.</td>
<td>10 min</td>
<td>0.5 hr</td>
<td>3 hr/cycle</td>
<td>30 hr/cycle</td>
</tr>
<tr>
<td>PM Colour Control @ Blend Chest</td>
<td>20 min</td>
<td>1 hr</td>
<td>6 hr/cycle</td>
<td>2.5 day/cycle</td>
</tr>
<tr>
<td>Digester K.No.</td>
<td>1 hr</td>
<td>3 hr</td>
<td>20 hr/cycle</td>
<td>8 day/cycle</td>
</tr>
</tbody>
</table>
regulator is able to reduce variability by up to 30% at the cut-off frequency (–3db) and also cases where the control loop oscillates by different amounts and amplifies variability. Figure 2 and Table III illustrate how relatively limited basis weight and K.No. control actually are in reducing fast variability. This limitation is solely due to the very slow process dynamics. Figure 2 also illustrates that a cycling loop can amplify variability extensively (by as much as 100% and more at a given frequency).

4.3 Level Control Loops - Common Cause of Cycling

Tank levels are integrating processes and usually prove to be extremely difficult to tune by feel alone, without the loop cycling as a result. The human tendency is to use far too much integral or reset action. Critically damped tuning methods do exist to prevent level controller cycling. However in practice these techniques are not widely known and hence most level controllers tend to cycle. The impact on variability is serious, because cycling stock and white water
chests badly destabilize consistency loops which in turn destabilize fibre delivery. Levels cycle with periods ranging from one minute for standpipes, through to over one hour for chests.

4.4 Variability Impact of Control Loop Tuning

The function of the 3000 to 5000 control loops in a typical pulp and paper mill (some mills have as many as 10,000 loops) is to reduce the impact of raw material variability as fibre flows through the pulp mill and on to the reel of the paper machine. Each loop will have a frequency response similar to that shown in Figure 2. Not only does some residual variability get through to the reel, but a good deal of unnecessary variability is often generated by loops which cycle, loops which interact with each other and loops with sticky valves. From Figure 2 it is clear that consistency variability faster than about 1.5 minutes/cycle (ideal consistency tuning) will get past the machine chest consistency regulator and go straight to the reel. Figure 2 also shows that the basis weight feedback control loop is powerless to remove variability that is much faster than about 20 minutes/cycle and only becomes 90% effective at 3 hours per cycle. This represents four jumbo reels or 12 sets of paper. Hence it is critically important to ensure that each of the 300 paper machine control loops upstream of the dry end is tuned as well as it can be, since basis weight and moisture control can only reduce very slow variability. By the same token the effectiveness of digester K.No. control is even slower with a cut-off of about 20 hours/cycle, while it becomes 90% effective at 8 days/cycle.

5.0 MARKET PERSPECTIVE - VARIABILITY AUDIT RESULTS

From a market quality demand point of view it is important to be sensitive to the specifications set by paper buyers. For instance Gannett Supply Corporation which buys all the newsprint for USA Today has set a basis weight variability specification of +/- 1% [4]. Linerboard suppliers are expected to meet roll moisture variability specifications of +/- 1% on a 2-sigma basis [5,6]. Variability audits of paper machines often yield results that are generally similar from mill to mill. This is symptomatic of generally similar process design, process control design, controller tuning and valve maintenance practice throughout the industry. The variability results presented are extracted from actual paper machine audits and include results ranging from newsprint machines to linerboard machines. These results are also very similar to results previously reported [7].
A) Basis Weight vs Time

B) Power Spectrum

Mean = 48.800  2-Sigma = 0.475  (.974%)

FIGURE 3 SCAN AVERAGE BASIS WEIGHT - ON CONTROL

5.1 Basis Weight MD Variability

Figure 3 shows a typical plot of basis weight scan averages (at approximately 40 seconds per scan) and shows over an hour of operation with basis weight on automatic control running at 48.8 gsm. Figure 3a shows the actual data versus time with each point being the average over one scan of the sheet. The '2-sigma' is 0.48 gsm or 0.97% of the average. This is a typical result that would be considered satisfactory by most paper makers. Figure 3b shows the power spectrum of the same data. This classifies the total variability by frequency content or cyclic period. This particular plot shows that the variability slower than some 26 minutes per cycle has been largely removed. This is because the cut-off frequency of the basis weight control has a period of about 26 minutes per cycle. There is, however, extensive variability at higher frequencies, of which 8 minutes/cycle and 1.2 minutes/cycle are the most obvious. These components are far too fast for basis weight control, which can do nothing to reduce their impact. The reasons for this lack of
FIGURE 4 BASIS WEIGHT VARIABILITY IN FIXED POINT

effectiveness is due primarily to the relatively slow process dynamics of the paper machine while scan times and algorithms have only a secondary influence. Figure 4 shows basis weight variability on the same paper machine collected directly off the beta gauge voltage at 4 readings per second over a 4 minute run. Figure 4a shows the actual data over time, while 4b shows the power spectrum. This data shows very fast basis weight variability of 3.9%, virtually all of which was previously averaged out by the scan averaging process as illustrated in Figure 3. This result is also typical of many paper machines, yet few mills are aware of either the extent or significance of such variability. The four minute run shown in Figure 4 represents more than 4 Km (2.3 miles) of paper.

Figure 5 contains a composite plot of the variability for the above paper machine data. The index is used by EnTech Control Engineering Inc. to classify paper machines and will be referred to as EnTech Variability Index (EVI). It classifies a paper machine in two ways: basis weight
variability by standardized frequency bands and also total variability in basis weight. The standard frequency bands are grouped by decades with periods of:

0.1 - 1 second/cycle  
1 . - 10 seconds/cycle  
10 - 100 seconds/cycle  
100 - 1000 seconds/cycle  
1000 - 10000 seconds/cycle

The plot shows the above paper machine basis weight variability per decade expressed as '2-sigma' percent of mean contribution in that decade. The bar graph on the right of Figure 5 shows the total variability for this example paper machine expressed as a '2-sigma' percent of mean. These values are also shown against the average of all paper machines surveyed to date. The minimum and maximum values of the variability in each category are also shown. From

FIGURE 5 ENTECH VARIABILITY INDEX (EVI) FOR BASIS WEIGHT VARIABILITY ShOWN IN FIGURE 3 AND 4
Figure 5 it is clear that overall variability of over 4% in basis weight is quite common in the industry and that it ranges from 2% on a brand new paper machine to nearly 10% on a relatively old linerboard machine with a number of badly tuned loops. In the five frequency ranges between less than 1 second per cycle and over an hour per cycle, the variability extends from 1% and less to over 6% in each band. Hence the example paper machine is about average. This index has considerable potential for assessing competitive position.

5.2 Causes of Variability on the Paper Machine

Why is the variability in basis weight so high? A close examination of some of the other 300 process variables on the paper machine often reveals an abundance of clues. Figure 6 shows six examples taken at random from various paper machine audits. Figure 6a shows a machine chest consistency over a 30 minute period. The variability is 2.2% of fibre flow and contains strong components at 15 minutes per cycle (caused by a white water chest cycle due to poor tuning) and at 5 minutes/cycle, caused by the saveall discharge tank oscillation (due to poor tuning). Figure 6b shows a saveall discharge tank level oscillation (due to poor tuning). The major disturbance at 2 minutes/cycle is caused by the saveall rotation speed. Figure 6c shows a white water header pressure. This header supplies all of the consistency regulators with dilution water. The 4 minute disturbance is caused mainly by the dry end pulper operation and indicates a piping design error. Figure 6d shows a TMP stock flow into a blend chest. The oscillation of nearly +/- 12% at 8 seconds per cycle is caused by aggressive tuning using the quarter amplitude tuning rule, together with a valve positioner with excessive gain. Figures 6e and 6f show groundwood and sulphite flows also going to another blend chest. Again both loops have been tuned for quarter amplitude damping but they oscillate at different frequencies (one at 1 minute/cycle the other at 25 seconds). Hence they upset the blend proportions on a continuous basis. All three of the above examples (Figures 6c, d, f) illustrate how the furnish mix can be so easily upset on a continual basis for trivial reasons.

In addition to the causes of variability illustrated above, a major cause of variability is pulp supply. Variability in consistency, brightness, strength, viscosity and freeness can often be traced back to the pulp mill. In one paper machine audit a four minute consistency cycle was traced all the way back to a disturbance in the pulp mill. In non-integrated mills, the operation of continuous pulpers is a common and serious cause of consistency variability which is often beyond the cut-off frequency of all the consistency controls. For batch pulpers, the consistency variability is both slower and larger. Although consistency control can be reasonably effective at a batch pulping frequency of 20 minutes per batch, this is still beyond the cut-off of basis weight control.
A) Machine Chest Consistency vs Time
Mean = 3.15  2-Sigma = 2.2%

B) SaveAll Tank Level vs Time
Mean  65.27  2-Sigma = 4.9%

C) White Water Header Pressure vs Time
Mean = 36.38  2-Sigma = 17.8%

D) TMP Stock Flow to Blend Chest vs Time
Mean = 590  2-Sigma = +/- 11.6%
E) Groundwood Flow to Blend vs Time
Mean = 685.6  2-Sigma = 6.2%

F) TMP Flow to Blend vs Time
Mean = 549  2-Sigma = 5.3%

FIGURE 6  A FEW CAUSES OF HIGH PAPER MACHINE VARIABILITY

6.0 PULP MILL VARIABILITY AND ITS IMPACT ON THE PAPER MACHINE

Figure 7 shows the K.No. variability in a continuous digester operation in a Kraft mill cooking softwood species. The data represents two and a half weeks of operation with K.No. samples collected every hour. Note that the highest variability occurred at a period of about 130 hours or over 5 days per cycle. This is much slower than the theoretically possible cut-off period of 10 - 30 hours of Table II. Unfortunately pulp quality data was only available based on hourly K.No. samples taken after the screens. Figure 8 illustrates the behaviour of the digester counter wash flow on the same digester. This shows a clear cycle at about 80 seconds and suggests strongly that other digester variables including K.No. also are likely to have variability at the same frequency even though this is not visible in the hourly results.
FIGURE 7  CONTINUOUS DIGESTER K.NO. VARIABILITY BASED ON 1 HOUR SAMPLES

Mean = 24.5  2-Sigma = 5.3  (19.9%)

FIGURE 8  DIGESTER COUNTER WASH FLOW 1 SECOND READINGS

Mean = 511.3  2-Sigma = 7.63  (1.49%)
Figure 9 shows a chlorination stage brightness control loop in manual. The brightness variability over 1000 seconds shows a slow drift with a period of about 15 minutes, likely caused by K.No. changes in the digester. With at least 10 seconds of signal filtering in the sensor electronics it is important to realize that the brightness signal cannot show significant variability much faster than 1 cycle per minute. Figure 10 shows the same brightness control loop on automatic control. There is a significant brightness cycle with a period of about 1.5 minutes and the variability is nearly 40% higher on automatic. The cause of the problem is due mainly to the relative tuning of the chlorine flow control and the brightness control loops. The net result is pulp with the correct brightness on the average. However, 50% of all production is over-chlorinated, has poor viscosity and likely has contributed to the discharge of over-chlorinated organics. The other half of the production is under-chlorinated, contains excessive shives and dirt count but has good viscosity. It is highly unlikely that this brightness variability can be eliminated in subsequent bleaching stages. The cycle period of 1.5 minutes is slightly faster than the bale drop rate of 2 minutes/drop on the pulp dryer lay-boy. This guarantees that there is bale-to-bale brightness variability leaving with the finished product. When these bales are used in the customer's paper mill, they will likely destabilize brightness and colour control at rates much faster than feedback control can correct, as paper machine colour and brightness control typically have cut-off periods well in excess of one hour.

As in the digester, there are many reasons to suspect that actual product variability in the bleach plant is much faster than that indicated by the pulp brightness sensors. Figure 11 shows chlorine gas pressure control under normal conditions. The oscillation at about 40 seconds per cycle is caused by interaction with the chlorine flow loop. Undoubtedly this destabilizes pulp strength and brightness within a bale and will only be partially removed by subsequent mixing. Figure 12 shows the flow of chlorine dioxide solution to a chlorine dioxide bleaching stage. The $\pm$ 7% oscillation in chlorine dioxide solution flow is caused by a 'sticky valve'. The flow setpoint is about 57 USGPM, however the controller is never able to position the valve to the correct setting to achieve this flow. The resulting oscillation has a period of about 3 minutes per cycle and will also destabilize bale-to-bale pulp brightness.

An additional consideration concerning the operation of integrated and non-integrated mills is that in an integrated mill one can consider the pulp mill and paper mill to be essentially one process. Variability present in the pulp mill will be felt in the paper mill some time later. It will contain the same frequency content that was present in the pulp mill except that the higher frequency content would have been attenuated through process mixing, while the lower
FIGURE 9  CHLORINATION STAGE BRIGHTNESS - ON MANUAL 5 SECOND READINGS

FIGURE 10  CHLORINATION STAGE BRIGHTNESS - ON AUTOMATIC 5 SECOND READINGS
frequency content should have been attenuated through process control hopefully, without any variability being amplified as a result of cycling loops. The paper mill controls hence have to deal with this variability to the extent that they can.

In a non-integrated mill, the same pulp mill variability is also present in the pulp bales. These however are usually stored in a warehouse, shipped by train and stored in another warehouse. Eventually they are selected for input into the pulper. However, the sequence in which they are run on a paper machine is probably fairly random out of the total pulp production from a given pulp mill. As a result they contain samples from the total variability that may have existed in several pulp mills over an extended period of time. This total variability is then injected into the paper machine at the bale pulping rate. This rate can vary from a bale every 30 seconds in some continuous bale pulping operations to a batch pulper every 20 minutes for batch operation. The resulting disturbances however, occur at a rate much faster than most of the control loops on a paper machine can actually cope with. Consider for example the nearly 2 weeks of pulp production illustrated by the K.No. variability of Figure 7. The '2-sigma' is 19% and the highest variability occurs at about 130 hours per cycle. If this digester were feeding a paper machine in an integrated mill, there would likely remain

![Graph](image)

**FIGURE 11 CHLORINE GAS PRESSURE - ON AUTOMATIC 1 SECOND READINGS**

Mean = 82.6 2-Sigma = 0.546 (0.661%)
some residual disturbances in pulp quality at this dominant frequency, as the stock is fed to the paper machine. If the same pulp is purchased in bale form and consumed over a 2 week period, then the majority of the 19% ‘2-sigma’ will likely exist primarily at the disturbance period of the bale pulper of say 20 minutes per cycle. A much more damaging frequency, since fewer control loops will be able to cope and the machine will have more difficulty absorbing this effect.

6.1 Variability Impact on Press Operation

A modern press is a web-fed machine which often has several controls, including tension control in the MD sense and side register control to maintain CD alignment. Presses with four and five colour printing may also have automatic colour register control which dynamically adjusts individual press roll speeds in order to maintain colour register. Like on the paper machine, these control loops will respond differently to the incoming variability as a result of design, maintenance and tuning. Should they resonate at a given frequency, then the presence of that specific frequency in a given paper roll can induce resonance and potentially cause amplification, just like a vehicle without shock absorbers
will bounce at its natural frequency when the road unevenness contains components at the natural frequency.

Hence a press is likely to be specifically sensitive to certain paper roll variability signatures and not to others. As a result some rolls will run better than others, even though their absolute variability may be of similar magnitude.

7.0 PAPER VARIABILITY AND CONVERTING EQUIPMENT RUNNABILITY

The unit of production in the paper mill is the roll of paper. It typically takes 15 minutes to manufacture one set of paper from which several rolls are cut. These have similar MD variability and different CD variability present. The MD variability extends from 15 minutes per cycle through 1Hz and continues to the very high frequencies as determined by furnish properties. Typical variability below 1Hz was discussed in the previous section. Paper machines differ on the EnTech Variability Index (EVI) scale not only in their total variability but also in their dominant frequency content. Most machines have total basis weight variability ranging from about 2% to in excess of 4% on a ± 2-sigma basis although variability as high as +/- 10% is possible in extreme cases. The dominant resonant frequencies vary extensively. Some are at 15 seconds/cycle, others at 1 minute/cycle others at 5 minutes and others at 17 minutes/cycle. Individual causes of variability also vary, but often include the control cycling of such variables as: consistencies; white water header pressures; white water chest levels; deculator overflows; cleaners; broke pulpers; savealls; instrument air headers. Each machine has a unique signature which the EVI characterizes. The index can offer both a competitive comparison for the papermaker and a sensitivity index for the pressroom, cut-size sheeter or corrugator operator.

7.1 Pressroom Operation

A large modern pressroom can easily consume in excess of 500 rolls of newsprint in a single daily run. Rolls are usually supplied by several manufacturers, off various paper machines and CD positions. The likelihood of rolls from one paper machine being run on a press in the same order in which they were manufactured is virtually zero. Typically a roll will run-out in about 20 minutes or so. The press is thus subjected to the in-roll variability for about 20 minutes followed by a massive discontinuity to the next roll. The discontinuity exists in both the MD and CD sense.

Figure 13 attempts to illustrate the impact of in-roll variability on press runnability. The variability exists not just in basis weight but also in furnish mix, moisture, modulus of elasticity and essentially all other paper
FIGURE 13 ILLUSTRATION OF MACHINE AND ROLL VARIABILITY IMPACT ON PRESSROOM RUNNABILITY

parameters. The frequency content for basis weight is likely common for many other variables.

7.2 Converting Operations

Converting operations such as cut-size sheeters, off-machine coaters, polyethylene extruders are similar to the newsprint press to the extent that they are web fed operations. They have control loops for at least tension control and may have more complicated operations such as coater control and moisture control. The variability of the incoming raw stock hence provides the disturbance energy for the operation of the converting operation. The control loops of the converting operation may or may not be suitably effective to handle these disturbances. In the worst case the incoming variability may be coupled into their specific resonant frequencies and cause extensive destabilization of the product uniformity.

7.3 Box Plants and Corrugators

Corrugators typically manufacture boxboard by gluing corrugated medium to both top and bottom liners. This requires two rolls of linerboard and one roll of corrugating medium. These products typically are manufactured on at least two if not three paper machines. A predominant problem faced by the box plant is to manufacture warp-free board. Warp is induced when there is excessive differential between the moisture of the three rolls. To minimize this problem the packaging industry normally specifies an average roll moisture of 6% +/- 1.5% or a ‘2-sigma’ of 1% [5,6]. By this standard
the linerboard moisture variability shown in Figure 14 would be acceptable. This shows over 2 hours of scan average moisture data where each point is a 1 minute scan average. At least 3 sets could be cut from this data with an average value of 5.9% moisture and a '2-sigma' of 0.5%. However, if the moisture sensor were put in fixed position at any point on the sheet the resulting moisture variability is as shown in Figure 15 in which 20 minutes of 5 second moisture averaged values are shown. Twenty minutes is about equivalent to one set. The averaged value is 6.2% but the '2-sigma' is +/- 1.5%. This high variability is the direct result of an oscillatory wet end control loop which was cycling at about 80 seconds per cycle. This high frequency disturbance was largely averaged out by the 1 minute scan time of scanning sensors. In 80 seconds a typical linerboard machine makes about 500 meters of linerboard which in turn will make 500 meters of boxboard with ample opportunity for significant warping given the moisture variability present.
8.0 THE UNTAPPED COMPETITIVE EDGE

From the audit data presented it is clear that paper machine operation is frequently destabilized by badly tuned control loops which cycle. U.S.A. Today’s published specification for basis weight is +/- 1%, as are linerboard moisture specifications. Are these levels of uniformity actually being achieved? The audit results indicate that 2.5% to 4% for basis weight is a more realistic good current level of performance. Is it not likely then that as much as 50% of the paper is actually outside the specification when the relatively fast variability is considered? Worse still, this fast variability may in turn induce press resonant oscillations at the same frequency or similarly destabilize other converting operations such as the box plant. In a similar way the variability present in the pulp can destabilize the paper making process. How much could the variability be improved if all loops were well tuned? It is certain that a large reduction in variability is achievable but only with a disciplined and structured approach to process control, loop tuning and process variability. Whether or not 1% basis weight variability can ever be reached is debatable. A good process control effort applied on an ongoing basis can likely reduce variability from say the 3% level to below the
2% level. A substantial competitive advantage. A reduction
to below 1% will require much more effort on several fronts.
Why have the 3000 control loops in the mill not been tuned
more effectively? In many cases, mills seldom, if ever, tune
control loops — historically it has not been considered a
priority. Instrument departments have historically been
staffed to only fix things when they do not work. Loop tuning
hence is a luxury seldom enjoyed. Typically the gain and
reset adjustments of each controller are initially set at
‘default’ values in the ‘comfort zone’ shown in Figure 16.
Once set, the tuning values are hardly ever changed unless the
operator complains. The operator, however, has not been
trained to understand variability or to expect any specific
performance. The default gains vary between 0.3 and 3.0 and
reset times between 0.3 and 3.0 minutes/repeat. Such settings
are likely to cause level loops to cycle and to cause flow
loops to be slow to reach setpoint. Some loops have been
tuned to be deliberately oscillatory using the quarter
amplitude damping method of tuning in the belief that this is
the most effective tuning available. Such tuning in the noisy
environment of the pulp and paper mill will likely cause
sustained oscillations as illustrated in Figures 6 d, e, and
f. Quarter amplitude damping may be effective for servo
control or for manufacturing a liquid or gas product.

![Controller Tuning Chart]

**Figure 16** Controller Tuning — Popularly Accept Gain/Reset
"Comfort Zone"
However, in manufacturing a solid product like paper, oscillatory loops cause permanent deviation of the product from target. This deviation then continues over many thousands of meters of product.

A chief cause for ineffective loop tuning throughout the industry is believed to be due to a lack of management awareness and sensitivity to process variability, loop tuning and the extent to which these issues relate to product uniformity and customer satisfaction. It is ultimately a management problem. Only management can steer the organization towards a more competitive position by changing priorities, staffing, training, organization and corporate climate. Fortunately for the paper industry, paper buyers are equally insensitive, at the present time, although this is likely to change fairly quickly as the marketplace becomes still more quality conscious.

A second cause for ineffective loop tuning is a lack of effective training of both instrument technicians and process control engineers in modern loop tuning methods which reduce variability. Currently, instrumentation training usually only covers the quarter amplitude method. This produces oscillatory loops. Most people have rejected this method preferring to just tune by 'feel'. This however often produces sluggish flow loops and oscillatory level controllers. The training received by engineers in control engineering is usually not very practical and seldom includes any tuning rules for reducing variability. The training programs initiated by CPPA in 1984 were intended as a stop gap measure and have been successful [8] in partially bridging this gap, although in limited numbers. There have also been management seminars intended to sensitize management to the issues relating process control and product quality. Currently the CPPA/TS Process Control Subcommittee on Education is active in exploring both the engineering and technical education issues with Canadian universities and technical colleges.

9.0 CONCLUSIONS

Present levels of product variability are often in the 3% range and much greater than +/- 1% as demanded by the marketplace. To a large extent the 3000 or so automatic control loops in a modern pulp and paper mill are seldom tuned in such a way that they actually reduce variability. In fact the poor tuning of these loops often contributes to increased product variability through control cycling. The competitive potential exists for most pulp and paper manufacturers to substantially improve uniformity with an expected improvement in converting performance and customer satisfaction. It is estimated that such effects can likely reduce variability below the 2% level. However, industry management must first choose to reshape priorities and organizations in order to
bring this about. A new index (EVI) has been presented which
classifies paper machine basis weight variability. It uses
standardized frequency bands together with an overall index to
classify paper machines. This index will allow an assessment
of competitive position. Also the lack of good control loop
performance has been linked to a lack of training and
education in modern loop tuning methods for both the
instrument technician and process control engineer. This
issue is currently being further investigated by the CPPA/TS
Process Control Subcommittee on Education.

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and support of the paper companies where EnTech audits have
been carried out, on the basis of which this paper has been
written. Their identity has been kept anonymous and mill data
has been presented in such a way that it cannot be traced to a
specific mill.

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OPTIMIZATION OF POWERHOUSE OPERATION AT CHAMPION INTERNATIONAL INC.


1 Now with Weyerhaeuser Inc, Longview, WA

ABSTRACT

The Champion International Inc Quinnesec mill has experienced boiler instability problems since start-up, to the extent that boiler trips occurred as a result of boiler steam drum level swings. Steam demand changes of only 22.7 tonnes/hr (50,000 lbs/hr) were often sufficient to induce upsets large enough to cause boiler trips. Many attempts were made to correct this problem without success. The problem has now been resolved as a result of a recent project which involved a thorough analysis of the dynamic behavior of the boilers, steam header, fuel system, equipment and instrumentation; the building of an accurate dynamic simulation model of the powerhouse, and Lambda tuning key control loops. The simulation model allowed both equipment sizing and control tuning changes to be assessed. The Lambda tuning methods used to tune the boiler steam drum level controls, accounted for the steam drum shrink/swell effects, and allowed the drum level controls to be both responsive and stable. Similar tuning was used to settle down the bark delivery system the steam header controls and the boiler water conductivity control. As a result of the project, there have not been any boiler trips resulting from drum level upsets.

INTRODUCTION

The Champion International Inc produces 1478 tonnes per day of fine paper and market pulp at the Quinnesec Mill in Norway, Michigan. The powerhouse consists of a recovery boiler, a waste fuel (bark) boiler and a natural gas fired package boiler, all of which feed a common 4136 kPa (600 psi) steam header. The recovery boiler is base loaded. The pressure of the 600 psi steam header is controlled by a plant master controller, which in turn sets the boiler master controllers on the bark and natural gas boilers. There is a 448 kPa (65 psi) steam header that is fed by a 28 megawatt turbine and a PRV. There is also a 1137 kPa (165 psi) steam header which is fed by small turbines and a PRV. All steam users are fed from either the 65 psi or the 165 psi steam header.

Problem Statement

Since start up of the mill, there has been a tendency for the boilers to apparently interact with the 600 psi steam header. When a sudden steam demand was placed on the header, the boilers would react to the change and increase or decrease the firing rate in order to maintain the header at 600 psi. The boilers would then appear to interact through the 600 psi steam header in a cyclic manner, with the drum level in one boiler going up initially and another going down. Steam demand changes of as little as 22.7 tonnes/hr (50,000 pounds per hour) were enough to cause severe cycling, which would continue for some time. There have been many boiler "trips", due to low or high drum level during such upsets. The powerhouse operators had to be very alert in order to take corrective action quickly enough when such "bumps" struck the boilers.

As a result of these problems, it was decided to mathematically model the boiler and powerhouse dynamics to provide a basis for developing a boiler control tuning strategy. Such a model would not only allow control tuning changes to be evaluated, but would also allow physical equipment changes, such as re-sizing of the steam header, or the boiler drums to also be evaluated with equal ease.

This paper describes the development of the model, the control tuning steps that were taken to correct the interaction, and what was learned in the process. The work was done jointly by mill staff at Champion International and EnTech Control Engineering Inc. Site work involved dynamic testing and controller tuning. Off-site work was done at the EnTech office in Toronto, Canada and involved analysis of dynamic results, the development of the dynamic simulation model, design of control strategies and calculation of controller tuning settings. The majority of the on-site bump tests were done by mill staff, with the resulting data being transmitted off-site by modem for analysis.

PROCESS

Figure 1 shows a schematic of the powerhouse operation at the Quinnesec mill. The recovery boiler is a B&W unit.
sized to burn 1524 Tonnes/day (3.3 million pounds) of black liquor dry-solids per day, with a steaming capacity of 222 tonnes/hr (490,000 pounds per hour). The waste fuel boiler is a B&W moving grate type boiler, designed to burn bark, coal, oil or gas, at a steaming capacity of 170 tonnes/hr (375,000 pounds per hour). Under normal conditions, bark is the only fuel used. The bark is transported from the reclaim pile by conveyor. The gas fired package boiler has a capacity of 136 tonnes/hr (300,000 pounds per hour). This capacity is needed during the winter months, and then the package boiler becomes the swing boiler. During the summer months the header is maintained with the waste fuel boiler, with the package boiler on minimum fire in a standby mode. The total power house capacity is 528 tonnes/hr (1,165,000 pounds per hour). The power house is equipped with a Foxboro Spectrum Distributed Control System (DCS).

BOILER DYNAMICS

Data Gathering Procedures

The boiler dynamics were studied by collecting analog signals from the Foxboro DCS termination connectors under various conditions and tests. During the initial one week site visit by EnTech personnel, this was done using EnTech's data acquisition equipment (EnTech Toolkit) to collect data at high frequency. Typically, this equipment collects data every 0.025 seconds or faster and uses appropriate anti-aliasing filters to ensure data integrity. There were up to 16 separate analog input channels available with data collection rates up to 200 samples per second. Since the project was likely to last several months, greater flexibility was provided by writing a BASIC program in the DCS to collect data from a pre selected list of DCS variables. The Foxboro DCS updates its process measurements every 0.5 seconds but the program was not able to collect data this fast. However, the BASIC program was able to collect data at rates of up to 10 seconds. Detailed comparisons of data integrity were made between this data and the EnTech Toolkit data, and it was found that the DCS data was acceptable for most purposes. Naturally, fast data concerning valve or damper nonlinearity could not be studied in this way. Since such data should be collected at rates of at least 0.025 seconds. However, most process responses could be studied using the DCS data. This provided a significant breakthrough for the project, since it meant that most tests could be run by mill personnel. The additional benefit was that mill personnel would have a high level of involvement in the testing program and hence benefit from the learning opportunities that this provided. The data collected in this way was saved in ASCII format and transmitted via modem to EnTech's host computer in Toronto, Canada for analysis. This collection routine was used extensively for the remaining five months of the project.

Testing Methods

Three types of tests were used throughout the project: open loop "bump tests", manual tests, and automatic tests.

Open loop "bump tests" were conducted with the control loop in manual mode, in order to evaluate the process dynamics and the effectiveness of the final control element. Typically, the output of the controller was changed in a step fashion while the process value was monitored. These tests allowed the process dynamics to be evaluated and transfer function parameters such as process gain, deadtime, and time constant to be calculated. Also, the condition of the final control element, such as a control valve or damper drive, could be quantified with respect to the ability of the actuator to respond in a repeatable manner without backlash and stiction (mechanical lost motion). Open loop bump tests are very important as they provide the basis for loop tuning and also for building a dynamic simulation model.

Manual tests were run with the control loop in manual mode. These tests provided insight into the nature of the process disturbances to which the controller was typically subjected. Data from these tests is often used to construct a noise model which can mimic the particular type of random behavior which was observed during the test. This information is useful in certain types of control loop tuning where the intent is to tune the loop to nullify the random disturbances which were observed. The type of data is also useful for building realistic dynamic simulation models.
Automatic tests were conducted with the control loop in automatic mode and provided information about how well the control loop actually worked. These tests were used to characterize the performance of existing control configurations and tuning parameters.

Process Systems Studied

To understand the overall dynamic behavior of the whole powerhouse, the testing methods described were applied to the following process systems in turn: the feedwater header and pumps, boiler drum levels on all three boilers, 600 psi steam header and plant master, waste fuel boiler bark feed system; and boiler water conductivity.

Feedwater Header and Pumps

The feedwater header is supplied by three feedwater pumps. Two are driven by steam turbines while the third is electric. The speeds of the turbines are controlled by local analog controllers. Bump tests on the turbine speed controllers showed that these were not tuned very well. It was also discovered that these analog controllers only provided a relative adjustment capability for their tuning parameters, and could not be placed in manual mode. As a result, tuning would be difficult. Current tuning was such that a change in turbine speed setpoint would typically result in an initial fast 'kick', followed by a slow response. As a result, the header pressure could not be tuned to be adequately fast, which in turn caused the pressure to "droop" when any of the three boiler feed water valves opened. Although this was important, it was decided to solve the problem by only running two pumps at maximum speed, thereby flattening the effective flow pressure droop characteristic of the feedwater header and thereby minimizing the feedwater coupling between all three boilers.

Boiler Drum Levels

Boiler drum level is maintained on all three boilers by a three-element control arrangement, using feedwater, steam flow and drum level as inputs to the control strategy. Logic exists to switch the control strategy to a single-element strategy for start-up or shut-down, when the boiler steam flow is below some pre-determined minimum value. Bump tests were carried out on all three boilers. This was done by putting the steam drum level and feedwater controls in manual mode and doing a series of bumps to the feedwater valve. Great care was taken when doing this. The tests were done with working with the operator at the control console. Each series of tests started with very small bumps of perhaps 2% or 3% then waiting for the feedwater flow and drum level to respond. The flow response is normally very fast while the level response is typically very slow as a result of the delay caused by the feedwater shrink effect, as well as the fact that the drum level is an integrating process and moves in a ramp-like fashion. The feedwater flow controller was tuned based on the feedwater bump test results. It was then placed on automatic cascade mode and drum level bump tests were carried out by stepping the drum level controller output. This series of bump tests usually consisted of ever larger size bumps being done in the opposite direction to the one immediately preceding. Figure 2 shows the latter part of such a series of drum level bumps.

![Figure 2: Package Boiler Drum Level Bump Test Sequence](image)

on the package boiler drum level, where 13FIC125 is the feedwater flow, and 13LC102A is the drum level. The series of increasing bumps is important, as it provides an opportunity for both the operator and the control engineer to work together and build confidence in the procedure. The larger the bumps that can be performed, the greater the value of the dynamic information. During each test, both the operator and control engineer anticipate the moment when to initiate the next test so that the drum level does not go outside agreed-upon limits. In the case shown in Figure 2, the drum level had to be kept within 5 inches of setpoint. This procedure ensures that the tests are done safely. Figure 3 shows the details of one such bump test, together with the resulting analysis of the process dynamics.
Figure 3 also shows the results of the parameter fitting performed by the analysis software for this one bump test. The net result for this series of bump tests was the process is integrating, with a process gain of 0.0050 level (%) per feedwater flow change (%) per second, and there is deadtime of 59 seconds due to the shrink effect of the cold feedwater. Note that the shrink effect for the package boiler is not the classic “wrong way” response. Instead, the additional feedwater seems to cause the level to rise right away. However, this effect is not sustained, resulting in an apparent delay or deadtime. Additional analysis showed that an integral and pure deadtime model provided a good dynamic approximation in this case.

**Drum Level Transfer Function Analysis**

From such bump tests it is possible to derive the linear transfer function that defines the dynamic reaction of the drum level to changes of feedwater flow. The general form of the transfer function is

\[ G_p(s) = \frac{K_p e^{-T_d s}}{s} \]

The results for all three boilers were as follows:

**Table I - Boiler Drum Level Process Dynamics**

<table>
<thead>
<tr>
<th>Boiler</th>
<th>Process Gain ( K_p )</th>
<th>Deadtime ( T_d )</th>
</tr>
</thead>
<tbody>
<tr>
<td>Package</td>
<td>0.0050</td>
<td>59</td>
</tr>
<tr>
<td>Waste Fuel</td>
<td>0.0094</td>
<td>60</td>
</tr>
<tr>
<td>Recovery</td>
<td>0.0104</td>
<td>53</td>
</tr>
</tbody>
</table>

The boiler parameters vary quite significantly. The drum level process gain is primarily a function of the free surface area inside the boiler drum, together with the instrument span of the feedwater flow loop. The apparent deadtime of the shrink effect is determined by the internal baffling arrangement which separates the cold feedwater entering the drum and downcomers from the two-phase mixture entering the drum from the generating tube bank. Some boilers do not exhibit a shrink effect. This data can only be confirmed by the bump test procedure.

**600 psi Steam Header and Plant Master**

Bump tests for the 600 psi header were performed as follows. With the plant master controller in manual mode the controller output was stepped and the 600 psi steam header pressure was monitored. Tests were conducted for two separate cases, either the waste fuel boiler as swing boiler or the package boiler as swing boiler. In both cases the resulting dynamics were integrating with deadtime, as in Eqn 10, and similar in general form to the boiler drum level results. To simplify the results only the dynamic parameters are presented below.

**Table II - 600 psi Header Pressure Dynamics**

<table>
<thead>
<tr>
<th>Swing Boiler</th>
<th>Process Gain ( K_p )</th>
<th>Deadtime ( T_d )</th>
</tr>
</thead>
<tbody>
<tr>
<td>Waste Fuel</td>
<td>0.000914</td>
<td>209</td>
</tr>
<tr>
<td>Package</td>
<td>0.001</td>
<td>21</td>
</tr>
</tbody>
</table>

Of most significance was the large difference in deadtime between the two boilers. The speedy package boiler, firing natural gas, is able to respond quickly to fuel demand signal changes, exhibiting a deadtime of only 21 seconds. On the other end of the spectrum, the waste fuel boiler responds to a change in fuel demand with a deadtime of over 3 minutes (209 seconds, ten times longer). This is due to the transport delay of the waste fuel, the relatively slow heat release dynamics of the bark, and varying moisture content of the waste fuel.

**Waste Fuel Boiler Bark Feed System**

In the project the waste fuel bark delivery system was found to be a major cause of disturbances to the waste fuel...
boiler drum level and hence to the steam header pressure control. Figure 4 illustrates how the bark and chip screen.

![Image of Waste Fuel Delivery System](image)

**Waste Fuel Delivery System**

Waste fuel delivery systems transport fines by a 300-meter long conveyor to two consecutive waste fuel bins. Chip screen fines, which have a more uniform heat value, are fed at a constant rate to the conveyor while bark and hog fuel are fed to the belt by hydraulic rams beneath the reclaim pile. The first bin, or surge bin, is under level control which manipulates the speed of the bark pile reclaim rams to maintain setpoint. From the surge bin, bark is transported to the feed bin which directly supplies the furnace bark feeders. The feed bin is also under level control which adjusts the speed of the discharge screws at the outlet of the surge bin.

Bump tests on the surge and feed bin level loops revealed the following integrating plus deadtime process transfer functions.

<table>
<thead>
<tr>
<th>Bin</th>
<th>Process Gain $K_p$</th>
<th>Deadtime (sec) $T_d$</th>
</tr>
</thead>
<tbody>
<tr>
<td>Surge Bin</td>
<td>0.00108</td>
<td>210</td>
</tr>
<tr>
<td>Feed Bin</td>
<td>0.00434</td>
<td>64</td>
</tr>
</tbody>
</table>

**Table III - Bark Surge and Feed Bin Dynamics**

Boiler Water Conductivity

Another case of integrating plus deadtime process dynamics involved boiler conductivity and continuous blowdown. Conductivity or solids content is controlled by manipulating the continuous blowdown of boiler water from the steam drum. Control of boiler water suspended solids is critical to the prevention of deposition on the water side heating surfaces. Large excursions in blowdown flow also represent disturbances to the feedwater/steam balance, and ultimately can destabilize drum level control. Additionally, unnecessary blowdown of boiler water represents a loss of expensive treatment chemicals and heat energy. From bump tests, the conductivity loop process transfer function was determined to be integrating plus deadtime with the following parameters:

<table>
<thead>
<tr>
<th>Process Gain $K_p$</th>
<th>Deadtime (sec) $T_d$</th>
</tr>
</thead>
<tbody>
<tr>
<td>-0.00004</td>
<td>1211</td>
</tr>
</tbody>
</table>

**Table IV - Boiler Water Conductivity Dynamics**

The deadtime of 20 minutes (1211 seconds) is due to the conductivity analyzer sample line, which is of small diameter and very long. Note that the process gain is negative. This is due to the fact that an increase in blowdown flow causes a decrease in boiler water conductivity.

**DYNAMIC SIMULATION**

Dynamic simulation of the boilers was based as much as possible on thermodynamics and heat transfer first principles, together with actual boiler dimensions. The primary source was the boiler model from Morton and Price [1] which considers the boiler drum as a vessel operating at a vapor/liquid equilibrium established by the drum vapor pressure. Evaporation takes place at a rate determined by the vapor pressure. Hence the boiler will always satisfy the steam demand, but the drum level, boiler water inventory, and stored heat will gradually drop if the net heat input derived from the hot flue gas, less the enthalpy of the incoming feedwater, does not balance the net heat loss. Assuming a constant heat input, an increase in steam demand causes the vapor pressure to be lowered, hence resulting in an increase in evaporation rate and a gradual drop in drum level. An increase in feedwater flow upsets the heat balance, lowers evaporation rate, and causes buildup in drum level. The model did consider the impact of having multiple boilers connected to a common header. The steam flow from each boiler being determined by the differential pressure between each boiler and the common header, which in turn was modeled as a vapor space capacity - hence an integrator. The model did not consider the natural circulation of each boiler, resulting in the model not handling the feedwater shrink effect correctly.

The first boiler which was modeled in this way was the package boiler. The software program used to accomplish the modeling was TUTSIM [2]. The package boiler was transferred to VisSim software [3] to improve the man
machine interface in Microsoft Windows®. The modeling approach taken was to supplement the first principles thermodynamic model by using bump tests results. In the case of the missing feedwater shrunk effect, this was done by adding a transfer function model with the right parameters to the thermodynamic model. The responses of the thermodynamic model were checked against bump test results and, where necessary, parameters were adjusted slightly to account for discrepancies. Figure 5 shows the response of this simple package boiler model to step changes in both steam flow and feedwater flow using the VisSim based model. In this simulation, the package boiler is operated on manual, with fixed firing. Note the "swell" effect which occurs when the steam flow is increased from 83 tonnes/hr (183 kpph) to 92 tonnes/hr (203 kpph). This effect is quite small and short lived, and is caused by the steam flow induced vapor pressure reduction. After this effect is over, the drum level drops at a constant rate due to the fact that the steam flow exceeds the feedwater flow. In the middle of the simulation run the feedwater flow is increased from 83 tonnes/hr (183 kpph) to 107 tonnes/hr (235 kpph), a larger amount than the steam flow was increased previously. This causes a shrink effect, which in this simulation is modeled as pure deadtime as explained earlier. After this effect is over, and a new thermal equilibrium has been established, the boiler drum level starts to rise at a constant rate, as there is now more feedwater being added then steam being taken off.

![Graph showing steam drum level and feed water bump test](image)

**FIGURE 5 VisSim**

**SIMULATION OF PACKAGE BOILER BUMP TEST**

After the package boiler model was verified, it was copied and appropriately customized to form equally accurate models for the other two boilers. All three were then connected to a model of the 600 psi header via individual de superheater sections (modeled as flow resistors with de superheater water flow added). The 600 psi header was modeled as a vapor space with three sources of steam and one steam user. The integration rate of the steam header was matched to actual bump test data.

The feedwater system was modeled by starting with the actual pump curves, pipe flow resistances and valve feedwater characteristics. Combustion control and heat transfer dynamics were grossly simplified by using transfer function models only, since the main focus of the model was on the steam/water side.

The dynamic model confirmed that the boiler drum level controls tended to cycle when the drum level controllers were tuned using the same controller parameters as in the Foxboro DCS. Increasing the size of the drums altered the behavior very slightly. Also, increasing the 600 psi header volume did not change the result. At this point, it was clear that the dynamic problems present in the powerhouse were not of a physical process equipment sizing nature, but rather in the domain of controller tuning, especially the type of Lambda tuning which pertains to an integrating type process with deadtime.

**LAMBDA TUNING**

In the pulp and paper industry, the control loop tuning concept known as Lambda tuning is thought to be the most effective tuning method for uniform manufacturing. The concept started with the work of Dahl in 1968. It has been brought into wide use by extensive training [5, 6, 7] of pulp and paper engineers and technicians. In the last few years it has been enriched by the general concept of internal model control (IMC) [8], and as a result, an extended set of tuning rules for PID controllers has come into general use [9]. In this boiler project almost all of the difficult tuning problems were in the class of loops where the process is integrating with deadtime, as in Eqn. 10. The recommended tuning rule for this case [9] is presented below.

**Integrator plus Deadtime Tuning Rule**

**Algorithm:** PI

\[ G_c(s) = \frac{K_C}{1 + \frac{1}{T_R s}} \]

Select closed loop time constant \( \lambda \). For a level controller \( s \) is the time for the level control to arrest a level change caused by a sudden increase in demand out of the vessel. Based on the value of \( \lambda \) that has been selected, and the
process transfer function parameters $K_p$ and $T_d$ the tuning parameters are

$$T_R - 2\lambda + T_d \quad 30$$

$$K_C = \frac{T_R}{K_p(\lambda + T_d)^2} \quad 40$$

Package Boiler Drum Level -- Old Tuning

The initial tuning of the package boiler had a PI controller with a PB of 75% and a reset time of 5 minutes. This corresponds to a controller gain of 1.33 and a reset time of 300 seconds/repeat. For this case, the loop transfer function is

$$G_{\text{loop}}(s) - G_C(s)G_P(s) - 1.33 \left[ \frac{300s + 1}{300s} \right] 0.005e^{-59s} \quad 50$$

The loop transfer function allows the root locus plot to be calculated. This predicts what will happen to the closed loop poles of the system as the loop gain is varied. The results are shown in Figure 6. The calculation was made with Program CC [10] and was based on using a first order Padé approximation for the deadtime term. Only the area closest to the origin is shown in Figure 6 in order to focus on the most important issues. The plot shows two integrator poles at the origin one from the integrating process and the other from the PI controller. There is a zero from the PI controller reset time at a location of $01/(5 \text{ min} \cdot 60) - 0.0033$. The dominant poles initially try to encircle this zero, but there already is a root locus branch terminating on the zero from the deadtime pole. As a result, the dominant poles swing outwards and move eventually into the right half-plane (RHP). For the chosen gain the dominant poles are at a location which is just starting to become oscillatory (damping coefficient of 0.7), with a slight tendency to cycle (period of 6.28\times10^{-6}/0.0046 \approx 22.9 \text{ minutes/cycle})

The package boiler was chosen as the first boiler to work on for several reasons -- one of which was that it was thought that its tuning was 'pretty good'. Based on this analysis the package boiler tuning is "okay" in the sense that it did not tend to cycle too badly (the other two boilers were much worse). Figure 7 shows the predicted load response for the package boiler with the original tuning. The load response predicts how the drum level response would recover after a sudden increase in steam load. As expected, there is a slight tendency to cycle with a period of about 23 minutes/cycle. Note that the level control arrests the change in about 3.3 minutes.

![Package Boiler Load Response -- Old Tuning](image)

FIGURE 7
PACKAGE BOILER - OLD TUNING LOAD RESPONSE

Package Boiler Drum Level -- New Tuning

The final tuning for the package boiler was based on a $\lambda$ (Lambda), or closed loop time constant of 30 minutes. Attempts to use faster values all caused the level control to be marginally "robust". It was shown that it was not a good idea to select Lambda values much faster than about three times the process deadtime. Using Equations 3.0 and 4.0 and a Lambda value of 30 minutes produces the following tuning constants.
Reset time \[ T_R \quad 2\lambda + T_d \approx 7 \text{ 0 min} \]

Gain
\[ K_C \quad \frac{T_R}{K_P(\lambda + T_d)^2} \approx 1.4 \]

Prop Band \[ PB \quad 100 - 71.4\% \]

1.4

Note that these values are quite close to the original values of 5 minutes and a gain of 1.33. Figure 8 shows the root locus plot for this new tuning. Note that the dominant poles are now located on the negative real axis and as a result there can be no tendency to oscillate. Even if the loop gain should increase for whatever reason, the very worst that can happen is that the poles may move slightly off the real axis, but could not cause notable oscillatory behavior — in sharp contrast to the previous tuning, which would become completely unstable as the loop gain was increased indefinitely. Note also that the dominant pole is located at about 3 minutes, in keeping with the desire to have a closed loop time constant of 3 minutes. Figure 9 shows the load response for this new tuning. As expected the level is arrested in about 3 minutes without any tendency to cycle.

Lessons Learned

A major lesson learned was that almost all of the difficult loops in the powerhouse had essentially the same generic behavior (integator plus deatime dynamics), hence the same type of tuning approach should work for each of them. The tuning that was required to make this type of dynamic response stable involved a very tight interaction between both the proportional and integral control constants. To achieve this tuning by "feel" is almost impossible. Using the appropriate Lambda tuning method is essential. One major concern was the realization that the fastest tuning was constrained by the process deadtime (Lambda had to be three deadtimes or slower). What would happen if the resulting speed of response was not fast enough for the process conditions which prevailed? The answer to this question is to implement and make effective use of feedforward control. If effective feedforward control could do "most of the work", the resulting feedback control could afford to be slower. This was especially true for the boiler drum levels themselves, which use feedforward from steam flow.

Summary of Feedback Controller Tuning

The key control loops which made a critical difference to the powerhouse operation all had integrating dynamics with deatime. A summary of their tuning is listed below.
Table V - Summary of Tuning

<table>
<thead>
<tr>
<th>Loop</th>
<th>Lambda (minutes)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Package Boiler Steam Drum</td>
<td>3.0</td>
</tr>
<tr>
<td>Waste Fuel Boiler Steam Drum</td>
<td>3.0</td>
</tr>
<tr>
<td>Recovery Boiler Steam Drum</td>
<td>3.0</td>
</tr>
<tr>
<td>600# StmHdr-Package Blr Swing</td>
<td>1.05</td>
</tr>
<tr>
<td>600# StmHdr-Waste Fuel Blr Swing</td>
<td>10.5</td>
</tr>
<tr>
<td>Waste Fuel Surge Bin</td>
<td>10.5</td>
</tr>
<tr>
<td>Waste Fuel Feed Bin</td>
<td>3.2</td>
</tr>
<tr>
<td>Boiler Water Conductivity</td>
<td>60</td>
</tr>
</tbody>
</table>

RESULTS

In all cases the variability was reduced substantially. Space prohibits a detailed review of each case, hence only the highlights will be presented.

600 psi Header Pressure Control

In the original design of the 600 pound header, pressure control was performed by a single PI plant master controller. During the project, this scheme was found to have significant dynamic interaction problems, resulting in cyclic behavior of the 600 psi steam header. Operation of the controls also tended to be complicated and confusing. In an effort to improve the performance of the header pressure control, it was decided to simplify the strategy by modifying the plant master controller to include individual PI boiler master controllers for the waste fuel and package boilers. The operator is free to select which boiler controller will swing automatically in order to control the 600 psi steam header pressure. The potential exists to add feedforward control from steam demand. This was not done but would be very useful, especially when the waste fuel boiler is used as the swing boiler, since the Lambda value is over 10 minutes.

Waste Fuel Delivery System

The long conveyor transport delays in the bark delivery system tended to de-stabilize both waste fuel bin level controllers, resulting in output limit cycles of 0-to-100% at periods of 15-to-25 minutes. Figure 10 (upper plot) shows the Surge Bin 21LIC 086 in which the variability is 76.9%.

Figure 10
WASTE FUEL SURGE BIN VARIABILITY

The PI controllers, as originally implemented, simply could not compensate for the large deadtimes following each output correction. Since chip fines are fed at a constant rate to the conveyor belt, while the reclaim controls the amount of bark and hog fuel delivered to the belt, steam speed swings caused the composition of, and the heating value of the fuel being delivered to the furnace to cycle. This cycle propagated to the boiler, negatively impacting drum level and header pressure control. The improvements made to the bark system control included the addition of a feedforward control strategy as shown in Figure 4, together with the tuning of the surge and feed bins. This resulted in a dramatic reduction in cyclic behavior as illustrated in Figure 10 (lower plot), which shows the Surge Bin variability reduced to 21.9%—less than one third of the variability before the project.

Boiler Water Conductivity

Figure 11 shows the boiler water conductivity variability before and after the project, over a period of about 48 hours. The upper plot shows the conductivity before, with a dominant 7 hour cycle and variability of about 13%. The lower plot shows the conductivity after the project. The variability in steady state is less than 1%. In the early part of the plot a 1% setpoint change was made, just to show that the loop was actually working.
FIGURE 11
BOILER CONDUCTIVITY VARIABILITY

Boiler Trips Due to Drum Level

The most important result of the project was that there have not been any boiler trips resulting from drum level upsets since the project has been finished. Of equal importance is the fact that the operators no longer have to face the added stress caused by the possibility that at any time the powerhouse could go down as a result of a boiler trip.

CONCLUSIONS

This project was initiated to prevent boiler trips by doing a thorough analysis of the powerhouse dynamics. This was done successfully. The methods used were general in nature. The dynamic responses of major process systems (boilers, headers) were systematically tested. At the same time a first principles dynamic model, based on thermodynamics and the differential equations of motion was built. The experimental and theoretical approaches were brought into agreement. This alone provided considerable confidence in the results. In addition, it was possible to work out the linear control theory based tuning rule for the chief cause for the poor behavior of the powerhouse - the deadtime and integrating type process dynamics. The Lambda tuning rule which applies to this case, together with robust select on of the Lambda values themselves, proved to be the chief elements for solving the boiler tripping problem.

NOMENCLATURE AND SYMBOLS

\(\text{alasing} \)  
\(\text{apparent short on of an analog signal when sampled too slowly} \) 
\(\text{anti aliasing} \) 
\(\text{BASIC} \) 
\(\text{DCS} \) 
\(G_C(s)\)  
\(G_P(s)\)  
\(\text{Kpph} \) 
\(K_C\)  
\(K_P\)  
\(\text{lbs} \) 
\(\text{Lambda tuning} \)  
\(\text{Lambda} \) 
\(\text{hr} \) 
\(\text{PB} \) 
\(\text{PI} \)  
\(\text{PID} \) 
\(\text{psi} \) 
\(\text{PRV} \)  
\(\text{shrink/swell effect} \) 
\(\text{RHP} \) 
\(T_d\)  
\(T_R\)  
\(\text{a special signal filtering technique to prevent aliasing} \) 
\(\text{a simple programming language} \) 
\(\text{distributed control system} \) 
\(\text{controller transfer function in the continuous (Laplace) domain} \) 
\(\text{process transfer function in the continuous (Laplace) domain} \) 
\(\text{thousands of pounds per hour} \) 
\(\text{controller gain (100%/PB%)} \) 
\(\text{process gain normally expressed as process measurement (% of span)/ controller output (%). For integrating process variables, this quantity is also per second pounds} \) 
\(\text{a method of tuning which requires the user to specify the desired closed loop time constant - Lambda} \) 
\(\text{the desired closed loop time constant - usually in seconds per hour} \) 
\(\text{Proportional Band (PB% 100%/controller gain)} \) 
\(\text{Proportional Integral controller} \) 
\(\text{Proportional-Integral Derivative controller} \) 
\(\text{pressure measurement, pounds per square inch (10 psi 689 kPa)} \) 
\(\text{pressure reducing valve} \) 
\(\text{the tendency of a boiler drum level to contract due to the addition of cold feedwater, and to expand when the drum pressure is reduced on a sudden increase of steam demand} \) 
\(\text{right half plane (of the s plane)} \) 
\(\text{the unstable region} \) 
\(\text{deadtime} \) 
\(\text{controller reset or integral time/repeat} \)
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2. PROCESS DYNAMICS
Process Modeling

It is useful to develop a mathematical model of the process which describes its behaviour with a certain degree of accuracy. Developing a process model will result in an improved understanding of the process, assist in optimizing the control strategy, controller tuning and operating conditions.

A prime concern in process control is the time dependent or dynamic characteristics of the process. The model should incorporate this time dependency. A dynamic model differs from a steady state model in that an accumulation term is required. Ideally, the model is simple and easy to obtain. It will also preferably be linear.
Methods of Process Modeling

Three basic methods used

1 *Mechanistic*
   - Derived from first principles, such as conservation laws for mass and energy.
   - Usually involves simplifying assumptions.
   - May be difficult and/or time consuming to develop

2 *Empirical*
   - Based upon input/output data
   - Limited range of validity.

3 *Semi-Empirical*
   - Combination of above two.
   - Model form generated from a mechanistic approach (this may yield some model parameters)
   - Use empirical data to fit unknown parameters
     Usually provides a larger range of validity than pure empirical model
Examples of Process Modeling Methods

1 Mechanistic
   - Modeling of stock chest mixing as a continuous stirred tank reactor

Fig. 2-1. Stock chest with agitation.

Assuming that the mixing in the stock chest is ideal and that \( F_{IN} - F_{OUT} \) \( F \) the instantaneous fiber balance around the stock chest is.

\[
\text{(Rate of Fiber Accumulation)} = \text{(Rate of Fiber In)} - \text{(Rate of Fiber Out)}
\]

This is equivalent to the differential equation.

\[
V \frac{dC_o}{dt} = F(C_i - C_o)
\]

Rearranging,

\[
\left( \frac{V}{F} \right) \frac{dC_o}{dt} + C_o - C_i
\]

where

\( V \) = Volume of stock in the chest
\( F \) = Flow through put
\( C_O \) = Consistency at chest discharge
\( C_I \) = Consistency at chest inlet

2 Empirical
   - Modeling of cell evaluation kinetics in digesters

3 Semi-Empirical
   - Modeling flow through towers in a bleach plant
   - Expect a combination of mixing and plug flow (channeling)
   - Don't know extent of either (don't know parameters)
Input/Output Nomenclature

Input/output nomenclature can be confusing. It depends on whether or not we are 'thinking' from the vantage point of the process or the vantage point of the control system.

Fig. 2-2 Signal flow information between the process and the control system
First Order System

When the input to a first order system is stepped, the output response would appear as follows.

![Graph of input and output over time for a first order system](image)

Fig 2-3. Bump test or step test for a first order system.

This is a very common response. It will usually be observed on flows, pressures and outlet concentrations of tanks.

It is referred to as a first order system because it can be described mathematically by a first order linear differential equation

\[
\tau \frac{dy}{dt}(t) + y(t) = K_p u(t)
\]

where
- \( u(t) \) is the process input
- \( y(t) \) is the output output
- \( \tau \) is the process time constant
- \( K_p \) is the process gain

The two parameters which must be determined to characterize a first order process model are:

1. Process time constant
2. Process gain.
Process Gain

Of these two parameters, the easiest to identify is the process gain. In response to a step input, the output eventually settles out to some 'steady state' value. The process gain is simply the ratio of the change in the output to the change in the input

\[ K_p = \frac{\Delta y}{\Delta u} \]

For example, if a valve is opened by 5% and the flow increases by 150 USGPM, then the gain is

\[ K_p = \frac{\Delta y}{\Delta u} = \frac{150 \text{ USGPM}}{5\% \text{ Output}} = 30 \text{ USGPM}/\% \text{ Output} \]

Notice that the gain has units. The units of the gain are expressed as measurement units over actuator units.

Note also that the gain can be positive OR negative. Consider a typical setup for consistency control.

Fig 2-4 Stock Consistency process

If the dilution valve is opened, the consistency would typically respond as follows

Fig 2-5 Typical consistency response to a step increase of the dilution valve

In this case, the process gain is negative.
Time Constant

The time constant of a first order process is its main dynamic characteristic and provides some indication of how long it takes for the system to move from one steady state to another steady state. This dynamic behaviour usually arises as a result of the inherent inertia of the system caused by its mass, for example, in flow. There are other causes of inertia, such as tank volume, which prevents the concentration at the tank outlet from changing instantaneously.

The time constant characterizes the 'non-flat' or transient portion of the output response. The output response of a first order system to a unit step change can be described by

\[ y(t) = K_p \left(1 - e^{-t/\tau}\right) \]

As can be seen, the term \( e^{-t/\tau} \) represents the time dependent behaviour.

The value of \( e^{-t/\tau} \) tends towards zero as time tends toward infinity. The smaller the value of the time constant, the faster this term tends to zero, and the quicker the output will reach its new steady state.

It is common to explain the effect of the time constant in the following ways:

1) The initial slope of the process response intersects the final value at \( t = \tau \)

Fig 2-6 Output response of a first order system
2) The output change, expressed as a per cent of its total change $\Delta y$, versus time, is given in the following table.

<table>
<thead>
<tr>
<th>Time (after step)</th>
<th>Output change as a % of $\Delta y$</th>
</tr>
</thead>
<tbody>
<tr>
<td>$1\tau$</td>
<td>63.2</td>
</tr>
<tr>
<td>$2\tau$</td>
<td>86.5</td>
</tr>
<tr>
<td>$3\tau$</td>
<td>95.0</td>
</tr>
<tr>
<td>$4\tau$</td>
<td>98.2</td>
</tr>
<tr>
<td>$5\tau$</td>
<td>99.3</td>
</tr>
<tr>
<td>$6\tau$</td>
<td>99.8</td>
</tr>
<tr>
<td>$7\tau$</td>
<td>99.9</td>
</tr>
<tr>
<td>$8\tau$</td>
<td>$\approx$100</td>
</tr>
</tbody>
</table>

Table 2-1 Output change as a % of total change after a multiple of the time constant

The output has effectively stopped changing after about 3 to 4 time constants. A common method of determining the time constant empirically is to divide the total 'response time' by 4. Other methods try to determine the 63% rise time to give the time constant.
First Order Impulse Response

Besides step tests, another process response is to an impulse in the process input. An ideal impulse is of infinite magnitude with an infinitely small duration. In practice, this may be reasonably approximated by a very short duration pulse of some finite magnitude.

The impulse response of a first order system is given as

\[ y(t) = \frac{K_p}{\tau} e^{-t/\tau} \]

It is important to note that for any system, the impulse response is simply the derivative of the step response. The impulse response of a first order system is shown in the following figure.

---

**Fig 2-7** Impulse Response of a first order system
First Order with Deadtime

The response of this type of system is very similar to a first order process except that a change in the output does not begin to occur until some time after a change has happened to the input.

![Diagram showing the response of a first order with deadtime system.](image)

**Fig 2-8** Output response of a first order with deadtime system

Deadtime is usually due to some type of transport phenomena. Examples of this are:

1. Plug flow through pipes
2. Mechanical transport of paper through the dryer sections

A first order plus deadtime process can be described by the following differential equation:

\[
\tau \frac{dy(t)}{dt} + y(t) = K_r u(t - T_d)
\]

where \(T_d\) is the deadtime.

We will see later that deadtime can be very destabilizing to a controller.
Second Order Systems

A second order system is one for which the differential equation governing its dynamics has a second order derivative:

$$\tau^2 \frac{d^2 y(t)}{dt^2} + 2\tau\zeta \frac{dy(t)}{dt} + y(t) = K_P u(t)$$

An additional parameter has been introduced: $\zeta$. This is called the damping coefficient, and will determine the nature of the response.

There are three forms of the solution to the differential equation depending upon the value of $\zeta$. These are:

1. **Underdamped** $\zeta < 1$.
   
   In this case, the step response will overshoot the final value and eventually approach the final steady state value in a decaying oscillatory fashion.

2. **Critically Damped**: $\zeta = 1$.

3. **Overdamped**: $\zeta > 1$.
   
   In the case of overdamped and critically damped, the step response will be somewhat similar to a first order response but the first part of the response will be somewhat slower than that of a first order model.

The process gain, $K_P$, is determined in the same way as with first order systems. However, the other parameters may be harder to determine.

It is very hard to distinguish between first order or first order plus deadtime versus critically damped or overdamped second order systems. These difficulties become greater in the presence of noise in measurements, disturbances or nonlinearities. Very often higher order systems are modeled as first order or first order with deadtime.
Second Order Overdamped and Critically Damped

Critically damped and overdamped systems may be considered to be two first order systems in series. The only difference between the two is that the two time constants of a critically damped system are equal.

**Example** Consider two stock chests in series.

![Diagram of two stock chests in series](image)

**Fig 2-9** Stock Chests in series

The responses from the inlet consistency to the first stock chest to the outlet consistency of the first stock chest and the outlet consistency second stock chest are shown in the following figure.

![Diagram of stock consistency responses](image)

**Fig 2-10** Stock consistency responses for two stock chests in series

Notice that the outlet consistency of the first tank exhibits first order characteristics. The outlet consistency of the second tank is initially more 'sluggish'. Thus, \( C_2 \) exhibits a second order response to \( C_0 \) as an input.

Neither a critically damped or overdamped system will oscillate unless driven by an oscillatory input.
Second Order Underdamped

Example. Consider a mass spring system immersed in a viscous liquid.

![Mass spring damper system diagram](image)

Fig. 2-11. Mass spring damper system.

The input to the system is the vertical displacement of the upper end of the spring and the output is the vertical displacement of the mass.

If the upper end of the spring is stepped upwards, the spring will pull the mass upwards. After a while, the spring will no longer be providing any upwards force, and gravity will start to pull the mass down again and the mass will oscillate. Due to the viscosity of the liquid, some damping of the oscillations will occur, and they will eventually die away. See Fig. 2-12.

![Step response graph](image)

Fig. 2-12. Mass position response due to a step change in spring upper end position.
Describing Second Order Underdamped Systems

For underdamped second order systems, the step response appears as follows.

Fig 2-13 Step Response of a second order underdamped system.
There are several common terms used in describing this type of response

**Overshoot:** Factor by which the response overshoots the final value

\[ O = \frac{A}{B} - \exp \left( -\frac{-\pi \zeta}{\sqrt{1 - \zeta^2}} \right) \]

**Decay Ratio** \( D \): Ratio by which successive peaks decay

\[ D = \frac{C}{A} - \exp \left( -\frac{-2\pi \zeta}{\sqrt{1 - \zeta^2}} \right) = (\text{Overshoot})^2 \]

**Rise Time** \( t_R \): Time required to first reach the final value

**Period of Oscillation** \( T \)

\[ T = \frac{2\pi \tau}{\sqrt{1 - \zeta^2}} \]

**Frequency of Oscillation** \( \omega \):

\[ \omega = \sqrt{1 - \zeta^2} \ \text{(radians/unit time)} \]

\[ f = \frac{1}{T} = \frac{\sqrt{1 - \zeta^2}}{2\pi \tau} \ \text{(cycles/unit time)} \]

**Undamped Natural Frequency** Frequency at which the system would oscillate if damping were removed (\( \zeta = 0 \))

\[ f_n = \frac{1}{T_n} = \frac{1}{2\pi \tau} \]

\[ \omega_n = \frac{1}{\tau} \]
Step responses of a second order underdamped system for various values of the damping coefficient are shown in the following figure.

Fig 2-14. Second order underdamped step responses for various values of the damping coefficient.
Note that the overshoot $O$ depends only on the damping coefficient $\zeta$ and not on the parameter $\tau$. The percent overshoot as a function of $\zeta$ is shown below.

![Graph showing percent overshoot vs damping coefficient $\zeta$.]

**Fig 2-15 Plot of Percent Overshoot vs Damping Coefficient.**

Solving for the damping coefficient $\zeta$ in terms of the overshoot $O$ yields

$$\zeta = \frac{a^2}{\sqrt{a^2 + \pi^2}} - \frac{1}{\sqrt{1 + \left(\frac{\pi}{a}\right)^2}}$$

where $a$ is the natural log of the overshoot, i.e. $a = \ln \left( \frac{A}{B} \right)$. 
Integrating Systems

An integrating system is one that integrates its input.

Example: Tank level

\[
\frac{dV}{dt} = F_{IN} - F_{OUT}
\]

or

\[
\frac{dL}{dt} = \frac{1}{A} (F_{IN} - F_{OUT})
\]

Separating variables and integrating both sides yields an equation for the level

\[
L = \frac{1}{A} \int (F_i - F_o) dt + L_o
\]

Suppose the tank is initially empty \((L_o = 0)\) and there is no flow out of the tank \((F_o = 0)\).
The level will be

\[
L = \frac{1}{A} \int F_i dt
\]

If \(F_i\) is positive (flow going into tank), then the tank will fill up

An integrating system can be described by

\[
y(t) = K_p \int u(t) dt
\]

To keep a tank level constant the difference between all inlet and outlet flows must be zero. To change the level, the flow difference must be non-zero for some period of time
For a level process, the bump test will appear as follows:

\[ u(t) = \text{FLOW IN} \]

\[ \Delta u_1 \]

\[ \Delta u_2 \]

\[ \text{FLOW OUT} \]

\[ \Delta y_1 \]

\[ \Delta y_2 \]

\[ \Delta t_1 \]

\[ \Delta t_2 \]

\[ \text{LEVEL L} \]

\[ \text{slope } s_0 \]

\[ \text{slope } s_1 \]

\[ \text{slope } s_2 \]

Fig 2-17  Bump test for a level process.

The process gain for a level process is calculated from the bump test as follows

\[ K_p = \frac{(\Delta y_1 / \Delta u_1)}{\Delta t_1} = \frac{(\Delta y_2 / \Delta u_2)}{\Delta t_2} \]

\[ \frac{s_1 - s_0}{\Delta u_1} = \frac{s_2 - s_1}{\Delta u_2} \]

The level process will be covered in more detail later when level controller tuning is discussed.
PROCESS DYNAMICS EXAMPLES FOUND IN PULP AND PAPER

The focus of this course will be to model a process response by either a first order with deadtime model or an integrator with deadtime model wherever possible. However, certain processes found in pulp and paper often have interesting dynamics which cannot adequately be modeled by first order with deadtime or integrator with deadtime models. Some of these will now be examined.

Paper Machine Dryer Steam Pressure

A sketch of a typical Paper Machine dryer steam pressure process configuration is shown in the following sketch.

![Diagram of Paper Machine Dryer Steam Pressure System]

Fig 2-18 Typical Paper Machine Dryer Section

The Dryer pressure response to a bump test may look like either of those shown on the following two pages.
Fig. 2-19. Dryer Steam Pressure 1 bump test overview.

Fig. 2-20. Dryer Steam Pressure 1 bump 3 close-up
Fig 2-21  Dryer Steam Pressure 2 bump test overview

Fig 2-22  Dryer Steam Pressure 2 bump 1 close-up
Stock Chest Level With Two Stock Chests in Series

Consider the stock chest level control scheme shown below

![Diagram of stock chest level control scheme](image)

Fig 2-23 Stock blending system with gravity flow from Blend Chest to Machine Chest.

A sample bump test is shown on the following page.
Fig 2-24. Machine Chest Level bump test overview

Fig 2-25. Machine Chest Level bump 1 close-up
Bleach Plant Steam Mixer Temperature

Consider the Bleach Plant Steam Mixer Temperature control loop shown below

STOCK FROM CD WASHER

EO STEAM MIXER

FT 44

65# STEAM

TIC 28

LT 25

LIC 25

EO MC PUMP

TT 28

PI 29

TO D TOWER

Fig 2-26 Bleach Plant Steam Mixer process area

A sample bump test is shown on the following page
Fig 2-27. Steam Mixer Temperature bump test overview

Fig 2-28 Steam Mixer Temperature bump 2 close-up
Boiler Steam Drum Level

Consider the Boiler Steam Drum Level control loop shown below.

![Diagram of Boiler Steam Drum Level control loop](image)

Fig. 2-29 Three-element Boiler Steam Drum Level

The 'liquid' in the bottom of the drum is actually composed of both water and steam bubbles existing in equilibrium at the saturation temperature.

If the feedwater flow is increased, the liquid temperature will initially drop and many of the bubbles will collapse, with the result that the level will initially drop (shrink-swell effect).

Eventually, the temperature will increase to the saturation point and the level will rise as the bubbles once again begin to form.

These systems are characterized by a response that initially goes the 'wrong' way. Such dynamic systems are called non-minimum phase systems.

A sample bump test is shown on the following page.
Fig. 2-30. Steam Drum Level bump test overview.

Fig. 2-31 Steam Drum Level bump 3 close-up
Waste Fuel Boiler Bark Surge Bin Level

Consider the Waste Fuel Boiler Bark Surge Bin Level control loop shown below.

Waste Fuel Boiler
Bark Fuel Transport System

Fig 2-32  Waste Fuel Boiler Bark Fuel Transport System

A sample bump test is shown on the following page
Fig 2-33 Bark Surge Bin Level bump test overview.

Fig 2-34 Bark Surge Bin Level bump 1 close-up
Consistency Loop with Recycle Dynamics
The presence of a recycle path can dramatically alter a loop's dynamic response. Consider the following process and discuss the expected dynamics of the consistency loop.

White Water Header

Stock Chest

Refiner

To Blend Chest

RSP from Level Controller
Sample bump test results are shown below.

1st Order
\[ t = 296 \quad U = 55.6 \quad DU = -4.09 \quad Y = 33.4 \quad DY = 4.16 \]
\[ Kp = 1.02 \quad Td = 7.00 \quad Tau = 53.8 \]
APPENDIX 2A
SAMPLE BUMP TESTS
3. NONLINEARITY IN PROCESS DYNAMICS
Non-Linearities

Up to now we have been dealing with systems that are completely linear. Although we have made some simplifying assumptions in certain instances, the system was still considered to be linear.

The systems that we will be dealing with in practice will generally exhibit some non-linear characteristics. It is important to recognize the nature of these non-linearities and what, if anything, may be done to deal with them so that their impact upon system performance is minimized.
Nonlinearities in the Final Control Elements

Control Valve Backlash & Stiction

These two effects are a result of irregularities in actuator motion.

Backlash or hysteresis is a directional effect. It manifests itself as 'lost motion' when the direction of movement of the actuator changes and is due to such factors as gearing and linkage, as well as other sources of 'slop'. Figure 3-6 shows a control valve assembly which illustrates potential sources of backlash.

Example

Fig 3-1
In this example, the actuator did not move immediately after the direction of motion was changed. The second upwards step only gave a partial response. Both of these irregularities were due to backlash. All of the backlash after about 1 1/2 steps and the following upwards steps provide full response. The similar response on the downward steps is also due to backlash.
Fig. 3-2

Stiction is a result of static friction. When the actuator has been at rest, it will require a certain force to start the actuator moving. This force is required to overcome the stiction.

Example

Fig 3-3

With the valve initially at rest, the initial step was insufficient to overcome the stiction. The second step provided enough force to overcome the stiction and move the actuator to its desired position. The same thing happens with the next two steps. This time a
third step occurs before the actuator has come to rest and a 'single' response results. The two down pulses also show stiction.

Note that stiction is not dependant upon a change in direction of motion.

The EnTech specification for control valve performance allows a combined backlash and stiction of 1% as indicated in Figure 3-5 below.
Figure 3-6 Spring Return Diaphragm with Motion Balance Positioner
Comments on Backlash and Stiction

- These non-linearities may be non-linear with actuator position (possibly more noticeable the longer spent in service)
- Stiction and backlash may occur together
- These effects may be 'stochastic' in nature (unable to determine when and by how much they exist)
- They may be asymmetric (easy to start opening a quick opening valve, but difficult to finish closing it)
Compensating for Backlash and Stiction

It is theoretically possible to compensate for backlash and stiction by adding an extra 'kick' to the control action. Caution should be exercised in the application due to the non-repeatable characteristics. Compensation for these effects may aggravate the situation.

Example:
- Providing compensation for backlash
- Tests show backlash = 1% of output span

Situation:
- Direction of changes
- Magnitude of change = 0.1%
  \[ \Delta U = 1.0\% + 0.1\% = 1.1\% \]

If the magnitude of the actuator's backlash had decreased to 0.5%, then an effective change of 0.6% in process input occurs. This represents 600% of the desired change!

* Often better to ignore compensation or to only provide partial compensation.

* Best solution is to remove the cause at source.
Control Valve Flow Characteristics

Example: Valves

Valve Curves

A QUICK OPENING
B LINEAR
C EQUAL PERCENTAGE

Fig 3-7
An important point to remember is that these curves are determined with a constant pressure drop across the valve. These may be considered as normalized $C_v$ curves. With a constant $\Delta P$, the slopes give $K_P$.

Fig 3-8
From this it would appear as though a linear valve is best. But $\Delta P$ usually changes as the valve opens or closes. We often find

$\Delta P$

% TRAVEL

Fig 3-9
Thus, we need to determine installed characteristics. These will vary according to installation.
Control Valve Speed Of Response

Consistent speed of response of a control valve is crucial for the operation of a control loop. For small step changes in input signal, the plug or trim should track the input signal with at least the speed of response listed in Table 1 and illustrated in Figure 3-11, over a range of step changes ranging from 10% of travel, down to the combined backlash/stiction limit achieved above, plus 1%. (If combined backlash/stiction limit is 0.5%, then down to 1.5%)

![Response Window Diagram](image)

**Figure 2**

**Figure 3-10 EnTech Specification for Speed of Response**

\[
\begin{align*}
T_d &= \text{dead time before any response is visible (seconds)} \\
T_{63} &= \text{time to reach 63\% of a step change in input signal (seconds)} \\
T_{98} &= \text{time to reach 98\% of a step change in input signal (seconds)} \\
BW &= \text{bandwidth (Hz), the frequency at which the plug or trim position amplitude is attenuated by -6 dB} \\
&\text{Bandwidth has been included in the specification for completeness, even though it is difficult to test for in most cases}
\end{align*}
\]
### Speed of Response

<table>
<thead>
<tr>
<th>Valve Size (inches)/cm</th>
<th>Td (sec)</th>
<th>T63 (sec)</th>
<th>T98 (sec)</th>
<th>BW (Hz)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0-2</td>
<td>0.1</td>
<td>0.3</td>
<td>0.7</td>
<td>1.6</td>
</tr>
<tr>
<td>&gt;2-6</td>
<td>&gt;5-15</td>
<td>0.2</td>
<td></td>
<td></td>
</tr>
<tr>
<td>&gt;6-12</td>
<td>&gt;15-30</td>
<td>0.4</td>
<td>0.6</td>
<td>1.4</td>
</tr>
<tr>
<td>&gt;12-20</td>
<td>&gt;30-50</td>
<td>0.6</td>
<td>1.2</td>
<td>2.8</td>
</tr>
<tr>
<td>&gt;20+</td>
<td>&gt;50+</td>
<td>0.8</td>
<td>1.8</td>
<td>4.2</td>
</tr>
</tbody>
</table>

The above specification allows for large valves to be progressively slower than small ones. The Td, T63, T98 and BW specifications are a compatible set of numbers for a second order critically damped system with deadtime, representing positioner hesitation equal to one time constant. All the responses have the shape illustrated in Figure 3-10. Responses for larger valves are identical, except that time is rescaled appropriately.
Electric Actuators

High Torque Electric Actuators
High torque electric actuators have become available for both rotary valves and damper drives as a replacement for pneumatic actuators. These actuators are driven by fixed speed electric motors under the control of an internal positioner. Due to the fact that the motor is of fixed speed, the apparent 'time constant' is nonlinear with actuator move size (in most cases this is unlikely to be serious since the speed is relatively fast). Also, there is a small inherent backlash. Naturally, linkage problems between the actuator and valve or damper can still cause serious problems. However in general, these actuators provide a far superior performance over pneumatic actuators and represent an attractive alternative for some rotary valves and most damper drives.

Low Torque Electric Valves - Basis Weight Stock Flow
Many paper machines use electric valves for stock flow control. These actuators consist of small low torque motors with large gear-downs. Typically, these valves have stop-to-stop times of about 100 to 300 seconds (slew rate). While these valves move slowly and produce rather slow control loop dynamics, this may be adequate for stock flow regulation. However there are a number of potential problems with this type of valve.

The minimum pulse duration for such valves is typically less than 100 mS. This is the minimum time that the motor must be energized in order to actually move. This together with the slew rate will determine the actual control resolution. Often this is well below 1:1000 and may result in a nonlinear limit cycle.

Another potential problem is backlash, which typically occurs in the spline coupling between the gear-down output shaft and the valve plug. For large valves with high torque requirements this appears to be a bigger problem. Backlash of as much as four seconds of actuation time has been reported.

An associated problem, although not the fault of the valve, is that in many basis weight control systems, the thick stock flow control loop is being executed at only 5 seconds per sample. This leads to problems with aliasing and inaccurate stock contro.
ON COMPUTER CONTROL THE OPEN/CLOSE CONTACTS ARE ENERGIZED BY TIMED OUTPUT OF OPEN/CLOSE DO'S

- THESE VALVES MOVE AT FIXED SPEED WHICH IS TYPICALLY BETWEEN 180 SECONDS TO 300 SECONDS FOR FULL TRAVEL (0 5%/SEC TO 0 3%/SEC)

- START/STOP DYNAMICS
  TO ENERGIZE RELAY AND CLOSE CONTACT ≈ 50ms
  ACCELERATE MOTOR TO FULL SPEED ≈ 100ms

Figure 3-11 Electric Actuators
A Special Case

Many presently available actuators are velocity mode devices. They are usually driven by stepper motors and have the advantage that they are fail-safe as regards their input from the controller. Output to such devices is usually in the form of a pulse train or time duration pulse, which will open or close the valve.

Example

![Diagram of motor output and valve position](image)

Fig 3-12

A special form of stiction may arise with this type of actuator. When the duration of the pulse is extremely short, relays may not have time to (de)energize and/or the motor may not have time to start in motion, with the result that the pulse has no effect on the process output.

In this instance, a deadband may be used to prevent unnecessary wear on the elements. This deadband should be applied to the output and not the error or a limit cycle may result. It should also be linear.

Also, compensation for stiction could be applied. The compensation should be made equal to the minimum required actuation time.
VARIABLE SPEED DRIVES FOR FANS AND PUMPS

- DC Drives - proven technology,
  - essentially linear dynamics,
  - time constant of 0.1 to 0.5 seconds.

- Variable Frequency Drives,
  - newer technology,
  - energy efficient,
  - fast (velocity limited) dynamics,
    (small signal slew rates in 0.25 seconds, full range),
  - very cost effective - can use existing motor (providing
    frame size OK for extra heat dissipation).

In fact in many cases a VFD can be installed without replacing any field gear at all. The
original control valve or damper can be kept in place as a mid-ranging device if needed.
From an energy efficiency point of view, it should be fully opened however. This
knowledge helps in 'selling VFD's' to production management who might be reticent to
try 'something new'.

- Controller output Bit Resolution.
Since the internal bit resolution of most digital drive systems is at least 12 bits, it is
critically important that the controller output have the same level of output resolution.
Significant limit cycling problems have been reported when the controller output was
only 10 bits while the drive had 12 bits.
THE DRIVE EIGE RELANCE ETC
TYP CALLY TUNE AS FOLLOWS

CURRENT LOOP \( \lambda_{\text{current}} = \frac{1}{30} \) 0.03SEC
SPEED LOOP \( \lambda_{\text{speed}} = \frac{1}{30} 0.03\text{SEC} \)

NORMAL OBJECTIVE IS TO GET A SPEED STEP RESPONSE LIKE THIS

NOTE THAT TH S UNDERDAMPED (\( \zeta = 0.8 \)) AND CAN BE THE CAUSE OF OSCILATORY BEHAVIOR

Figure 3-13 Variable Speed Drives
Final Control Element - Recommendations

Insist on near linear operation of final control elements from a backlash/stiction point of view. If this cannot be achieved, then insist on being able to meet performance as defined by the EnTech Valve Specification. For damper drives, the specified dynamic performance of a large valve, such as a 20 inch valve would be appropriate. If this still cannot be done, then consider the following action.

1) Replace valves with VS drives. Replace dampers with VS fans. (backlash & stiction no longer possible and open loop dynamics down to less than 1 second.)

2) Do not insist on tight shut-off for control valves, (removes largest source of static friction) and install separate shut-off valves

3) Replace rotary valves with sliding stem valves, if the application will permit. The combined backlash/stiction for sliding stem valves can typically be 20% of that for quarter turn rotary valves.
Process Non-Linearities

Besides non-linearities in the actuators (which are usually considered as part of the process), non-linearities in the 'actual' process are very common.

Considering the process to be linear is usually valid over a narrow region near the operating point. Many processes will exhibit non-linearities over the available operating range. These usually show up as:

- non-linear gain
- non-linear deadtime
- non-linear time constant(s)

Example Variable Time Constant

\[
\begin{align*}
C, F & \quad \rightarrow \\
& \quad h \\
& \quad \downarrow \\
& \quad C_0, F_c \\
\end{align*}
\]

\[V = h \cdot A_x \quad (A_x \text{ cross sectional Area})\]

\[\tau = \frac{V}{F} \quad \text{(assume } F_i, F_s)\]

The time constant will vary with flow as well as level.

\[\tau \quad \text{vs. } F\]

\[\tau \quad \text{vs. } h\]

Fig 3-14

Fig 3-15
**Example:** Variable Gain with pH

![Titrination Curve](image)

**Fig 3-16**

The process gain is proportional to the slope of the titration curve.

![Gain vs pH](image)

**Fig 3-17**

The process gain is also related to the flowrate of the controlled flow.

![Gain vs Flow](image)

**Fig 3-18**
Example  Variable 'Rate of Integration'

\[ L = \int_0^t \frac{1}{A} (F_i - F_o) dt + L_o \]

Note  \( \frac{1}{A} \) cannot be taken outside since it may change with time

'Rate of Integration' = \( \frac{1}{A} \)

Copyright EnTech 3-21
Example: Variable Time Delay

Pure transport delay of the web down the machine varies with speed.

Fig. 3.21
Example: Variable Gain, Time Delay, Time Constant

Consistency Control

Deadtime decreases with Stock flow

Process Gain decreases with Stock Flow

Mixing time decreases slightly with Stock Flow

Figure 3-22 Non-Lineanties in Consistency loop
**Example:** Variable Gain

Drying Rate on Paper Machine

Temperature differential is the driving force for drying

![Graphs showing the relationship between parameters](image)

**Fig 3-23 Non-Linear dynamics of Moisture Loop**

Many other factors affect $K_p$

- machine speed
- 'rimming' characteristics
- press moisture
4. LAPLACE TRANSFORMS AND TRANSFER FUNCTIONS
Dynamics in Other Domains

The time domain has the advantage that it represents quantities in terms of variables with which we are most familiar (most notably time). The main disadvantage when working in the time domain is the cumbersome nature of dealing with differential equations.

This difficulty can be overcome by transforming into the LAPLACE DOMAIN. In the Laplace Domain, differential equations are solved algebraically, which makes them much easier to solve. The main disadvantage of working in the Laplace Domain is that the dynamic behavior is not as readily visualized.

Laplace Transform

In the time domain, we represent variables as functions of time, \( t \). In the Laplace Domain, we represent variables as functions of the Laplace variable, \( s \).

Define the Laplace Transform

\[
\mathcal{L}(y(t)) = Y(s) = \int_{0}^{\infty} y(t)e^{-st}dt
\]

The Laplace Transform transforms a function of \( t \), \( y(t) \) into a function of the complex variable \( s \), \( Y(s) \).

Tables are available to perform the transforms. Virtually all transforms of interest in process control are available.

Define the inverse Laplace Transform

\[
\mathcal{L}^{-1}(Y(s)) = y(t)
\]
# Common Laplace Transform Pairs

<table>
<thead>
<tr>
<th>Time Function</th>
<th>Laplace Transform</th>
</tr>
</thead>
<tbody>
<tr>
<td>$f(t)$</td>
<td>$F(s) = \mathcal{L}(f(t))$</td>
</tr>
<tr>
<td>$\delta(t)$ (unit impulse)</td>
<td>1</td>
</tr>
<tr>
<td>$u(t)$ (unit step)</td>
<td>$\frac{1}{s}$</td>
</tr>
<tr>
<td>$e^{at}$</td>
<td>$\frac{1}{s-a}$</td>
</tr>
<tr>
<td>$t$ (unit ramp)</td>
<td>$\frac{1}{s^2}$</td>
</tr>
<tr>
<td>$te^{at}$</td>
<td>$\frac{1}{(s-a)^2}$</td>
</tr>
<tr>
<td>$e^{at} t^n / n!$</td>
<td>$\frac{1}{(s-a)^{n+1}}$</td>
</tr>
<tr>
<td>$\sin(\omega t)$</td>
<td>$\frac{\omega}{s^2 + \omega^2}$</td>
</tr>
<tr>
<td>$\cos(\omega t)$</td>
<td>$\frac{s}{s^2 + \omega^2}$</td>
</tr>
<tr>
<td>$e^{at} \sin(\omega t)$</td>
<td>$\frac{\omega}{(s-a)^2 + \omega^2}$</td>
</tr>
<tr>
<td>$e^{at} \cos(\omega t)$</td>
<td>$\frac{s-a}{(s-a)^2 + \omega^2}$</td>
</tr>
</tbody>
</table>

Table 4-1  Common Laplace Transforms
Properties of the Laplace Transform

1. Multiplication by a constant does not affect the transform.

\[ \mathcal{L}(cy(t)) = c \mathcal{L}(y(t)) \]
\[ = cY(s) \]

2. The transform of a sum of time functions is the sum of the individual transforms of each time function

\[ \mathcal{L}(y_1(t) + y_2(t)) = \mathcal{L}(y_1(t)) + \mathcal{L}(y_2(t)) \]
\[ = Y_1(s) + Y_2(s) \]

3. The transform of the derivative of a time function is the transform of the time function multiplied by \( s \).

\[ \mathcal{L}\left(\frac{dy(t)}{dt}\right) = s \cdot Y(s) \]
\[ \mathcal{L}\left(\frac{d^2y(t)}{dt^2}\right) = s^2 \cdot Y(s) \]

etc.

4. The transform of the integral of a time function is the transform of the time function divided by \( s \).

\[ \mathcal{L}\left(\int_0^1 y(t) \, dt\right) = \frac{Y(s)}{s} \]

The above properties of the Laplace transform allow us to transform a differential equation in the time domain into an algebraic equation in the Laplace domain. This tremendously simplifies the analysis of the dynamic system.
Dynamic Representation in the Laplace Domain

Example  First Order System

The differential equation is

$$\tau \frac{dy(t)}{dt} + y(t) = K u(t)$$

Take the Laplace Transform of both sides

$$\mathcal{L}\left( \tau \frac{dy(t)}{dt} + y(t) \right) = \mathcal{L}(K u(t))$$

$$\mathcal{L}\left( \tau \frac{dy(t)}{dt} \right) + \mathcal{L}(y(t)) = K \mathcal{L}(u(t))$$

$$\tau \mathcal{L}\left( \frac{dy(t)}{dt} \right) + Y(s) = KU(s)$$

$$\tau sY(s) + Y(s) = KU(s)$$

$$(\tau s + 1)Y(s) = KU(s)$$

$$Y(s) = \frac{K}{\tau s + 1} U(s)$$

$$Y(s) = G(s)U(s)$$

where

$$G(s) = \frac{K}{\tau s + 1}$$

$G(s)$ is called the transfer function of the dynamic system.
Transfer Functions

Define the transfer function as that function which relates the output transform to the input transform. Denote the transfer function by $G(s)$.

The transfer function provides information about the dynamic nature of the system. It describes how the information in the input signal is transferred to the output signal. A transfer function is sufficient to describe the input/output relationship of any given system.

For a first order dynamic system, the transfer function is

$$G(s) = \frac{K}{\tau s + 1}$$

Note that the transfer function contains the same two parameters, $K$ and $\tau$, which characterize the dynamic system model. Also note that $K$ can be characterized by

$$K = G(s)_{s=0}$$

The quantity $G(s)_{s=0}$ is called the steady-state gain of the transfer function $G(s)$. 
Block Diagrams

It is convenient to represent the transfer function pictorially. They are typically represented with *block diagrams*.

![Block Diagram 1](image1)

**OR**

![Block Diagram 2](image2)

Fig 4-1 Block diagrams representing dynamics using transfer functions.
Block Diagram Reduction

**Rule 1:** Blocks in series

\[ X \rightarrow G_1(s) \rightarrow G_2(s) \rightarrow Y \]

IS EQUIVALENT TO

\[ X \rightarrow G_1(s) \cdot G_2(s) \rightarrow Y \]

Fig. 4-2. Block diagram reduction of blocks in series.

**Rule 2** Blocks in parallel

\[ X \rightarrow G_1(s) \leftarrow \bigoplus \rightarrow Y \]

\[ X \rightarrow G_2(s) \leftarrow \bigoplus \rightarrow Y \]

IS EQUIVALENT TO

\[ X \rightarrow G_1(s) + G_2(s) \rightarrow Y \]

Fig. 4-3 Block diagram reduction of blocks in parallel.
Transfer Functions For Common Process Models

Using the same procedure as that used for the first order process model we obtain transfer functions to represent the dynamics of other commonly occurring process models. This is shown in Table 4-2.

<table>
<thead>
<tr>
<th>Process Model</th>
<th>Transfer Function</th>
</tr>
</thead>
<tbody>
<tr>
<td>1. Pure gain</td>
<td>$G_p(s) = K_p$</td>
</tr>
<tr>
<td>2. Pure delay</td>
<td>$G_p(s) = e^{\tau_d s}$</td>
</tr>
<tr>
<td>3. First order</td>
<td>$G_p(s) = \frac{K_p}{\tau s + 1}$</td>
</tr>
<tr>
<td>4. First order with deadtime</td>
<td>$G_p(s) = \frac{K_pe^{\tau_d s}}{\tau s + 1}$</td>
</tr>
<tr>
<td>5. Second order</td>
<td>$G_p(s) = \frac{K_p}{\tau_s^2 s^2 + 2\tau \zeta s + 1}$</td>
</tr>
<tr>
<td>6. Integrating</td>
<td>$G_p(s) = \frac{K_p}{s}$</td>
</tr>
</tbody>
</table>

Table 4-2 Transfer Functions For Common Process Models
Non Process Transfer Functions

Transfer functions have been obtained for process models for common processes such as stock flows and pressures, stock consistencies and stock chest levels. Transfer functions will also be useful to represent other parts of the control loop which have dynamics, not just the process. Transfer functions for other elements in the control loop will be further discussed later in these notes.

![Block Diagram of a control loop.](image)

Fig. 4-4  Block Diagram of a control loop.

1. **Control loop filters.** Signal filters used to reduce noise also have dynamic characteristics. A simple first order filter, can be modeled by a first order transfer function:

   \[ H(s) = \frac{1}{\tau_f s + 1} \]

   where \( \tau_f \) is the filter time constant. Note that the gain is 1.

2. **Feedback Controllers:** Feedback controllers also have dynamic behavior. For example, a simple PI (Proportional-Integral) controller will be described by the following transfer function.

   \[ G_C(s) = K_P + K_I s - \frac{K_P K_I}{s} \]

   where \( K_P \) (proportional gain) and \( K_I \) (integral gain) are controller parameters.

3. **Other control system blocks.** A commonly occurring dynamic element often used in conjunction with a feedforward control scheme is a lead-lag block which has the transfer function:

   \[ G(s) = \frac{\beta s + 1}{\tau s + 1} \]

   where \( \beta \) (lead time constant) and \( \tau \) (lag time constant) are parameters.
Complex Plane Representations

The Laplace variable, \( s \), is a complex variable and can be written as

\[
s = A + Bj
\]

where \( j = \sqrt{-1} \).

Any given value of \( s \) can be plotted in the complex plane.

![Complex Plane Diagram]

**Fig 4-5** Representing a complex number as a point in the complex plane

The polar representation of \( s \) is given by \( s = Me^{j\theta} \). \( M \) is called the absolute value of \( s \) and \( \theta \) is called the angle of \( s \). The polar representation is useful for describing the frequency response of a dynamic system.
Poles and Zeros of Transfer Functions

Most of the dynamic information contained in the transfer function is represented by the values of $s$ that represent the poles and zeros.

A pole is defined as a value of $s$ which makes the magnitude of the transfer function at that point infinite. Usually, the transfer function is a rational function having a numerator and a denominator, and a pole is simply a value of $s$ which makes the denominator of the transfer function equal to zero.

A zero is defined as a value of $s$ which makes the magnitude of the transfer function at that point equal zero. For a rational transfer function a zero is simply a value of $s$ which makes the numerator of the transfer function equal to zero.

A pole-zero plot for a transfer function consists of the complex plane with the transfer function poles represented by $\ast$ and the transfer function zeros represented by $\circ$.

Fig 4-6 Geometric interpretation of transfer function poles and zeros
Poles and Zeros of a First Order System

\[ G(s) = \frac{K}{\tau s + 1} \]

A first order transfer function has no zeros and one pole located at \( s = -1/\tau \).

**Fig 4-7** Pole-Zero Plot of a first order system
Poles and Zeros of an Integrator

\[ G(s) = \frac{K}{s} \]

An integrator has no zeros and one pole located at \( s = 0 \). A pole located at the origin of the complex plane is commonly called an integrator pole.

Fig 4-8 Pole-Zero Plot of an integrator
Poles and Zeros of a Second Order Overdamped System

A second order overdamped system has a transfer function that can be written as

\[ G(s) = \frac{K}{(\tau_1 s + 1)(\tau_2 s + 1)} \]

where \( K, \tau_1 \) and \( \tau_2 \) are all positive parameters.

There are no zeros and two real poles at \( s_1 = 1/\tau_1 \) and \( s_2 = -1/\tau_2 \).

Fig 4-9  Pole-Zero Plot of an second order overdamped system

Recall that the step response will be somewhat similar to a first order response but the first part of the response will be somewhat slower than that of a first order system. A step response without overshoot or oscillatory behavior is associated with poles located on the real axis.
Poles and Zeros of a Second Order Underdamped System.

A second order underdamped system has a transfer function that can be written as.

\[ G(s) = \frac{K}{\tau^2 s^2 + 2\tau \zeta s + 1} \quad \zeta < 1 \]

where \( K \), \( \tau \) and \( \zeta \) are all positive parameters.

There are no zeros and there are two poles which are the two solutions of the characteristic equation:

\[ \tau^2 s^2 + 2\tau \zeta s + 1 = 0 \]

By the quadratic formula the two solutions to this equation are:

\[ s_1 = \frac{\zeta + \sqrt{\zeta^2 - 1}}{\tau} = -\frac{\zeta + \sqrt{1 - \zeta^2}}{\tau} j \quad \text{and} \]

\[ s_2 = \frac{\zeta - \sqrt{\zeta^2 - 1}}{\tau} = -\frac{\zeta - \sqrt{1 - \zeta^2}}{\tau} j \]

Fig 4-10 Pole-Zero Plot of an second order underdamped system

Recall that the step response will overshoot the final value and eventually approach the final steady state value in a decaying oscillatory fashion. A step response with overshoot and oscillatory behavior is associated with poles which do not lie on the real axis, namely, with complex conjugate poles with non-zero imaginary parts.
Stability

What is meant by stability?

"A system can be considered stable if its output response to a step input is bounded."

How do we determine stability?

Several methods
- Examine time domain response
- Location of poles in s-plane
- From Frequency Response

Example

The time response of a first order system to a unit step input is

\[ y(t) - K \left( 1 - e^{t/\tau} \right) \]

Case 1 \( \tau > 0 \)

Determine the final value of \( y(t) \) as \( t \to \infty \)

\[
\lim_{t \to \infty} y(t) = \lim_{t \to \infty} \left[ K \left( 1 - e^{t/\tau} \right) \right] - K
\]

Bounded output Stable

Case 2 \( \tau < 0 \)

Determine the final value of \( y(t) \) as \( t \to \infty \)

\[
\lim_{t \to \infty} y(t) = \lim_{t \to \infty} \left[ K \left( 1 - e^{t/\tau} \right) \right] - \infty
\]

Unbounded output Unstable
Stability in the Laplace Domain

Since it is cumbersome to work in the time domain, we should examine how stability can be determined in the Laplace Domain.

From the previous example, plot the system poles

![Diagram showing pole locations for a stable and unstable system.]

Fig 4-11 Pole locations for a stable system and unstable system.

- $\tau > 0$ gives stable poles
- $\tau < 0$ gives unstable poles

$\Rightarrow$ A SYSTEM WITH POLES IN THE RIGHT HALF PLANE IS UNSTABLE

The stability of a system can thus be determined by the locations of its poles. The locations of the poles are determined by the denominator of the transfer function.

Define the characteristic equation of a system to be the equation generated by equating the denominator of its transfer function to zero.

**Example** First Order System

Transfer function

$$G_p(s) = \frac{K_p}{\tau s + 1}$$

Characteristic Equation

$$0 = \tau s + 1$$

Root of Characteristic Equation

$$s = \frac{1}{\tau}$$

**Problem** The roots of the characteristic equation are not easily determined when the order of the equation is greater than 2

**Example** $As^3 + Bs^2 + Cs + D = 0$

In such cases, the Routh test can be used, see Appendix 2A
Pole Position Interpretation

Poles occur either as single poles on the real axis of the s-plane or as complex conjugate pairs. The location of the pole determines the key features of the dynamic response such as settling time, whether or not there is overshoot and if the is response is oscillatory, etc. This will now be discussed in detail.

Pole Position Interpretation of a Real Pole

Consider the first order transfer function

\[ G(s) = \frac{K}{\tau s + 1} = \frac{K}{s - \alpha} \]

where \( K \) is the gain of the transfer function and \( \tau \) is the time constant. There is a single pole at \( s = -1/\tau - \alpha \).

A pole on the real axis is associated with a first order response. A pole on the negative real axis at \( s = \alpha \) is associated with a first order response with time constant \( \tau = 1/\alpha \).

Step responses for various real pole positions are shown in the following figure.

Fig 4-12  Step responses for various pole positions on the real axis.
Pole Position Interpretation of Purely Imaginary Poles

Consider the Harmonic Oscillator or Linear Oscillator with transfer function

\[ G(s) = \frac{K\omega^2}{s^2 + \omega^2} \]

and purely imaginary poles located at \( s_1 = \pm j\omega \) and \( s_2 = \pm j\omega \). Purely imaginary poles are associated with a sustained oscillatory response. Poles located at \( s = \pm j\omega \) correspond to a sustained oscillatory response with angular frequency \( \omega \) (cyclic frequency \( f = \omega/2\pi \) and period \( T = 2\pi/\omega \)). This is illustrated below.

Fig. 4-13  Step responses for various pole positions on the imaginary axis.
Pole Position Interpretation of Complex Conjugate Pole Pairs

Consider the second order underdamped transfer function

\[ G(s) = \frac{K_p}{\tau^2 s^2 + 2\tau\zeta s + 1} - \frac{K_p \omega_n^2}{s^2 + 2\omega_n\zeta s + \omega_n^2} \]

where \( \tau \) is the second order time constant, \( \omega_n = 1/\tau \) is the undamped natural frequency and \( \zeta \) is the damping coefficient.

There are two complex conjugate poles located at:

\[ s_{1,2} = \omega_n \zeta + j\omega_n \sqrt{1 - \zeta^2} - \sigma + j\omega \]

where the real part of the pole is \( \sigma = -\omega_n \zeta \) and the imaginary part of the pole is \( \omega = \omega_n \sqrt{1 - \zeta^2} \).

Complex conjugate poles are associated with a damped oscillatory response. The pole location determines the features of the response as follows:

1. The angular frequency of the oscillation will be equal to \( \omega \) (imaginary part of the pole). Thus the cyclic frequency will be \( \omega = \omega / 2\pi \) and the period will be \( T = 2\pi / \omega \).

2. The amplitude of the oscillation will decay with time in a first order fashion with time constant \( \tau = -1/\sigma \) (where \( \sigma \) is the real part of the pole).

3. The amount of overshoot of the response will be determined by the damping coefficient \( \zeta \). In turn, \( \zeta \) is determined by the angle \( \theta \) that the pole makes with the negative real axis according to the formula \( \zeta = \cos \theta \).

This is illustrated in the following figures.
Fig. 4-14. Effect on step response of varying the real part of complex conjugate poles.
Fig 4-15  Effect on step response of varying the imaginary part of complex conjugate poles
Fig 4-16. Effect on step response of varying the angle $\theta$ of complex conjugate poles
Dominant Poles Concept

The response of a system is the superposition of the responses due to each of its poles (the zeros also effect the response). The contribution of the poles closest to the imaginary axis (called the dominant poles) determine the basic shape of the response.

Example

**System 1**

(Process model)

\[ G_1(s) = \frac{K_p}{(10s + 1)(2s + 1)} \]

**System 2**

(Simplified model determined by dominant pole)

\[ G_2(s) = \frac{K_p}{10s + 1} \]

Fig 4-17  Step responses of a second order overdamped system and its dominant dynamics
The Effect of an Additional Real Pole on Dominant Dynamics

The effect on the step response of adding a single real pole to a transfer function is shown in Fig. 4-18.

Fig 4-18  The effect of an additional real pole on the step response.

The added pole slows down the initial part of the response relative to the dominant dynamics.
The Effect of an Additional Real Zero on Dominant Dynamics

The effect on the step response of adding a single real zero to a transfer function is shown in Fig 4-19

Effect of Adding an Extra Zero

Output

Dominant Dynamics

Time

Fig 4-19  The effect of an additional real zero on the step response

The added zero speeds up the initial part of the response relative to the dominant dynamics

The lead lag transfer function considered on the following page exhibits these features
Lead-Lag Transfer Function

The transfer function of a lead-lag block is

\[ G(s) = \frac{\beta s + 1}{\tau s + 1} \]

where \( \beta \) is the lead time constant and \( \tau \) is the lag time constant. Thus there is one pole at \( s = -\frac{1}{\tau} \) and one zero at \( s = -\frac{1}{\beta} \). The lead-lag can be considered to be a first order transfer function with a zero added.

A common mechanism for the creation of transfer function zeros is that the system can be regarded as a parallel combination of two systems (of different form). By long division, the lead-lag transfer function can be represented as follows:

\[ G(s) = \frac{\beta s + 1}{\tau s + 1} = \frac{\beta}{\tau} + \frac{1 - \beta/\tau}{\tau s + 1} \]

Thus the lead-lag system can be regarded as a parallel combination of a pure gain system with a gain of \( \beta/\tau \) and a first order system with a gain of \( 1 \) and \( \beta/\tau \) as shown below.

![Lead-Lag Block Diagram](image)

Fig 4-20  Representation of a lead-lag as a parallel combination.

The step response of the lead-lag will be the sum of the step responses of the two blocks in parallel as shown in the following figure. Thus the total response time (time to reach steady-state) will be determined by the first order part namely the lag time constant \( \tau \). The effect of the system zero will be to speed up the initial part of the response, the larger the ratio \( \beta/\tau \), the larger will be the effect on the speeding up effect on the initial part of the response.
Fig 4-21. Step response of a lead-lag transfer function with $\beta < \tau$.

Fig 4-22. Step response of a lead-lag transfer function with $\beta = \tau$.

Fig 4-23. Step response of a lead-lag transfer function with $\beta > \tau$. 
Overshoot Caused By Transfer Function Zeros

The Lead-Lag example shows that effect of the zero on the dynamic response can be quite significant. When the zero location is slower (closer to the imaginary axis) than the dominant pole(s), the result is likely to cause an overshoot in the step response of the system, as shown below.

Fig. 4-24. Overshoot caused by a zero slower than the dominant pole(s)
Wrong Way Response Caused By Transfer Function Zeros

The effect of a transfer function zero in the right half plane, is to make the step response exhibit a "wrong way response", as shown below.

![Diagram of output and time response]

**Fig. 4-25 Wrong way response caused by a right half plane zero**

This wrong way response behavior is exploited by the Pade approximations for deadtime in a transfer function. Pade approximations will be covered later in the course.

The remainder of this section consists of a number of process dynamics examples exhibiting behavior more complicated than simply first order or integrating. The principles developed here can be used to identify an appropriate process model.
Fig. 4-26. Sample step response.

Dominant Dynamics Transfer Function: \( G(s) = \frac{k}{\tau_s s + 1} \)

Fig. 4-27. Pole Zero Plot for sample dynamics

Model Transfer Function: \( G(s) = \frac{k}{\tau_s s + 1} \)
Fig 4-28  Dryer Steam Pressure step response

Dominant Dynamics Transfer Function  $G(s) = \frac{1}{1.s}$

Fig 4-29  Pole Zero Plot for Dryer Steam Pressure dynamics

Model Transfer Function  $G(s) = \frac{1}{1.s+1}\frac{1}{1.s+1}$
Fig 4-32  Dryer Steam Pressure step response

Dominant Dynamics Transfer Function:  \( G(s) = \frac{1}{(s+1)(s+2)} \)

Fig 4-33  Pole-Zero Plot for Dryer Steam Pressure dynamics
Fig 4-34  Boiler Steam Drum Level step response.

Dominant Dynamics Transfer Function \( G(s) = \frac{\frac{1}{s}}{\frac{s}{\zeta} + 1} \).

Fig 4-35  Pole-Zero Plot for Boiler Steam Drum Level dynamics

Model Transfer Function: \( G(s) = \frac{k}{s} \left(1 - \frac{\beta}{s}\right) \left(\frac{1}{s\zeta}\right)\).
5. FREQUENCY RESPONSE
Sinusoidal Inputs

The frequency response of a system (e.g. valve, process, filter, etc) is the response of the system to a sinusoidal input.

A sinusoidal input can be described by specifying its amplitude $A$, and its period, $T$.

Fig 5-1

Alternatively, we may specify instead of the period $T$, either:

(i) the cyclic frequency

$$f = \frac{1}{T} \quad \text{(cycles/time unit)}$$

(ii) or, the angular frequency

$$\omega = 2\pi \cdot f \quad \text{(radians/time unit)}$$
Fig 5-2

- frequency stays the same
- phase change between input and output occurs
- amplitude of sinusoid changes

\[ T_o = T_i \]
\[ AR = \frac{A_o}{A_i} \]
\[ \phi = \frac{t_p}{T_i} \times 360 \text{ (degrees)} \]

AR is called the amplitude ratio (gain)
\( \phi \) is called the phase shift
Example 1  Chest consistency from inlet to outlet

Input

Fig 5-3
Step Response

Output

Kp=1
\( \tau = 5 \text{ min} \)

Fig 5-4
**Frequency Response**

**Frequency Response of Stock Consistency from Inlet to Outlet**

![Graph](image)

**Fig 5-5**

**Example 2**  
First order filter, with filter time constant \( T_f \)

**Step response**

![Step response graph](image)

**Fig 5-6**

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Frequency response

Frequency Response of a First Order Filter

\begin{align*}
T_c &= 2\pi f_c \\
f_c &= \frac{1}{2\pi \tau_f} \\
\omega_c &= \frac{1}{\tau_f}
\end{align*}

Fig 5-7
First Order System

\[ K_p \quad \text{process gain} \]
\[ \tau \quad \text{process time constant} \]

The amplitude ratio is given by.

\[ AR = \frac{K_p}{\sqrt{\omega^2 \tau^2 + 1}} \]

where \( \omega \) is the input angular frequency

The cutoff frequency \( \omega_c \) is

\[ \omega_c = \frac{1}{\tau} \quad \text{(rad/time unit)} \]

At the cut-off frequency

\[ AR = \frac{K_p}{\sqrt{2}} = 0.707K_p \]

Using cyclic frequency \( f = \frac{\omega}{2\pi} \)

\[ AR = \frac{K_p}{\sqrt{(f_c)^2 + 1}} \]

where

\[ f_c = \frac{\omega_c}{2\pi} - \frac{1}{2\pi \tau} \approx \frac{1}{6.28\tau} \]

is the cyclic cut-off frequency
Using period of the input \( T = \frac{1}{f} \)

\[
AR = \frac{K_p}{\sqrt{\left(\frac{T_c}{T}\right)^2 + 1}}
\]

where

\[
T_c = \frac{1}{f_c} = 2\pi\tau \approx 6.28\tau
\]

is the cut-off period.

For \( T \gg T_c \) (low frequency)

\[
AR \approx \frac{K_p}{\sqrt{1}} = K_p
\]

For \( T \ll T_c \) (high frequency)

\[
AR \approx \frac{K_p}{\sqrt{\left(\frac{T_c}{T}\right)^2}} = \left(\frac{T}{T_c}\right)K_p
\]
Example  First order process, $K_P - 1$, $\tau = 5.0$ min.

Find the $AR$ when (a) $T = 25$ min  (b) $T = 5$ min

cut-off period

$$T_c = 2\pi \times \tau = 6.28 \times 5 = 31.4 \text{ min}.$$  

(a) when input period is 25 min

$$AR = \frac{K_P}{\sqrt{\left(\frac{T_c}{T}\right)^2 + 1}} = \frac{1}{\sqrt{\left(\frac{314}{25}\right)^2 + 1}} \approx 0.62$$

(b) $AR$ when input period is 5 min:

$$AR = \frac{K_P}{\sqrt{\left(\frac{T_c}{T}\right)^2 + 1}} = \frac{1}{\sqrt{\left(\frac{314}{5}\right)^2 + 1}} \approx 0.16$$

Note

$$AR \approx \left(\frac{T}{T_c}\right)K_P \approx \frac{5}{314} = 0.16$$

<table>
<thead>
<tr>
<th>$T$ (min)</th>
<th>$AR$</th>
<th>Note</th>
</tr>
</thead>
<tbody>
<tr>
<td>input period</td>
<td>amplitude ratio</td>
<td></td>
</tr>
<tr>
<td>314</td>
<td>100%</td>
<td>slower than the cutoff</td>
</tr>
<tr>
<td>314</td>
<td>70%</td>
<td>at the cutoff</td>
</tr>
<tr>
<td>314</td>
<td>10%</td>
<td>10 times faster than the cutoff</td>
</tr>
<tr>
<td>0.314</td>
<td>1%</td>
<td>100 times faster than the cutoff</td>
</tr>
</tbody>
</table>
Bode Diagrams

Bode diagrams provide a graphic depiction of the system's frequency response characteristics.

A Bode diagram consists of two portions. One plot shows log(AR) versus log(freq) and the other phase versus the log(freq).

Bode diagrams sometimes plot the AR in decibels. The relationship is

\[
\text{decibels (dB)} = 20 \log(AR)
\]

Also, there are usually distinct 'straight line' portions which give the asymptotic nature of the system. The AR plot will often only show the asymptotes

<table>
<thead>
<tr>
<th>AR</th>
<th>AR (dB)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>0</td>
</tr>
<tr>
<td>10</td>
<td>+20</td>
</tr>
<tr>
<td>0.1</td>
<td>-20</td>
</tr>
<tr>
<td>0.707</td>
<td>-3</td>
</tr>
<tr>
<td>2</td>
<td>+6</td>
</tr>
<tr>
<td>0.5</td>
<td>-6</td>
</tr>
</tbody>
</table>
Example Bode Plot of a first order system with process gain $K_p$ and process time constant $\tau$

![Bode Plot Diagram]

**Fig. 5-8**

The frequency $\omega = 1/\tau$ is referred to as the breakpoint. At this frequency

- the $AR$ asymptotes intersect
- $AR = 0.707$
- $20 \log(AR) = -3$ dB
- $\phi = -45^\circ$

The maximum phase angle is $90^\circ$
Example. Second order system.

\[ G(s) = \frac{1}{\tau^2 s^2 + 2 \tau \zeta s + 1} \]

\[ \zeta = \text{damping coefficient} \]

\[ \zeta = 0.1 \]

\[ \zeta = 1.0 \]

\[ 20 \log (\text{AR}) \] (dB)

\[ \text{Phase angle} \]

\[ \omega \] (rad./s or rad./time)

Fig. 5-9

A second order underdamped system may exhibit a resonant peak. This will occur for \( \zeta < 0.707 \). In these situations, the AR will exceed the steady state.
gain at the resonant frequency with the result that input variations may drastically increase in amplitude (resonance)

Example  PI Controller
6. PID CONTROLLERS
Closing the Loop

Thus far, we have considered processes operating in open loop (manual). Automatic control effectively closes the loop.

In closed loop operation, two additional components are added.

1) Comparator
2) Controller

The comparator, which is sometimes called the error summing junction, compares the process measurement with a reference signal (the setpoint).

It can be diagrammed as

\[
\begin{align*}
Y_{SP} & \rightarrow E \\
E & \rightarrow Y
\end{align*}
\]

Fig 6-1

where:

- \( Y \) is the process measurement
- \( Y_{SP} \) is the setpoint or reference signal
- \( E = Y_{SP} - Y \) is the control error

The controller acts upon the control error to determine the control output which becomes the input to the process.

The closed loop system may be represented as.

\[
\begin{align*}
Y_{SP} & \rightarrow E \\
E & \rightarrow U \\
U & \rightarrow Y
\end{align*}
\]

Fig 6-2
The response from the setpoint $Y_{SP}$ to the PV $Y$ is given by

$$G_{SP}(s) = \frac{G_C(s)G_P(s)}{1+G_C(s)G_P(s)}$$

and is called the closed loop transfer function.

In addition, we are greatly concerned with any load disturbances which may affect the output. Denote the load by $L(s)$ and its effect on the output by $G_L(s)$ which now gives

![Block diagram of closed loop system with load disturbance](image)

Fig 6-3

When we do not have any particular load disturbance in mind we represent the net effect of all disturbances by an additive disturbance as follows

![Block diagram of closed loop system with additive disturbance](image)

Fig 6-4

\[ Y - D = Y(GC \times GP) \]

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Assuming that the setpoint remains constant, the response from the load $D$ to the PV $Y$ is given by

$$G_{LD}(s) = \frac{1}{1 + G_C(s)G_P(s)}$$

This transfer function will be referred to as the load transfer function.

Both the closed loop transfer function and the load transfer function are obtained as special cases of the following block diagram reduction rule.

**Rule 3:** In a negative feedback control loop, the closed loop transfer function between two points 'A' and 'B' is

$$G_{CLOSED\_LOOP_{AB}} = G_{AB\,(OPEN\_LOOP)} \cdot \frac{1 + G_{LOOP}}{1 + G_{LOOP}} = G\,(FORWARD)$$

---

**Diagram Description:**

- The diagram represents a control system with feedback loops.
- The transfer functions $G_2(s)$, $G_3(s)$, and $G_1(s)$ are shown.
- The loop is highlighted, indicating the feedback path.
- The equation for the closed loop transfer function is derived from the open loop transfer function.

---

**Figure 6-5**
The PID Control Algorithm

By far the most common control algorithm used is the PID algorithm. It is composed of three portions which provide:

1) P - Proportional Action
2) I - Integral Action
3) D - Derivative Action

Examine the rationale of each term.

P - This term generates a control action proportional to the control error. This is perhaps the most typical of what would occur with continual manual adjustment.

I - This term generates a control action proportional to the integral of the control error. It can be shown that for most systems, proportional only control will not bring the control error to zero for a step change in setpoint (leaves an 'offset'). This will be overcome by Integral Action.

D - This term contributes a control action proportional to the derivative of the control error. This can be used when we want the controller to respond quickly to a sudden change in the control error, such as would result from a step change in a load disturbance.

Forms of the PID Controller

There are a variety of ways in which the PID controller may be represented mathematically. The three common forms are:

(1) Standard, Noninteracting or 'ISA' Form,
(2) Parallel Form and
(3) Classical, Interacting or Series Form

Each of the three forms has a slightly different way of producing the proportional, integral and derivative actions in the controller output.

There are also three major forms of the PID equations. These are the ISA STANDARD form, the PARALLEL PATH and the CLASSICAL form. The controller output equation differs for each of the three forms, as is illustrated below.
Parallel Form

The parallel form of the PID controller can be represented as follows:

\[ u = K_P e + K_I \int e \, dt + K_D \frac{de}{dt} \]

The controller transfer function is:

\[ G_c(s) = K_P + K_I s + K_D s^2 + \frac{K_P s + K_I}{s} \]

The tuning parameters are.

- \( K_P \) called the proportional gain,
- \( K_I \) called the integral gain and
- \( K_D \) called the derivative gain.

Note In this chapter only, \( K_P \) will be used to denote proportional gain and not process gain

Although the parallel form presents the P, I and D actions in a straight forward manner it has the disadvantage that a change of any of the three gains will change the location of the two controller zeros. This problem is overcome by using the standard form of the PID controller.
Standard Form of the PID Controller

The standard form of the PID controller can be represented as:

\[ u = K_c \left( e + \frac{1}{T_R} \int e \, dt + T_D \frac{de}{dt} \right) \]

The controller transfer function is

\[ G_c(s) = \frac{K_c \left( 1 + \frac{1}{T_R s} + T_D s \right)}{T_R \left[ T_R T_D s^2 + T_R s + 1 \right]} \]

The tuning parameters are

- \( K_c \) called the controller gain,
- \( T_R \) called the reset time and
- \( T_D \) called the derivative time

\( T_R \) and \( T_D \) are usually expressed in minutes in North American control equipment and in seconds in European and Japanese equipment.
Classical Form of the PID Controller

The classical form of the PID controller can be represented as:

\[
\begin{align*}
    u &= K_C' \left[ \left(1 + \frac{T_D'}{T_R'}\right) e + \frac{1}{T_R'} \int e \, dt + T_D' \frac{de}{dt} \right]
\end{align*}
\]

The controller transfer function is:

\[
G_c(s) = K_C' \left( T_D's + 1 \right) \left( 1 + \frac{1}{T_R's} \right)
= \frac{K_C'}{T_R'} \left[ \frac{(T_D's + 1)(T_R's + 1)}{s} \right]
\]

The tuning parameters are:

- \( K_C' \): called the controller gain,
- \( T_R' \): called the reset time and
- \( T_D' \): called the derivative time

\( T_R' \) and \( T_D' \) are usually expressed in minutes in North American control systems and in seconds in European and Japanese systems.

Although the tuning parameters of the classical and standard forms have the same names, they are not identical.
**PROPORTIONAL**

The proportional term will move the output PROPORTIONAL to the error

This component is equal to the following,

\[ P = K_c \times \text{error} \]

where \( K_c \) = proportional gain, normally expressed in units of \( \frac{\%\text{OUTPUT}}{\%\text{SPAN}} \)

\( \text{error} = (\text{Setpoint} - \text{Process}) \), normally expressed in units of \% Span

Consider the setpoint bump shown below

Assume

\( K_c \) is equal to 0.5 \( \frac{\%\text{Output}}{\%\text{Span}} \) and the Setpoint bump is equal to 2\% Span

On the setpoint bump the error is equal to 2 \% Span

Therefore, the proportional term will move by:

\[ P = (K_c \times \text{error}) = 0.5 \times \frac{\%\text{Output}}{\%\text{Span}} \times 2\% \text{Span} = \%\text{Output} \]

This control action taken at the setpoint bump or in response to a disturbance is often referred to as the proportional \( K_c \)
In the example above, the error is fixed. The proportional term makes a single adjustment. As long as the error does not change, the proportional term will remain constant.

However, our normal expectation is that when the controller output changes, the process changes and the error decreases. Consider the setpoint bump below, where the process changes in response to the output change. In this example, the initial proportional 'kick' would eventually bring the process to the new setpoint. However, as the process responds, the error decreases and the proportional term decreases. This will result in an 'offset' from the setpoint. This offset represents the primary limitation of the proportional term.
PROPORTIONAL ONLY CONTROL

Figure 6-10  Proportional Only - Offset from Setpoint

Proportional gain can also be expressed as PROPORTIONAL BAND The relationship is as follows.

$$PB = \frac{100\%}{Kc}$$
The INTEGRAL term is proportional to the integral of the error over time. In the ISA standard form, the integral term is equal to the integral of error over time multiplied by the Proportional gain and divided by the Reset Time.

\[ I = \frac{K_c}{T_R} \int_0^t \text{error}(t)\,dt \]

where \( T_R \) = Reset Time, normally expressed in minutes,

and, \( \int_0^t \text{error}(t)\,dt \) = Integral of the Error over time

The integral of the error over time can be thought of as the accumulated value of the error multiplied by the change in time. The integral term will keep accumulating as long as the error is not equal to zero. Thus, the integral term will eventually bring the error to zero.

Consider the following PI controller response to a Setpoint bump. In this example the process (PV) does not respond to the controller output changes and therefore the error is constant.
PI CONTROLLER RESPONSE TO A SETPOINT BUMP
NO PROCESS CHANGE

Figure 6-11 PI Controller Response to Setpoint Bump - Error is Fixed

In this special case where the error is fixed, the integral term accumulates at a fixed rate. After one Reset period, the integral term has repeated the proportional kick. This is the definition of the Reset (or Integral) time.

The normal expectation however is that the process will respond to a controller output change. As the process moves toward setpoint the error decreases. The integral of the error multiplied by time increases but at an ever decreasing rate. A more typical controller output response is illustrated in the figure below.
Figure 6-12 PI Controler Response to Setpoint Bump - Error not Fixed

Reset can be expressed as a Reset Time (or Integral Time), $T_R$ (with units MINUTES PER REPEAT), or as RESET GAIN, $K_I$ (with units REPEATS PER MINUTE) The relationship is

$$K_I = \frac{1}{T_R}$$
DERIVATIVE

The DERIVATIVE corrective action is proportional to the rate of change of the error. This term is equal to the following:

\[ D = K_c \cdot T_D \cdot \frac{de}{dt} \]

where \( T_D \) = Derivative Time, normally expressed in minutes

and \( \frac{de}{dt} \) = rate of change in the error

The FASTER the error grows, the MORE DERIVATIVE action there is. The DERIVATIVE TUNING PARAMETER (\( T_D \)) is usually expressed in minutes.

Consider the setpoint bump below. The derivative term will change in the following manner on a setpoint change.

![Diagram showing the operation of a PID controller with a setpoint bump, illustrating the proportional, integral, and derivative terms over time.]

Figure 6-13 Sum Total of the Proportional, Integral and Derivative Control Terms
Before the setpoint bump, the rate of change in the error is zero. Immediately at the setpoint bump, the rate of change of error is high and therefore the derivative term is high. Following the setpoint bump, the rate of change of error returns to zero and the derivative term returns to zero. Note that the rate of change of the error is filtered to prevent an infinite derivative term on setpoint change or sudden disturbance.

DERIVATIVE can be useful when the PROCESS DYNAMICS are.

1. very quiet (no noise)
2. predictable or repeatable
3. slow and gradual

Derivative can be thought of as introducing a "lead" factor in the controller's corrective action. Derivative action is sometimes useful in temperature loops to 'charge-up' the second order dynamics. In the example below, a second order open loop response is compared with the closed loop PV response of an aggressively tuned PID controller. Note that the closed loop PV response more closely resembles a first order response, due to the Derivative lead action. However, note the very aggressive controller (valve) action associated with derivative.

Figure 6-14 Open Loop Second Order Response Versus Closed Loop PID Tuned Response

DERIVATIVE MAKES A CONTROLLER VERY SENSITIVE TO A NOISY PV
CONVERSION FORMULAS
The tuning rules presented in this course are based on the classical form of the PID algorithm. Formulas to convert these tuning rules to the Parallel path form and the ISA Standard form are presented below.

CLASSICAL FORM TO ISA STANDARD

The Classical form of the PID equation is identical to the Standard form if Derivative action is not being used. Therefore, entering the same tuning constants into a Classical form or the ISA Standard form will produce the same controller response to a load disturbance or setpoint change. However, if Derivative is being used, the conversion formulas shown below must be applied.

CLASSICAL FORM

Typical users: HONEYWELL, FISHER, FOXBORO, TAYLOR

To convert from CLASSICAL TO STANDARD FORM

\[ K_{C\,\text{Standard}} = K_{C\,\text{Classical}} \left( \frac{T_{R\,\text{Classical}} + T_{D\,\text{Classical}}}{T_{R\,\text{Classical}}} \right) \]

\[ T_{R\,\text{Standard}} = T_{R\,\text{Classical}} + T_{D\,\text{Classical}} \]

\[ T_{D\,\text{Standard}} = \frac{T_{R\,\text{Classical}} \cdot T_{D\,\text{Classical}}}{(T_{R\,\text{Classical}} + T_{D\,\text{Classical}})} \]

Also NOTE the following

For Foxboro

\[ K_C = \frac{100\%}{\text{PB}} \]

where PB is called a Proportional Band

For Fisher and Taylor

\[ K_I = \frac{1}{T_R} \]

where \( K_I \) is in repeats per minute
CLASSICAL FORM TO PARALLEL PATH FORM

In the parallel form of the PID equation, the Integral and the Derivative terms are independent of the proportional gain. The Parallel path Integral constant, normally expressed in units of %Span / %Output / Minute, does not have the same physical meaning as the Integral constant of the ISA standard form.

It is important to use the conversion formulas below when using a Parallel Path PID equation.

Typical users - BAILEY

To convert from the CLASSICAL to PARALLEL PATH FORM

\[ K_c_{\text{Parallel}} = K_c_{\text{Classical}} \left( \frac{TR_{\text{Classical}} + TD_{\text{Classical}}}{TR_{\text{Classical}}} \right) \]

\[ K_I_{\text{Parallel}} = \frac{K_c_{\text{Classical}}}{TR_{\text{Classical}}} \]

\[ K_D_{\text{Parallel}} = K_c_{\text{Classical}} TD_{\text{Classical}} \]

* Note that Bailey terminology for the proportional gain is \( K_p \), this should not be confused with the PROCESS GAIN \( K_p \)
PI control

Classical Form

\[ G_c(s) = K_c'(1 + \frac{1}{T_R's}) = K_c'\left(\frac{T_R's+1}{T_R's}\right) \]

Standard Form

\[ G_c(s) = K_c(1 + \frac{1}{T_R's}) = K_c\left(\frac{T_R's+1}{T_R's}\right) \]

= \frac{K_c + \frac{K_c}{T_R}(\frac{1}{s})}{T_R's}

Parallel Form

\[ G_c(s) = K_{PROP} + \frac{K_{INT}}{s} \]

= \frac{K_{PROP}s + K_{INT}}{s}

6-8a
PI only

To convert classical/standard to Parallel.

\[ K_{\text{PROP}} = K_c \]

\[ K_{\text{INT}} = \frac{K_c}{TR} \]
7. LAMBDA TUNING FOR FIRST ORDER PROCESSES
GENERAL TUNING PRACTICE

The general tuning procedure can be described as follows:

- If the loop is too slow, then the tuner would increase the proportional gain. The proportional component represents the majority of the corrective action.
- Add some reset in order to remove any offset.
- Very few loops (if any) use derivative action.

This approach is essentially Tuning "by feel" and tends to result in relatively high controller gains and slow reset times. After years of tuning "by feel", certain ranges of tuning parameters are regarded as ACCEPTABLE. This range of tuning parameters is called the "COMFORT ZONE", shown below. Unfortunately, it is hard to achieve excellence in manufacturing through 'feel' alone.

Figure 7-1 Typical Tuning Parameters vs Lambda Tuning Parameters

Copyright EnTech
The Lambda tuning method is the focus of this course. Two other tuning methods, namely the Ziegler-Nichols method and the Cohen-Coon method are briefly described below.

The objective of the Ziegler-Nichols and the Cohen-Coon methods is to achieve a Quarter Amplitude damped process response to a setpoint bump. These have been the most widely taught methods since the 1940s and 1950s. It is likely that people with previous instrumentation training have been exposed to these methods or similar variations. Lambda tuning has its origins in more modern ideas (1968) and has been gaining popularity in recent years.
The Ziegler-Nichols method was first developed in 1942. It is based on concepts first used for ant-aircraft gun dynamics. The objective then was to get the guns pointing in the right direction as fast as possible. There was very little concern about oscillations or an underdamped response.

The Ziegler-Nichols method produces a quarter-amplitude damped (underdamped) response to a setpoint change in which each successive overshoot is 1/4 of its predecessor in magnitude. The initial overshoot is approximately 50% of the setpoint bump. The Ziegler-Nichols method has been the chief tuning method taught at colleges and universities for many years. It has been automated in several loop auto-tuners that are currently available.

In the process industries where product variability is very important, this is a destabilizing and damaging tuning method.

\[
\frac{B}{A} = \frac{1}{4}
\]

Figure 7-2: Quarter-Amplitude Damped Response Using Ziegler-Nichols Tuning Method.

The tuning results are recommended only as a reasonable starting point.
PROCEDURE

The tuning procedure is as follows:

1. Remove any RESET and DERIVATIVE from the controller by making $T_R$ as large as possible or making $K_I$ equal to zero.

2. Put the loop on AUTO and increase the controller gain $K_C$ (or decrease PB) until a SUSTAINED OSCILLATION is recorded. The amplitude of the OSCILLATION must not be increasing or decreasing. In order to cause the loop to cycle, a small setpoint change may be required.

   ![Diagram of Tu (Ultimate Period)]

Figure 7.3: Measure the Ultimate Period From the Sustained Oscillation

3. The following parameters are then recorded:

   - $K_C$ or PB, this is called the ULTIMATE GA N (the gain which caused the oscillation)
   - $T_U$, this is called the ULTIMATE PERIOD (the period of the oscillation)

4. Use these numbers to calculate the tuning parameters according to the Table 7-1.
### Ziegler-Nichols Tuning Method

<table>
<thead>
<tr>
<th>TYPE OF CONTROLLER</th>
<th>PROPORTIONAL</th>
<th>UNITS</th>
<th>RESET</th>
<th>DER</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>$K_P$</td>
<td>$PB(%)$</td>
<td>$TR$ (min/rpt)</td>
<td>$KI$ (rpt/min)</td>
</tr>
<tr>
<td>P ONLY</td>
<td>0.5 $K_P$</td>
<td>2 $PB_u$</td>
<td></td>
<td></td>
</tr>
<tr>
<td>PI</td>
<td>0.45 $K_P$</td>
<td>2.22 $PB_u$</td>
<td>$T_u/1.2$</td>
<td>1.2/$T_u$</td>
</tr>
<tr>
<td>PID</td>
<td>0.6 $K_P$</td>
<td>1.67 $PB_u$</td>
<td>$T_u/2$</td>
<td>2/$PB_u$</td>
</tr>
</tbody>
</table>

Table 7-1: Formulate a Table for Ziegler-Nichols Tuning Method

### Conclusions

1. The method does not provide a knowledge of the important non-linearities (i.e., backlash and friction) in the loop.

2. The method requires that the proportional gain be increased until the process oscillates, which may be unacceptable for operational reasons.

3. It is difficult to determine if the sustained oscillation is a result of true process dynamics or nonlinearities in the loop.

4. The controller's speed of response is not defined in the procedure. It is thus difficult to tune two loops to have an identical speed of response. This is a common tuning objective.
The Cohen-Coon Method, also known as the reaction curve method, was devised in the early 1950's. Though the procedure is different, its objective is also to produce a quarter-amplitude DAMPED response as with Ziegler-Nichols.

Its development was due to the fact that it was not practical to make loops oscillate during the tuning procedure. Imagine making the boiler feed water flow cycle!

The procedure is simple. It consists of performing a series of bump tests on the loop, while in MANUAL (as was discussed in Lecture 1) and recording the following process parameters:

- Deadtime: \( T_d = \)
- Time Constant: \( \tau = \)
- Process Gain: \( K_p = \)

The tuning parameters are calculated using the appropriate formula as in the Table 7.2.
### COHEN-COON TUNING METHOD

<table>
<thead>
<tr>
<th>Type of Controller</th>
<th>Proportional</th>
<th>Reset</th>
<th>Derivative</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>$K_C$</td>
<td>$PB$</td>
<td>$TR$</td>
</tr>
<tr>
<td>P ONLY</td>
<td>$\frac{Tau}{K_pT_d}\left(1+\frac{T_d}{3\text{Tau}}\right)$</td>
<td>$300K_pT_d$</td>
<td>(3Tau + Td)</td>
</tr>
<tr>
<td>PI</td>
<td>$\frac{Tau}{K_pT_d}\left(0.9 + \frac{T_d}{12\text{Tau}}\right)$</td>
<td>$1200K_pT_d$</td>
<td>(108Tau + Td)</td>
</tr>
<tr>
<td>PD</td>
<td>$\frac{Tau}{K_pT_d}\left(1.25 + \frac{T_d}{6\text{Tau}}\right)$</td>
<td>$600K_pT_d$</td>
<td>(75Tau + Td)</td>
</tr>
<tr>
<td>PID</td>
<td>$\frac{Tau}{K_pT_d}\left(1.33 + \frac{T_d}{4\text{Tau}}\right)$</td>
<td>$400K_pT_d$</td>
<td>(532Tau + Td)</td>
</tr>
</tbody>
</table>

$TD = \text{Deadtime}$

$TD = \text{Derivative Time}$

Table 7.2
CONCLUSIONS

1. The formulas shown all heavily rely upon deadtime. They are not applicable if $T_d$ is zero. This reliance on deadtime is mainly due to the fact that in the early 1950s most controllers were pneumatic, and pneumatic lines introduce DEADTIME. A general rule of thumb for estimating DEADTIME is that 100 feet of 1/4" pneumatic tubing introduces about 1.5 seconds of deadtime.

2. The testing time is usually quicker than the Ziegler-Nichols procedure as it allows the tuned to identify potential maintenance problems (e.g., BACKLASH, STICKING, etc.), and causes fewer process upsets.

3. As with the Ziegler-Nichols method, the tuning will cause the process to cycle on a setpoint bump or a random disturbance.
BACKGROUND

The ideas behind LAMBDA Tuning originated in the synthesis design method whereby the controller must "cancel out" the process dynamics. This method was applied in the pulp and paper industry in 1967 by E B. Dahl. In his paper on a model reference controller (not a PID controller), he used the Greek letter LAMBDA (λ) to signify the time constant of the control loop on automatic, hence the name LAMBDA TUNING.

The idea of using the Lambda tuning method in a PID controller (98% of all controllers in the pulp and paper industry are PID controllers) was put forward in 1975. However, it was not applied extensively in the pulp and paper industry until the early 1980's.

LAMBDA TUNING is still not a very well known tuning method but this will likely change as the demand for highly uniform products increases. It produces a first order (NON-Oscillatory) response to a setpoint change. The tuner decides how fast the PV will reach or return to setpoint. This is done by choosing the desired closed loop time constant, LAMBDA (λ). Even though certain restrictions do apply when CHOOSEING LAMBDA, it does offer great FLEXIBILITY and REPEATABLE RESULTS.
RESPONSE

By choosing LAMBDA, the tuner decides how fast the PV will reach the SP.

Figure 7-4 Lambda Tuned Response

THE PV ESSENTIALLY REACHES THE SP WITH N FOUR CLOSED LOOP TIME CONSTANTS LAMBDA

EXAMPLE IF THE TUNER CHOOSES A FLOW LOOP TO HAVE A LAMBDA OF 9 SECONDS, THE FLOW 'PV' WILL REACH THE SP, AFTER A STEP CHANGE IN 36 SECONDS.

NOTE THAT LAMBDA IS SIMILAR TO τ IN THE FOLLOWING WAY

\[ \frac{\tau}{\lambda} \]

CONTROLLER IN MANUAL

CONTROLLER IN AUTO
**PROCEDURE**

The procedure is quite simple.

1. Perform a series of bump test and obtain the following results
   - PROCESS DEADTIME $T_d = \underline{\phantom{0}}$
   - PROCESS TIME CONSTANT $\tau = \underline{\phantom{0}}$
   - PROCESS GAIN $K_p = \underline{\phantom{0}}$

   The results from each individual bump test will not be identical. It is safer to choose relatively conservative values. Choose a relatively high $K_p$ and a typical value of $\tau$.

2. Choose the LAMBDA value

   **REMEMBER, IT TAKES FOUR LAMBDA'S FOR THE PV TO REACH SP IN AUTOMATIC**

3. Using $\tau$, $T_d$, $K_p$ and LAMBDA, calculate the PI tuning parameters as follows,

<table>
<thead>
<tr>
<th>$\tau$ (min/rpt)</th>
<th>$K_i$ (rpt/min)</th>
<th>$K_C$ (%Out/%Span)</th>
<th>PB %PB</th>
</tr>
</thead>
<tbody>
<tr>
<td>$T_\tau$</td>
<td>$\frac{1}{\tau}$</td>
<td>$\tau$</td>
<td>$100 \times \frac{K_p(\lambda + T_d)}{T_\tau}$</td>
</tr>
<tr>
<td>$\frac{1}{\tau}$</td>
<td>$K_p(\lambda + T_d)$</td>
<td>$\tau$</td>
<td>$100 \times \frac{K_p(\lambda + T_d)}{T_\tau}$</td>
</tr>
</tbody>
</table>

Table 7-3 Lambda Tuning Formulas

The Reset Time is always set equal to the open loop time constant $\tau$. The calculation of the proportional gain is a function of the process gain, the desired Lambda value, the open loop time constant and the dead time.
EXAMPLE:
The process dynamics of a flow oop are as follows

\[ K_p = 2 \frac{\text{%Span}}{\text{%Output}} \]

\[ T_d = 2 \text{ seconds} \]

\[ \tau = 5 \text{ seconds} \]

Assuming the desired lambda were 15 seconds, \( K_C \) and \( T_R \) would be calculated as follows

\[ T_R = \tau \quad 5 \text{ seconds} = 0.083 \text{ minutes} \]

\[ K_C = \frac{T_R}{K_p(\lambda + T_d)} = \frac{5}{2(15+2)} = 0.147 \frac{\text{%Output}}{\text{%Span}} \]
Transfer Functions.

First order process. No deadtime.

\[ G_p(s) = \frac{K_p}{\tau s + 1} \]

PI controller (classical Form)

\[ G_c(s) = K_c \left( \frac{TRs + 1}{TRs} \right) = K_c \left( 1 + \frac{\lambda}{TRs} \right) \]

Select \( \lambda \) and \( \lambda \)-tune the controller.

\( TR = \tau \)

\[ K_c = \frac{1}{K_p} \cdot \frac{TR}{\lambda} \]

\[ G_c(s) = \frac{\tau}{K_p \lambda} \left( \frac{\tau s + 1}{\tau s} \right) \]
closed-loop transfer function

\[ G_{sp}(s) = \frac{G_c(s)G_p(s)}{1 + G_c(s)G_p(s)} \]

\[ G_c(s) \cdot G_p(s) = \frac{\tau}{K_p \lambda \left( \frac{\tau s + 1}{\tau s} \right)} \cdot \frac{K_p}{\tau s + 1} \]

\[ = \frac{1}{\lambda s} \quad \text{(loop transfer function)} \]

Therefore

\[ G_{sp}(s) = \frac{1}{\lambda s} \cdot \frac{1}{\frac{1}{\lambda s}} \]

\[ = \frac{1}{\lambda s + 1} \]

First order with gain of \( K \)

and time constant \( \lambda \).
Choosing the Lambda value is a key decision since this determines the speed of the process response in Auto. The primary criteria in selecting Lambda must be the operating requirements for the loop. However, the requirement for robust (non-oscillatory) control performance over the operating range of the loop in the presence of process deadtime and nonlinearities effect very limit the controller speed (Lambda value).

Typically, the Lambda value should be constrained to be no faster than an appropriate multiple of the open loop time constant or process deadtime for the loop. The appropriate multiple should reflect the following factors:

- confidence in the process dynamics
- process deadtime
- valve nonlinearities

For example, if we have a limited knowledge of the process dynamics (due perhaps to valve nonlinearities), it would be appropriate to use a relatively high Lambda to τ multiple. Conversely, if we were very confident in the process dynamics and there was a small amount of deadtime a relatively small Lambda to τ ratio would be appropriate.

Since process deadtime tends to destabilize the controller, this must also be considered in the selection of Lambda. It is appropriate to limit the speed of the controller to a multiple of the process deadtime.

The general guidelines that should be used in selecting Lambda are described below.

**VERY AGGRESSIVE TUNING**

\[ \text{Lambda} \leq \frac{1}{2} \tau \]

When tuning for a value of Lambda within this range (or faster), the controller becomes very sensitive to miscalculations or changes in Kp, Td, and τ. The end result might be an oscillatory response and high risk of controller instability.
The following example shows the process and controller responses with lambda set equal to τ. The process dynamics are as follows,

\[ K_p = 2 \% \text{ Span/\% Output} \]
\[ T_d = 0 \text{ seconds} \]

**Figure 7-5 Lambda Tuned Controller - Lambda = \( \tau \)**

In this case, the controller output steps to its final position at the setpoint change. Thus the closed loop controller response is identical to the open loop bump - which is the only means whereby the closed loop time constant (\( \text{Lambda} \)) can be identical to the open loop time constant (\( \tau \)).

The proportional and integral terms are also shown. Note that the proportional 'kick' immediately brings the output to its final position. As the process responds, the decrease in the proportional term is exactly offset by the increase in the integral term.
If the process dynamics are not well known, choosing an aggressive Lambda value is likely to result in an oscillatory response. For example, let's assume that the process deadtime in the above example increases from 0 seconds to 3 seconds, perhaps because of wear in the control valve. The results are shown below.

**LAMBDA TUNING**

Lambda Tuning for Lambda = Tau
Assumed Td = 0 seconds  Actual Td = 3 seconds

![Graph showing Lambda tuning with 3-second deadtime](image)

Figure 7-6 Lambda = \( \tau \) with 3-second deadtime

Because the PV does not change over the deadtime period, the controller output continues to increase and overshoots the final steady state position. This causes the PV to overshoot the setpoint.
Let us further assume that the process gain changes from 2 \% Span / \% Output to 4 \% Span / \% Output perhaps due to an operating regime change. The process response to a setpoint bump is as follows.

**LAMBDA TUNING**, \( \text{Lambda} = \text{Tau}, \text{Td} = 3 \text{ Seconds}, \text{Kp} = 4\%, \text{Span}/\%\text{Out} \)

*Note Tuning based on Kp = 2\% Span/\%Out, Td = 0 Seconds*

---

**Figure 7-7** \( \text{Lambda} = \tau \) 3 seconds deadtime, Kp Increased

Under these conditions, the response to a setpoint bump is very oscillatory.
It is generally recommended that the Lambda value be equal to or slower than $3\times\tau$. Though the closed loop response is relatively slow compared with the open loop response, the risk of controller instability and oscillatory responses is very low. The controller is forgiving even if the process dynamics $K_p$, $T_d$ and $\tau$ are significantly different from those assumed.

The following examples illustrate this point. In the figure immediately below the response achieved by setting Lambda equal to $3\times\tau$ is shown. The process dynamics are the same as in the examples above. If $K_p = 2 \% \text{ Span}/\% \text{ Output}$, $T_d = 0 \text{ seconds}$, Note that the proportiona 'kick' achieves one third of the total controller output change.

Figure 7-8: Lambda = $3\times\tau$

The process dynamics are changed as shown in the example above. The dead time is increased to 3 seconds and the process gain is increased to 4 \% Span/\% Output. When Lambda is equal to $3\times\tau$, the tuning is more forgiving to changing process dynamics as shown below.
LAMBDA TUNING, \( \text{LAMBDA} = 3^*\text{Tau} \ Td - 3 \ \text{SECONDS} \ Kp \ 4\% \text{Span}/\%\text{Output} \\
(\text{TUNING BASED ON} \ Kp = 2\% \text{Span/Output} \ Td = 0 \ \text{SECONDS}) \\
\begin{array}{c}
\text{SP} \\
\text{PV} \\
\text{OUTPUT} \\
\text{INTEGRAL} \\
\text{PROPORTIONAL} \\
\end{array}
\begin{array}{c}
% \\
\text{TIME} \\
\end{array}

Figure 7-9 LAMBDA = 3 * \tau, Kp, Deadtime Dynamics Altered

In summary though it might be possible to go faster,

\text{LAMBDA} = 3 * \tau \text{ or Slower IS A SAFE STARTING POINT}
DEADTIME LIMITATION

The performance of a control loop is very strongly dependent on the process deadtime. It is recommended that Lambda is chosen to be at least 3 times higher than the deadtime. Thus, if the deadtime is equal to or greater than $\tau$, the deadtime constraint should control the selection of Lambda. The objective of this constraint is to extend the range of applicability for the Lambda tuning rules presented above.

LAMBDA = \(3 \times T_d\) OR SLOWER IS A SAFE STARTING POINT
As was mentioned in Lecture 1, excessive use of filtering will result in masking the true dynamics. This filtering, which can be applied at several locations in the control loop, is often referred to in the literature as dampening or snubbing. The following rules should be used when deciding how much filtering is acceptable:

a) When performing a BUMP TEST, the filtering should be kept to the minimum. The measured process response to the controller output bump should reflect the true process response and not the upstream ‘filtering’. The total filter time constant (which determines how much dampening there is) should be at least 2 to 5 times faster than the actual PROCESS time constant:

\[ \text{T}_{\text{FILTER}} = \frac{T_{\text{u}}}{2 \text{ to } 5} \]

b) The FILTER time constant \( T_{\text{FILTER}} \) should be 5 to 10 times faster than the LAMBDA value chosen for the control loop. Excessive dampening will destabilize the loop and lead to cycling:

\[ \text{T}_{\text{FILTER}} = \frac{\text{LAMBDA}}{5 \text{ to } 10} \]

Consider the following example. The controller has been tuned with a Lambda value of 10 seconds, which is equal to \( 1 \times T \). Following return of a 10 second PV filter, a signal at the controller. The unfiltered and filtered responses to a setpoint bump are compared in the figure below.
PI CONTROLLER - 10 SEC PV FILTER VERSUS UNFILTERED

Figure 7-10 Oscillatory Setpoint Response due to excessive filtering

In the example above, the controller does not see the true process response and winds up causing the process to cycle.

From the point of view of simplicity, it is best to apply filtering at only one location. Since a precise filter time constant can be entered in a digital controller, this is usually considered the best choice. However, it is important to note that the process must be filtered at each point of sampling in a digital control system in order to avoid signal aliasing. Anti-aliasing filters will be discussed later.
EXAMPLE 7-1
The time series plots shown below were taken from a stock consistency loop in a fine paper mill. The process dynamics and original tuning parameters for this loop were as follows:

**Process Dynamics**
- \( K_p = 3.0 \% \text{ Span} / \% \text{ Output} \)
- \( \tau = 6.6 \text{ seconds} \)
- \( T_d = 13.8 \text{ seconds} \)

**INITIAL TUNING**
- \( K_c = 3.7 \% \text{ Output} / \% \text{ Span} \)
- \( T_R = 45 \text{ seconds} \)
- \( T_D = 30 \text{ seconds} \) (Derivative)
- \( T_F \text{ Iter} = 120 \text{ seconds} \)

In the past, the operations group had complained about the amount of noise in the consistency signal and also to the slow response to setpoint bumps.

**Questions:**
1. Calculate the Lambda tuning constants for this loop. How do the Lambda tuning constants compare with the existing tuning constants?
2. What might be causing the strong consistency cycle?
3. Why is a PV filter of 120 seconds used?

**Tuning Checklist**
1. What is an appropriate starting point for Lambda:
   - \( 3 \times T_d = \) ________ seconds
   - \( 3 \times \tau = \) ________ seconds
2. If Lambda is set equal to ________ seconds
   - the total time to reach the new setpoint is ________ seconds
   - the total filter time in the loop should be less than ________ seconds
EXAMPLE 7-2

A series of setpoint bumps were performed on the Blend Chest consistency controller (span 2 - 4 % consistency) shown below. The first bump shows the 'Tune by Feel' response. The following setpoint bumps show the Lambda tuned response with increasingly faster Lambda values.

Questions:

1/ Compare the 5 responses? Which is fastest? Which is best?
2/ Estimate the process gain of this controller. Is this process gain reasonable?
EXAMPLE 7-3

Retention Aid Flow - Variable Speed Drive

Retention aid is added to the machine whitewater in order to increase first pass fiber retention on the wire. The following time series plot shows an open loop bump and a closed loop setpoint bump.

Questions:
1/ Calculate Lambda tuning constants for Lambda = 3 * τ
2/ Estimate the actual lambda value, based on the closed loop setpoint bump
3/ Is it reasonable to tune this loop more aggressively than Lambda=3 * τ.
EXAMPLE 7-4

The following figure shows successive setpoint bumps on a Paper Machine Headbox Total Head controller. The first is the Tune by Fee response, and the following are Lambda tuned responses.

Questions:

Discuss the relative merits of the Tune by Fee response versus the Lambda tuned response.
APPENDIX 7A
DEVELOPMENT OF LAMBDA TUNING RULE
FOR FIRST ORDER PROCESSES
Consider the standard control loop configuration

\[ Y_{SP} \quad E \quad \text{CONTROLLER} \quad G_C(s) \quad U \quad \text{PROCESS} \quad G_p(s) \quad Y \]

The closed loop transfer function (as always) is given by

\[ G_{SP}(s) = \frac{G_C(s)G_p(s)}{1 + G_C(s)G_p(s)} \]

regardless of \( G_C(s) \).

Consider designing an arbitrary controller such that any particular desired closed loop transfer function \( G_{SP}(s) \) can be achieved. Specifying the closed loop transfer function \( G_{SP}(s) \) will require the controller transfer function to be (by algebraic manipulation)

\[ G_C(s) = \frac{G_L(s)}{G_p(s)} \frac{G_M(s)}{1 + G_M(s)G_{SP}(s)} \]

Suppose that we want the closed loop system to respond as a first order process to a setpoint change. Thus \( G_{SP}(s) \) is chosen to be

\[ G_{SP}(s) = \frac{1}{\lambda \cdot s + 1} \]

where \( \lambda \) (lambda) is the desired closed loop time constant. Note that the gain is 1. This is to ensure that the change in the output will match the setpoint change in steady state while giving the required dynamics.

To obtain a closed loop system which is first order with time constant \( \lambda \), will require
\[ G_C(s) = \frac{1}{G_p(s)} \cdot \frac{G_{SP}(s)}{1 - G_{SP}(s)} \]

\[ = \frac{1}{G_p(s)} \cdot \frac{1}{\lambda s + 1} \]

Note that the controller consists of an inverse of the process times \(1/\lambda s\). Hence all process dynamics are cancelled out, leaving an integrator with a gain of \(1/\lambda\).

Suppose that the process to be controlled is adequately described by first order dynamics with no deadtime, i.e.

\[ G_p(s) = \frac{K_p}{\tau s + 1} \]

Now using the first order transfer function for \(G_p(s)\) results in

\[ G_C(s) = \frac{1}{G_p(s)} \cdot \frac{1}{\lambda s} \]

\[ = \frac{1}{\tau s + 1} \cdot \frac{1}{\lambda s} \]

\[ - \frac{\tau}{K_p \lambda} + \left( \frac{1}{K_p \lambda} \right) s \]

\[ - \frac{\tau}{K_p \lambda} \left[ 1 + \frac{1}{\tau s} \right] \]

For a first order process, the designed controller has the form of a PI controller. Compare the above controller transfer function with the transfer function of a PI controller in standard form.

\[ G_C(s) = K_C \left[ 1 + \frac{1}{T_R s} \right] \]

Hence, to tune a PI controller in standard form for a 1st order process set...
Note that the choice of the reset time is determined strictly from the open loop time constant. Once set, it need not be changed. The speed of response is determined strictly from the controller gain.

This method is called 'pole placement' because it tries to manipulate the location of the closed loop poles. Remember that the PI controller has one pole at the origin (always) and one zero whose location is determined by the reset time.

A first order process has one pole. As a result, the closed loop system has two poles and one zero. What Lambda Tuning does is shift the zero to lie directly on top of the pole generated by the process, thereby effectively cancelling both the pole and zero. We are now left with a single closed loop pole whose location we are free to choose by varying the proportional gain (see Root Locus example).

This method is meant primarily for processes that can be reasonably approximated as first order.

\[
T_r, \tau \\
K_c = \frac{1}{K_p} \left( \frac{\tau}{\lambda} \right) \\
\text{(reset time)} \\
\frac{1}{K_p} \left( \frac{I_r}{\lambda} \right) \quad \text{(controller gain)}
\]
8. NONLINEARITY IN CONTROL LOOPS
NON-LINEARITIES RESULTING FROM THE CONTROL STRATEGY

EXAMPLES

Split Ranging Strategy
A split ranged controller has one output but more than one manually variable. This strategy is most commonly used in order to extend the operating range of the process system. Common examples from the Pulp and Paper industry are Repulper Level control, Dryer Pressure and Differential control, Air Padded Headbox level control, Steam line pressure control and others.

An example of split ranging is shown in Figure 8-1. The Headbox controller output is split ranged, opening the supply valve from 0 to 50%, and closing the Bleed valve from 50 to 100%. Thus, the Supply valve and Bleed valve are fully open at a Level controller output of 50% and the Bleed valve fully closed at an output of 100%.

This control strategy permits a wide range of operation. However, because the controller manipulates two different control valves, the range in the process dynamics can be expected to be relatively high. Control performance will suffer as a result. It is particularly important to understand the process requirements of the loop in order to make sound tuning decisions.

Figure 8-1 Air Padded Headbox Using Split Ranged strategy to control Level
Low / High Select Strategy

The low / high select strategy configuration perm ts a manipulated variable to be shared by two or more controllers. This strategy is typically employed to extend the control range and/or to a low safe operation of a process system.

On a paper machine, a Dryer section which recirculates blowthrough steam with a Thermocompressor (TC) most always uses a Low-se lect strategy. The TC is available to both the Differential controller and the Pressure controller via the Low Select strategy. Normally, the Pressure controller adjusts the Makeup Steam valve and the Differential controller adjusts the TC. There are occasions when this will not be true. If the condensing rate is lower (as on sheet breaks or a low demand grade) than the Motive Steam flow at a given differential target, then the Pressure controller output will decrease until it assumes control of the TC. At this point, the Differential controller must adjust the vent valve to achieve the differential target.

The split ranging / ow select strategy is required from a process point of view since it provides the means for extending the Pressure and Differential control range. However, from a control perspective, the strategy carries with it the potential for control deadbands if both controller outputs fail below 50% and nonlinear response to setpoint changes or load disturbances if the pressure controller output decreases below 50%. The process dynamics will be substantially affected when the final control element(s) changes.

1140 kPa (150 PS G) STEAM HEADER

REMOTE SP FROM MO STURE CONTROLLE

620 kPa (75 PS G) STEAM HEADER

MAKEUP

6 m MAKEUP VALVE
9-15 PS A/O

D FFERENT AL OR
BLOWTHRU
CONTROL

DIFFERENT AL OR
BLOWTHRU
CONTROL

3 N VENT VALVE
9-15 PS A/O

OR F CE
PLATE

STEAM
SEPARATOR

Figure 8.2 Dryer Section using Low Select Strategy
Other Control Strategy Examples From Pulp And Paper Which Introduce Nonlinearity

Stock Chest Blending Control Configuration

1/ Dry Stock Blending strategy implementation results in Level process gain being dependent on stock consistency (i.e. Blend Chest controller output scaled to Dry Tons target rather than a Wet flow target)
2/ Broke addition strategy effects the Blend Chest level controller process gain (i.e. Broke is ratioed to virgin fibre (rather than directly based on Level controller demand) making the Level process gain dependent on the Broke ratio)

Logic Controllers

Refriner Control
1/ Refiner plates move at fast speed or slow speed depending on the magnitude of the error
2/ Hysteresis deadband employed for control of Refiner plate position
Nonlinearities Resulting From The Control Algorithm

Deadbands / Error Characterization
A deadband is a function that treats small signals as zero. Due to discrete representations of continuous numbers, computers effectively implement deadbands. This effect is usually so small as to be negligible. A deadband is non-linear because the ratio of input to output is not constant over the signal range.

Although all deadbands are non-linear, there are two types which are usually referred to as 'linear' and 'non-linear'.

![Linear and Non Linear Deadbands](image)

Figure 8-3 Linear and Non Linear Deadbands

Perhaps the worst effect that can be caused by a deadband is a limit cycle.

Deadbands are a non-linearity which we induce. Because they are non-linear and can result in limit cycles, they should preferably not be used.

A deadband in an inner cascade loop can either cause a limit cycle in the inner loop or a slower limit cycle in the outer loop.

A Special Case

Many presently available actuators are velocity mode devices. They are usually driven by stepper motors and have the advantage that they are fairly safe regarding their input from the controller. Output to such devices is usually in the form of a pulse train or time duration pulse which will open or close the valve.
Example

**Fig 8-4 Velocity Mode Actuator**

A special form of stiction may arise with this type of actuator. When the duration of the pulse is extremely short, relays may not have time to (de)energize and/or the motor may not have time to start in motion, with the result that the pulse has no effect on the process output.

In this instance, a deadband may be used to prevent unnecessary wear on the elements. This deadband should be applied to the output and not the error or a limit cycle may result. It should also be linear.

Also, compensation for stiction could be applied. The compensation should be made equal to the minimum required actuation time.
Error Characterizers

Fig 8.5 Error Characterization
Strategies for Dealing with Non-Linearities

1/ Fix them
This strategy is particularly applicable to control valve design and maintenance. The single biggest sources of short term variability are control valve backlash and stiction. These can be minimized first by specifying the dynamic performance requirement prior to the valve purchase and second by employing a pro-active maintenance approach.

2/ Hide Them in An Inner Loop
Cascade control refers to a configuration where the controller output of one loop (outer or master controller) becomes the setpoint of another controller (inner or slave controller). This is shown in block diagram format below.

![Block Diagram of Cascaded Loops](image_url)

Fig 8-6 Block Diagram of Cascaded Loops

There are several motivations for designing a cascade configuration. First, the inner flow control loop will take control action to attenuate an inner process disturbance before it has an effect on the outer process. Secondly, installing an inner control loop will linearize (to some degree) the dynamics of the outer control loop. This is because a unit step in the outer controller output will result in a constant change to the inner control loop setpoint. The nonlinearity in the control valve will be effectively passed on to the inner controller.
In the example below, the Brightness controller cascades a setpoint to the \( CL_2 \) flow controller. Therefore, the steady state change in the \( CL_2 \) flow for a unit step in the Brightness controller output will be constant from 0 to 100% of the Level controller output. The nonlinearities resulting from the control valve flow characteristic and/or backlash and stiction will be largely eliminated.

Fig 8.7 Brightness Control
3/ Tune for the ‘worst’ case.
When the non-linearities need to be considered, but non-linear control is not
going to be applied then the loop must be appropriately tuned. If the main
consideration is ensuring stability over the entire operating range, then the loop
tuning should be based on the most ‘destabilizing’ conditions.
The main ‘destabilizing’ conditions are:
- Highest Process Gain
- Longest Process Deadtime
- Longest Process Time Constant
- Highest Resonant Peak

**Example** Consistency Control
Max \( K_p \) at Min \( F \)
Max \( T_d \) at Min \( F \)
Max \( \tau \) at Min \( F \)
⇒ Tune loop at Minimum flow

**Example** pH Control
Max \( K_p \) Centre’ pH
⇒ Tune loop at ‘Centre’ pH

**Example** Moisture Control
Max \( K_p \) at Min \( P \)
⇒ Tune loop at Minimum pressure

Another possibility which may be available on certain distributed control systems
is ‘Grade’ Dependent Tuning. This would allow for different tuning on different
grades or at different operating points. This alternative may be considered as a
subset of non-linear control.
Defining The Process Model From Bump Test Results

The process dynamics results vary due to a wide variety of nonlinearties such as valve flow characteristics, backlash/stiction, the size of the bump and others. These nonlinearties often result in a significant spread in the dynamic response. Performing a series of bump tests will assist us in understanding the nature and degree of the nonlinearities. The bump test results must be interpreted with respect to a tuning strategy to develop the 'best' controller tuning.

Some typical results from a bump test series are shown below:

<table>
<thead>
<tr>
<th>$K_P$</th>
<th>$T_d$</th>
<th>$\tau$</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.8</td>
<td>30</td>
<td>20</td>
</tr>
<tr>
<td>0.5</td>
<td>25</td>
<td>30</td>
</tr>
<tr>
<td>1.3</td>
<td>40</td>
<td>20</td>
</tr>
<tr>
<td>1.5</td>
<td>30</td>
<td>40</td>
</tr>
<tr>
<td>1.1</td>
<td>20</td>
<td>40</td>
</tr>
<tr>
<td>0.9</td>
<td>30</td>
<td>30</td>
</tr>
</tbody>
</table>

Using the most conservative process dynamics will ensure closed loop stability under all operating conditions. The concession which accompanies this tuning strategy is sluggish control performance under some operating conditions. Adaptive tuning may be required to ensure adequate control performance under all conditions.

Consequences of $K_P$ Changing

$$\lambda = \frac{T_r}{K_cK_p}$$

$K_C$ calculated based on high $K_p$.

As the process gain gets smaller, $\lambda$ will get longer and the loop will become more sluggish. Since $K_C$ was probably calculated based on a high value of $K_p$, $\lambda$ will vary from the value chosen to slower values. Should the process gain increase above that used to calculate $K_C$, the loop will speed up. This may cause the fast dynamics to oscillate.
**4/ Design a Non-linear Controller**

This technique attempts to account for non-linear characteristics and then base the control on a 'central' linear controller. The concept is best explained through examples.

**Example: Moisture Control**

![Moisture Control Diagram](image)

*Fig 8-8 Non-Linear Moisture controller*

The controller determines output in terms of temperature which is converted to pressure. The controller views the process as the dashed box which is linear.

**Example: pH Control**

![pH Control Diagram](image)

*Fig 8-9 Non-L near pH controller*

The controller sees a modified error which is the actual error multiplied by a factor which is dependent upon the actual pH. This factor is high at low and high pH and low in the middle range. Ideally, this factor is inversely related to the actual process gain.

The error, $e^*$, is, thus, non-linearly related to the actual error, but the controller can be tuned as if the titration curve was a straight line (i.e. a linear system).
Other examples of applications of non-linear control are:

- vary estimated $T_d$ as function of machine speed for weight control with Dahl n or Smith
- characterize a valve for flow control. Then compensate proposed $U$ for valve behavior
- vary $K_p$ & $T_R$ according to the estimated $\tau$ on a Lambda Tuned PI controller.

The key is to determine how the non-linearity affects the system and then devise a scheme to counteract or compensate for the effect.

5/ Use an Adaptive Controller
Adaptive control is not covered in this course.
9. SETPOINT RESPONSE AND LOAD RESPONSE
Controller Performance

There are several requirements of a typical controller, some of which are listed below

1) Steady state tracking accuracy (zero offset)

2) Setpoint response (also called servo response)
   Transient Accuracy
   - rise time
   - overshoot
   - settling time

3) Load response (also called regulator response)
   Disturbance Reject on
   - steady state
   - transient

4) Required control effort
   - maximum magnitude of controller output
   - variance of controller output
   - none or very little output saturation

5) Robustness
   - insensitivity to process parameter changes

6) Stability

In general, it is unlikely that any given controller will be able to achieve all of these objectives. A primary concern is that the system be stable in closed loop. All too often a controller is considered to be adequately tuned if it is stable. The other objectives need to be considered also.

Sensitivity is usually concerned with providing a certain margin of stability. This means that for small changes in the process parameters, the system will remain stable. Sensitivity should also be studied for its effect on the other objectives of performance.
Tuning for Setpoint Response and Regulation

A control loop can perform two basic functions:

- Respond to setpoint changes
- Regulate out the effects of process disturbances

SETPOINT

\[ \text{DISTURBANCE (LOAD)} \]

\[ Y(s) \]

\[ Y_{sp}(s) \rightarrow G_c(s) \rightarrow U(s) \rightarrow G_p(s) \rightarrow Y(s) \]

Fig 9-1 Block diagram of a feedback control loop

Note that there are two external inputs to the loop: the setpoint and the disturbance. The loop responses to these two inputs are fundamentally different.

**Setpoint Response**

The loop is intended to track the setpoint. The loop tuning will determine how and when the process will settle once the new setpoint is reached.

- \( \lambda_1 = \tau \)
- \( \lambda_2 = 3 \tau \)
- \( \lambda_3 = 10 \tau \)

Fig 9-2 Setpoint responses for various controller tunings

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The transfer function from the setpoint $Y_{Sp}$ to the PV $Y$ is

$$G_{SP}(s) = \frac{G_C(s) \cdot G_P(s)}{1 + G_C(s) \cdot G_P(s)}$$

Example

For a 1st order process with a PI controller that has been λ tuned

$$G_{SP}(s) = \frac{1}{\lambda s + 1}$$

The gain is 1 0 because the process is expected to be identically equal to the setpoint in steady state. If the setpoint is moved more and more rapidly, however, the process will track less and less well. This can be shown best by the frequency response.

Fig 9-3  Setpoint tracking Bode plots for various controller tunings

Note

The period at which frequency response starts to decay (cut-off frequency cut-off period) is given by the following formula as

$$f_c = \frac{1}{2\pi \lambda}$$

$$T_c = \frac{2\pi \lambda}{\lambda}$$

Note that Ziegler Nichols and other oscillatory tuning will amplify certain frequencies (resonant frequency).
CASCADED LOOPS

In a cascaded loop situation, the outer loop (or master loop) updates the setpoint of an inner loop (or slave loop). The inner loop then adjusts its process to achieve the new setpoint. This in turn affects the outer loop's process dynamics.

Figure 9-4  Block Diagram of Cascaded Loops

In the pulp and paper industry, there are many examples of cascaded loops. In the paper machine area, for instance, the sheet moisture controller (OUTER LOOP) cascades a setpoint to the steam pressure controller (INNER LOOP).

It is important to tune the INNER controller so that it is fast enough to reasonably track the setpoint from the OUTER loop. A good rule of thumb is that the OUTER loop should be tuned at least 5 times SLOWER than the INNER loop. This means

\[ \lambda_{\text{OUTER}} > 5 \times \lambda_{\text{INNER}} \]

If this rule is not kept, then both the Inner and Outer processes may tend to oscillate.
BRIGHTNESS CONTROL EXAMPLE

![Diagram of Brightness Control System]

Figure 9.5 Typical C Stage Brightness Control

The dynamics of the Brightness and CL₂ loops are as follows:

<table>
<thead>
<tr>
<th>Brightness</th>
<th>CL₂ Flow</th>
</tr>
</thead>
<tbody>
<tr>
<td>Kp 1 %Span/%Out</td>
<td>1 %Span/%Out</td>
</tr>
<tr>
<td>Tau 10 sec</td>
<td>5 sec</td>
</tr>
<tr>
<td>Td 10 sec</td>
<td>0 sec</td>
</tr>
</tbody>
</table>

If the CL₂ flow controller is tuned with a Lambda value of 15 seconds, the Brightness controller should be tuned with a Lambda value of 75 seconds or slower.

\[
\text{LAMBDA}_\text{FLOW} = 15 \text{ SECONDS}
\]

\[
\text{LAMBDA}_\text{BRIGHTNESS} = 75 \text{ SECONDS OR SLOWER}
\]

The brightness example described above is illustrated in the following figures.
In the first figure, the Inner flow loop is tuned with a Lambda value of 15 seconds and the Brightness controller is tuned with a Lambda value of 75 seconds.

**CASCADE LOOPS - INNER LOOP 5 TIMES FASTER**

![Graph](image)

**Fig 9.6** Brightness, Flow Responses to a Brightness SP Bump

In the second figure, the Lambda value of the Outer loop are both equal to 75 seconds. The Inner flow loop is unable to effectively track the Brightness controller setpoint, and the response is oscillatory.

**CASCADE LOOPS - INNER & OUTER LAMBDA = 75 SECONDS**

![Graph](image)

**Fig 9.7** Brightness, Flow Responses to a Brightness SP Bump
Regulation or Load Response

The loop is intended to maintain the process variable on setpoint. When a disturbance or load change occurs, the loop will bring the process variable back to setpoint. The loop tuning will determine how and when the process will settle out.

\[
\lambda_1 = \tau \\
\lambda_2 = 3\tau \\
\lambda_3 = 10\tau
\]

Quarter Amplitude Tuning

Fig 9-8 Load recovery for various controller tunings

The transfer function from the disturbance \( D \) to the PV \( Y \) is

\[
G_{LD}(s) = \frac{1}{1 + G_C(s) \cdot G_P(s)}
\]
Example For a 1st order process with a PI controller that has been $\lambda$ tuned

$$G_{LD}(s) = \frac{\lambda s}{\lambda s + 1}$$

Since there is an 's' or differentiation in the numerator, the steady state gain from $D$ to $Y$ is zero. This is so because no matter what $D$ does, $Y$ comes back to the setpoint value. However, as $D$ changes more and more rapidly, the loop will be less and less able to bring the process back to setpoint in time. Hence, for very rapid disturbances no effective correction will be done and the gain will be unity. The frequency response for the regulator is shown in Fig. 9-9.

![Regulator Bode plots for various controller tunings](image)

**Fig. 9-9** Regulator Bode plots for various controller tunings

**Note** The lower the frequency of the disturbance, the more effective $Y$ will the loop regulate or reduce the amplitude of the fluctuations in $D$. Attenuation starts at the cut off frequency

$$f_c = \frac{1}{2\pi\lambda}$$

$$T_c = \frac{1}{2\pi\lambda}$$

"EnTech"
As the disturbance frequency is reduced, attenuation will occur at 1 decade/decade (-20 dB/decade). This corresponds to:

<table>
<thead>
<tr>
<th>Disturbance Period</th>
<th>(1000T_c)</th>
<th>(100T_c)</th>
<th>(10T_c)</th>
<th>(T_c)</th>
<th>(0.1T_c)</th>
<th>(0.01T_c)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Variability Remaining</td>
<td>0.1%</td>
<td>1%</td>
<td>10%</td>
<td>70%</td>
<td>100%</td>
<td>100%</td>
</tr>
</tbody>
</table>

Note that Ziegler-Nichols and other oscillatory tuning will amplify variability at certain frequencies.

Example: Flow loop \(T_d = 0\) sec, \(\tau = 3.0\) sec, \(\lambda = 9.5\) sec

Cut-off period \(T_c - 2\pi\lambda \approx 59.7\) sec, \(\equiv 1.0\) min

<table>
<thead>
<tr>
<th>Disturbance Period</th>
<th>100 min</th>
<th>10 min</th>
<th>1 min</th>
<th>6 sec</th>
</tr>
</thead>
<tbody>
<tr>
<td>Variability Remaining</td>
<td>10%</td>
<td>10%</td>
<td>70%</td>
<td>100%</td>
</tr>
</tbody>
</table>
Trade Offs in Choosing Lambda

Fig 9.10 Setpoint tracking and regulator Bode plots for a Lambda tuned first order system.

- cut-off frequency \( f_c = \frac{1}{2\pi\lambda} \)
- cut-off period \( T_C = \frac{2\pi}{\lambda} \)

Trade offs in choosing \( \lambda \):

\( \lambda \) smaller
- Higher bandwidth
- Faster setpoint response
- Better disturbance rejection
- Less robust

\( \lambda \) larger
- Smaller bandwidth
- Slower setpoint response
- Less disturbance rejection
- More robust
The Effect of Process Deadtime

The presence of process deadtime makes the setpoint and load frequency responses more complicated.

Typical regulator and setpoint tracking frequency responses for a Lambda tuned PI controller controlling a first order with deadtime process are shown in the following figure.

![Graph showing Bode plots with deadtime](image)

Fig 9.11 Setpoint tracking and regulator Bode plots with deadtime present

The frequency responses over the low frequency range up to the cutoff frequency are similar to the case of no deadtime, with a low frequency asymptote having a slope of 1 decade per decade (+20 dB/decade). However, due to the presence of deadtime, the regulator frequency response clearly exhibits resonance (amplification of disturbances over a frequency range) and the bandwidth of the controller is decreased. The cutoff of the control loop in the case of process deadtime is given by

\[
f_C = \frac{1}{2\pi(\lambda + T_d)}
\]

\[
T_C = \frac{2\pi(\lambda + T_d)}
\]

Note that the cutoff period depends on both Lambda and the process deadtime.
Regulator Performance in the Presence of Process Deadtime

The regulatory frequency response for a first order process with 5 seconds deadtime and Lambda values of 5, 8, 10 and 15 s shown in the following figure.

![Graph showing AR vs Frequency for different Lambda values]

Fig 9.12  Regulator Bode plots for various values of Lambda

Due to the presence of resonance the meaning of cutoff should be clarified. The cutoff frequency is defined as the frequency at which the low frequency asymptote intersects the line of unity amplitude ratio namely AR 1.

Note that higher disturbance attenuation in the low frequency range can only be obtained at the expense of higher resonance over a range of frequencies faster than the cutoff.

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The dependency of the resonant peak $M_p$ on the Lambda to deadtime ratio $\lambda/T_d$ is shown in the following figure.

![Graph](image)

Fig 9-13 Regulator Resonance for a First Order with Deadtime Process

The dependency of the resonant period to deadtime ratio $T_p/T_d$ on the Lambda to deadtime ratio $\lambda/T_d$ is shown in the following figure.

![Graph](image)

Fig 9-14 Regulator Resonant Period for a First Order with Deadtime Process

From the above graph the resonant period can be very roughly approximated by

$$T_p \approx 6 \times T_d$$
Lambda Selection and Control Loop Resonance

In order to make the controller robust, \( \lambda \) must not be selected too small relative to the process deadtime \( T_d \). A safe (but conservative) suggested requirement is to choose

\[
\lambda > 3 \times T_d
\]

As the ratio of deadtime to lambda decreases, the controller will exhibit more control loop resonance, resulting in increased amplification of disturbances in a frequency range faster than the cutoff. The following table gives the lower limit for choosing \( \lambda \) relative to the deadtime for a PI controller controlling a first order with deadtime process in order to cap the resonance at various values:

<table>
<thead>
<tr>
<th>Resonant Peak AR = ( \frac{A_{\text{AUTO}}}{A_{\text{MAN}}} )</th>
<th>Resonant Peak in dB</th>
<th>Lower limit for ( \lambda ) to cap the Resonant Peak</th>
</tr>
</thead>
<tbody>
<tr>
<td>1.26</td>
<td>20</td>
<td>( 3 \times T_d )</td>
</tr>
<tr>
<td>1.35</td>
<td>26</td>
<td>( 2 \times T_d )</td>
</tr>
<tr>
<td>1.41</td>
<td>30</td>
<td>( 16 \times T_d )</td>
</tr>
<tr>
<td>1.59</td>
<td>40</td>
<td>( T_d )</td>
</tr>
</tbody>
</table>

An increased certainty in the model parameters may justify faster tuning, however faster tuning will increase control loop resonance.
Limits to Control Loop Performance

The process dynamics present in the control loop ultimately limit the best control loop performance that can be achieved. The selection of Lambda must take into account the process deadtime \( T_d \) and process time constant \( \tau \) according to both of the following constraints

\[
\lambda > 3 \times \tau \\
\lambda > 3 \times T_d
\]

Once Lambda has been chosen the cutoff period is then fixed by the equation

\[ T_C = 2\pi(\lambda + T_d) \]

Example  Consider a stock consistency control loop with a process time constant of 25 seconds and various process deadtimes as shown in the following table

<table>
<thead>
<tr>
<th>Process Deadtime ( T_d ) (sec)</th>
<th>Closed-Loop TC ( \lambda ) (sec)</th>
<th>Cutoff Period ( T_C ) (sec)</th>
<th>Cutoff Period ( T_C ) (min)</th>
</tr>
</thead>
<tbody>
<tr>
<td>5</td>
<td>15</td>
<td>126</td>
<td>2.1</td>
</tr>
<tr>
<td>10</td>
<td>30</td>
<td>251</td>
<td>4.2</td>
</tr>
<tr>
<td>15</td>
<td>45</td>
<td>377</td>
<td>6.3</td>
</tr>
<tr>
<td>20</td>
<td>60</td>
<td>502</td>
<td>8.4</td>
</tr>
<tr>
<td>25</td>
<td>75</td>
<td>628</td>
<td>10.5</td>
</tr>
<tr>
<td>30</td>
<td>90</td>
<td>754</td>
<td>12.6</td>
</tr>
</tbody>
</table>

There is much benefit from keeping the process deadtime in a stock consistency control loop to a minimum by locating the sensor as close as practically possible to the dilution point.

In general all control loops should be designed to keep process deadtime to a minimum since it degrades control loop performance.
The Effect of Process Dynamics on Regulation

- \( \tau \) process time constant
- \( T_d \) process dead time
- \( \lambda \) closed loop time constant
- \( T_c \) cut off period for disturbance rejection

\[ \text{log AR} \]

\[ \text{og f} \]

<table>
<thead>
<tr>
<th>Process</th>
<th>( \tau )</th>
<th>( T_d )</th>
<th>( \lambda )</th>
<th>( T_c )</th>
</tr>
</thead>
<tbody>
<tr>
<td>HB Total Head</td>
<td>1 sec</td>
<td>0</td>
<td>3 sec</td>
<td>20 sec</td>
</tr>
<tr>
<td>Stock Flow</td>
<td>3 sec</td>
<td>0</td>
<td>10 sec</td>
<td>1 min</td>
</tr>
<tr>
<td>Stock Consistency</td>
<td>3 sec</td>
<td>5 sec</td>
<td>15 sec</td>
<td>20 min</td>
</tr>
<tr>
<td>Pulp Brightness</td>
<td>20 sec</td>
<td>20 sec</td>
<td>1 min</td>
<td>8.4 min</td>
</tr>
<tr>
<td>PM Basis Weight</td>
<td>45 sec</td>
<td>100 sec</td>
<td>4 min</td>
<td>36 min</td>
</tr>
<tr>
<td>PM Moisture</td>
<td>2 min</td>
<td>2 min</td>
<td>6 min</td>
<td>50 min</td>
</tr>
</tbody>
</table>

Period of sine wave disturbance

\( 30 \text{ min} \), \( 3 \text{ min} \), \( 20 \text{ sec} \), \( 2 \text{ sec} \)
\( 100 \text{ min} \), \( 10 \text{ min} \), \( 10 \text{ min} \), \( 6 \text{ sec} \)
\( 200 \text{ min} \), \( 20 \text{ min} \), \( 20 \text{ min} \), \( 12 \text{ sec} \)
\( 14 \text{ hrs} \), \( 84 \text{ min} \), \( 84 \text{ min} \), \( 50 \text{ sec} \)
\( 60 \text{ hrs} \), \( 6 \text{ hrs} \), \( 36 \text{ min} \), \( 36 \text{ min} \)
\( 83 \text{ hrs} \), \( 8\text{.3 hrs} \), \( 50 \text{ min} \), \( 5 \text{ min} \)

Fig 9-15 Cutoff Periods for typical pulp and paper control loops

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APPENDIX 9A

REGULATORY CONTROL EXAMPLE
EXAMPLE:

The whitewater dilution header pressure is cycling. When the consistency controller is in manual mode (no control action), the machine chest consistency cycles from 3.0% to 3.2% consistency. The Lambda value of the consistency controller is 10 seconds. There is 5 seconds of process deadtime.

Figure 9A.1: Machine Chest Consistency Loop

The cutoff period of the consistency controller is equal to

\[ T_C = \frac{2\pi}{\lambda} \left( 1 + \frac{1}{T_d} \right) \]

- \[ 6.28 \times 15 \text{ seconds} = 94.2 \text{ seconds} \]
Case 1:

The Dlution Header cycles at a period of 942 seconds. This is 10 times slower than the Cut-Off period of the consistency controller.

FREQUENCY RESPONSE REGULATORY CONTROL

![Diagram showing frequency response and regulatory control with constants and periods marked.]

Fig. 9A.2 Frequency Response - Slow Disturbance Cycle

In this case, the controller is successful in reducing the impact of the disturbance cycle by 90%. The amplitude of the process cycle is only 0.01% consistency.

\[
AR = \frac{0.01\%\text{Consistency}}{0.1\%\text{Consistency}} = 0.1
\]
CASE 2

In this case, the period of the disturbance cycle is equal to 94.2 seconds which equals the loop's cut-off period.

FREQUENCY RESPONSE  REGULATORY CONTROL

Fig. 9A-3. Frequency Response - Disturbance Cycle at Cutoff

The controller is only capable of reducing the disturbance by only 30%. The amplitude of the consistency cycle is 0.07% consistency.

\[
AR = \frac{0.07}{0.1} = 0.7
\]
CASE 3

In this case, the period of the disturbance cycle is 9.42 seconds or 10 times faster than the Cut Off period of the loop.

**FREQUENCY RESPONSE - REGULATORY CONTROL**

![Diagram](image)

**Fig 9A-4** Frequency Response - Fast Disturbance Cycle

In this case, the disturbance passes through the controller completely unattenuated.

\[
AR \quad \frac{0.1}{0.1} = 1
\]

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10. TUNING INTERACTING CONTROL LOOPS
INTERACTIVE LOOPS

In the majority of process systems, the control actions of an individual control loop has some effect on adjacent control loops. On the paper machine, an example of interactive loops is the Total Head and Liquid Level loops in an air-baffled headbox.

Fig 10-1 Headbox Total Head and Liquid Level Process Diagram

In this example the level is controlled by adjusting the air bleed valve, and Total Head (headbox pressure) is controlled by adjusting the fan pump speed. A level controller output change will affect Total Head and a Total Head controller output change will also affect the Liquid Level. The control actions of one loop generates a disturbance which the other loop will try to attenuate. This carries the potential for the controllers to fight each other, causing both process variables to cycle.

The simplest way to solve this potential problem is to sacrifice the performance (or speed of response) of one of the loops. In the previous example, Total Head is the most important process variable because it has more impact on product quality than the Liquid Level. By tuning the Liquid Level controller much slower than the Total Head controller the Total Head controller will be capable of adequately maintaining setpoint even when disturbed by the action of the Liquid Level controller.

The power of the Lambda tuning method is that it allows us to quantify “making one loop slower than the other” by comparing the Lambda values. This also allows for the systematic development of a tuning strategy for a group of several interacting loops as we will see below.
A block diagram representing control loop interaction is shown in Figure 10.2. Notice how the controller output of one controller creates a disturbance for the other controller by driving through the coupling dynamics.

Fig 10-2 Block Diagram Illustrating Controller Interaction

The general tuning procedure is as follows:

1. Select one loop to be tuned faster (the more important one) - say $Y_1$

2. Put both loops in Manual mode and perform bump tests - in particular this gives process dynamics for loop $Y_1$

3. Tune the more important loop $Y_1$ selecting the closed loop time constant $\lambda_1$ to give good performance (typically $\lambda_1 = \max(3\tau_1, 3T_i)$)

4. Put loop $Y_1$ in Auto mode (with the above Lambda tuning) and keeping loop $Y_2$ in Manual mode perform bump tests on loop $Y_2$

5. Using the process dynamics for $Y_2$, tune $Y_2$ with a closed loop time constant $\lambda_2$ at least 5 times slower than $\lambda_1$, namely choose

$$\lambda_2 > 5 \times \lambda_1$$

In the Headbox example choose $\lambda_{LEVEL} > 5 \times \lambda_{TH}$
TUNING A SYSTEM OF INTERACTING LOOPS

Consider the Stock Refining system shown in Figure 10.3.

Fig 10-3 Stock Refining System with Interacting Loops
Process Objectives

The objective of this system is to safely produce high quality refined Kraft stock with the required strength characteristics. The control strategy involved in achieving these objectives includes the following:

1. The Kraft Surge Chest consistency (KIC002) is controlled to an operator setpoint by manually adjusting the flow of dilution whitewater to the pump suction.

2. The Refiner inlet pressure (PIC003) is controlled to setpoint by manipulating the main stock line pressure valve (0 to 50% A/O) and the recirculated flow control valve (50 to 100% A/C). During normal operation, the controller output is above 50% and is manipulated through the recirculation flow valve.

3. The stock flow through the refiner (FIC004) is controlled to a setpoint cascaded from the Refned Chest level controller. The flow control valve is positioned downstream of the refiner.

4. A specific energy controller (HPD/Ton) manipulates Refiner plate position in order to maintain the specific energy at an operator entered setpoint. The controller PV is calculated from the Refiner KW load, the Stock Flow and Stock Consistency using the following formula:

   \[
   \text{HPD/Ton} = \frac{(k \times \text{KW})}{(\text{Flow} \times \text{Cons})}
   \]

5. The Refined Stock consistency (KIC007) is controlled to setpoint by manually adjusting the dilution valve to the Refned Kraft pump suction. In this example, the dilution whitewater for both consistency oops is supplied from a common header.
Process Variable Dynamic Interactions

Clearly, there are many examples of controller interactions in this system. These are discussed below.

1. Stock Flow (FIC004) and Refiner Inlet Pressure (PIC003) Controllers

These two controllers are coupled via the Kraft Surge Chest pump. The pump discharge head will decrease with increasing pump flow according to the pump Head/Flow curve.

2. Stock Flow (FIC004), Refiner Inlet Pressure (PIC003) and Specific Energy (HPD/Ton) Controllers

These controllers are coupled via the Kraft Surge Chest pump, the Refiner Plate position versus Refiner Differential Pressure relationship and directly via the control strategy. For example, a stock flow controller output change will immediately change the calculated HPD/Ton value. Adjusting the Refiner plate position will change the Refiner differential, and therefore the available pressure drop across the flow control valve. This will also disturb the Refiner Inlet pressure. Adjusting the Pressure controller output will change the Refiner flow and therefore the specific energy value.

3. Kraft Surge Chest Consistency (KIC002), Stock Flow (FIC004) and Refiner Inlet Pressure (PIC003)

These variables are coupled via piping friction losses and flow load disturbance to consistency.

- Increasing the stock flow controller output will increase the stock flow from the Chest and increase the consistency.

- Increasing the consistency controller output will decrease the consistency and decrease frictional losses, acting to increase both the flow and pressure. It is common that the flow to consistency coupling is much stronger than the consistency to flow/pressure coupling.

- Increasing the Pressure Valve position will increase the stock flow through the pump, which will increase the consistency over the short term. In this process design, the recirculated stock flow change will reduce the chest consistency, bringing the discharge consistency back to its original value.

4. Kraft Surge Chest (KIC002) and Kraft Refined Chest (KIC007) Controllers

These are coupled via a common whitewater header. An increase in the Refined Chest consistency valve will decrease the Header Pressure, which decreases the dilution flow to the Surge Chest consistency, thereby decreasing the Surge Chest Consistency.

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Step 1 Bump Tests to Determine Coupling Dynamics

The first phase of tuning must focus on developing an understanding of the process dynamics and coupling dynamics in the system. The following tasks should be performed in the following order:

* All interactive controllers are placed into manual mode
* Bump tests are performed sequentially on interactive controllers
* Process Dynamics and Coupling Dynamics are defined. The coupling gains are particularly important since this will assist in defining the need for decoupling via tuning. The Step 1 results provide guidance regarding the relative need for decoupling, constraints on Lambda selection condition of the control equipment, etc.

Summarizing the results into a spreadsheet of the form shown below is a useful method of appreciating the nature of system interactions. In the Refining example, this would be a 4 by 4 matrix.

<table>
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</thead>
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<td>HPDT/FC004</td>
<td>KC002/FC004</td>
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<td>$K_{13}$</td>
<td>$K_{14}$</td>
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</table>
Step 2 Tuning Strategy Development

The second phase focuses on the development of the tuning strategy. The tuning strategy must support the process and manufacturing objectives. Developing the tuning strategy is often the most difficult part of the tuning process, requiring a great deal of understanding of both process dynamics and production requirements. The process objectives of this stock refining system are as follows:

3. Ensure safe Stock Refiner operation.
4. Ensure adequate Level without overflowing in the Refined Kraft Chest.

The specific energy controller is employed to support the first process objective. However, either a Surge Chest consistency or Stock Flow change affects the calculated HPD/Ton controller PV. The Stock flow setpoint will be expected to vary since it receives its setpoint from the Refined Kraft Chest level controller. In addition, the Kraft Surge Chest consistency loop is expected to be subjected to relatively severe upsets since this is close to the Stock Pulpers. The HPD/Ton controller will also be disturbed by fiber characteristics (freeness) from the Kraft Surge Chest.

The Pressure controller effectively controls the flow of stock through the pump. From this perspective, the pressure controller will act to minimize the disturbance of Refiner Flow changes (which are expected) to the consistency loop. If the Stock Flow increases, the Pressure controller decreases the recirculated flow. With the recirculated flow reduced, the Kraft Surge Chest consistency will slowly increase but the consistency controller will be better able to deal with this slow disturbance. Thus, the flow controller should be tuned at least 5 times slower than the pressure loop, both in order to minimize the disturbance to the HPD/Ton controller and to decouple the Pressure and Flow controllers.

The HPD/Ton controller should be capable of attenuating the impact of stock flow changes. As a minimum, the HPD/Ton controller should be at least 5 times faster than the Level controller (which adjusts the flow setpoint). The Stock Flow controller dynamics are likely to be faster and more linear than the HPD/Ton controller, suggesting that the HPD/Ton controller be tuned at least 5 times slower than the Flow controller. Overall, this tuning strategy would support the objective of constant specific energy by making the HPD/Ton controller capable of attenuating Stock Flow setpoint changes and by decoupling the Stock flow and HPD/Ton controllers. Since the HPD/Ton controller will typically have slow tuning, to satisfy the safety objective, additional logic must be added.
The Surge Chest consistency loop would ideally be tuned as fast as possible in order to maximize the attenuation capability of the loop. This assumes that the interaction between the consistency, flow and pressure is relatively small and decoupling is therefore not a requirement.

The interaction between the two consistency loops may be significant however. Since the disturbances from the upstream chest are likely to be more severe, and we want to minimize disturbances to the HPD/Ton controller, it may be desirable to tune this loop faster than the Kraft Refined Chest consistency.

In summary then the following tuning strategy is recommended to best achieve the process objectives:

**Pressure Loop (PIC003)** - Tune this loop to be the fastest with Lambda selected subject to the usual constraints regarding process time constant and process deadtime.

**Flow Loop (FIC004)** - Select Lambda for the flow loop at least 5 times slower than the pressure loop.

**Specific Energy Controller** - Select Lambda for the Specific Energy controller at least 5 times slower than the flow loop.

**Refined Kraft Chest consistency loop** - Tune the level loop to be at least 5 times slower than the inner flow loop.

**Kraft Surge Chest Level loop** - Tune this loop as fast as possible subject to the process deadtime constraint.

**Refined Kraft Consistency Loop** - Possibly tune this loop 5 times slower than the Surge Chest consistency.
Step 3 Implementing the Tuning Strategy

1. The Pressure loop will be the fastest in the system. Using the bump test results from Step 1, tune this controller as fast as possible, consistent with normal Lambda selection constraints. Tune the controller and place it into Auto mode.

2. Tune the Kraft Surge Chest consistency controller as fast as possible, consistent with the normal Lambda selection constraints. Place the controller in Auto mode.

3. Perform open loop bumps on the Flow controller with the Pressure controller in Auto mode. The dynamics of this controller may be substantially altered with the Pressure controller in Auto mode. The Flow will reach steady state after 4 Pressure controller Lambda values (the time required to bring the pressure back to setpoint following the Flow controller output bump). The process gain of the Flow loop will be higher, since the pressure controller will keep the pressure near its setpoint.

4. Select a Lambda value for the flow loop at least 5 times slower than the pressure controller Lambda. Tune the Flow controller and place it into Cascade mode.

5. Perform open loop bump tests on the Refiner HPD/Ton controller with the Pressure and Flow controllers in Auto mode. Tune the controller to be at least 5 times slower than the Flow controller.

6. Tune the Refined Kraft Chest Level controller to be as slow as possible, consistent with maintaining the level within an acceptable operating range. The level controller Lambda should be at least 5 times slower than the Refined Stock Flow controller.

Step 4 Validate the Tuning

A series of bump tests should be performed to ensure that the system of loops is performing as expected. For example, a setpoint change to the Refined Kraft Level should be made. The upstream controlled variables should maintain setpoint reasonably well during the transitional period.
EXAMPLE FAST PRESSURE

REFINER FLOW AND PRESSURE INTERACTION

INTERACTIVE LOOPS - FLOW AND PRESSURE IN MANUAL

INTERACTIVE LOOPS - FLOW AND PRESSURE
LAMBDA OF PRESSURE CONTROLLER = LAMBDA OF FLOW CONTROLLER
INTERACTIVE LOOPS - FLOW AND PRESSURE
PRESSURE CONTROLLER 10 TIMES FASTER THAN FLOW CONTROLLER
11. TUNING LEVEL CONTROLLERS FOR LOAD RECOVERY
Integrating Process

As discussed in Section 2, an integrating process is different from a self regulating process. The process response to a controller output bump is a change in slope. The process does not reach a new steady state as is the case for self regulating processes.

The most common integrating processes are LEVEL controllers. Consider the basic LEVEL CONTROL LOOP.

![Diagram of a basic level control loop]

Fig 11-1 Basic Level Control Loop
Bump Tests

Open Loop Bump tests on a Level controller look like the following. Two different methods can be used to calculate the process gain: the slope method (Fig 11-2) and the graphical method (Fig 11-3). These two methods will give the same answer.

LEVEL CONTROLLER BUMP TESTS

\[ \Delta U_1 \]
\[ \Delta U_2 \]
\[ \text{SLOPE 1} \]
\[ \text{SLOPE 2} \]
\[ \text{SLOPE 3} \]
\[ \text{OUTPUT} \]
\[ \% \]
\[ \text{PV} \]

Fig 11.2 Level Controller Bump Tests (Slope Method for Kp)

SLOPE METHOD

\[ K_{p1} = \frac{\text{SLOPE 2} \times \text{SLOPE 1}}{\Delta U_1} \]

\[ K_{p2} = \frac{\text{SLOPE 3} - \text{SLOPE 2}}{\Delta U_2} \]
Fig. 11-3  Level Control Bump Tests (Graphical Method for Kp)

**GRAPHICAL METHOD**

\[
K_{P1} = \frac{\Delta Y_1 / \Delta T_1}{\Delta U_1}
\]

and,

\[
K_{P2} = \frac{\Delta Y_2 / \Delta T_2}{\Delta U_2}
\]

Differences in the values of \(K_{P1}\) and \(K_{P2}\) (or any other subsequent bump tests) might indicate some valve problems or other non-linearities.
Proportional Only Controller

Since the process being controlled is an integrating process one could argue that no integration action (the "I" part in a PI controller) is required.

A proportional only controller moves the output by an amount proportional to the input error signal. Consider the following proportional on y level control loop below:

![Diagram of a proportional level control loop]

Fig 11-4 Level Control Tuning
Proportional Controller Response To A Setpoint Change

A. Assuming a steady state condition (i.e., flow-in $F_1$ is equal to flow-out $F_2$) and the level is constant.

B. The setpoint is changed, causing an error signal ($LSP$ minus $LPV$). The output moves by an amount proportional to the error. This causes the flow in ($F_1$) to increase, and $F_1$ exceeds $F_2$. The level will increase.

C. As the level increases, the error decreases, and the proportional term is reduced. The output to the controller moves back towards its initial value, reducing the flow-in. When the level reaches the new setpoint, the flow-in is equal to the flow-out.

**Fig 11.5 Level Controller - Proportional Only Response to Setpoint Bump**
Tuning An Integrating Process For Setpoint Changes

If only setpoint tracking is required, the following P controller tuning is required

**TUNING RULE**  
**SETPOINT TUNING FOR INTEGRATING PROCESSES**

- Choose $\lambda$ determine $K_P$
- $K_C = \frac{1}{\lambda K_P}$

This rule can also be used for near integrating processes that is processes which have very slow (but not quite true integrating) responses relative to the desired speed of response of the closed loop ($\lambda$ much faster than $\tau_p$).
Proportional Controller Response To A Load Disturbance

A load disturbance consists of a change in the flow-out \( f2 \). Since more stock is flowing out than \( n \), the level will drop. As the level decreases, the error increases, and the controller output moves \( n \) proportion to the error.

A block diagram representing the load disturbance is shown below.

![Block Diagram](image)

**Fig 11-6** Block diagram for level control.

As the level continues to drop, the error increases hence increasing the flow-in up to the point where the flow-in is equal to the flow-out. At this point, the level will have stopped dropping. However, the level will not return back to setpoint. The error has stopped changing and thus the proportional term will stop changing. An offset from setpoint will remain.

![Graph](image)

**Fig 11-7** Level Controller - Proportional Only Response to Load Disturbance.

**CONCLUSION** Reset is needed to remove the offset.

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Tuning For Load Recovery

A block diagram representing level control using a PI controller is shown below

Fig 11-8 Block diagram for level control

The closed loop transfer function from the disturbance $D = F_{ou}$ to the level $Y$ using a PI controller is

$$G_{LD}(s) = \frac{K_p}{1 + G_c(s) G_p(s)} = \frac{T_R}{K_C K_p (T_R s + 1)} - \frac{T_R}{s^2 + T_R s + 1}$$

The characteristic equation for the closed loop level control system is

$$0 = \frac{T_R}{K_C K_p} s^2 + T_R s + 1$$

The closed loop poles are the solutions of the above quadratic equation. In level control, it can easily happen that the closed loop poles resulting from 'tune by feel' are located off the real axis, resulting in oscillatory behavior. We want to select the controller tuning parameters $K_C$ and $T_R$ so that we have a repeated closed loop pole at

$s = \frac{1}{\lambda}$

resulting in a second order critically damped closed loop response

This can be accomplished by using the following tuning rule

CRITICALLY DAMPED LEVEL TUNING RULE

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Verification That The Level Tuning Rule Produces Critically Damped Response

Observe that when the above tuning rule is used

$$\frac{T_R}{K_C} - \frac{2\lambda}{\left(\frac{2}{K_p\lambda}\right)} - K_p\lambda^2$$

and

$$\frac{T_R}{K_C K_p} = \frac{2\lambda}{\left(\frac{2}{K_p\lambda}\right)} = \lambda^2$$

Using the above expressions, the closed loop transfer function from the disturbance $D - F_{out} - f_2$ to the level $Y$ becomes.

$$G_{LD}(s) = -\frac{T_R}{K_C} - \frac{s}{K_C K_p s^2 + T_R s + 1} - \frac{-K_p\lambda^2 s}{\lambda^2 + 2\lambda + 1} - \frac{-K_p\lambda^2 s}{(\lambda s + 1)^2}$$

Thus the above tuning rule does in fact result in a closed loop response which is critically damped with time constant $\lambda$, which we are free to select.

By substituting $\lambda - T_R/2$ in the equation for the controller gain, the tuning rule can be written alternatively as

$$\begin{array}{c}
T_R - 2\lambda \\
K_C - \frac{4}{K_p T_R}
\end{array}$$

CRITICALLY DAMPED LEVEL TUNING RULE
Lambda Interpretation in Level Controller Tuning

The question that remains is "What is the meaning of $\lambda$ in the above tuning rule and how should we choose $\lambda$?"

When the critically damped tuning rule is used, the load response is the step response of the transfer function

$$G_{LD}(s) = \frac{K_P \lambda^2 s}{(\lambda s + 1)^2}$$

The controller load recovery is shown in the following figure:

![Figure 11-9 Level Control Response When Tuned For Load Recovery](image)

In $\lambda$, the level change will be arrested and the flow in $f_1$ will equal the flow out $f_2$. In $6\lambda$, the level will be brought back to setpoint as a result of the flow in $f_1$, having been higher than the flow out $f_2$, between the times $t = 1\lambda$ and $t = 6\lambda$.

We refer to the time $t$ takes the controller to bring the level back to setpoint following a load step change as the recovery time, $T_{REC}$. We refer to the time it takes the controller to arrest the level from falling (or rising) after a load step change as the arrest time, $T_{ARR}$. $\lambda$ is identical to $T_{ARR}$, and thus

$$\lambda = \frac{T_{REC}}{6}$$

Thus the tuning rule can be expressed as

<table>
<thead>
<tr>
<th>$T_R$</th>
<th>$2T_{ARR}$</th>
<th>$T_{REC}/3$</th>
</tr>
</thead>
<tbody>
<tr>
<td>$K_C$</td>
<td>$K_P$</td>
<td>$T_R$</td>
</tr>
</tbody>
</table>
Level Controller Setpoint Response

The level response to a setpoint change is as follows:

**PI CONTROL - SETPOINT RESPONSE**

Fig 11-10 PI Controller Response to a Setpoint Bump

The important points to note are

1. The level first reaches the new setpoint at the Arrest Time ($-\lambda$)
2. There will be an overshoot of approximately 13% of the setpoint change
3. The level finally settles out at the new setpoint at the Recovery Time ($6\lambda$)
Tuning Rules For Integrating Plus Deadtime Processes

To tune an integrating plus deadtime process, the tuning rules are modified as shown below

\[
\begin{align*}
T_R &= 2\lambda + T_d \\
K_C &= \frac{T_R}{K_P(\lambda + T_d)^2}
\end{align*}
\]

CRITICALLY DAMPED LEVEL WITH DEADTIME TUNING RULE

EXAMPLE:

The level process gain, \(K_P\) from a series of bump tests was 0.2 \(\%\) Span/Min. The deadtime was 10 seconds or 0.167 minutes.

The desired Lambda value (ARREST) time for this level controller is 5 minutes.

The Reset Time is calculated as follows

\[
T_R = 2\lambda + T_d = 2 \times 5 + 0.167 = 10.167 \text{ minutes}
\]

The Controller gain is calculated as follows

\[
K_C = \frac{T_R}{K_P(\lambda + T_d)^2} = \frac{1016}{0.2(5+0.16)^2} = \frac{1016}{0.2(5.16)(5.16)} = 19 \%\text{Output} / \%\text{Span}
\]
Choosing Lambda

The most important decision in tuning integrating processes is the selection of the LAMBDA (or ARREST) time. The LAMBDA value should be based primarily on the system design (i.e., chest capacity, throughput etc.) and operating and control requirements.

For most level controllers, two important operating and control requirements are:

1. The chest should not overflow (for environmental and fiber savings reasons) and
2. The chest should not run dry (for production reasons)

Consider the following example.

![Diagram of level loop]

Fig 11-11 Operating conditions of a level loop

In this example, the level controller setpoint is set at 60%. At this level, there is 10 minutes of Retention Time. If the flow of stock into the chest was stopped, the chest would run DRY in 10 minutes. On the other hand, there is 6 minutes of capacity in the chest. That is, if the stock pump stopped, the chest would OVERFLOW in 6 minutes.

A reasonable expectation of the level controller tuning is that the most severe disturbance will be ARRESTED before the chest overflows. The LAMBDA value (ARREST time) in this case should be 6 minutes or faster. It may be possible to increase Lambda if interlocks safeguard against certain conditions.

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Once the LAMBDA value is selected the RECOVERY time is also selected since the RECOVERY time is equal to 6 TIMES LAMBDA. This means that the controller will bring the level back to setpoint in 36 minutes following a disturbance.

\[ \lambda = 6 \text{ Minutes} \]

OR

\[ T_{\text{Rec}} = 36 \text{ Minutes} \]

The process gain (Kp) is determined from the bump tests. Knowing Kp and \( \lambda \), the Proportional Gain (Kc) and RESET TIME (\( T_{R} \)) can be calculated. For a level controller, we would like to ensure that the level controller is capable of arresting a disturbance within the arrest time over the entire operating regime of the level controller.
Effect of Aggressive Level Tuning

You might wonder why a very fast Lambda (Arrest Time) wouldn't always be chosen. This would ensure that the controller will respond very aggressively to disturbances, thereby maintaining the level close to setpoint. The answer is that these very aggressive level controller actions will unnecessarily disturb upstream and downstream processes. It is very important to choose a LAMBDA value that will perform adequate level control and also minimize disturbances to upstream and downstream processes.

Consider the example below which shows two different level control responses to the same load disturbance. The Arrest time of the fast tuning is 10 times faster than the slow tuning.

**FAST VERSUS SLOW ARREST TIME**

![Graph showing fast versus slow arrest time](image)

**Fig 11-12 Fast Versus Slow Arrest Time**

There are two important points to observe.

1. With the fast tuning, the level controller output increases very quickly, which means that the flow rate into the chest increases very rapidly. With the relatively slow tuning, the flow rate changes much more gradually. The impact of the fast flow change on upstream process conditions (consistency, freeness, etc.) will be much more severe than with the more gradual change. The upstream controllers will obviously have a greater ability to eliminate the impact of a gradual flow change versus a fast flow change.

2. With the fast tuning, the load disturbance is arrested quickly and the level decrease is small. The fast tuning will hold a much tighter setpoint than the slower tuning.
Effects Of Dampening On Level Controllers

Since tuning level controllers for load recovery yields relatively high proportional gains, the controller may react significantly to noise present in the measurement signal. In fact, the level often contains fast variability due to the action of the chest agitator and surface wave sloshing. The high gain tuning can act to wear out valves, actuators and disturb upstream processes.

The following example compares fast to slow tuning in the presence of a disturbance cycle. Note the control action taken with the fast tuning versus the slow tuning. The impact of the tuning will be to generate large upstream flow disturbances while doing little to eliminate the level disturbance.

FREQUENCY RESPONSE PROCESS FILTERING

![Graph showing the impact of fast variability on level controller output.](image)

Fig 11-13 Impact of Fast Variability on Level Controller Output

Dampening can be used to attenuate the noise present while not significantly affecting the controller's performance. The Filter Time Constant \( \tau_f \) should be at least 10 times smaller than \( \lambda \).

\[ \tau_f < \lambda/10 \]

Note that the faster the LAMBDA, the faster the filter should be. Excessive use of filtering can result in level cycling.
Example 11-1

Calculate the process gain for LC101 in units of \( \frac{\% \text{ Span} / \text{Min}}{\% \text{ Output}} \). Please note that this is simulated data and there are no nonlinearities.

**Checklist**

1. The Process Gain for a Level controller can be calculated using the \( \frac{\Delta Y}{\Delta T} \) or the \( \frac{\Delta Y}{\Delta U} \) method.

2. Using the \( \Delta Y / \Delta T \) method, the process gain is \( \frac{\Delta Y}{\Delta U} \).

3. Using the \( \Delta Y / \Delta U \) method, the process gain is \( \frac{\text{SLOPE2-SLOPE1}}{\Delta U_1} \).
Example 11-2

The bump tests shown below were taken from an air padded headbox. The level controller manipulates the air supply valve to the air pad in order to maintain the level at setpoint.

Questions

Q1 Calculate the process gain for this level controller.

Q2 Can you identify any nonlinearities?

Q3 Can you explain the common cycle in the Headbox Level and Total Head?

---

[Graphs showing data series with time series plots and time in seconds.]
Checklist

1. Have enough bump tests been performed to properly evaluate back ash and stiction in the control valve.

2. For an even controller, control valve stiction will result in a ________ shape in the level.

3. In an air padded Headbox, the Level and Total Head controllers are ________.
Example 11-3

The simplified P&ID shows a pine High Density (HD) and level chest process and control system. The time series data shows the pine chest level, HD flow, HD consistency and coarse dilution flow.

Questions:

1. How important is the variability in the pine chest level?
2. What might be causing the 60 second cycle in the pine chest level?
3. Comment on the control strategy. Can you suggest any improvements?
APPENDIX 11A

SAMPLE BUMP TESTS
Time Series Analysis involves analyzing a series of data taken at equally spaced intervals of time. We are primarily interested in using time series analysis techniques to describe a data series statistically and to characterize its frequency content.

**Statistics**

Using statistics is a useful way of communicating information about the extent of the variability in a data series. Some typical statistical measures (defined below) are widely used in the process industries and are an effective way of measuring and communicating product quality and process variability. The following statistics are widely used and defined as follows:

1. The **mean** of the time series \( (x_k) \), also called the expected value, is simply the average value of the data points. It is denoted by \( x \) and is defined mathematically as follows:

\[
x = \frac{1}{N} \sum_{i=1}^{N} x_i
\]

where \( N \) is the number of data points.

2. The **variance** of the time series \( (x_k) \) is a measure of the spread in the data. It is denoted by \( \sigma^2 \) and is defined by

\[
\sigma^2 = \frac{1}{N} \sum_{i=1}^{N} (x_i - x)^2
\]

3. The **standard deviation** of the time series \( (x_k) \) denoted by \( \sigma \), is probably the most widely used statistic and is just the square root of the variance, namely

\[
\sigma = \sqrt{\text{var}(x_k)}
\]

If the data series is normally distributed (seldom the case in continuous process industries) approximately 95% of all the data points will fall within \( \pm 2\sigma \) of the mean.

4. The variability is equal to \( 2 \times \sigma \). This is a frequently used measure of the spread of data in a time series and is used in EnTech's data analysis software.

5. Expressing the variability as a percentage of the mean allows comparisons between the level of variability in different processes. This is equal to

\[
\% \text{ Variability} = \frac{2 \times \sigma \times 100}{\text{MEAN}}
\]
Power Spectrum

The power spectrum describes the frequency content in a data series by decomposing the signal into many sine waves, known as harmonics. It is particularly useful because it assists in identifying any significant cycles in the process. This is important for a number of reasons:

1. It will assist in diagnosing the source of product cycles. A product cycle at a given frequency is caused by a process cycle at the same frequency. Thus, it assists in linking process disturbances with product disturbances. Ultimately, this is an important tool in reducing product variability.

2. It will assist in analyzing control loop performance.

3. It will provide insights into control or process design modifications which might aid in reducing process variability.

The plot of variance versus frequency is known as the 'Power Spectrum' plot. The power spectrum indicates how much variance exists at a particular frequency or period. The sum of the individual variances at each frequency is equal to the total variance in the data series.

The power spectrum analysis has a number of properties which are very useful in characterizing a data series. These are described in detail below.
Power Spectrum Terminology

Consider Time Series data \((x_k) = (x_1, x_2, \ldots, x_N)\) consisting of \(N\) samples collected with a sample interval of \(T_s\). As mentioned previously, the power spectrum describes the frequency content in the time series by decomposing the signal into sine waves called harmonics. There may be many harmonics, however, since the data is sampled and not continuous, the power spectrum will only be able to resolve the data into a finite number of discrete frequencies or harmonics defining a frequency window.

The lowest frequency represented in the time series is defined by the fundamental period. The fundamental period \(T_1\) is defined by

\[ T_1 = N \cdot T_s \]

and is simply the time interval spanned by the time series is called the fundamental frequency.

The highest frequency represented in the series is defined by the Nyquist Period. The Nyquist Period \(T_N\) is defined by

\[ T_N = 2T_s \]

and the corresponding Nyquist frequency is then

\[ f_N = \frac{1}{2T_s} \]

The harmonics are sinusoidal time series with frequencies that are a multiple of the fundamental frequency. The \(m\)th harmonic of the time series has frequency

\[ f_m = m \cdot f_1 = m \cdot \frac{T_1}{NT_s} \]

and period

\[ T_m = \frac{T_1}{m} \]

Assuming that the number of samples \(N\) is even, the number of harmonics will be \(N/2\).
Example:
Consider Times Series data collected with $T_s = 0.2$ sec and $N = 1024$ samples.

**Fundamental Period**

$$T_1 = 1024 \times 0.2 = 204.8 \text{ sec / cycle}$$

**Fundamental Frequency**

$$f_1 = \frac{1}{204.8} = 4.88 \times 10^{-3} \text{ Hz}$$

**Nyquist Period**

$$T_N = 2T_s = 0.4 \text{ sec / cycle}$$

**Nyquist Frequency**

$$f_N = \frac{1}{2} \left( f_s \right) = \frac{1}{2} \left( \frac{1}{0.2} \right) = 2.5 \text{ Hz}$$

**Harmonics present**

There are $N/2 = 1024/2 = 512$ harmonics present given by

$$f_k = kf_1 = \frac{k}{204.8} \text{ cycles / sec}$$

$$T_k = \frac{T_1}{k} = \frac{204.8}{k} \text{ sec}$$

The period and frequency of the first few harmonics are shown in the following table:

<table>
<thead>
<tr>
<th>Harmonic $k$</th>
<th>Period $T_k$ sec</th>
<th>Frequency $f_k$ Hz</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>204.8</td>
<td>0.00488</td>
</tr>
<tr>
<td>2</td>
<td>102.4</td>
<td>0.00977</td>
</tr>
<tr>
<td>3</td>
<td>68.3</td>
<td>0.01465</td>
</tr>
<tr>
<td>4</td>
<td>51.2</td>
<td>0.01953</td>
</tr>
<tr>
<td>512</td>
<td>0.4</td>
<td>2.5</td>
</tr>
</tbody>
</table>

Notice that the frequencies are equally spaced.
**Variance versus Cycle Amplitude**

The power spectrum plot separates the total variability into its frequency components. At each analyzed frequency, the power spectrum will report a variance. The variance contributed at each frequency is directly related to the amplitude of the cycle at that frequency. The relationship between variance and cycle amplitude is

\[
\text{VARIANCE} = \frac{1}{2} a^2
\]

The variance contributed by an individual cycle is therefore independent of the frequency of the cycle. This important point is illustrated in the following simple examples.

**TIME SERIES PLOT**

<table>
<thead>
<tr>
<th>Time Series</th>
</tr>
</thead>
<tbody>
<tr>
<td>3200</td>
</tr>
<tr>
<td>0.0</td>
</tr>
<tr>
<td>Mean = 31049</td>
</tr>
</tbody>
</table>

**POWER SPECTRUM PLOT**

<table>
<thead>
<tr>
<th>Power Spectrum</th>
</tr>
</thead>
<tbody>
<tr>
<td>VAR = 0.5 a^2</td>
</tr>
<tr>
<td>VAR = 0.005</td>
</tr>
</tbody>
</table>

\[
\text{STD DEV} = \sqrt{\text{VARIANCE}} = 0.0707
\]

\[
\text{VARIABILITY} = 2 \times \text{STD DEV} = 0.141\%
\]

**Figure 12-1 Power Spectrum Plot of Single Sine Wave**

The amplitude of the single cycle in the time series plot shown above is 0.1% consistency. The expected variance contribution from this cycle is equal to \( \frac{1}{2} a^2 \), or 0.5 \((0.1)^2 = 0.005\). Note that the power spectrum plot indicates a single 'peak' at 25 seconds/cycle. The height of this peak reflects the expected variance contribution of this cycle (approximately 0.005).

The STANDARD DEVIATION is equal to the square root of the variance, which in this case is 0.0707%. Consistency. The Variability is equal to 2 times the Standard deviation or 0.141% Consistency. The variability expressed as a percentage of the mean is

\[
\text{Variability} \% \text{ of mean} = \frac{0.141}{3.1} \times 100 = 4.57\% \text{ of mean}
\]
The following example again shows a consistency signal containing a single cycle. The period of the cycle in this example is 100 seconds.

**TIME SERIES**

**POWER SPECTRUM**

![Graphs showing time series and power spectrum with labels and data points]

Fig. 12-2 Power Spectrum Plot of Single Sine Wave

This example illustrates that the variability is independent of the period of the cycle. In this example, the amplitude of the cycle is again 0.1% consistency. The power spectrum indicates that there is a single cycle at a period of 100 seconds/cycle and that the variance contributed by this cycle is approximately 0.005.
Complex Process Signals

Process signals almost always contain many cycles, ranging from very slow cycles to very fast cycles. The power spectrum analysis breaks these cycles down into their individual contributions. The sum of the individual variances is equal to the total variance.

Consider the example below, where the process signal consists of two sine wave cycles added together. The amplitude of each cycle again is 0.1% consistency. One cycle has a period of 100 seconds/cycle and the other cycle has a period of 50 seconds/cycle.

![Graph showing time series and power spectrum of two sine waves with equal magnitude](image)

Fig 12-3 Sum of two Sine Waves - Equal Magnitude

It is still possible to 'eyeball' the period of the two cycles, although more difficult to judge the respective amplitudes simply from the time series plot. The power spectrum plot clearly shows both the period and the amplitude of the two cycles.

Each peak is expected to contribute variance of 0.005 since the amplitude of each cycle is 0.1%. The individual variances are additive, and therefore the expected total variance is 0.01. The standard deviation is $\sqrt{0.01}$ or 0.1% consistency. The VARIABILITY is $2 \cdot \sigma$ or 0.2% Consistency. Expressed as a percentage of the mean, the variability is equal to $0.2 \cdot 100$ or 6.47% of mean. These expected values are indicated on the time series plot above.
The usefulness of the power spectrum becomes increasingly evident as the complexity of the data series increases. Consider the following process signal composed of three sine wave cycles. The amplitude and period of these cycles are as follows:

<table>
<thead>
<tr>
<th>Period</th>
<th>Amplitude</th>
<th>Variance</th>
<th>Variance %</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cycle 1</td>
<td>100</td>
<td>0.05</td>
<td>0.0013</td>
</tr>
<tr>
<td>Cycle 2</td>
<td>50</td>
<td>0.10</td>
<td>0.0050</td>
</tr>
<tr>
<td>Cycle 3</td>
<td>25</td>
<td>0.15</td>
<td>0.0113</td>
</tr>
<tr>
<td>Total</td>
<td></td>
<td></td>
<td>0.0175</td>
</tr>
</tbody>
</table>

In this case, each cycle does not contribute equally to the overall variance. The unequal peaks in the power spectrum indicate the relative contributions and are consistent with our expectations.

Fig 12-4 Power Spectrum - Sum of 3 Sine Waves Unequal Amplitudes

It would clearly be difficult to accurately judge the period and the amplitude of the three cycles in the process signal in this example. Yet, this is a relatively simple signal. Most process measurements are far more complex and are increasing difficult to assess in terms of frequency content.
Cumulative Spectrum

The cumulative spectrum plots the percent of total variance versus the frequency. It is very useful in indicating the percent contribution of the individual cycles in the process signal to the total variance.

For example, in the previous example the percentage contribution of the 100, 50 and 25 second cycles were 7.2, 28.6 and 64.3% respectively. The cumulative spectrum plot for this process signal is shown below.

![Cumulative Spectrum Diagram](image)

Fig 12-5 Cumulative Spectrum

The cumulative spectrum plot indicates that

1. 7.2% of the total variance is slower than 100 seconds/cycle,
2. 35.8% of the total variance is slower than 50 seconds/cycle, and
3. 100% of the variance in the process signal is slower than 25 seconds/cycle.
White Noise

A process signal that contains purely random variability is termed WHITE NOISE. Such a process signal is shown below.

![Graph showing time series and power spectrum of white noise.]

![Graph showing autocorrelation and cumulative spectrum of white noise.]

Fig. 12.6 Time Series Plot of White Noise

Since the white noise signal is purely random, the variance is divided equally among all frequencies included in the power spectrum. Thus, the white noise signal contains slow variability as well as fast variability (equal variance at all analyzed frequencies). Note that the magnitude of the average variance present in each harmonic is very small (about $0.5 \times 10^{-5}$) since the total variance is divided equally in this case into 512 frequency components.
Autocorrelation

The Autocorrelation plot is considered the best indicator of the randomness in a time series. If the data in a process signal is completely random (white noise) then all of the vertical lag lines in the autocorrelation plot will fall within the horizontal confidence limits.

White Noise is often associated with measurement noise. The process signal can be thought of as the combination of the true process value and the measurement noise. This combined signal is the only process measurement available for control purposes. However, it is clear that taking corrective action to compensate for measurement noise will increase the true process variability.

The concept of minimum variance control states that the controller should be tuned to eliminate only true process variability. Once a white noise signal is achieved, no further control improvement can be made. The autocorrelation function can be used to assess whether minimum variance control is being approached. Subject to certain assumptions, minimum Variance Control will be achieved when the autocorrelation function decays to zero (falls within the confidence limits) with one plus the number of lags representing to the process deadtime.
Filtered White Noise

An ideal white noise signal contains equal variance at all frequencies. It is of interest to discover how the white noise signal is 'shaped' by a first order filter.

The white noise signal shown in Fig 12-6 is passed through a first order filter. The time series plot of the resulting signal is shown in Fig 12-7 below.

Fig 12-7 Filtered White Noise

Note that the variability in this signal is much lower than in the original white noise signal.
Relation of Power Spectrum Plots to Bode Plots

Using a variance vs. period plot, it is difficult to determine how the power spectrum plot is affected in the higher frequency range. In the power spectrum below, the log variance is plotted against the log frequency.

Fig 12-8 Filtered White Noise using log-log Power Spectrum

The shape of this plot is similar to the Bode plot for a first order system. The variance is constant at lower frequencies, but then begins to decrease with a relatively constant slope. For a first order system, the cutoff period (at $2\pi$) defines the frequency where cycles begin to be attenuated. The AR decreases at a constant slope of -1 decade AR/decade frequency for frequencies greater than the cutoff period. This form of the power spectrum can therefore be quite useful in identifying the amount of filtering in a process signal.

It is interesting to note that since the variance of a cycle is proportional to the amplitude squared, the log variance/ log frequency slope is equal to -2 decade variance/decade frequency at frequencies faster than the cutoff frequency.
Aliasing

Digital controllers do not see the process as a continuous analog signal. Instead, they sample the continuous signal through an A/D converter at regular intervals. The sampling rate varies greatly depending on the system, ranging from as fast as 100 milliseconds to as slow as once per 30 seconds (a paper machine scanner for example).

The sampling rate can affect the controller's performance depending on how fast the process varies. Shown below is a signal that is sampled much faster than it is changing, the end result is that the sampled data (the controller "sees" the process as connected dots) is very close, if not identical to the true process signal.

![Diagram of a signal sampled at different rates.](image)

**Fig 12-9** Effect of Sampling a Signal

However, if the signal is sampled too slowly, as is shown below it is evident that the connected dots do not resemble the true signal. The controller essentially sees a 'ghost' which has nothing to do with reality, and if I attempt to control out this false variability, this is called ALIASING.

![Diagram showing aliasing effect](image)

**Fig 12-10** Effect of Aliasing
Digital Systems Signal Aliasing

A sine wave cycle must be sampled at least twice per cycle period (Nyquist period) if the cycle is to be recovered from the digital representation. If a signal contains a cycles faster than the Nyquist period, then the controller is vulnerable to aliasing.

**EXAMPLE**

The Headbox Total Head is sampled once per second for 1024 seconds. The Nyquist period is 2 seconds/cycle or 0.5 Hertz. There is no anti-aliasing filter installed.

**Fig 12-11**

The power spectrum contains a peak at the Nyquist period which accounts for the majority of the variance. A power spectrum peak at the Nyquist frequency is a strong indication of aliasing in the data. Also note the ragged pattern in the autocorrelation function which also indicates the possibility of aliasing.
As a result the test procedure is modified to sample the Total Head at 20 mlliseconds. The 2 second cycle disappears. Now, a very strong cycle at 0.4 seconds/cycle is obvious, accounting for approximately 90% of the total variance.

Fig 12.12 Sampling the Total Head every 20 milliseconds.
Anti-Aliasing Filtering

To minimize the effect of aliasing, an anti-aliasing filter should be installed at every point of sampling in the digital system. Knowing the sampling rate, the signal needs to be filtered prior to sampling to take out variability that is faster than the Nyquist frequency corresponding to the sampling operation. At the point where the analog signal first enters the digital system and is initially sampled, an analog filter (e.g., a hardware filter, not a software filter) must be used for anti-aliasing.

A minimum amount of anti-aliasing filtering will be achieved by using a first-order filter with the filter time constant set equal to 1.3 times the sampling period.
White Noise And Cycling

The majority of process signals contain true process disturbances as well as white noise. This is illustrated in the following example. The process signal consists of a 17 second cycle with an amplitude of 0.025% consistency, and white noise with a variance of 0.0025

<table>
<thead>
<tr>
<th></th>
<th>AMPLITUDE</th>
<th>VARIANCE</th>
</tr>
</thead>
<tbody>
<tr>
<td>WHITE NOISE</td>
<td>0.025%</td>
<td>0.0025</td>
</tr>
<tr>
<td>17 SEC/CYCLE SINE WAVE</td>
<td>0.025% Cons</td>
<td>0.0003</td>
</tr>
<tr>
<td>TOTAL VARIANCE</td>
<td>0.0028</td>
<td></td>
</tr>
</tbody>
</table>

Fig 12.13 White Noise Plus Single Sine Wave

The power spectrum shows a peak at a period of 17 seconds. This cycle can be detected, but is not obvious from the time series plot alone. Note that while the 17 second cycle dominates the power spectrum plot, it only represents 10.7% of the total variance.
13. INTERPRETING MILL PROCESS DATA
EXAMPLE 1: Fast Basis Weight Variability

The following data was collected from a coated linerboard machine. The data shown in Figure 1.1 was collected at 0.1 second intervals during production of the 68 lb grade. The data shown in Figure 1.2 was collected at 0.1 second intervals on the 96 lb grade.

Questions:

1. What is the dominant behavior in the BW variability data?
2. What area of the machine would be a likely source of the BW variability?
3. Why might the variability be higher on the heavy grade?

Figure 1.1 - BW Variability - Sampling at 0.1 second intervals, 1024 samples. - 68 lb grade
Figure 1.2 - BW Variability - Sampling at 0.1 second intervals, 1024 samples. 96 lb grade
EXAMPLE 2
Stock Chest Level Variability

The stock preparation area of a tissue machine is shown below. The Machine Chest (MC) level controller maintains level at setpoint by adjusting the stock flow rate from the Blend Chest. The Blend Chest Level controller maintains setpoint by cascading a setpoint to the Broke and Sulphite flow controllers.

Questions
1. What is the dominant cycle in the process data?
2. Can you explain why all of the process variables in this system contain the same cycle?
3. What tests would you conduct to determine the ultimate source of the process cycles?
Figure 2.1 Variability in the Blend Chest and Machine Chest Levels

Figure 2.2 Variability in the Stock Flows to the Blend Chest
EXAMPLE 3
Sensor Noise

The procedure used to collect process data often has a significant impact on the measured variability. Figures 3.1 shows BW variability results on a tissue machine. In Figure 3.1 (top) the BW was sampled at 1 second intervals for 1024 seconds. An anti-aliasing filter of 13 seconds was installed during the collection. In Figure 3.1 (bottom) the BW was sampled at 0.1 second intervals for 819.2 seconds. An anti-aliasing filter of 0.13 seconds was installed during this collection.

Figure 3.1 BW Variability under different test conditions.

Questions
1. Explain why the variability is so much higher on the second run.
2. What is the implication of this result in terms of quantifying the variability in a data signal?
EXAMPLE 4.1
Control Valve Performance

Control valve non-linearities are very often the most significant source of process variability. Backlash, Stiction, and Positioned Overshoot are examples of hard nonlinearities which will compromise control performance and can generate process cycles. Figures 4.1.1 and 4.1.2 show time series data from the Pressure Screen Rejects Flow loop in the Lean Stock system. The first plot shows the process and controller output with the loop in Auto mode. The following plot shows a series of open loop bump tests.

Questions
1. Identify the nature of the control valve non-linearity.
2. Discuss the impact of modifying the Lambda value on the period and amplitude of the flow cycle.

Figure 4.1.1 - Variability caused by Control Valve Non-linearity
Figure 4.1.2 Open Loop Bump Tests on Screen Rejects Flow Loop
EXAMPLE 4.2
Control Valve Non-linearities

Backlash is caused by slop in the linkages between the control signal and the valve plug. Backlash compromises control performance and can cause the process to cycle particularly in the case of integrating processes or cascaded loops.

Figure 4.2.1 shows a series of open loop bumps on a Thick Stock consistency loop. Figure 4.2.2 shows a series of setpoint bumps on a Refined Kraft flow loop.

Questions

1. Quantify the backlash shown in Figure 4.2.1
2. Discuss whether retuning this control loop is a good idea?
3. Quantify the backlash indicated in the Figure 4.2.2 testing.
4. Estimate the response dead time resulting from the backlash?

Figure 4.2.1 Open Loop Bump tests - showing Backlash.
Figure 4.2.2 Setpoint Bumps in the presence of Backlash
EXAMPLE 5
Time Series Analysis and Process Upsets

The following data was taken from Section 3 of the Dryer Section Pressure and Different a loops shown in the sketch below. The data was collected immediately following startup.

Questions

1. Characterize the pressure and differentiable variability during this startup period.
2. Discuss the possible benefits of the power spectrum plot analyzing the startup data.
3. Why does the differentiable decrease so dramatically following the sheet break? How does the differentiable controller compensate for the disturbance?

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Figure 5.1 Dryer Data gathered during startup period.
EXAMPLE 6
Process Inherent Cycle

In a typical Saveall the mesh covered disks rotate through a Vat filled with medium consistency stock (0.5 - 2.0%). The Vat level is controlled to setpoint (typically 50% level) by adjusting the disk speed. As the disk speed increases the filtrate flow rate increases. As the disk rotates through the Vat the filtrate flow rate varies causing a level cycle at a period equal to the disk revolution time. A time series plot of a typical Saveall Vat level is shown below.

Questions
1. Should the process inherent cycle be a consideration when developing a tuning strategy for this loop?
Figure 6.1 Saveall Level Variability
EXAMPLE 7
Long Term Variability

Figures 7.1 and 7.2 show scan average BW, Moisture and Conditioned Weight data collected from a linerboard machine. The Moisture controller was in Auto mode throughout both collections per odds.

Questions
1. Can the BW and Moisture control performance be improved?
2. Is the BW variability primarily a result of Moisture or Dry Fiber variability? Explain your conclusions?
3. What are some possible explanations regarding the power spectrum peak at approximately 500 seconds shown in Figure 7.2?

Figure 7.1 Scan Average BW, Conditioned Weight and Moisture Variability - 40000 second run
Figure 7.2 Scan Average Moisture Variability - 5000 second run
EXAMPLE 8
Signal Integrity

With the advent of digital transmitters microprocessors have found their way out to the field devices. This has been beneficial in terms of performing tasks such as remote re-spanning and calibration. However, because the signals sampled in the field there is the possibility of aliasing occurring at the field device. Furthermore, the signal often undergoes various forms of digital signal processing such as spectral suppress or moving average calculations. The resolution of the digital transmitter may be inadequate because of A/D resolution. The impact of these non-linearities on control performance is not understood by the software designer who have understood their design challenge as a real-time computing task only.

Figures 8.1 to 8.4 show several examples of digital sensor data.

Questions
1. Describe the non-linearity shown in Figure 8.1. Is this acceptable for this application?
2. Estimate the A/D resolution in the data signal shown in Figure 8.2. The even transmitter is spanned from 0 to 100% level.
3. A Fan Pump speed signal compared with simultaneously collected Total Head signal in Figure 8.3. A bump test showing the total head response to a fan pump speed bump shown in Figure 8.3a. Does the fast cycle in the Fan Pump speed signal appear to be real?
4. Expand the differences in the power spectrums shown in Figure 8.4. The top plot shows an analog transmitter signal. The simultaneously collected Smart transmitter data signal is shown on the bottom plot. The sensors are both flush mounted to the Headbox.

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Figure 8.2 Level response to Open Loop Bumps - using Smart Level Transmitter
8.3 Variability in the Total Head and Simultaneously collected Fan Pump Speed

**Figure 8.3a Total head response to Fan Pump Speed setpoint bump.**
Figure 8.4 Variability Reported by an Analog Pressure transmitter compared with a Smart transmitter.
EXAMPLE 9
Paper Machine Dryer Control

In the case of Paper Machine Dryers it is often difficult to achieve acceptable control performance because the process dynamics of the Pressure and Differential control loops vary widely depending on operating regime. Thus, while control performance appears to be acceptable at the grade where the tuning was performed, the processes will have high variability on a different grade.

Questions
1. Describe how the dynamics of the Blowthru flow controller vary with thermocompressor (TC) spool position.
2. Explain why the Blowthru variable begins to cycle so vigorously.
3. What this tells us the power spectrum plot.

![Diagram of Paper Machine Dryer Control System]
Figure 9.1 Variability Results from the Pressure and Blowthru loop of Dryer Pressure and Blowthru Loops.

Figure 9.2 Open Loop Bump test results from Blowthru Control Loop.
EXAMPLE 10
Fast Basis Weight Variability

The Basis Weight data shown in Figure 10-1 was collected from a Newsprint machine running at 4000 fpm. As can be seen from the power spectrum plot, there's a dominant cycle in the data at 16.5 seconds. The cumulative spectrum plot represents more than 50% of the variance.

Questions

1. What process area would be a key candidate as the source of this cycle?
2. What methods could be employed to determine the source of this cycle?
The time series plots shown in Figure 10.2 show the Basis Weight data compared to Total Head data. As can be seen from the power spectra, the 16.5 second cycle is dominant in both sets of data. This is an indication that the cycle in BW may be caused by the Total Head variability.

Figure 10.2 Basis Weight and Total Head variability

In order to determine if this is the definite source of the BW variability the Total Head control loop was placed into manual mode and data was collected on both the Total Head and Basis Weight. As can be seen from the time series plots shown in Figure 10.3 the 16.5 second cycle no longer visible.

Figure 10.3 Basis Weight and Total head variability with the Total Head controller in Manual mode.
14. INTEGRATED PROCESS AND CONTROL DESIGN
Process control consists of three main areas:

1. Process Design

At the Process design stage, the process system is defined (a flow sheet containing steady state flows and other operating conditions), the process and control equipment is selected (tanks, pumps, fans, screens, sensors, final control elements etc.), and isometric drawings will be prepared detailing the piping runs connecting the process equipment and positioning the instrumentation.

These design decisions will contribute in a fundamental way to the dynamics (and ultimately the controllability) of the installed process system. Yet, the design will usually be completed without any assessment of system dynamics.

2. Control Strategy Design

The Control Strategy design includes the selection of the controlled variables (flows, pressures, temperatures), the manipulated variables (location and type of control elements which are used to maintain the controlled variables at setpoint), the definition of the control algorithm (simple PID, Cascade Feedforward, Mode Switching, Tracking logic). These decisions are also fundamental to final process stability and product variability. The control strategy design should be based on optimizing the process objectives - usually involving minimizing product variability and maximizing machine efficiency.

Some key concepts articulated by 'Doss and Downs' of Tennessee Eastman Company are that the process design and control strategy design should be integrated. The objective of the integrated approach is to eliminate sources of variability where possible and to create variability pathways which direct variability from key process variables to less important processes.
3 Tuning Strategy

The Tuning Strategy should support the process objectives. In this course the tuning strategy is implemented through Lambda selection. Six different Lambda selection criteria have been discussed. These are,

- Select Lambda as fast as possible while limiting resonance to an AR of 1.41 (1.3 dB). The main objective is to maximize ability to attenuate process disturbances.

- Select Lambda to be the same as another loop in order to support a process objective. The best example of this is Stock Blending. Setting the same Lambda value for all flow controllers which participate in Stock Blending will support the objective of maintaining a constant fiber ratio in the chest.

- Select Lambda to be much slower than an interactive controller (5 to 10 times rule) to prevent controller induced cycling.

- Select Lambda to be faster than the next upper cascade Lambda to prevent inner/outer loop cycling (5 to 10 times rule).

- Select Lambda to be as slow as possible, consistent with allowing the process to function. For example, level controllers should generally be tuned slowly to minimize disturbance to upstream processes.

- Select Lambda in order to minimize variability. Set \( \text{Lambda} = \text{Corner Period}/2\pi \).
Efficient Manufacturing Of A Uniform Product

Key concepts to achieve efficient manufacturing of a uniform product are

1. **Eliminate Sources of Variability at the Design Stage**

Remember that Process Mixing and Control loops can only attenuate. There will still be variability remaining.

2. **Make Effective Use of Process Mixing**

- Improving the agitation in Mixing Caches is effective at attenuating high frequency variability. In the Pulp and Paper industry, inadequate attention is paid to optimizing stock chest mixing.

- One tank attenuates at -20 dB/decade after the cutoff period. Two tanks in series attenuates at -40 dB/decade for frequencies faster than the cutoff.

Fig. 14-1
3 Make Effective Use of Control

The control loop is effective at attenuating very slow variability. Generally, the controller should be tuned to have as little resonance as possible. It is impossible to achieve good control with poor control equipment. Backlash and stiction are particularly important nonlinearities which act to compromise control performance.

![Diagram](image)

*Fig 14-2*

<table>
<thead>
<tr>
<th>$\lambda$</th>
<th>Resonance</th>
</tr>
</thead>
<tbody>
<tr>
<td>1.6 Td</td>
<td>+30 dB = 14</td>
</tr>
<tr>
<td>2.0 Td</td>
<td>+26 dB = 13</td>
</tr>
<tr>
<td>3.0 Td</td>
<td>+20 dB = 12</td>
</tr>
</tbody>
</table>
Concept Of Variability Pathways

The process design and control strategy design determine the variability pathways. The controller algorithm and tuning determine how much of the variability is rectified. An integrated process and control design approach attempts to effectively redirect variability from key processes to non-critical process streams.

In the pulp and paper industry, it is common design practice to deliver dilution whitewater to many consistency loops via a single whitewater header, as shown below.

![Diagram of process flow]

**Fig 14-5**

This design does not effectively redirect variability away from key process streams (e.g., NC-4). Variability from the HD fiber stream (NC-1) is directed to the whitewater stream, and towards a key process variable (NC4). The benefits of process mixing are compromised since a pathway for high frequency variability from the HD chest to the farthest downstream chest has been created.
Dynam c Representation of Variability Pathways

The mechanism for variability pathways is shown in the following block diagram

HF VARIABILITY

NC-1

NC-2

NC 3

NC-4

LAST CHANCE

HT VARIABILITY

FDLN

PDLN ALLOWS ALL HF NO SE BACK

FROM THE FBER STREAM TO THE WW STREAM & BACK AGAIN

Fig 14-6
Eliminating the variability pathway from upstream consistency loops to the Tank 4 consistency loop disturbances to NC4 will now be limited by the Tank 4 pass filter.

Fig 14-7

14-10
Dynamic Representation of Elimination of Variability Pathways

HF VARIABILITY

NC-1b

NC-2

NC-3

NC-4

HF VARIABILITY

DISTURBANCES LIMITED BY TANK 4 LP FILTER & WW TANK LEVEL CONTROL

TANK 4 NC4

Fig 148
Tuning Strategy - How To Select Lambda

The process design and control strategy design determine the variability pathways. The Lambda selection defines how much of the process variability will be redirected. This decision requires a fundamental knowledge about the process objectives of the control loops.

Consider the Stock Blending system shown in the figure below:

![Stock Blending System Diagram](Image)
NC-1, NC-2

These are very important loops. Severe disturbances are expected, but there is much downstream mixing which will attenuate higher frequencies. The best approach is to tune these controllers as fast as possible while limiting resonance to less than an AR of 1.41 (3 dB)

FC-1, FC-2

These stock flows participate in the Stock Blending strategy. Since the primary objective of the Blending system is to produce a uniform finish, both flow loops should have the same Lambda value. The flows will track the cascaded setpoint (from the level controller) identically producing a constant ratio. In addition, since these are inner loops, they must be 5 to 10 times faster than LC-1 to avoid inner/outer loop interaction. It may also be desirable to tune FC-1 and FC-2 5 to 10 times slower than NC-1 and NC-2 respectively so that the consistencies can more effectively maintain setpoint

LC-1

This controller should be tuned as slow as possible, while ensuring that no stock ends up on the floor and adequate level is maintained for mixing and pumping. The tank should act to buffer the effect of downstream variability on upstream processes. Tight level control is not nearly as important to product variability and machine efficiency as is tight stock consistency and refiner control

NC-3

This is the final consistency control point. Hopefully the variability is reasonably low and random in nature. The minimum variance criterion can therefore be considered on this loop
EnTech Seminars include process control and dynamics training for all people who are affected by process design, control or instrument performance. With a focus on minimizing variability, EnTech training provides understanding of the issues and techniques needed to correct problems or affect performance. All training is based on case studies, examples from processes and providing usable methods.

**PROCESS CONTROL FOR ENGINEERS**

The PCE I & II courses have become pre-requisite training for process control, control systems or process engineers in many mills. They provide a hands-on learning experience that many have said is unmatched by any other courses they have ever taken. The usable information and the exposure to techniques that make a difference in a person's effectiveness will have an immediate impact on process performance.

PCE I - Control Engineering and Tuning Fundamentals is designed to give the participant working knowledge of PID, feedforward, decoupling and deadline compensation in a process setting. Both classical and modern control tuning methods, including IMC and robust control, are covered. A thorough grounding in Lambda tuning is practiced. Dynamic simulator exercises teach participants how the theory really applies.

PCE II - Minimizing Variability Through Control provides working knowledge of the control techniques required to minimize the effects of random noise and disturbances. The use of time series analysis techniques, including spectral analysis and autocorrelation functions, provides usable tools that can be directly applied to process problem solving. Identification of process dynamics, IMC and robust control concepts are strengthened. This course builds on the theoretical and simulation experience gained in PCE I.

<table>
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<tr>
<th>PCE I</th>
<th>January 27 - 31</th>
<th>March 10 - 14</th>
<th>June 2 - 6</th>
<th>September 15 - 19</th>
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<tr>
<td>PCE II</td>
<td>February 3 - 7</td>
<td>March 17 - 21</td>
<td>June 9 - 13</td>
<td>September 22 - 26</td>
<td>November 10 - 14</td>
</tr>
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**PROCESS DYNAMICS FOR ENGINEERS**

The PDE I & II courses are allowing a greater number of process professionals, in their design, operating- or optimization roles, to move to a new level of capability in being able to improve process performance. Gain a better understanding of the dynamic principles that result from process design, equipment selection, control, tuning and operating maintenance. Understanding the dynamics allows improvements to be made.

PDE I - Practical Concepts and Process Variability investigates the theory behind the dynamics governing process behavior. Topics which are covered through discussion, demonstration and simulation include mixing and agitation, process design, equipment, measurement, sampling, control, tuning and instrumentation.

PDE II - Dynamic Simulation and Modelling provides understanding of how to use the physical laws that govern process dynamics for building dynamic simulators using VisSim. The duplication of process functions allows the testing of design changes, control strategies, tuning or operating changes. PDE II requires knowledge of math and physics, PDE I is not a pre-requisite.

<table>
<thead>
<tr>
<th>PDE I</th>
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<th>April 7 - 11</th>
<th>September 8 - 12</th>
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<tr>
<td>PDE II</td>
<td>April 14 - 18</td>
<td>December 1 - 5</td>
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</tbody>
</table>
MODERN LOOP TUNING

Loop dynamics and Lambda tuning training for pulp & paper instrumentation specialists. Measurement, sampling, filtering, dynamics testing, response evaluation, PI controllers, loop tuning, interactive loops and area tuning strategies are included in the topic list. The MLT I & MLT II courses help technicians and engineers overcome many of the deficiencies inherent in current loop tuning practices and to begin a program aimed at improving loop performance, manufacturing efficiency and product uniformity.

MLT I - On-site Course for groups of up to 6 participants per instructor. Course combines classroom setting instruction with hands-on process dynamics and tuning exercises on the mill processes. Practical examples are discussed and simulation is used to demonstrate a variety of commonly occurring issues. Time series analysis techniques and process variability issues are introduced.

MLT II - Registration Course he d in EnTech's training center or at a mill chosen site. Uses computer based dynamic simulation of the actual mill processes for the hands-on exercises described above.

<table>
<thead>
<tr>
<th>E &amp; I Technicians</th>
<th>Instrument Specialists</th>
<th>Instrumentation Engineers</th>
</tr>
</thead>
<tbody>
<tr>
<td>MLT I</td>
<td>On-site scheduled to suit mill requirements</td>
<td>MLT II adds one week of hands-on training with the EnTech Toolkit</td>
</tr>
<tr>
<td>MLT II</td>
<td>February 18 - 21</td>
<td>April 15 - 18</td>
</tr>
</tbody>
</table>

PROCESS CONTROL APPLICATIONS

The PCA II, III & IV courses use the basics learned in PCE I or PDE I to continue the learning and investigation process in a specific area of the mill. These courses are based on the experience and knowledge that EnTech has gained in our process evaluations. Commonly occurring problems are discussed and explored, participants are encouraged to bring specific mill problems for analysis.

PCA II - Paper Machine Applications & Problem Solving takes a close look at process areas from the HD tower to the reel. Subject areas include consistency control, refining, headboxes, cleaners, savesalls, drying, basis weight control and moisture control.

PCA III - Bleach Plant Applications & Problem Solving deals with dynamic and kinetic relationships and variability issues that occur in mill bleaching areas. High fidelity dynamic simulators of various bleach plant designs are used as the basis for problem discovery and solution development in this sensitive area.

PCA IV - Boiler & Furnace Applications & Problem Solving explores the power house area and investigates the very different problems encountered by these units. A set of high fidelity dynamic simulators provides the basis for exercises and discussion.

<table>
<thead>
<tr>
<th>Control Engineers</th>
<th>Process Control Engineers</th>
<th>Instrument Engineers</th>
<th>Systems Engineers</th>
</tr>
</thead>
<tbody>
<tr>
<td>PCA II (Paper Machine)</td>
<td>June 26 - 20</td>
<td>PCA III (Bleach Plant)</td>
<td>April 28 - May 2</td>
</tr>
</tbody>
</table>

Request Form

Fax to: 416 626-7691 or 404 577-0063

Name __________________________ Title __________________________ Company __________________________

Address __________________________ Telephone __________________________ Fax __________________________ email __________________________

I am interested in attending a course, please fax ☑ / send ☑ me more information about

PCE I ☑ PCE II ☑ PDE I ☑ PDE II ☑ MLT I ☑ MLT II ☑ PCA I ☑ PCA III ☑ PCA IV ☑

More information is available through our website at www.entechcontrol.com
Competency in Process Control - Industry Guidelines

(Version 1.0, 3/94)

This is a summary of a generic set of recommendations aimed at improving process operations by managing process and product variability. It suggests a minimum set of standards for human awareness, knowledge, and skills in the various aspects of process control. Newer design approaches at the initial stages of a process design and start-up and ongoing operations require attention.

1.0 Competitive Marketplace Background

The global market for manufactured products continues to focus on product quality and uniformity. Today, product uniformity and specification are highly regulated, and paper products can be rejected when they deviate from these standards. The variability of the products adversely affects the operation of the customer's secondary manufacturing processes, such as printing presses. There is a need to pay more attention to the quality control of personnel engaged in the design and manufacture of products, particularly in the context of automated control systems. The use of Centa (TM) Control Engineering has provided substantial improvements in the process variables by enhancing the control of the final product. In many cases, this has led to improved control loops, which are a major cause of instability in process control.

1.1 Mill Audit Variability Results

To date, only about 20% of automated control loops surveyed in mill audits have been found to actually reduce process variability. This is due to control valve non-uniformity and control equipment design. The remaining 80% of loops were found to increase variability. Of these, some were found to oscillate due to control valve non-uniformity or equipment design issues. To address these, EnTech has developed a program to review and modify control systems, providing a significant improvement in the overall control of process equipment. Procedures for procedure testing are now being implemented, leading to increased process stability and efficiency.

1.2 The Knowledge Gap and The Control Engineer

Many control engineers are aware of the knowledge gap in design and maintenance practice. This has occurred in spite of the fact that most engineers study control theory. The current knowledge of the mathematics of control is not up-to-date, and they often forget. The teaching of control theory in undergraduate programs is a weakness, with a lack of foundation for future process procedures and practice. New graduate engineers are not adequately prepared in the field of control. The industry is aware of the need for new knowledge in design and control theory. The need for a well-trained professional is critical. There are very few people who can manage the process design and start-up adequately. Awareness of the fundamentals of process control is necessary. The industry needs to ensure that knowledge gaps are minimized.
1.3 The Fully Qualified, Competent Control Engineer – A Role Model

To become a "Fully Qualified" control engineer is not a formidably task requiring years of experience, much determination and a very strong focus. It is a serious amount of work that can be achieved over one’s career. The authors’ own experience is evidence of this. Even after graduation, the challenge continues, as more knowledge is gained, and the engineer must be open to learning and adapting to new technologies.

1.4 Impact on Industry Thinking

Not only the industry, but also the education sector, have seen a shift in the way engineering is taught and practiced. There is a growing emphasis on developing a more hands-on approach to education, with a focus on practical problem-solving skills.

1.5 Advances in Process Control – Background

In the last decade, modern control systems, such as the Lambda Tuning Method (LCM) and Interact Control (MC), have been developed. These systems provide a more intuitive and user-friendly approach to control engineering.

2.0 Summary of Control and Dynamic Problems Encountered

Based on years of experience, the types of problems encountered in control engineering include:

- Excessively slow system response
- Lack of system stability
- Gaining and set-point difficulties
- Non-linear relationships
- External disturbances
- Unpredictable behavior

To address these issues, control engineers must be trained in advanced control techniques, such as adaptive control and robust control.

2.1 Purpose

This document sets out to define the elements of the control system and the role of the control engineer. It also highlights the importance of continuous training and education for control engineers.

2.2 The Manufacturing Team

The manufacturing team consists of many people from diverse disciplines, including engineers, technicians, and operators. Each member has a specific role and is responsible for a particular aspect of the manufacturing process.
3.0 Awareness, Skills and Knowledge by Manufacturing Team Task

Having defined the control engineer as the person where "the buck stops" on the subject of pup and paper process, a ty and control, what does a systematic breakdown of this core control eng nee ng matter into important elements which are relevant to the plant does not have been uncovered during the audit tests. These in turn are classified by the chief sk or knowledge needed by each person or task with whom a pup and paper manufacturing framework interacts.

Definitions - Skill Levels: Awareness, Skill, Knowledge

Three gross levels of know edge or skill are: a) basic b) advanced. Each category can be rated n terms of the depth of sk required (e.g., awareness, skill, knowledge) using a scale of 1 to 5.

Awareness: This s mp represents the condition of understanding the situation. Awareness does not require the knowledge and skill associated with a task.

Skill: This level is achieved when an individual can perform the task, such as a process operator or a control engineer. The individual is able to execute the task with a level of expertise that is considered acceptable in the industry.

Knowledge: This level signifies deep understanding and mastery, often through years of experience. Individuals at this level can explain the underlying principles and theories, and can forecast outcomes based on their knowledge.

3.1 The Bump Test

Anyone working with a control system must be familiar with bump tests to check the "health" of the system. An example is a given flow control system. A bump test should be done at a 5% response time, which is less than ten seconds otherwise it is considered too high.

3.2 Deadtime

Deadtime is the most important factor affecting process control. Deadtime is the time it takes for a process variable to change in response to a disturbance. It is measured in seconds.

3.3 Process Mixing

Process variables can be reduced by using two methods: process control, which can only reduce ow frequency or ow rate, and process mixing, which can reduce hgh frequency or fast var ab ty. The two

are complementary. However, m audits have found that the actual m x ng character st cs of tanks are often far less effective than predicted by the designed residence time. The mathematical model of the process should be used to predict the effect of mixing on the tank operation and thus determine the impact of variables that influence the process.

3.4 Air Entrainment

Air entrainment can occur in many processes. Adequate surface area and turbulence can help reduce entrainment. The amount of air entrained can be calculated using the following equation:

\[ \text{Air Entrainment} = \frac{\text{Air Flow Rate}}{\text{Surface Area}} \]

3.5 Control Valve Selection and Sizing

Control valves are used to regulate the flow of a process medium. They are sized based on the maximum flow rate and the pressure drop that the valve can handle. The selection of a control valve must consider factors such as the medium to be controlled, the temperature, and the pressure drop that the valve can handle.

3.6 Control Loop Interaction

Control loops are interconnected and influence each other. It is important to understand the interactions between control loops to ensure optimal system performance.

3.7 Control Algorithm Selection and Tuning

A control algorithm is the logic that determines how the control system should respond to changes in the process. There are several types of control algorithms, such as PID, optimizing, and adaptive control. The choice of algorithm depends on the specific process and the desired performance.

3.8 Control System Selection

When selecting a control system, it is important to consider factors such as the system requirements, the level of control necessary, and the cost of the system. The selection process should consider the available technologies and the specific needs of the process.
3.10 Anti-Aliasing Filters

Antialiasing filters are necessary on each process measurement to prevent crosstalk from reducing process variability due to a lack of system integrity. The amount of filtering needed is a function of the sampling rate [3] and everyone involved in control or instrumentation should know how to determine these values.

3.11 Digital Transmitters

The advent of digital transmitters has moved the issue of sample rates, anti-aliasing, and digital signal processing into the design and implementation of the transmitter itself. Transmitter selection now involves knowledge of these issues [3] and everyone involved in control or instrumentation should know the criteria for selecting transmitters.

3.12 DCS Display Ranges

Often, process operators are unaware of process variations between process variable ranges. They may need to be trained in the relationship between process variable ranges and digital signals.

3.13 DCS Report-by-Exception

The advent of the DCS has brought report-by-exception as a method of transmitting process data in a more efficient manner.

3.14 Faulty Control System Design and Coding

With the advent of digital control systems, many people have become involved in designing and coding control strategies. A solid understanding of control system software is necessary to ensure that the system is reliable and efficient.

3.15 Process Trouble-Shooting Using Time Series Analysis

Troubleshooting process problems using time-series analysis requires knowledge of the heart of the process, understanding of the nature of the variables, and the capabilities of the control software.

3.16 Spotting/Reducing Variability to Optimize Pulp and Paper Manufacturing

Mak ng unform product efficiency one of the goals of production involves understanding the factors that affect the variability of the process. This requires knowledge of the control system and the capability of the software to identify and reduce variability.

4.0 Summary - Skills Matrix

The matrix shown on page 16 of the recommendations is a summary of the skills required for successful control systems design.

Table I - The Skills Matrix

<table>
<thead>
<tr>
<th>Category</th>
<th>Proc Des'n</th>
<th>Cont SysSel</th>
<th>Instn Des'n</th>
<th>Cont Des'n</th>
<th>Loop Tune</th>
<th>Ops Staff</th>
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</table>

References and Bibliography

4. EnTech: Competency in Process Control: Industry Recommendations (version 1.0 3.09/94)

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EnTech Process Control Inc.
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Te 404-577-6640 Fax 404-577-0063
Automatic Controller Dynamic Specification
(Summary of Version 1.0, 11/93)

This is a summary of a generic specification which applies to the design and dynamic performance of automatic feedback controllers suitable for use in optimising pump and paper machine process variability. The full document is available upon request.

1.0 Competitive Marketplace – Background

The growing need for manufactured products to be produced efficiently and to a high quality has led to an increased focus on process control equipment and its condition. EnTech Contm Control Engineering Inc. has developed a number of analysis and optimisation tools for pump and paper manufacturing where product form fits specifications are now approached 1%. Paper products can be rejected when they deviate from specifications, or when the variability of the product's characteristics affects the performance of the customer's secondary manufacturing systems such as corrugated board press. These changes have shown that product variability is caused mainly by the combined behavior of the many upstream process variables. In many cases, it is possible to trace these causes to individual control loops. One of the most important factors affecting the impact of process variability on manufacturing efficiency is which small and significant improvements in the process variability of key variables can have a major impact. Audits have identified that the cycle behavior of automatic control loops is a major cause for destabilizing product on the pump and paper industry.

To date, only about 20% of automatic control loops surveyed were found to actually reduce process variability. Some 20% were found to be ineffective directly due to control valve issues and another 30% were found to oscillate due to control parameter tuning. It is recognized that control tuning is to a great extent a human training issue (a separate issue from that of control design). Most people are only familiar with the original control tuning methods known as Quarter Amplitude Dampening (published by Ziegler and Nichols in 1942) or, they just “tune by feel.” These methods often produce processes which tend to oscillate and are unstable, reducing product variability. By nature, trial and error methods accept the capabilities of existing control systems without questioning. For this reason, control systems have not advanced significantly in terms of performance over the years despite the major advances in control system technology from pneumatic control to electronic control with 64-bit microprocessors. In fact, the trend has been for each new generation of control systems to match the performance of the previous generation on most of the key parameters of power, performance, and flexibility. For this reason, most digital control systems are “analog like”, in that they do on what analog control systems could do in the past. What of deadtime compensation, vector mode control, feedforward control? On the other hand, as digital control systems have become prevalent in sampling rates have been relaxed and in some cases, in their operating error variables. The result has been implemented digital control systems which have been used to improve product uniformity for reasons of efficiency and accuracy in the pump and paper industry.

In the last decade, modern model-based methods such as Lambda Tuning and Iterative Model Control (IMC) have been used extensively in pump and paper mills with considerable success in reducing variability. The use of these methods has “pushed” existing control design to the limit. For example, it has been very difficult to discover that the recommended tuning parameters based on these methods could not be...
entered into some commercially available controllers because they were outside the controler mits or that a particular feature of the modern control solution could not be implemented because it was not part of the available controller algorithm.

Summary of Controller Problems Encountered

The types of problems encountered included:
- Slow control sampling rates
- Lack of anti-asymmetric filters
- Lack of control structure flexibility
- Too narrow parameter ranges
- Lack of feedforward control
- Lack of velocity mode control
- Lack of deadtime compensation
- Lack of adequate controller documentation

Controller Specification - Purpose

Controllers provided by pulp and paper industry suppliers provided a wide range of equipment needed for strubuted control systems (DCS's), such as controllers, programmable logic controllers (PLC's), variable speed drives, computer control systems, and more. A automated control system was designed to meet the needs of the majority of pulp and paper mills, process control problems needed for modern control systems. The specific control systems needed for any application, where general-purpose process control systems attempted to maintain an automation feedback control system or provide a feedback to the controller. This controller is needed for the proper functioning and assessing the capabilities of control equipment and provides a design guide for control equipment suppliers.

2.0 The Pulp and Paper Control Problem

Various pulp and paper mills use PID controllers for "analog type" or continuous control (those used digital control with adequate fast sampling rates) with 4 to 20 mA outputs to control valves and other actuators. Derivative action has been turned off in 98% of cases. Most commercial controllers have controller gain adjustments ranging from about 0.1 to 10. A number of important control loops are motor driven or "velocity mode" type actuators such as electric valves, referentially connected and headbox vertica slopes. These loops areNature use discrete or velocity mode a gor thms with d g t a output log c to hand e increase/decrease motor action on. Add tion, some loops have extended dead time and require deadtime compensation a gor thms. Other loops require feedforward control in order to achieve acceptable dynamic performance.

3.0 Specification

3.1 Bump Tests

The controller must have a "manual" mode in which feedback control is suspended. Should safety considerations preclude using normal operation as equivalent "test mode" will be used as safe for use under the supervision of a qualified person.

It will be possible to initiate small step changes in the (0.1% to 10%) controller output under the control of the user. These step changes will be as near as possible to pure steps.

To be used with the controller output and the process variables as seen by the controller at the controller update frequency. This could be done at times vary as save software ut ty or a as g n e d g t a to ana og converters having at east 12 b t resoluton.

3.2 Gain Scheduling as a Function of Error

Whereas gain scheduling is a useful feature which control systems should have, it is a means of scheduling the controller gain ("Kc", for SA standard form and a cass ca form PD) as a function of controller error. The provided天鹅 a ow a pre se nctg g a n s tab e for m m mum var ance control to be used as long as the absolute error is b w o ow ve. Above this value, a different gain is used for more aggressive control. Care should be taken to avoid the pos t b ty of a m t c y c e between these values. For a par a path PD m p ementat on the same method of gain scheduling must be arranged for a three parameters.

3.3 Control Structure - Controller Filter

To permit the MC type control structures which may require the existence of a process zero (8 value) with a controller error, there could be a first order filter implemented in the controller. The filter can be one of the controller error (before the PD) or after PD. The filter must be properly tuned (bumpless transfer). The transfer function of the filter is given by:

\[ G_F(s) = \frac{1}{(\tau_F s + 1)} \]

The time constant $\tau_F$ is to be selected between 0 seconds and 60 minutes.

3.4 Control Structure

The available control structures used are those of a configuration, controller, and PD, PID (both real and digital compensation) theorem.

Each of these will have a series of terms (see 3.3 above).

The derivation for $T_D$ will be normally a tenth and no more than one eighth of $T_D$.
3.5 Alternative PID Structures

Alternative structures such as "D" on PV and "P" on PV are offered; they must only be as optional. There must be an alternative driven by controller error.

3.6 PID Parameter Ranges

Parameters \((K_C, T_R, \tau, K_D, K_T)\)
These must be adjustable continuously (not in steps) and calibrated to at least 2 significant figures.

\[
\begin{align*}
K_C & = 0.001 \text{ to } 20 \% \text{ (output)/span}, \text{ (ISA standard or custom controller PID)} \\
T_R & = 1 \text{ second/step to } 10 \text{ hours/step} \\
T_D & = 0 \text{ seconds to } 60 \text{ minutes,} \\
K_I & = 0 \text{ to } 50\% /\text{sec}, \\
K_D & = 0 \text{ to } 1500 \% /\text{m units} \\
\tau & = 0 \text{ seconds to } 60 \text{ m units}
\end{align*}
\]

3.7 Process Variable Filter

The process variable will have a filter prior to the error summation junction of the form

\[
G_{PV}(s) = \frac{1}{(\tau_P s + 1)}
\]

The time constant \(\tau_P\) should be adjustable from 0 to 10 m units.

3.8 Feedforward Control

A dynamic feedforward controller on \(w\) is provided as per

\[
G_{CF}(s) = K_{CF} \left( \frac{\tau_{Lead} S + 1}{\tau_{Lead} S + 1} \right) e^{-sT_{FF}}
\]

The feedforward controller output \(w\) is added to the output of the feedback controller due regard for bumpless transfer, 0 ms and reset windup. The user selectable adjustments are:

\[
\begin{align*}
K_{CF} & = 0 \text{ to } 100, \tau_{Lead}, \tau_{Lead} \text{ 0 to } 10 \text{ m units} \\
T_{FF} & = 0 \text{ to } 10 \text{ m units}
\end{align*}
\]

3.9 Setpoint Filter

There will be a proportional filter to filter the setpoint through the form

\[
G_{SP}(s) = \frac{1}{(\tau_{SP} S + 1)}
\]

with a user selectable time constant from 0 to 100 m units.

3.10 Deadtime Compensation

There will be a deadtime compensation as a filter in the form

\[
G_D(s) = \frac{(\tau s + 1)}{K_p(\lambda S + 1 - e^{-sT_D})}
\]

Parameter values will include:
\(\tau\) 1 second to 10 m units, \(K_p\) 0.01 to 50,
\(\lambda\) 2 seconds to 30 m units, \(T_D\) 0 to 10 minutes

3.11 Controller Output Integrity

The controller output should not be a steady state change by rate or other means.

3.12 Sampling and Controller Execution Rates \((T_s)\)

Analog signal sampling and controller execution should be at least 10 times faster, and no slower than three times faster, than the process time constant. Process times constant faster than 0.5 seconds need not be considered in the "general" process control problem. Sampling can be slowed down to one sample per second when process times are equal to three seconds or slower.

Analog signal sampling and controller execution should be at 0.1 seconds/sample. However, it can be slowed down to between 0.3 seconds for time constants of 1 second \((T_s = \tau/3)\) to as slow as 1.0 second/sample for time constants of 3 seconds or longer.

3.13 Anti-Aliasing

Anti-aliasing filters shall be present at every point of signal sampling, from raw analog input to the point of use of the feedback signal at the controller. At each point of sampling, the anti-aliasing filter shall provide a minimum of 12 dB of attenuation at the Nyquist frequency of the next sample.

3.14 Bit Resolution

The minimum acceptable bit resolution for both inputs and outputs is 12 bits. Report by exception reporting should never be used inside a control loop.

3.15 Motor Driven Actuators - Velocity Mode Control

A velocity mode \(P\) a thm should be available of the form

\[
\Delta u_i = K_e(e_i - e_{i-1}) + K_r e_i
\]

- \(K_r\) \((e_i - e_{i-1}) + (K_r T_e) / T_r\)

The "time duration" of output hander should execute at a minimum update time of 0.05 seconds. Preferably, it should not execute at such a rate which would overload the processor. To achieve a quantization equivalent to about 12 bits of resolution on or 1 4000 Parameter ranges will be as defined for PD parameter ranges above. Output imbalances and windup issues must be considered.
3.16 Controller Documentation

Controller documentation will include the following:
controller transfer function for each option and signal path,
parameter units and, parameter ranges available.

Nomenclature and Symbols

- **a**: Or 1 exponent on integrator (0-none, 1 present n transfer function).
- **β**: Ead time constant (+ve or -ve).
- **e_k**: Setpoint error.
- **G_C(s)**: Control transfer function.
- **G_F(s)**: Fter transfer function.
- **G_FF(s)**: Feedforward transfer function.
- **G_FF(s)**: Process variable transfer function.
- **G_P(s)**: Process transfer function.
- **G_FSP(s)**: Setpoint fter.
- **l**: Integral a gor thm.
- **k**: Current time index for discrete time difference equat on.
- **K_C**: Controller gain.
- **K_FF**: Feedforward gain.
- **K_P**: Proportional gain.
- **K_D**: Proportional-Derivative gain.
- **K_PI**: Proportional-Integral gain.
- **K_PID**: Proportional-Integral-Derivative gain.
- **T**: Process time constant.
- **T_FF**: Feedforward deadtime compensation.
- **T_PV**: Feedforward lead/lag time constants.
- **T_d**: Deadtime.
- **T_D**: Deadtime (standard and custom PD).
- **T_PV**: Deadtime (standard and custom PD).
- **T_R**: Reset or integrator time.
- **T_SP**: Setpoint time constant.
- **ΔU**: Current steady state change in controller output.
- **ΔU_b**: Damping coefficient.

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Digital Measurement Dynamics – Industry Guidelines

(Version 1.0, 8/94)

Keywords: sensor, transmtter, samp ng, samp ng rate a as ng, f ter ng, anti-a as ng, de ay

Note: n ths document the words sensor and transmtter are used interchangeably to mean a measurement device which makes a ca brated s gna avaible e

1.0 Competitive Marketplace Background

The going market for manufactured products continues to focus on product quality and uniformity and more attention on s being paid to process control equipment and ts condition. EnTech Control Engineering Inc. has specified in the analysis and optimization of pulp and paper manufacturing where product uniformity specifications are now approaching 1%. Paper product can be rejected when it deviates out of specified limits or when the variables character of the product adversely affect the operation of the customer's secondary manufacturing such as a packaging press. Audits have shown that product variability is caused many by the combined behavior of the many upstream processes. Audits have denoted that the cyclical behavior of automated control loops is a major cause for destabilizing production in the pulp and paper industry and that the measurement equipment

Digital Measurement Problems Encountered in Mill Audits

- slow d g ta samp ng rates,
- nabl ty to change samp ng rates,
- s gna a as ng
- n o a t a as ng fters,
- nabl ty to change anti-a as ng fter settings
- ow b t resolution

- non near filter ng & s gna condit on ng,
- excess ve y s ow d g ta s gna transm ss on
- d g ta transm tter tme deays
- nabl ty to access data v a propretary d g ta protoco
- nabl ty to access the measured signa ahead of signa filtering
- nadequate s gna condit on ng documentat on

Digital Measurement Dynamics – Recommendations – Purpose

Control in the pulp and paper industry is accomplished by a wide range of d g ta equpment nc udng d str buted control systems (DCS's) s ng e oop contr ol ers, programmable logic controllers (PLC's), v abe speed de vices computer control systems and more. Data archiving systems. A of these systems common cate d g ta y in the near future, most transmitters and contro valves w a so common cate d g ta y w th a of the above equpment v a the F edbus protocol. Thus there is an even greater need to ensure that any problems related to d g ta s gna process ng are avo ded.

These guidelines define the minimum requirements needed to provide the pulp and paper industry with measurements which are needed useful as m s tr ve towards more efficient manufacture of gher un form ty product. Th s document s a med at pr ovid ng d g a des gn gu de for sensor transm tter and d g ta s gna transm ss on equpment n order to ensure that the des gns are capable of meeting the needs of process control tue shee t ng and data acquisit on of spec a y at a tme when there are ncresas ng y urgent needs in the pulp and paper industry for greater product uniformity and operating efficiency. Th s document has two intended purposes as a gu de ne for pulp and paper companies when specifying or buying process measurement and data transmission equipment and as a des gn gu de for equpment suppl ers to the pulp and paper industry.
2.0 Pulp and Paper Control and Digital Measurements Problem Definition

The Analog Signal

Until recently most sensors and transmitters were of analog design. Analog sensors obey Newton's law governing physical systems and can be modeled as linear, first-order systems with a dominant time constant. Most transmitters and sensors produce an analog output signal, which may be filtered to remove high-frequency components. The bandpass filter is often used to filter out unwanted frequencies and to shape the signal for further processing. The output of the bandpass filter is then amplified and sent to the control system.

Digital Signals

These signals are essential for modern control systems. Digital signals are discrete and can be easily processed by computers. They are also immune to noise and interference. However, digital signals must be converted to analog signals for display and control. This conversion is done using analog-to-digital (A/D) converters. The A/D converter takes an analog signal and converts it into a digital signal that can be processed by a computer. The digital signal is then converted back to an analog signal using a digital-to-analog (D/A) converter. The Nyquist frequency is based on Shannon's sampling theorem, which states that a signal can be perfectly reconstructed from its samples if the sampling rate is at least twice the highest frequency component of the signal. This is known as the Nyquist rate.

Signal Requirements for Control

Control of a motor system as in P&ID is an essential part of the control system. The presence of non-linear elements becomes very difficult to implement in a control system. However, there are a number of important properties of a digital system that must be considered. The Nyquist theorem is a fundamental concept in signal processing that states that a signal can be perfectly reconstructed from its samples if the sampling rate is at least twice the highest frequency component of the signal. This is known as the Nyquist rate.

Signal Requirements for Process Troubleshooting

Process troubleshooting requirements are more demanding than those for control, as the nature of the problem cannot be pre-defined. In this case, troubleshooting the process can be an iterative process of trial and error. The Nyquist theorem is a fundamental concept in signal processing that states that a signal can be perfectly reconstructed from its samples if the sampling rate is at least twice the highest frequency component of the signal. This is known as the Nyquist rate.

3.0 Digital Signal Properties

Digital信号s have two important properties and per od c samp ng ife ther of these are bad y chosen they can make the s gna of the use.

Quantization

A digital signal is quantized by some form of analog to digital converter (A/D) in a process. The resolution of these devices is limited. An A/D with 10 bits can resolve the signal to 1/1024. A 12-bit converter to 1/4096. This quantization limits the resolution of the signal, and hence the system's ability to accurately represent the analog signal in the control system.

Sampling Rate

The act of sampling a signal at some rate results in the bandpass dth of the signal to an upper frequency known as the Nyquist frequency. The Nyquist frequency is based on Shannon's sampling theorem, which states that a signal can be perfectly reconstructed from its samples if the sampling rate is at least twice the highest frequency component of the signal. This is known as the Nyquist rate.

![Figure 1 - Example of Signal Aliasing](image)

Signal Aliasing

If the original analog signal contains frequency content beyond the bandpass dth of the sampler (beyond f_n), then the signal is aliased. As the signal is aliased, the frequency components above the Nyquist frequency (f_n) are folded back into the lower frequency range. This is known as the Nyquist frequency, and it is a fundamental concept in signal processing that states that a signal can be perfectly reconstructed from its samples if the sampling rate is at least twice the highest frequency component of the signal. This is known as the Nyquist rate.
samped. The only solution is to prevent signal aliasing from occurring.

![Diagram of signal processing]

**FIGURE 2 - THE IMPACT OF SIGNAL ALIASING ON CONTROL**

**Random Variability**
The previous example, with S gna as ng used s neg e s ne waves to illustrate the point. In practice, process s gna s are usually affected by noise and change over time. Experiments have shown a high degree of accuracy in various settings. A key metric is the impact of a second on the process, and this impact is shown to be greater in certain situations. Each of these techniques is key to measuring ng and understand ng var bel by and its impact. Then y ng pr npe of these techn qua es s that any random s gna c can be decomposed into its harmonic content, hence many s ne waves ex st ng and each at a different frequency. The impact of an as ng can be read y understood. Each harmonic with s faster than the Nyquist frequency w produce its own unique signal as ng. Ths phenomenon is known as "flying over" and involves the corrupt of the power spectrum as a result of a ng

**Anti-Aliasing**
There are two ways to prevent as ng. One is to samp e faster than the fastest frequency component of the input data. These are known as high frequencies, and they are not. The other is to insta an ant as ng filter before the samp ng device. This way, the ng out of the s gna content wch ch s faster than the Nyquist frequency of the samp er at the chosen samp ng rate. Hence preventing as ng from occurring. The ant as ng filter should provide a minimum of 12 dB of attenuation at the Nyquist frequency. Ths can be provided by a filter order, which pass filters of time constant T/f. 1.7 for a can be provided by an aor met c average, which is T/avg ng tme. The 12 dB recommendation represents a bare minimum requirement. Where poss b e more effectve ant as ng shou d be considered by aphem y ng a filter with a sharper cut-off and more attenuation at the Nyquist frequency.

**Sub-Sampling**
ng data contro systems not on y are ana og s gna s samp ed to produce d gta vers ons of the input s gna s, but the d gta s gna s are usually samp ed many times later. For instance, in modern DCS systems, the output sub-systems do the A/D conversion may run at say 100 mwh en the controller er or e process may have var ab e samp ng rates depend ng on the ndv e needs of each contro er. Typica contro er ection rates may be 1 second per s. ng can occur between the input sub-system and the controller er even if as ng was prevented at the input sub-system. **This is the final sampling rate which governs aliasing.**

ng DCS s there are many other uses for the input s. For instance, the process va uses from the contro ler w be samp ed aga n to update the operator's conso e. Typ cay th s s done at a sower rate than the controller update frequency. The mew de arch vng system w also samp e the same data at a much sower rate. Typ cay on e per mume. Other contro oops may samp e the data at a sower samp ng rates for feedforward compensator. Al of these operations are subject to data a as ng Anti-aliasing filters are required before each sampling operation and must be matched to the samp ng rate. This is usually neglected in most systems today.

**4.0 Digital Signal Transmission**
Signals can be transmitted in a variety of ways, including RS-232, RS-422, Ethernet, LAN's, and other communication protocols. The goal is to transmit signals reliably and efficiently. A proper design should be transparent to the contro ler and troubleshoot app. cat ons by prov ing 100% rela e on met data transfer. It means that each data samp e is transferred at the agreed-upon samp ng rate wout the intro duct of excessive e tme de ays. (say ess than 5% of samp e per od)

**Report-by-Exception**
To prevent data from overloading many devices, an exception report is provided on a regular basis, which can be updated at any time. Often, deadbands are set to 1%. This seriously limits the signal quantization and should never be used for control or troubleshoot.

**5.0 Digital Processors and Transmitters**
Digital sensors and transmitters are becoming more prevalent due to their increased ease of use and are being used more frequently. This is because they provide many advantages, such as increased reliability and reduced maintenance costs. Often, it is more cost-effective to use digital transmitters, which can be configured to provide more detailed information than traditional analog devices. Digital transmitters are also more durable and can handle a wider range of conditions. Some digital transmitters can even be configured to provide real-time data, allowing users to monitor and control processes in real-time. These devices are essential for modern industrial applications, and their use is expected to continue growing in the future.

Some digital transmitters are sensors and transmitters, which can be used to measure various parameters, such as temperature, pressure, and flow. These devices are typically connected to computer systems, which provide a range of benefits, such as easier data management and analysis. Digital transmitters are also more accurate and reliable than traditional analog devices, which can be affected by noise and other factors. Overall, digital transmitters are an important tool for modern industrial applications, and their use is expected to continue growing in the future.

**7.0 Conclusion**
Digital signal transmission and processing are becoming increasingly important in modern industrial applications. Digital transmitters, which are used to measure various parameters, are becoming more prevalent due to their increased ease of use and are being used more frequently. These devices provide many advantages, such as increased reliability and reduced maintenance costs. Digital transmitters are also more durable and can handle a wider range of conditions. Some digital transmitters can even be configured to provide real-time data, allowing users to monitor and control processes in real-time. These devices are essential for modern industrial applications, and their use is expected to continue growing in the future.
6.0 Fieldbus Implications for Digital Sensor/Transmitter Design

Once F e dibus has become a ready process control system architecture undergoes a major change on users have access to the source of measurement information at the transmitter. However, each user has different needs with regards to sampling rate and data integrity. Control requires anti-aliasing signal processing, anticipating every 0.3 seconds typically. The operator's console may require anti-aliasing data every second, but one would think the measurement's relevant on the screen. The trend package may want anti-aliasing data every 1.0 seconds. The -40 decades magnitude may want the anti-aliasing smear every 10.0 seconds, but on a short burst extending over a few minutes due to uncontrolled time and variable these sampling rate, a transponder may so want access to the raw sensor data, it should be regularly sampled.

7.0 Sensor/Transmitter Requirements

7.1 Fieldbus Multi-Channel Design

Cons dcr ng E f e dibus sensor/transmitter designs and the needs and requirements stated are as follows: the sensor/transmitter design for the future should not include the following features:

Field Measurement Verification Capability

The ability to interrogate and verify the integrity of a measurement is of paramount importance. It often happens during process troubleshooting, attempting to verify the integrity of a measurement, a problem with the transducer must be taken to verify the measurement by comparing the raw and the returned signal. This comparison, which represents the difference between the raw and the returned signal, must be done accurately.

1. An analog output signal (4 to 20 mA or 1-5 V) which has not been filtered or distorted, and which represents the measurement as near as a calibrating form as possible. The signal can only be obtained from an inherently analog sensor.

2. A non-linear detector or sensor should provide an analog output (4 to 20 mA or 1-5 V) of the signal at the fastest sampling rate available.

3. There should be a FieldBus connector available so that any of the FieldBus signals can be assessed as we.

Fieldbus Data Access

Data access on the Fieldbus should be taken into account the multiplicity of users for each measurement. At a different sampling rate with anti-aliasing, the signal can be obtained from a digital signal. However, the data sampled can be filtered or distorted, and which represents the measurement as near as a calibrating form as possible. The signal can only be obtained from an inherently analog sensor.

8.0 Recommendations Summary

The following recommendations apply to most of the digital measurements in the process and paper industry:

- **Sampling Rates**
  - Process Trouble shooting: 10 to 100 ms, seconds/samp e
  - Process Control: 0.3 seconds/samp e or faster, 1 second/samp e can be tolerable for some with time constants of 3 seconds or slower

- **Operator Console**
  - Typ cal y 1 second down to 10 second update

- **Time delays** (Measurement Processing or Data Transmission): Less than 5% of the sampling rate per operating cycle.

- **Anti-aliasing**: A minimum of 12 dB of attenuator on at the Nyquist frequency of each signal: 63 Hz

- **Filtering**
  - Sensor's internal filter should be adjustable beyond the needs of anti-aliasing. The minimum filter time should be the expected process time constant (e.g., a filter for a low-pass filter with a time constant of 10 seconds).

- **Quantization**
  - The minimum acceptable signal quantization is ±160 for 14-bit A/D conversion.

Measurement Data Access

Field Access:

- Each sensor should provide digital access to the analog output port to the nearest possible point in the control room.
- Pure digital sensors should provide a digital output of the raw signal. The signal should be a digital signal to provide access to a sensor outputs.

- Digital Output:
  - Digital sensors should support additional digital signals (e.g., through potential or interface with analog channels). Each channel should provide anti-aliasing to the digital filter to avoid signal distortion.

Nomenclature and Symbols

- A/D: analog to digital converter
- dB: decibel
- DCS: Distributed Control System
- f_s: Nyquist frequency
- PLD: Proportional-Integral-Derivative Controller
- t_c: control time
- t_m: control time
- t_p: control time

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Control Valve Dynamic Specification

(Version 2.1, 3/94)

1.0 Competitive Marketplace Background

As the global market for manufactured products continues to focus on product quality and uniformity, more and more attention is being paid to process control equipment and its condition. EnTech Control Engineering has specified in the analysis and optimization of pulp and paper manufacturing, where product uniformity is critical. In this document, we will discuss the importance of control valves and the impact of process variables on manufacturing efficiency. Improvements in the process variables key values can have a large impact. Audits have found that the cyclical behavior of control loops is a major cause for instability in pulp and paper production. To date, about two-thirds of a control loops surveyed were found to oscillate in automatic mode. Of these, well over half were found to oscillate directly due to the behavior of pneumatic control valves. The undesirable behavior of control valves is hence the biggest single contributor to poor control loop performance and the destabilization of product uniformity.

Control Valve Dynamic Specification - Purpose

This specification is intended as a guide for end users to gauge control valve performance. It is also intended to provide a guide for both control valve purchase and selection. Whereby no means the "latest and greatest" in competitive control valve performance, the specification is intended to fit a void in the industry which has essentially no guidelines in this critically important area.

About Version 2.1

The original EnTech Control Valve Specification was issued in early 1992. Version 2.0 (11/93) [1] was or further to provide the end user with a "how-to" guide for control valve testing. Version 2.1 is a major revision to Vers on 2.0 intended to stress three points which have caused some confusion on. a) The Backlash/Stiction on specification by far the most important part of Version 2.1 and is critical that the process variables move slowly in response to small input step changes (1% and less) indicating that an actual change in the valve Cv has occurred. Valve stem movement alone is not an adequate indicator. b) There is a need to clarify terms used as in both Vers on 2.0 and 2.1 regarding backlash/stiction on. These terms are defined in [2] and result in a phenomenon known as signal deadband in the valve travel. The terms "backlash/stiction" emphasize the mechanical cause of the problem and are used throughout Vers on 2.1 to apply "deadband". c) Finally, ISA Specification S72-13 [3] is not equivalent to Vers on 2.1. The former applies to a test stand procedure for valve positioning only and is not adequate for today's in-situ process control performance requirements. Version 2.1 specifies in-situ control valve performance requirements for the whole valve assembly, under process operating conditions, in order to meet end-use performance and final product uniformity requirements.
2.0 Control Problem Definition - Linear Control

Most control valves are used as final control elements in a process control system. The feedback control loop essentially a near device. The purpose of this feedback is to improve the accuracy of the process. The feedback control loop is a simple device that improves the accuracy of the process. The feedback control loop is a simple device that improves the accuracy of the process. The feedback control loop is a simple device that improves the accuracy of the process.

Definitions:

Process dynamics the way a process variable responds to an input. The system's response is modeled by a transfer function and expressed in terms of the process gain and time constant.

Process Gain the ratio of process variable change to controller output change after a step change in input variable.

Time constant the time required for a first order process to reach 63% of the final change when a step change is made.

Deadtime the time after a step change prior to the start of the process reaction on the process variable.

Linear Dynamics a simple set of dynamic parameters (gain, time constants) that are used to model the process.

Nonlinear Dynamics a set of dynamic parameters that describe the process's behavior at different operating conditions.

Limit Cycle a sustained oscillation in the process output caused by a feedback loop.

3.0 Control Valve Nonlinearities

Control valves have several forms of nonlinearity. The most common are:

- Valve tracking nonlinearity
- Flow characteristics
- Valve wear, maintenance, age, installation, and other related factors
- Process dynamics

Valve tracking nonlinearity includes behavior such as 'stiction' and 'seat friction'.

3.1 Control Valve Tracking Nonlinearities

Valve tracking nonlinearity is caused by the relationship between the input signal and the valve position. The valve position is calculated using a linear or nonlinear model. The valve position is then compared to the desired position to determine the control signal.

3.1.1 Backlash/Stiction

Backlash and stiction are caused by friction and wear. The control signal is limited by the backlash and stiction to prevent overshoot and undershoot.

Specification

The combined Backlash and Stiction system is not to exceed 1.0%, as installed, during normal process operation. [Reference 2].
In-Situ Testing for Backlash/Stiction

n-Situ testing for backlash/stiction on or deadband can be done for most control loops with reasonably fast non-integrating process dynamics such as flow or pressure.

Procedure:

1) Connect the channels of a data collection device to the controller output and process variable signal and at the termination point. Scribe the step changes for the test sequence with black and the process signal is key to be noisy, an analog input filter with 40 Hz constant of a few seconds and in a stable. Set up to collect data at a rate of about 1 sample per second and start data collection on sampling time can be one third of the filter time constant.

2) Put the loop on manual and allow the process to stabilize. Observe the process variable change from 'auto' to 'manua'.

3) At the control console or controller face-plate, initiate a series of small step changes or bump tests each time waiting for the process to stabilize before making the next move. Each step should be near to a clean step as possible. This is best done by entering a new value through a keyboard, rather than using a soft button. If a soft button or manual output adjustment must be made by hand, then it should be a small move of approximately the height, made as rapid as possible. Start with a series of 0% bump to begin with and move one direct on (e.g., up or down) if no response is observed. Increase the size of the bump test until the process starts reacting. The bump test is repeated until the response occurs. For improved accuracy, repeat the test procedure with similar bump tests. It is important that these bump tests be less than the natural motion on threshold (Reg A) in order to properly measure backlash and 'stiction on'.

4) For the same bump test used in 3) above (say 0% bump), initiate a series of bumps in one direction and continue in that direction to observe the effects of combined backash and 'stiction'. For small bumps, the valve will not move initially and then move for the next bump effect of a previous bump. If not moved, then move with the next bump effect. Once initiated and moved, the valve will be stuck and several more bumps will be necessary to cause another movement. Record the input signal and the process variable (PV) output and make values at each bump. Continue the test at least two 'stiction' cycles as described above. Then reverse the direction on first applying the number of bumps needed to take up the backlash as estimated from 3) above. Continue the test in the new direction until the process value has returned to the natural value at the start of the test. Draw an XY plot of PV vs. input signal to that of F figure 1. Estimate the combined backlash/stiction or deadband by extracting the maximum width of the resulting hysteresis plot at the backash/stiction limit plus 1%.

3.1.2 Speed of Response and Overshoot

Consistent speed of response of a control valve is crucial for the operation of a control loop. For small step changes in input signal, the plug or trim should track the input signal with as much speed as possible. The speed of response listed in Table 1 and illustrated in Figure 2, over a range of step changes ranging from 10% of travel down to the normal backash/stiction limit, achieved above plus 1%. (Down to 15% for combined backash/stiction on 0%)

Definitions:

- **Td**: deadtime before any response (seconds)
- **T63**: time to reach 63% of a step change (seconds)
- **T98**: time to reach 98% of a step change (seconds)
- **BW**: bandwidth (Hz), the frequency at which the peak or trim position is attenuated by 6 dB.
- **Drawn**: as described above the speed of response for the specific condition on for completeness even though the specific condition on for n most cases.

### Table 1

<table>
<thead>
<tr>
<th>Valve Size (inches)</th>
<th>Td (sec)</th>
<th>T63 (sec)</th>
<th>T98 (sec)</th>
<th>BW (Hz)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.2</td>
<td>0.5</td>
<td>0.3</td>
<td>0.7</td>
<td>1.6</td>
</tr>
<tr>
<td>&gt;2.6</td>
<td>&gt;5:15</td>
<td>0.2</td>
<td>0.6</td>
<td>1.4</td>
</tr>
<tr>
<td>&gt;6.12</td>
<td>&gt;15:30</td>
<td>0.4</td>
<td>1.2</td>
<td>2.8</td>
</tr>
<tr>
<td>&gt;12.20</td>
<td>&gt;30:50</td>
<td>0.6</td>
<td>1.8</td>
<td>4.2</td>
</tr>
<tr>
<td>&gt;20</td>
<td>&gt;50</td>
<td>0.8</td>
<td>2.4</td>
<td>5.6</td>
</tr>
<tr>
<td>&gt;50</td>
<td>&gt;50</td>
<td></td>
<td></td>
<td>0.2</td>
</tr>
</tbody>
</table>

The above specification on a low range valve to be progressive over a wide range of small valves. The valve Td T63 T98 and BW specification for a compact valve set of numbers for a second order controller system with deadtime represented as the percentage of the steady state change.

Overshoot

Overshoot will be present if the step responses do not increase by more than 8%. This overshoot occurs due to the return to a steady state value. The % overshoot is the absolute value of each overshoot expressed as a percentage of the steady state change.

Specification

The percent overshoot should be less than or equal to 20% of the step magnitude for steps ranging from 10% of travel down to steps equal to the backash/stiction limit plus 1%.
In-Situ Testing for Speed of Response and Overshoot

In situ testing for speed of response is very similar to that for backash/stiction. The differences are as follows:

1. The data collection equipment should collect data at a rate which is 5 times faster than the expected time constant of the valve, typically 0.02 to 0.03 seconds. The input terminal should be reduced to a time constant of 0.02 seconds approximately.

2. To measure the valve speed of response, it may not be possible to use existing process instrumentation, such as a magnet for meter, since this may have a time constant of 3 seconds or longer. Hence, it may be necessary to install additional instrumentation, such as a pneumotransmitter, to measure as close as possible to the valve plug position. Alternatively, it may be possible to measure the pressure some distance after the valve, or the differential pressure across the valve itself. In any case the time constant of the measurement must be at least three times as fast as the expected time constant of the valve.

3. Finally, steps of 10% of travel should be done at the process variable response as ngure 2. Calculate the % overshoot on the basis of the final process value. Tests can vary in direction to suit process conditions at the time. Small steps should be done in the same direction as not to occur backlash. Now repeat the tests for successively smaller steps, down to steps equal to 1% of the previous value achieved combined backlash/stiction limit plus 1%. Measure % overshoot for each test. The results should comply to Table 1 and Figure 2.

3.2 Sizing and Flow Characteristic Nonlinearities

Control valves are sized for a given plant capacity. Normally an allowance is made for future throughput increases, together with a safety margin. Most valves have non-linear characteristics. This results in the fact that the 'process gain', the relationship between input signal change and the resulting process change, is often not the same as the operating point. The process gain combined with the control loop gain determines how a control loop behaves. These changes in process gain are a major source of process tuning problems and often result in loop instability. Over-sizing the valve may make the problem much more severe. To minimize this effect, the overall process gain for non-integrating process variable is decreased by about 10% and transmitter span should have a nominal value of 1.0 and be kept in the range.

0.5 to 2.0 (% of span) / (% output)

For a bump test, the controller output was stepped by 5% the flow changed by 7.2% of span.

Process Gain - 7.2% / 5% 1.44% /%

Summary of Control Valve Dynamic Performance Specification:

1. Combined backlash/stiction:
   - Not to exceed 1% of input span

2. Speed of Response:
   - Rate of change of valve position in response to step changes ranging from 10% of input signal (see Table 1 and Figure 2) down to the backlash/stiction limit plus 1%

3. Overshoot:
   - Percent overshoot in valve position to be less than 20% for all steps made

4. Sizing and Flow Characteristics (Engineering Design):
   - Loop Process Gain = 1.0 (% of transmitter span) / (% controller output)
   - Nominal Range: 0.5 to 2.0 % /%

References

1. EnTech Control Valve Specifications (Vers on 2.0, 3/94).
2. ANSI/ISA S51.1 1979 Process Instrumentation on Term
