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EnTech Seminars

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 wel' 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

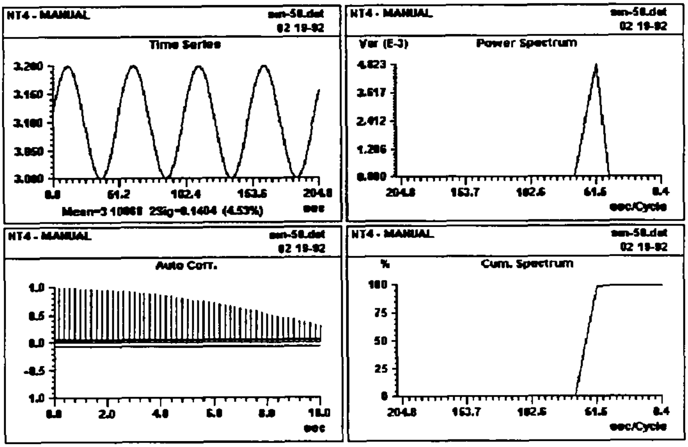
Identifying the Source(s) of Varriability

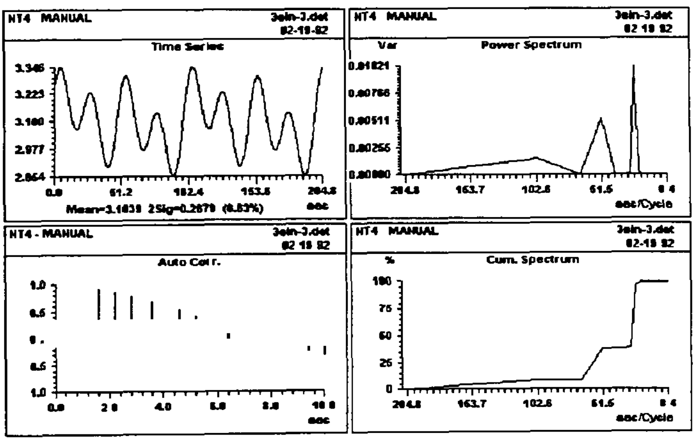
The process variability generated upstream at a given frequency will be felt at the reel at the **same** frequency and at an amplitude **depending** on:

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

Time Series Analysis

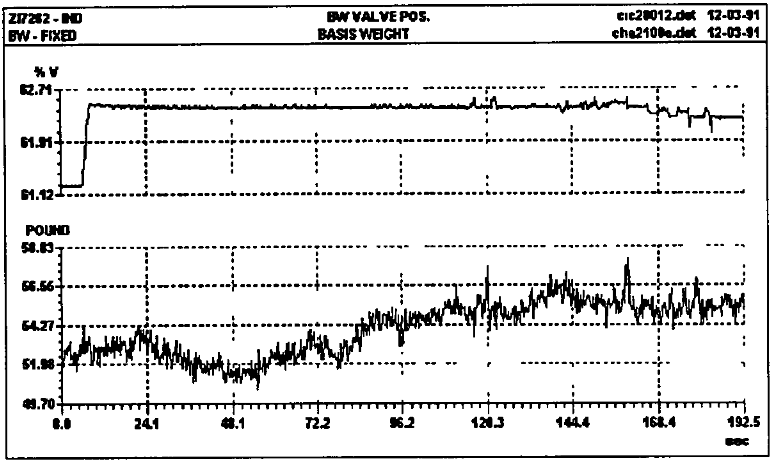
Examples





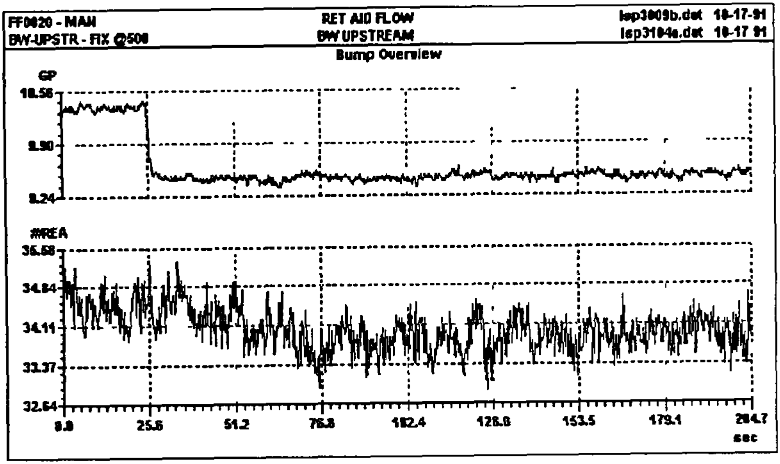
Coupling Gain Example:

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

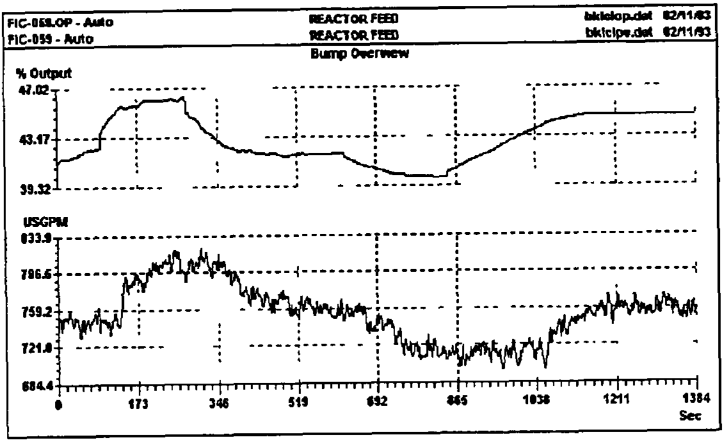
- Determine product variability and compare to market needs
- Determine process variability by process variable and link to product variability
- Identify causes of process variability
- Identify requirements for: maintenance, modifications, in order to improve variability
- Raise awareness of variability/process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

Elements of the Audit

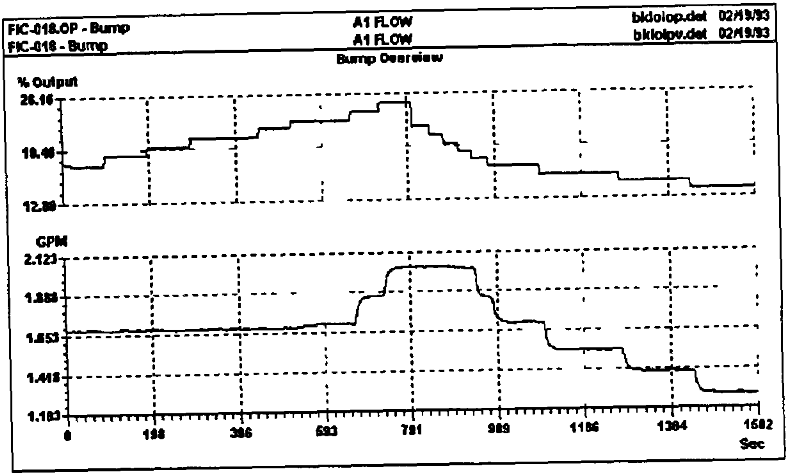
1. Preparation - P&ID Review, Planning
2. Field Data Collection and Preliminary Analysis
3. Data Analysis and Interpretation
4. Tune where Possible
5. Report Preparation and Presentation
6. Follow Up

Audit Examples

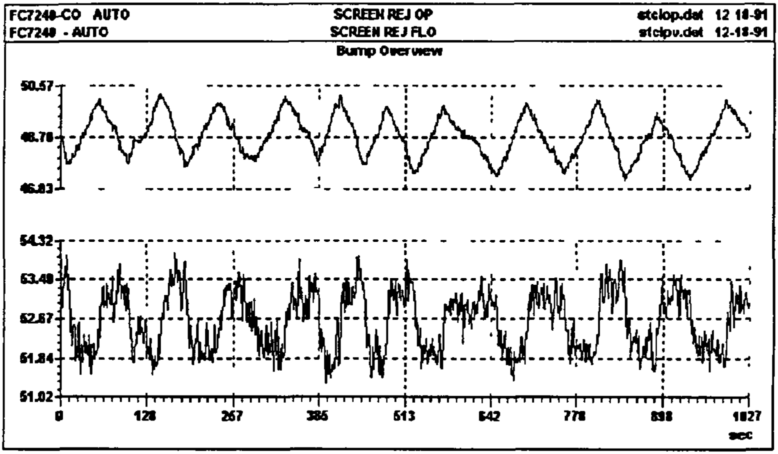
Closed-loop example of backlash



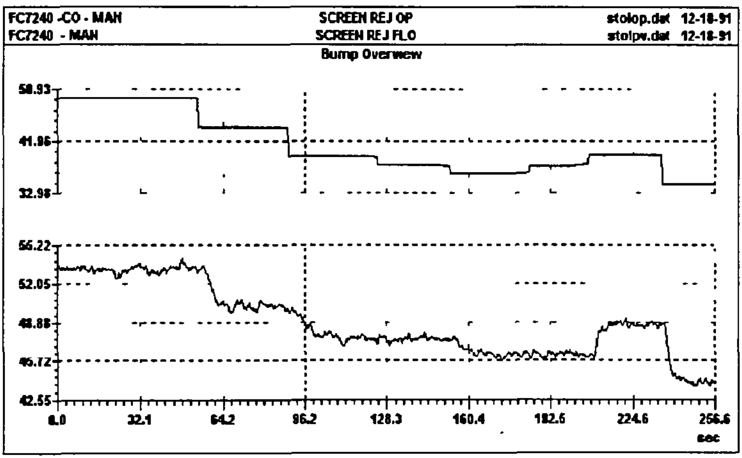
Open-loop bump test to determine amount of backlash. How much backlash is in this valve assembly?



Closed-loop example of stiction.

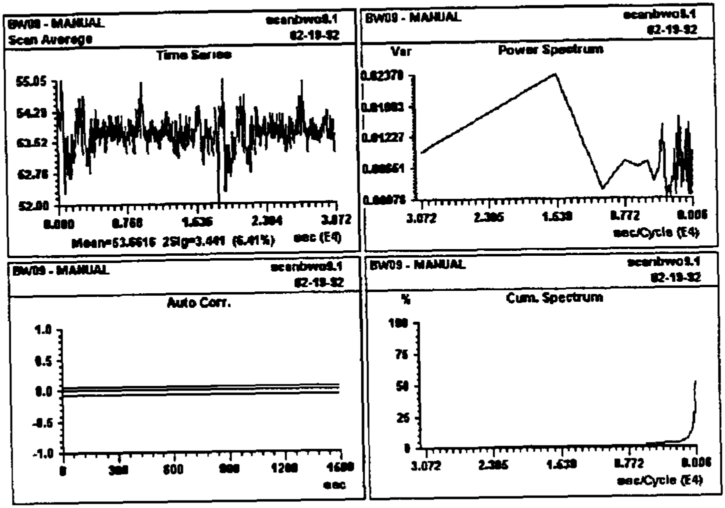


Open-loop bump test to determine amount of stiction How much stiction is in the valve assembly?

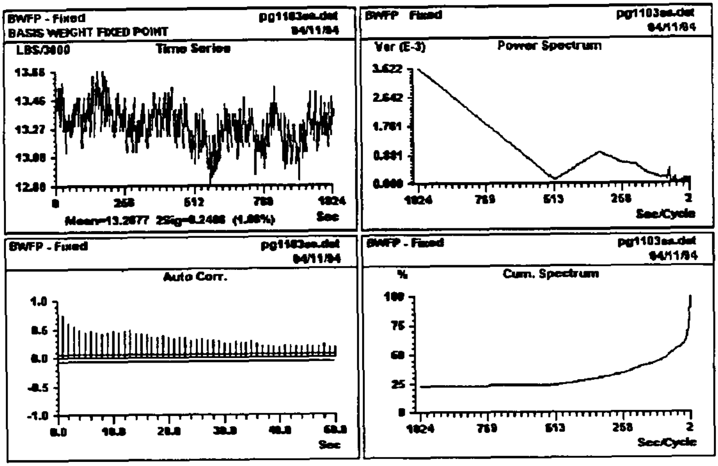


Effect of Sampling Rate

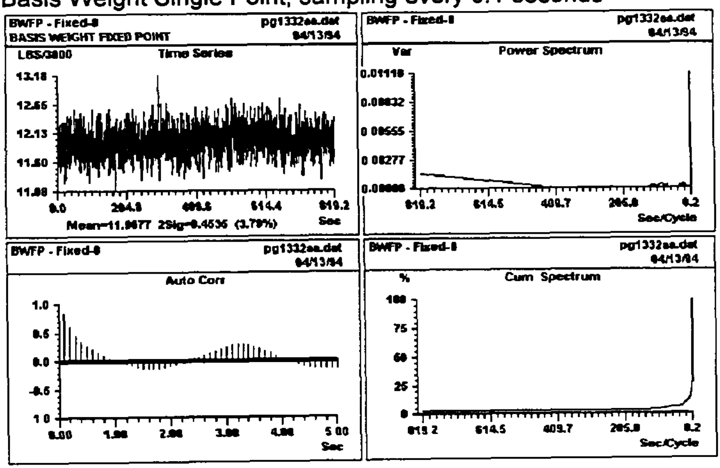
Basis Weight Scan Average, sampling every 30 seconds



Basis Weight Single Point, sampling every 1 second



Basis Weight Single Point, sampling every 0.1 seconds



APPENDIX 1A
AUDIT LITERATURE

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PROCESS CONTROL AUDITS HAVE MAJOR IMPACT ON UNIFORMITY

By William L. Bialkowski

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 1,300 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 fourdrinier 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

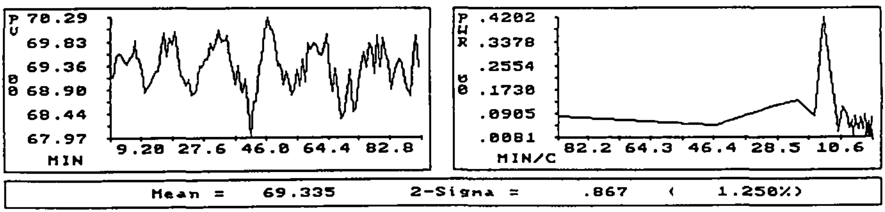
About 40% of the linerboard produced on this machine is sold to one large corrugated box manufacturer. In 1989, this manufacturer installed two new high-speed corrugators at its largest box plant.

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.

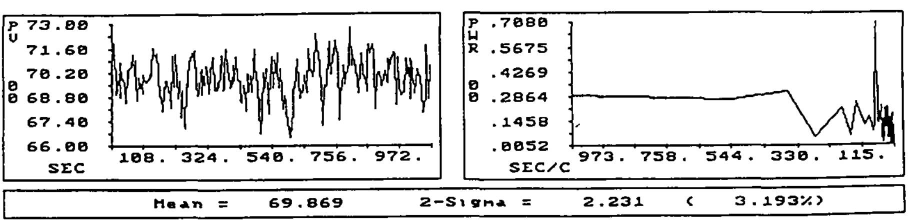
The mill undertook pulsation and vibration studies on the machine but did not identify any operating problems except for "random low frequency variability."

The paper machine control system

▸ Figure 1: Scan average basis weight variability on 69-lb linerboard. Measured value (left) and power spectrum (right).

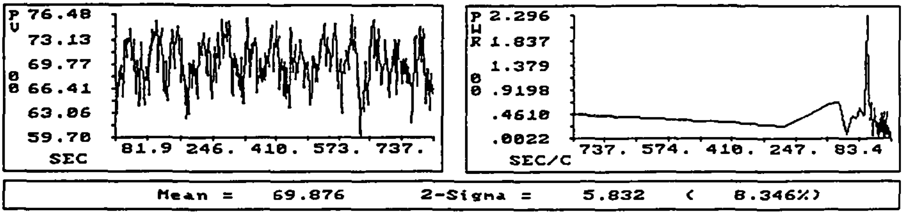


▸ Figure 2: Fixed point basis weight variability based on 5 second averages for 69-lb linerboard.



> *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.



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.

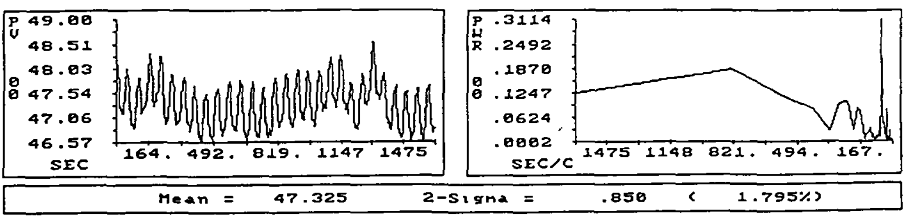
The audit team next examined basis weight variability in single point with the scanner positioned at the center of the sheet for about 18 minutes. The data consisted of 5 second average samples collected by the computer. This was done because experience has shown that a 45-second-scan average tends to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

Figure 4: Dilution header pressure measurement.



▸ *During the audit, all consistency and stock flow controls were evaluated.*

seen as a major destabilizing influence on the paper machine

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 retuned 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 cycled 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.

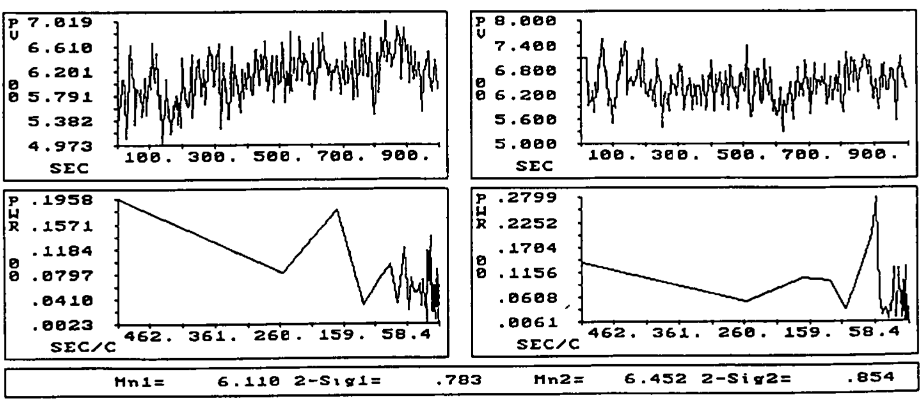
AUDIT ANALYSIS

The 65-second and 14-minute cycles had a significant impact on linerboard uni formity. 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

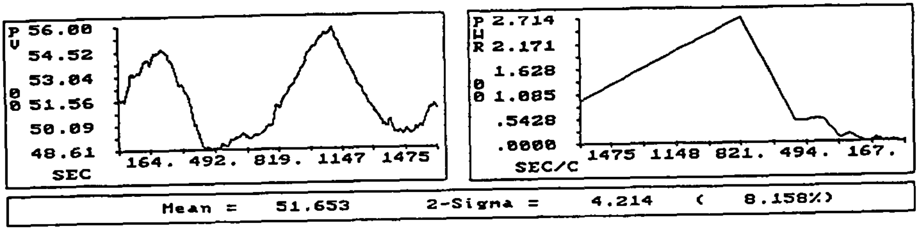
The results of this thorough investigation revealed that of the 60 loops, five had no significant problems, seven required minor tuning, and 32, including LC-203, required extensive retuning to minimize variability. Twenty-three required instrument repair or maintenance before the loop could be tuned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

▸ **Figure 5: Fixed point moisture variability with LC-203 on Manual (left) and Automatic (right).**

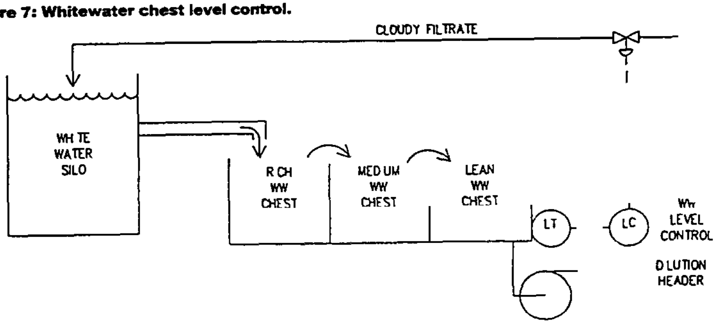


▸ *The process audit also had a major impact on mill management and staff.*

▸ **Figure 6: Whitewater chest level variability.**



▸ **Figure 7: Whitewater chest level control.**



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 pro-active 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. Bialkowski 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. Bialkowski 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 dampened 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.

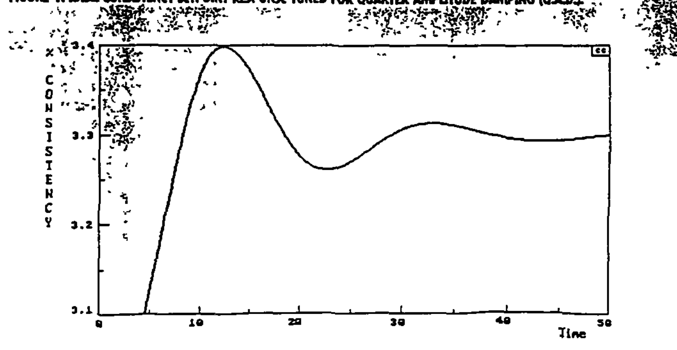
METHODOLOGY

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 pulp and paper industry engineers and technicians have better methods to insure product uniformity.

FIGURE 1: IDEAL CONSISTENCY SETPOINT RESPONSE TUNED FOR QUARTER AMPLITUDE DAMPING (Q.A.D.)



ANALYSIS

In a setpoint response Bode plot of a quarter amplitude damped, tuned consistency loop related to the frequency 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 retuned 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.

FIGURE 2: IDEAL Q.A.D. TUNED CONSISTENCY CONTROL OPERATING UNDER NORMAL NOISY MILL CONDITIONS. (L) TIME TRACE (R)POWER SPECTRUM

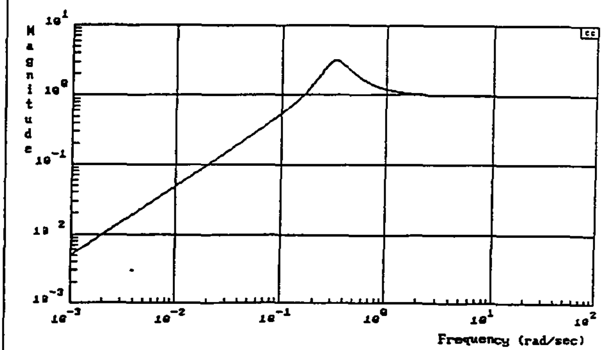
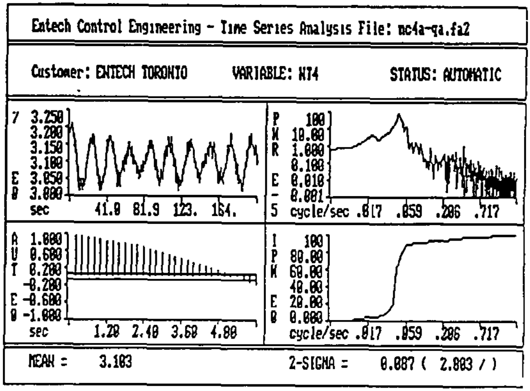


FIGURE 3: IDEAL CONSISTENCY LOOP LAMBDA TUNED TO MINIMIZE VARIABILITY. (L) TIME TRACE (R) POWER SPECTRUM



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 insure product uniformity and maintain market share.

The only alternative to Q.A.D. was to tune "by the seat of the pants."

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.

PRESENTED AT CONTROL SYSTEMS 90 - SEPT. 17-20, 1990 HELSINKI, FINLAND

PROCESS CONTROL RELATED VARIABILITY AND THE LINK TO END-USE PERFORMANCE

W.L. Bialkowski

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.

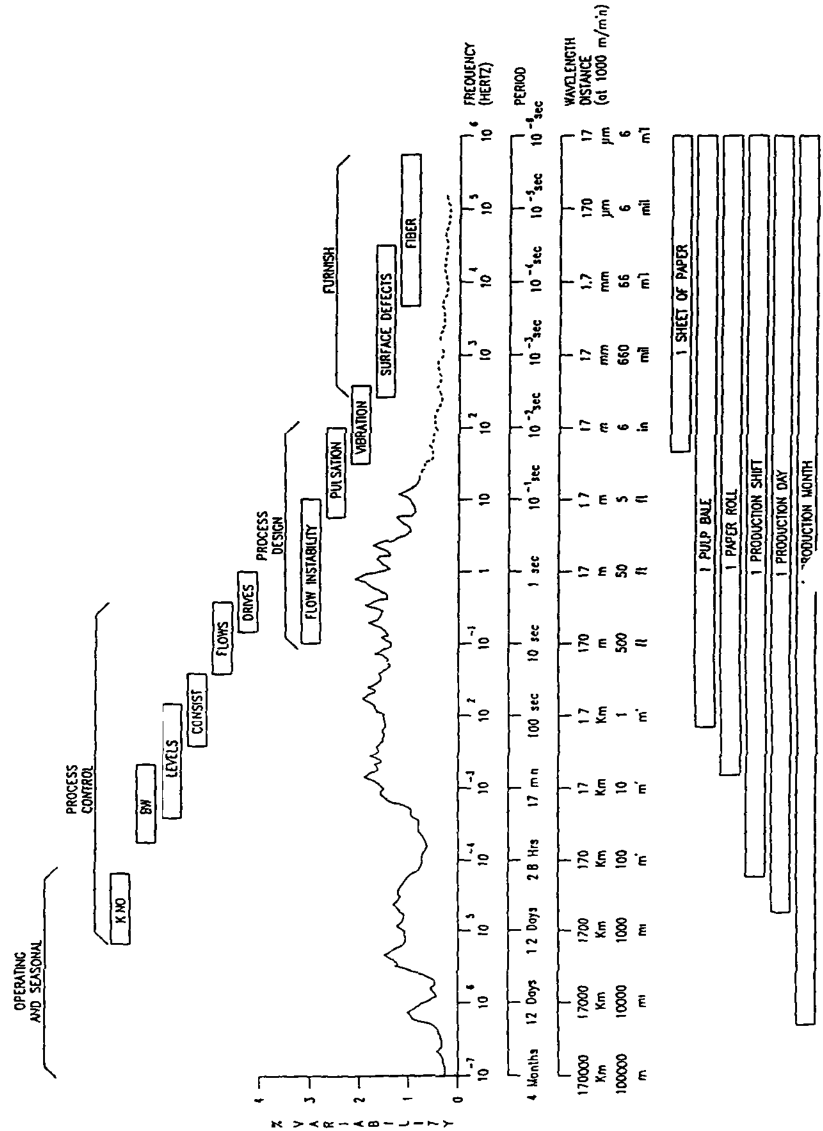
2.1 Process Variability Spectrum - As Seen at the Paper Machine

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.

TABLE 1

Paper Machine MD Variability from Very High Frequency to Very Low Frequency Range

ITEM	SIZE/LENGTH WAVELENGTH	AT 1000 M/MIN (3300 FT/MIN) PERIOD	FREQUENCY
Sheet Surface Micro Structure/Furnish Properties			
Width-Softwood Fibre	30-50um (1-2thou)	1.8-3μs	0.33-.56mHz
Length-Softwood Fibre	3-5mm (.1-.2in)	0.18-0.3ms	3.3-5.6kHz
Wire Mark	1.3-2.5mm (0.05-0.1in)	.08-0.15ms	6.7-13kHz
Floc Size	0.5-1cm (.2-0.4in)	3-6ms	0.17-0.33kHz
PM Design/Maintenance Problems			
Mechanical Vibration	0.08-1.7m (3.1-65in)	5-100ms	10-200Hz
Sheet of Paper	.3m (1ft)	.02s	50Hz
Hydraulic Pulsations	0.3-3m (12-120in)	.02-.2s	5-50Hz
Turbulence & Eddies	3-30m (10-100ft)	.2-2s	.5-5Hz
Clothing Rotation	30-100m (100-300ft)	2-6s	0.16-0.5Hz
PM Process Control Problems			
Beta Gauge Raw Signal Bandwidth	1.7-8m (5-27ft)	100-500ms	2-10Hz
Headbox and Fan Pump	30-300m (100-1000ft)	2-20s	.05-.5Hz
Flow Controller Problems	50m-1.7km (160ft-1mi)	3-100s	0.01-.3Hz
Consistency Controller Problems	500m-5km (1600ft-3mi)	30s-5min	.003-.03Hz
Bale Pulping Rate	1-20km (1.6-13mi)	1-20min	17-0.8mHz
C-Stage Brightness Control	3-15km (2-9mi)	3-15min	6-1mHz
Basis Weight Feedback Control	12-40km (7-25mi)	12-40min	0.4-1mHz
Level Controller Problems	800m-60Km (2700ft-40mi)	48s-1hr	0.3-21mHz
K.No. Control Problems	600-1800km (400-2200mi)	10-30hr	9-28μHz
Sampling/Testing Frequencies			
Set of Paper - period/set	15km (9mi)	15min	1mHz
Jumbo Reel - period/reel	45km (28mi)	45min	0.4mHz
SPC Sample Frequencies			
per reel - period c/2 reels	90km (56mi)	90min	0.19mHz
per hour - period c/2 hours	120km (75mi)	2hr	0.14mHz
per shift - period c/2 shifts	960km (600mi)	16hr	17μHz
per day - period c/2 days	2880km (1800mi)	48hr	6μHz
per month - period c/2 months	87000km (55000mi)	2months	0.2μHz


OPERATING AND SEASONAL
PROCESS CONTROL
K NO
BW
LEVELS
CONSIST
FLOWS
DRIVES
PROCESS DESIGN
FLOW INSTABILITY
PULSATION
VIBRATION
FURNISH
SURFACE DEFECTS
FIBER
% VARIABILITY
0
1
2
3
4
FREQUENCY (HERTZ)
10^-7
10^-6
10^-5
10^-4
10^-3
10^-2
10^-1
1
10
10^2
10^3
10^4
10^5
10^6
PERIOD
4 Months
12 Days
1.2 Days
2.8 Hrs
17 min
100 sec
10 sec
1 sec
10^-1 sec
10^-2 sec
10^-3 sec
10^-4 sec
10^-5 sec
10^-6 sec
WAVELENGTH DISTANCE (at 1000 m/min)
170000 Km 100000 mi
17000 Km 10000 mi
1700 Km 1000 mi
170 Km 100 mi
17 Km 10 mi
1.7 Km 1 mi
170 m 500 ft
17 m 50 ft
1.7 m 5 ft
17 cm 6 in
17 mm 660 mil
1.7 mm 66 mil
170 µm 6 mil
17 µm 6 mil
1 SHEET OF PAPER
1 PULP BALE
1 PAPER ROLL
1 PRODUCTION SHIFT
1 PRODUCTION DAY
1 PRODUCTION MONTH

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 0.3m (1ft 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 II

Effect of Time Constant Choice on Data Filtering

Typical Sensor	Time Constant τ	Cut-off Period $2\pi\tau$	Period at 90% Attenuation (-20 db)
o Fast Beta Gauge o Fast Pressure	20 ms	120 ms/cycle	12 ms/cycle
o Slow Beta Gauge	100 ms	0.63 s/cycle	0.063 s/cycle
o Pressure o Flow	1 sec	6 s/cycle	0.6 s/cycle
o Fast Consistency	3 sec	20 s/cycle	2 s/cycle
o Slow Consistency o Slow Flow o Fast Pulp Brightness Sensor	10 sec	60 s/cycle	6 s/cycle
o Heavy Snubber or Filter	1 min	6 min/cycle	40 s/cycle
o Extreme Snubber or Filter	10 min	60 min/cycle	6 min/cycle

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

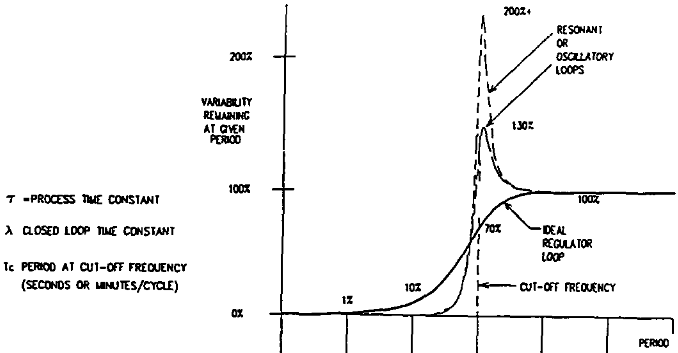
The cut-off frequency of a regulatory loop is a function of: process dynamics (time constant, deadtime); type of algorithm used; and tuning employed - the faster the tuning, the faster the cut-off frequency. Unfortunately, as the loop tuning becomes more aggressive, the more likely it is that the loop 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 2π (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

	Process Time Constant	Closed-Loop Time Constant	Cut-Off Period	Period at 90% Attenuation (-20 db)
o Fan Pump Speed	0.1 sec	0.3 sec	2 s/cycle	20 s/cycle
o Total Head	1 sec	3 sec	20 s/cycle	3 min/cycle
o Consistency	5 sec	15 sec	1.5 min/cycle	15 min/cycle
o C-Stage Brightness	.5 min	1.5 min	9 min/cycle	1.5 hr/cycle
o Basis Weight	1 min	3 min	20 min/cycle	3 hr/cycle
o Moisture	2 min	6 min	40 min/cycle	7 hr/cycle
o PM Colour Control @ Fan Pump	5 min	15 min	1.5 hr/cycle	15 hr/cycle
o O_2 Delignitation K.No. F.B.	10 min	0.5 hr	3 hr/cycle	30 hr/cycle
o PM Colour Control @ Blend Chest	20 min	1 hr	6 hr/cycle	2.5 day/cycle
o Digester K.No.	1 hr	3 hr	20 hr/cycle	8 day/cycle



TYPICAL DYNAMICS	τ	$\lambda \approx 3\tau$	$T_c = 2\pi\lambda$						
HEADBOX TOTAL HEAD	1sec	3sec	20sec	300min 5hrs	2000s 30min	200s 3min	20sec	2sec	0 2sec
CONSISTENCY	5sec	15sec	1 5min	1500min 1day	150min 2.5hrs	15min	1 5min	9sec	0 9sec
BASIS WEIGHT	1min	3min	20min	20000min 14days	2000min 1 4days	200min 3hrs	20min	2min	12sec
DIGESTER K.NO	1hr	3hrs	20hrs	20000hrs 2 3years	2000hrs 2 7months	200hrs 8 3days	20hrs	2hrs	12min

FIGURE 2 TYPICAL REGULATORY LOOP FREQUENCY RESPONSE

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

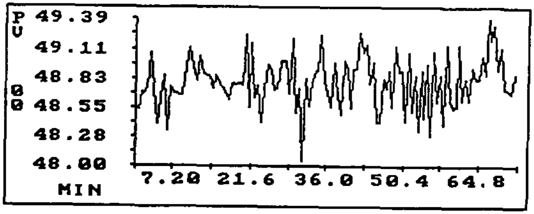
Tank levels are integrating processes and usually prove to be extremely difficult to tune by feel alone, without the loop cycling as a result. The human tendency is to use far too much integral or reset action. Critically damped tuning methods do exist to prevent level contróller cycling. However in 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

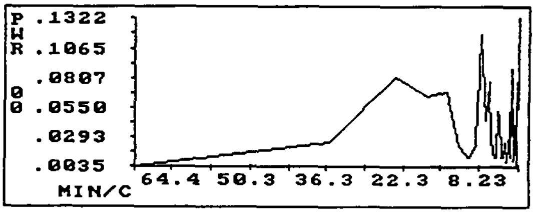
The function of the 3000 to 5000 control loops in a typical pulp and paper mill (some mills have as many as 10,000 loops) is to reduce the impact of raw material variability as fibre flows through the pulp mill and on to the reel of the paper machine. Each loop will have a frequency response similar 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



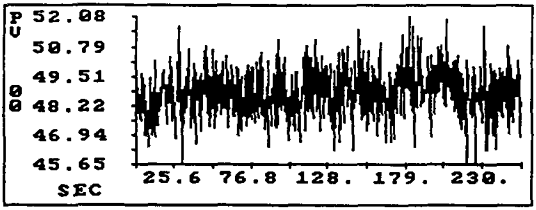
B) Power Spectrum

Mean = 48.800 2-Sigma .475 (.974%)

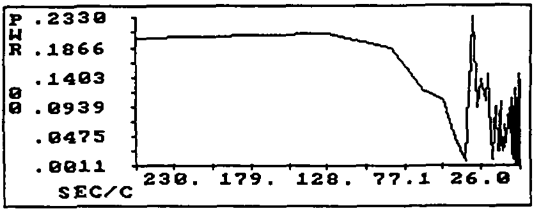
FIGURE 3 SCAN AVERAGE BASIS WEIGHT - ON CONTROL

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



A) Basis Weight vs Time



B) Power Spectrum

Mean = 48.900 2-Sigma - 1.907 (3.900%)

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

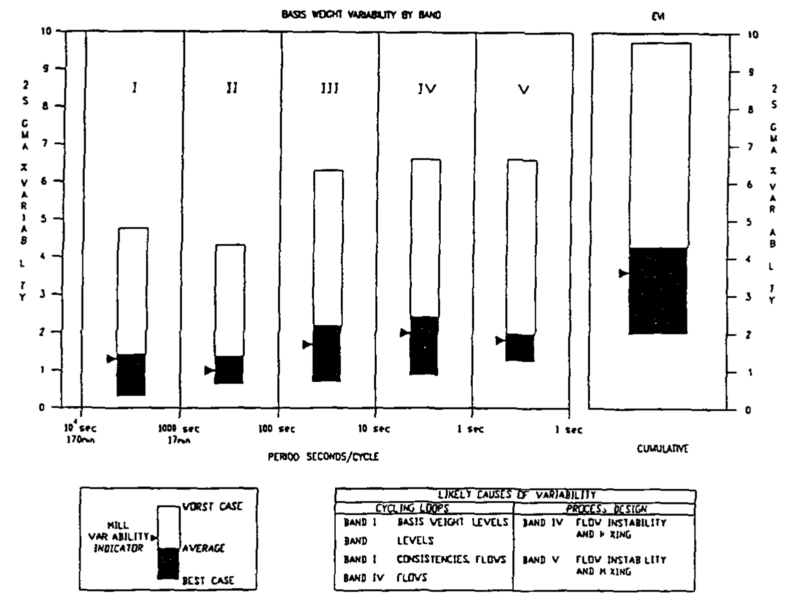


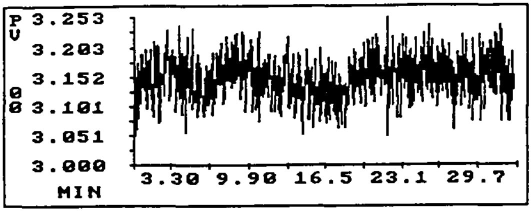
FIGURE 5 ENTECH VARIABILITY INDEX (EVI) FOR BASIS WEIGHT VARIABILITY SHOWN IN FIGURE 3 AND 4

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.

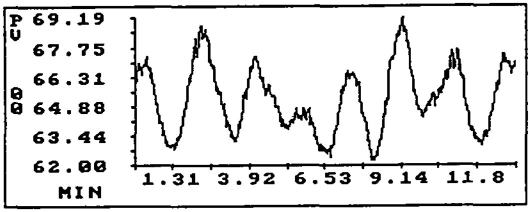
5.2 Causes of Variability on the Paper Machine

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.

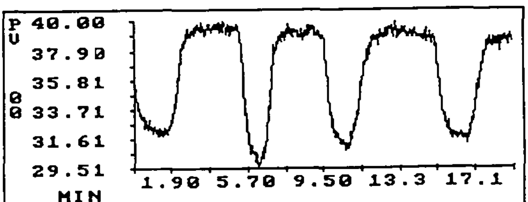
In addition to the causes of variability illustrated above, a major cause of variability is pulp supply. Variability in consistency, brightness, strength, viscosity and freeness can often be traced back to the pulp mill. In one paper machine audit a four minute consistency cycle was traced all the way back to a disturbance in the blow tank. 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.



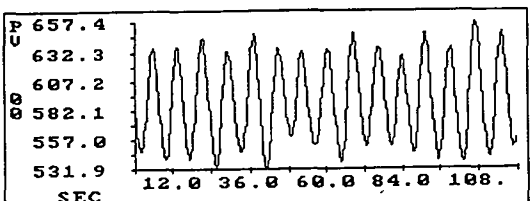
A) Machine Chest Consistency vs Time
Mean = 3.15 2-Sigma = 2.2%



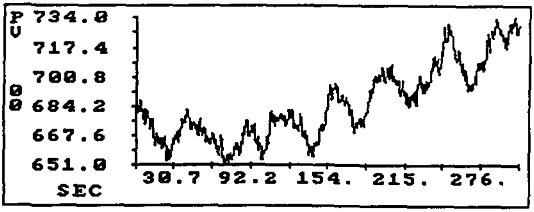
B) SaveAll Tank Level vs Time
Mean 65.27 2-Sigma = 4.9%



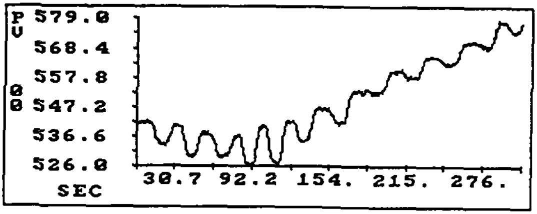
C) White Water Header Pressure vs Time
Mean = 36.38 2-Sigma = 17.8%



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%

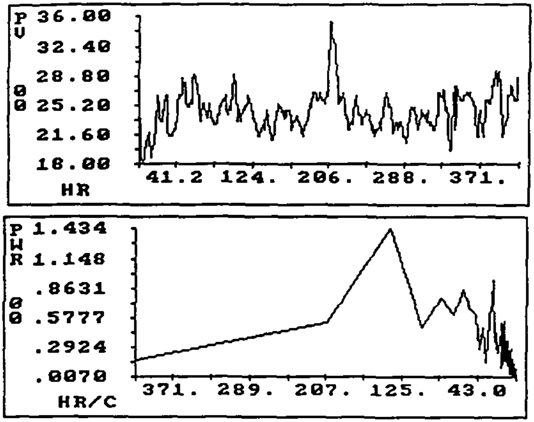


F) TMP Flow to Blend vs Time
Mean = 549 2-Sigma - 5.3%

FIGURE 6 A FEW CAUSES OF HIGH PAPER MACHINE VARIABILITY

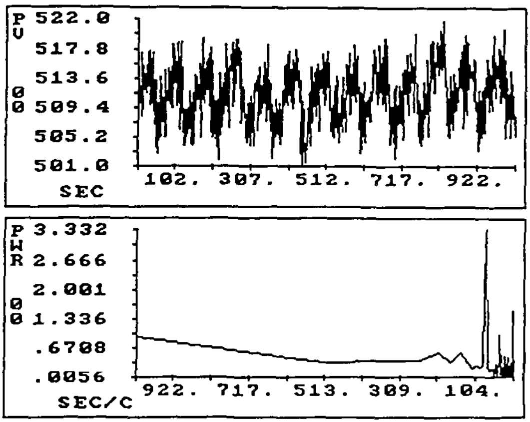
6.0 PULP MILL VARIABILITY AND ITS IMPACT ON THE PAPER MACHINE

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.



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

FIGURE 7 CONTINUOUS DIGESTER K.NO. VARIABILITY BASED ON 1 HOUR SAMPLES



Mean = 511.3 2-Sigma - 7.63 (1.49%)

FIGURE 8 DIGESTER COUNTER WASH FLOW 1 SECOND READINGS

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 +/- 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

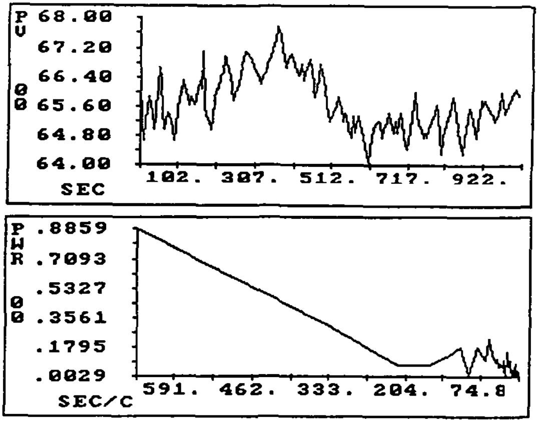


FIGURE 9 CHLORINATION STAGE BRIGHTNESS - ON MANUAL
5 SECOND READINGS

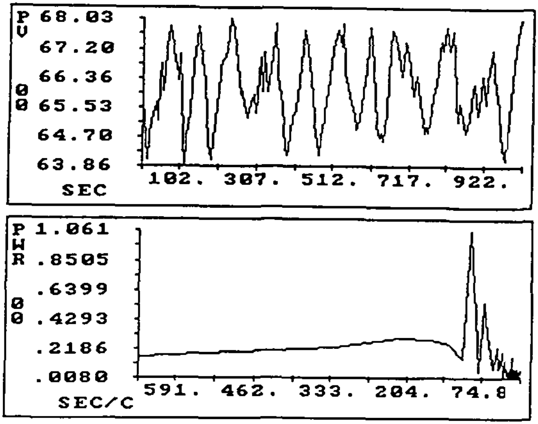
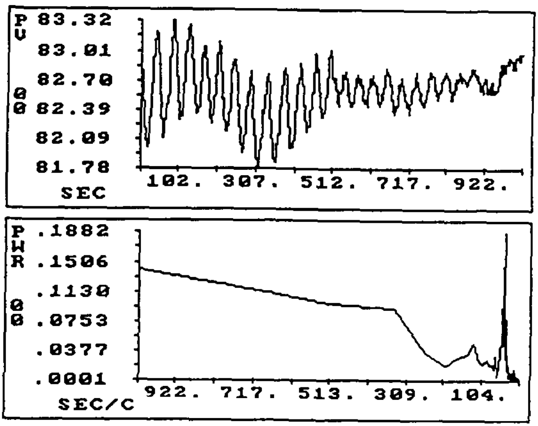


FIGURE 10 CHLORINATION STAGE BRIGHTNESS - ON AUTOMATIC
5 SECOND READINGS

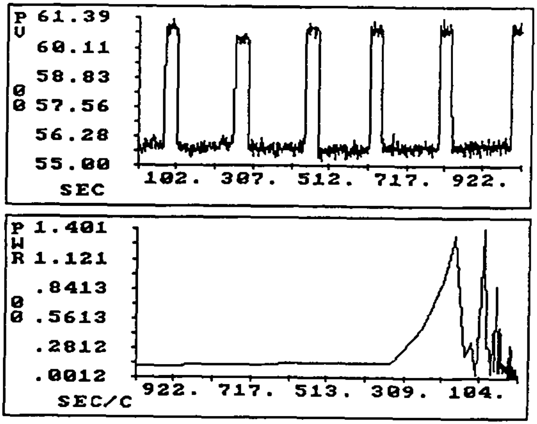
frequency content should have been attenuated through process control hopefully, without any variability being amplified as a result of cycling loops. The paper mill controls hence have to deal with this variability to the extent that they can.

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



Mean = 82.6 2-Sigma = 0.546 (0.661%)

FIGURE 11 CHLORINE GAS PRESSURE - ON AUTOMATIC
1 SECOND READINGS



Mean = 56.8 2-Sigma = 4.01 (7.053%)

FIGURE 12 CHLORINE DIOXIDE SOLUTION FLOW - ON AUTOMATIC
1 SECOND READINGS

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

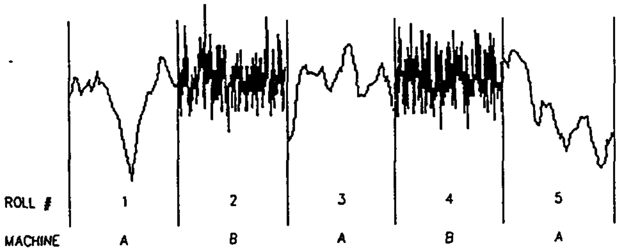


FIGURE 13 ILLUSTRATION OF MACHINE AND ROLL VARIABILITY IMPACT ON PRESSROOM RUNNABILITY

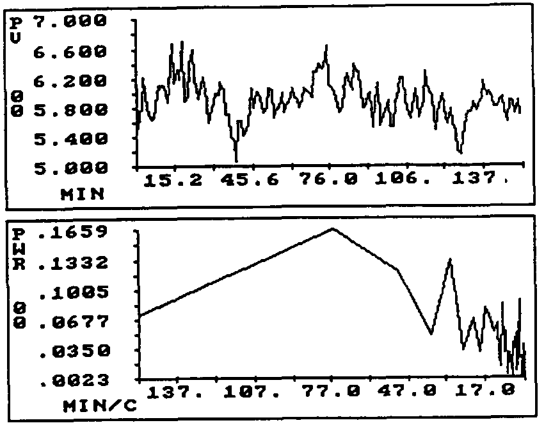
parameters. The frequency content for basis weight is likely common for many other variables.

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

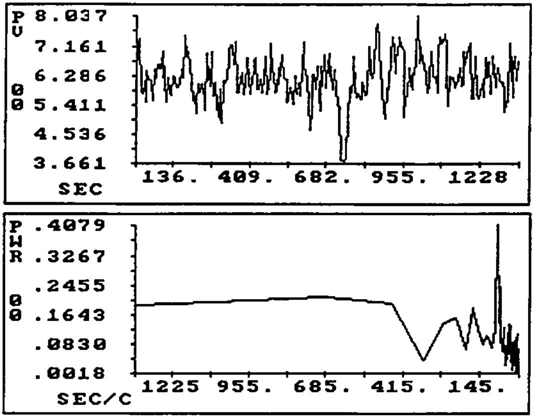
Corrugators typically manufacture boxboard by gluing corrugated medium to both top and bottom liners. This requires two rolls of linerboard and one roll of corrugating medium. These products typically are manufactured on at least two if not three paper machines. A predominant problem faced by the box 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



Mean = 5.9 2-Sigma = 0.525

FIGURE 14 LINERBOARD MACHINE 1 MINUTE SCAN AVERAGE MOISTURE

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.



Mean = 6.2 2-Sigma = 1.516

FIGURE 15 LINERBOARD MACHINE MOISTURE VARIABILITY IN FIXED POSITION - 5 SECOND AVERAGES

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.

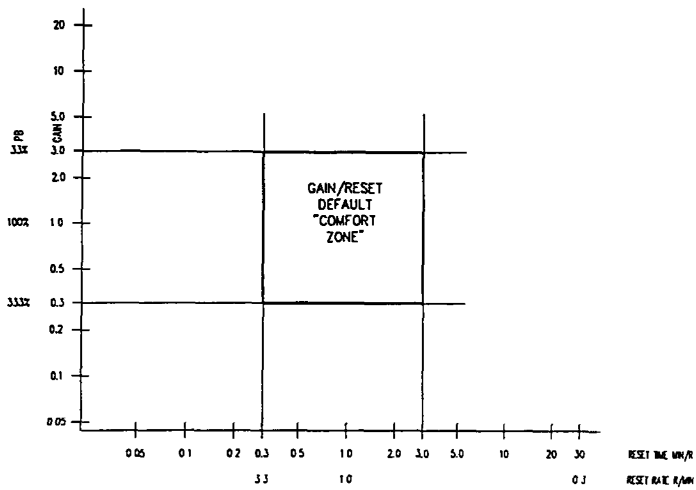


FIGURE 16 CONTROLLER TUNING - POLULARLY ACCEPT GAIN/RESET "COMFORT ZONE"

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.

ACKNOWLEDGEMENT

The author wishes to acknowledge the work of his associates at EnTech Control Engineering Inc. who have carried out many paper machine and pulp mill variability and control audits. The author also wishes to acknowledge the cooperation 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.

REFERENCES

[1] W.L. Bialkowski, Newsprint Uniformity and Process Control - The Untapped Competitive Edge. Presented at the 76th Annual Meeting of the CPPA, (1990).
[2] L.C. Ingman, Quality Report "Predicting End-use Performance". Pulp and Paper Canada, (1988).
[3] J.R. Parker, Vibration Isolation of Headboxes for the Control of Machine Direction Grammage Variations. Pulp and Paper Canada, 90:6, (1989).
[4] W.J. Hee, Jr., Specifications of Certification. Purchasing Manager, Gannett Supply Corporation, (1988).
[5] G.S. Rennie, Linerboard: Does your Product Meet the Needs of the Corrugating Industry? Pulp and Paper Canada, 91:3, (1990).
[6] R. Gries, Nekoosa Nixes Bad Boxes with Quality Reports. PIMA, (1990).
[7] J. Veyre, Headbox Approach - Flow Survey Analysis of Causes and Corrections of MD Basis Weight Variations. Pulp and Paper Canada, Vol. 82, No. 7, (1981).
[8] W.L. Bialkowski, Process Control Management and Technical Training Since 1984. Pulp and Paper Canada, 90:9, (1989).

23 31 12 04 10 07

23 31 19 04 10 07

OPTIMIZATION OF POWERHOUSE OPERATION AT CHAMPION INTERNATIONAL INC.

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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 resu t 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 power house 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 power house 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 power house 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,

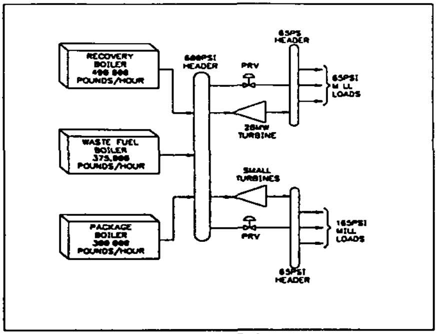

FIGURE 1
POWER HOUSE STEAM SYSTEM

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 stand by 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 1 0 second 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 nonlinearities 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 the 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 This 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 by 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

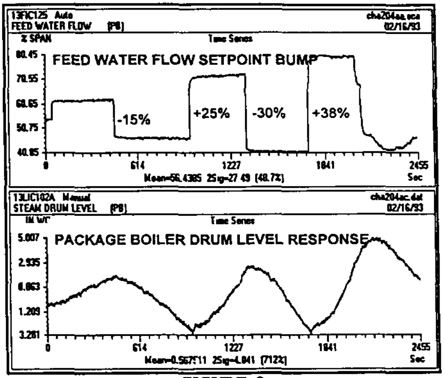

FIGURE 2
PACKAGE BOILER DRUM LEVEL BUMP TEST SEQUENCE

on the package boiler drum level, where 13FIC125 is the feed water 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

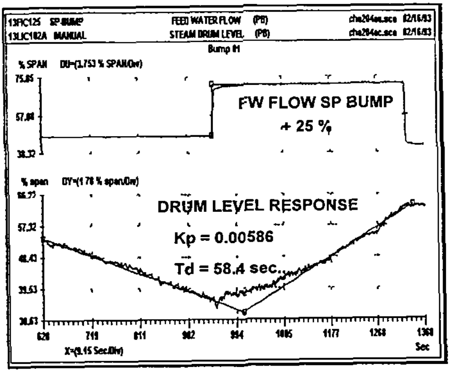

FIGURE 3
PACKAGE BOILER DRUM LEVEL BUMP

Figure 3 also shows the results of the parameter fitting performed by the analysis software for this one bump test. The net results for this series of bump tests were: 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 integrator 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:

$$G_p(s) = \frac{K_p e^{-T_d s}}{s} \qquad 1.0$$

The results for all three boilers were as follows:

Table I Boiler Drum Level Process Dynamics

Boiler	Process Gain K_p	Deadtime (sec) T_d
Package	0.0050	59
Waste Fuel	0.0094	60
Recovery	0.0104	53

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, with 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. 1.0, 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

Swing Boiler	Process Gain K_p	Deadtime (sec) T_d
Waste Fuel	0.000914	209
Package	0.001	21

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

Early in the project the waste fuel bark delivery system was found to be a major cause of disturbances to the waste fue

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boiler drum level and hence to the steam header pressure control Figure 4 illustrates how the bark and chip screen

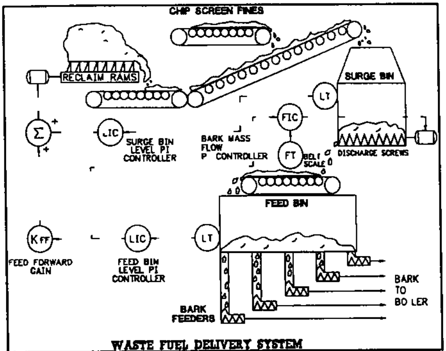

FIGURE 4
WASTE FUEL DELIVERY SYSTEM

fines are transported 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 III - Bark Surge and Feed Bin Dynamics

Bin	Process Gain K_P	Deadtime (sec) T_d
Surge Bin	0.00108	210
Feed Bin	0.00434	64

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 de-stabilize 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 IV - Boiler Water Conductivity Dynamics

Process Gain K_P	Deadtime (sec) T_d
-0.00004	1211

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 a 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] After the project, the model 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 the bump tests results In the case of the missing feedwater shrink 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

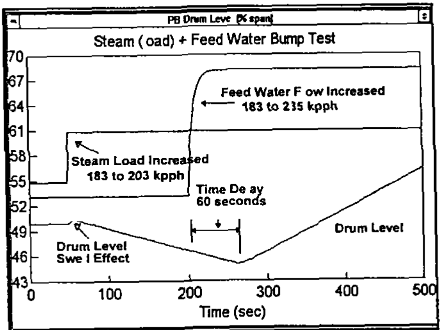

FIGURE 5 VisSim
SIMULATION OF PACKAGE BOILER BUMP TEST

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

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 desuperheater sections (modeled as flow resistors with desuperheater 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 Dahlin [4] 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 1 0 The recommended tuning rule for this case [9] is presented below

Integrator plus Deadtime Tuning Rule

Algorithm: PI

$$G_C(s) \quad K_C\left[1+\frac{1}{T_R s}\right] \qquad 2\ 0$$

Select closed loop time constant λ For a level controller this 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 λ that has been selected, and the process transfer function parameters K_P and T_d the tuning parameters are

$$T_R = 2\lambda + T_d \qquad 3\ 0$$

$$K_C = \frac{T_R}{K_P(\lambda + T_d)^2} \qquad 4\ 0$$

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

$$G_{loop}(s) = G_C(s)G_P(s) = 1.33\left[\frac{300s+1}{300s}\right]\frac{0\ 005e^{-59s}}{s} \qquad 5\ 0$$

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 Pade 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

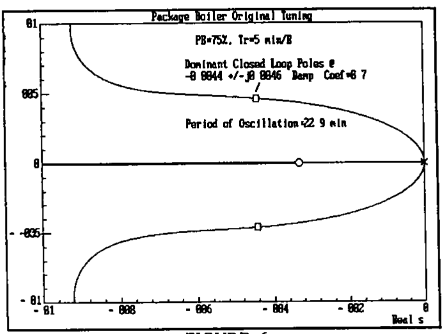

FIGURE 6
PACKAGE BOILER OLD TUNING ROOT LOCUS

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 min*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*60/0 0046 22 9 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

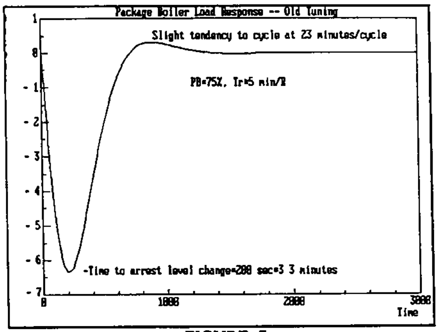

FIGURE 7
PACKAGE BOILER - OLD TUNING LOAD RESPONSE

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 Drum Level -- New Tuning

The final tuning for the package boiler was based on a λ (Lambda), or closed loop time constant of 3 0 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 3 0 minutes produces the following tuning constants

Reset time $\quad T_R \quad 2\lambda + T_d \approx 7.0 \text{ min}$

Gain $\quad K_C \quad \dfrac{T_R}{K_P(\lambda + T_d)^2} \approx 1.4$

Prop Band $\quad PB \quad \dfrac{100}{1.4} - 71.4\%$

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

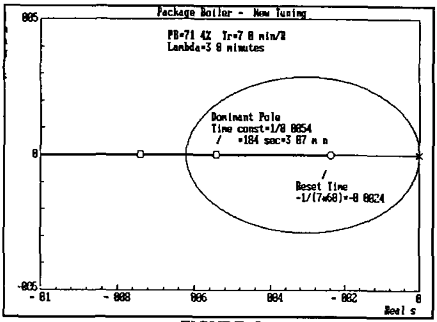

FIGURE 8
PACKAGE BOILER NEW TUNING ROOT LOCUS

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

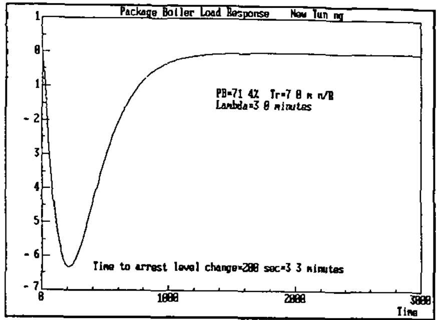

FIGURE 9
PACKAGE BOILER - NEW TUNING LOAD RESPONSE

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 (integrator plus deadtime 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 deadtime A summary of their tuning is listed below

Table V - Summary of Tuning

Loop	Lambda (minutes)
Package Boiler Steam Drum	3.0
Waste Fuel Boiler Steam Drum	3.0
Recovery Boiler Steam Drum	3.0
600# Stm Hdr-Package Blr Swing	1.05
600# Stm Hdr-Waste Fuel Blr Swing	10.5
Waste Fuel Surge Bin	10.5
Waste Fuel Feed Bin	3.2
Boiler Water Conductivity	60

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%

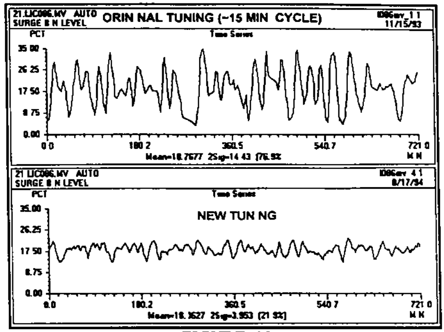

FIGURE 10
WASTE FUEL SURGE BIN VARIABILITY

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, ram 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 variabilty 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

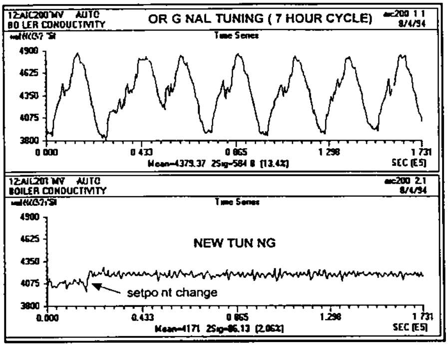

FIGURE 11
BOILER CONDUCTIVITY VARIABILITY

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 selection of the Lambda values themselves, proved to be the chief elements for solving the boiler tripping problem

NOMENCLATURE AND SYMBOLS

aliasing	apparent distortion of an analog signal when sampled too slowly
anti aliasing	a special signal filtering technique to prevent alising
BASIC	a simple programming language
DCS	distributed control system
$G_C(s)$	controller transfer function in the continuous (Laplace) domain
$G_P(s)$	process transfer function in the continuous (Laplace) domain
kpph	thousands of pounds per hour
K_C	controller gain (100%/PB%)
K_P	process gain normally expressed as process measurement (% of span)/ controller output (%) For integrating process variables, this quantity is also / second
lbs	pounds
Lambda tuning	a method of tuning which equires the user to specify the desired closed loop time constant - Lambda
Lambda	the desired closed loop time constant - usually in seconds
hr	hour
PB	Proportional Band (PB% 100%/controller gain)
PI	Proportional Integral controller
PID	Proportional-Integral Derivative controller
psi	pressure measurement, pounds per square inch (1 0 psi 6 89 kPa)
PRV	pressure reducing valve
shrink/swell effect	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
RHP	right half plane (of the s plane) the unstable region
T_d	deadtime
T_R	controller reset or integral time/repeat

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REFERENCES

1 Morton A J and Price P H , " *The Controllability of Steam Output Pressure and Water Level in Drum Boilers* ', Instit of Mech Eng Journal, 1977

2 TUTSIM (Dynamic simulation software) by TUTSIM Products, 200 California Avenue, Palo Alto, CA 94306

3 VisSim, (Dynamic simulation software) by Visual Solutions Inc 487 Groton Road, Westford, Massachusetts, USA 01886

4 Dahlin E B , *"Designing and Tuning Digital Controllers"*, Instrum and Control Systems, 41 (6), 77, 1968

5 Bialkowski W L., Haggman B C and Millette S K , *"Pulp and Paper Process Control Training Since 1984 '* Pulp and Paper Canada 95 4 (1994)

6 EnTech™ - Process Control for Engineers I - Course Curriculum Application of Model Based Control Concepts - Theory and Practice for Engineers

7 EnTech™ - Process Control for Engineers II - Course Curriculum Application of Stochastic Control Concepts to the Reduction of Process Variability

8 Morari M. and Zafiriou E , *"Robust Process Control"* Prentice Hall Inc , Englewood Cliffs, NJ, (1989)

9 Chien I-L and Fruehauf P S , *"Consider IMC Tuning to Improve Controller Performance"*, Hydrocarbon Processing, Oct 1990

10 Program CC (Control Engineering Software) by Peter M Thompson, Systems Technology, Inc ,13766 South Hawthorn Boulevard, Hawthorn, California, USA 90250 7083

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2. PROCESS DYNAMICS

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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

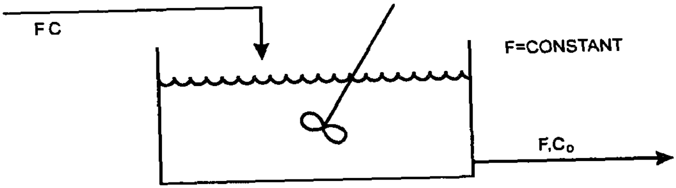

Fig. 2-1. Stock chest with agitation.

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.

(Rate of Fiber Accumulation) = (Rate of Fiber In) - (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 d scharge
C_I = Consistency at chest inlet

2 *Empirical*

- Modeling of del gnification k netics 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

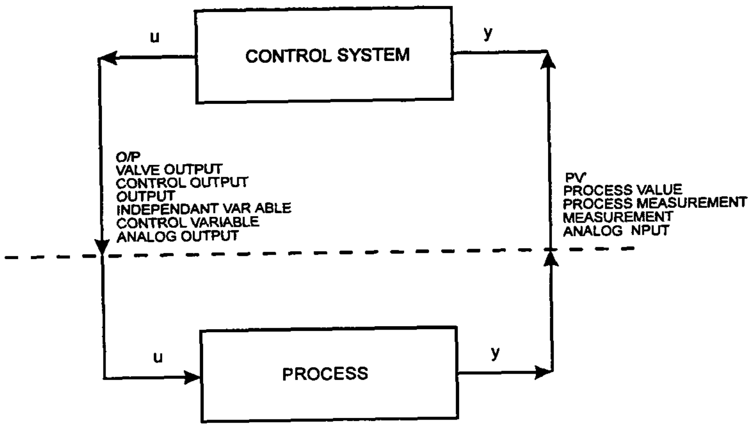

Fig. 2-2 Signal flow information between the process and the control system

First Order System

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

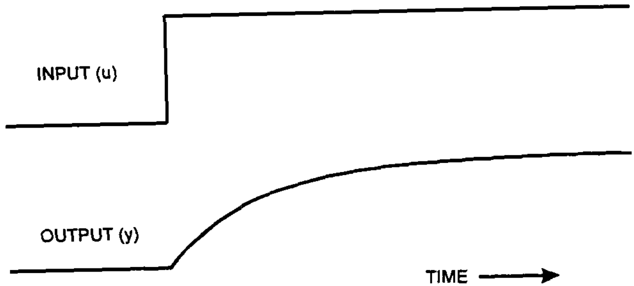

Fig 2-3. Bump test or step test for a first order system.

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*
τ 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.

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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

$$\begin{aligned} K_P &= \frac{\Delta y}{\Delta u} \\ &= \frac{150 \text{ USGPM}}{5\% \text{ Output}} \\ &= 30 \text{ USGPM/\% Output} \end{aligned}$$

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 <u>OR</u> negative Consider a typical setup for consistency control.

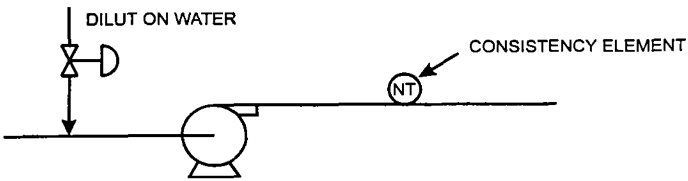


Fig 2-4 Stock Consistency process

If the di ution valve is opened, the consistency wou d typically respond as fol ows

Fig 2 5 Typical consistency response to a step increase of the dilut on 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 $t = \tau$

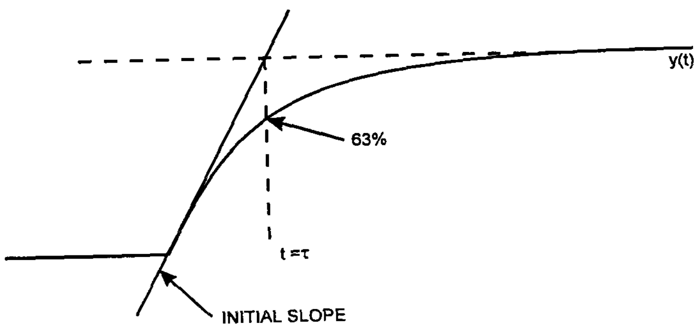


Fig 2-6 Output response of a first order system

2) The output change, expressed as a per cent of its total change Δy, versus time, is given in the following table.

Time (after step)	*Output change as a % of Δy*
1τ	63.2
2τ	86.5
3τ	95.0
4τ	98.2
5τ	99.3
6τ	99.8
7τ	99 9
8τ	≈100

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.

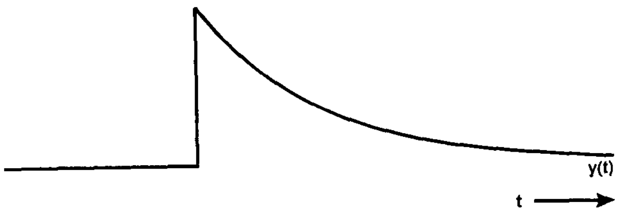


Fig 2-7 Impulse Response of a first order system

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

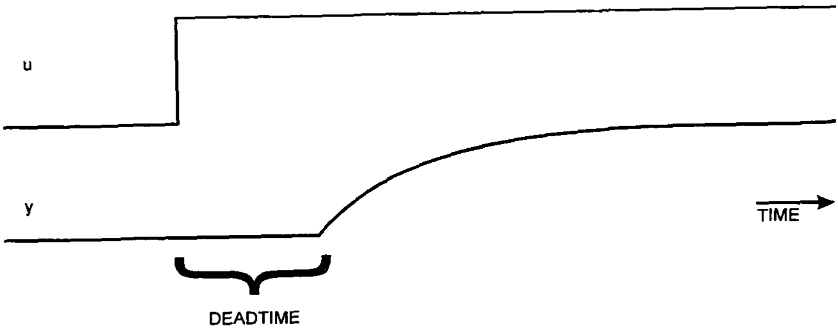


Fig 2-8 Output response of a first order with deadtime system

Deadtime is usual y 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 deadt me process can be described by the fol ow ng d fferent al equation

$$\tau \frac{dy(t)}{dt} + y(t) \quad K_P u(t \quad T_d)$$

where T_d is the deadtime.

We wi l see later that deadt me can be very destab lizing to a control er

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: ζ. 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 ζ. 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 crit cally damped or overdamped second order systems These diff culties 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.

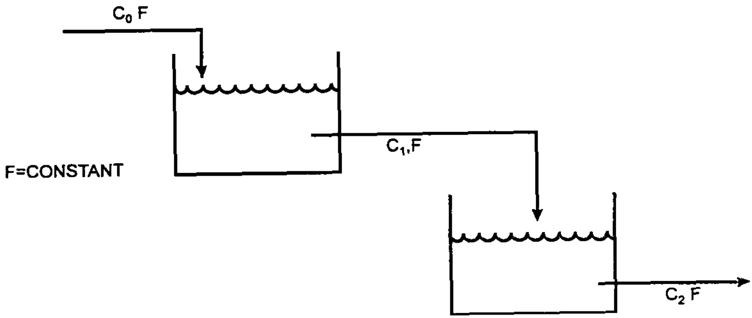


Fig 2-9 Stock Chests in series

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

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

<u>Example</u>. Consider a mass spring system immersed in a viscous liquid.

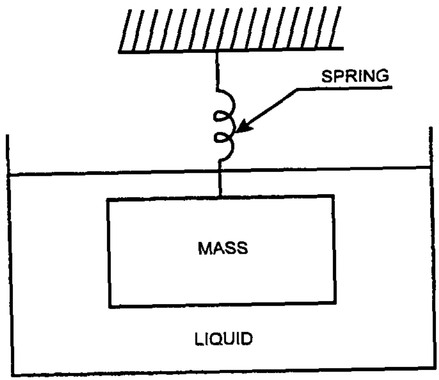


Fig. 2-11. Mass spring damper system.

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

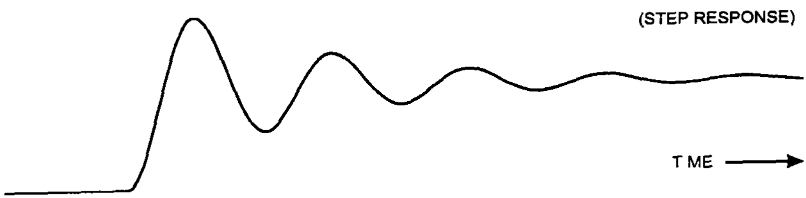


Fig 2-12. Mass position response due to a step change in spring upper end position

Describing Second Order Underdamped Systems

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

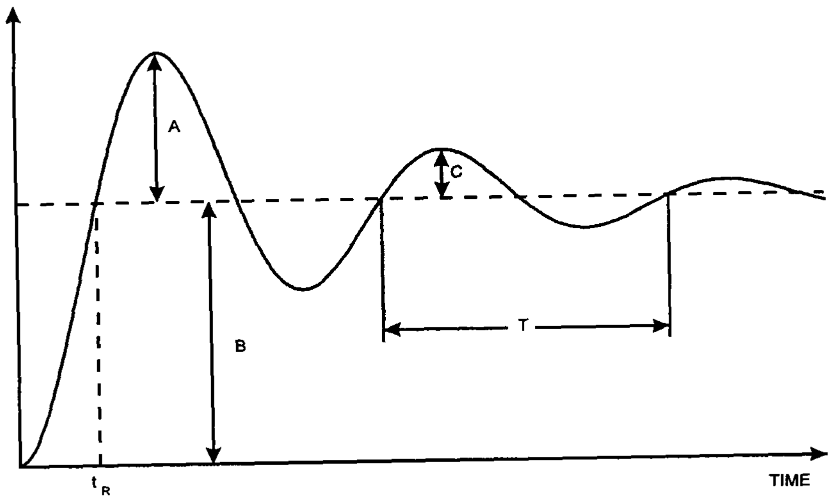


Fig 2-13 Step Response of a second order underdamped system.

There are several common terms used in describing this type of response

Overshoot: Factor by which the response overshoots the final value

$$
\begin{aligned}
O &= A / B \\
&- \exp\left(\frac{-\pi\zeta}{\sqrt{1-\zeta^2}}\right)
\end{aligned}
$$

Decay Ratio D: Ratio by which successive peaks decay

$$
\begin{aligned}
D &= C/A \\
&- \exp\left(\frac{-2\pi\zeta}{\sqrt{1\ \ \zeta^2}}\right) \\
&= (\text{Overshoot})^2
\end{aligned}
$$

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 ω:

$$
\omega = \frac{\sqrt{1-\zeta^2}}{\tau} \quad \text{(radians / unit time)}
$$

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

Undamped Natural Frequency: Frequency at which the system would oscillate if damping were removed ($\zeta \quad 0$)

$$
f_n - \frac{1}{T_n} = \frac{1}{2\pi\tau}
$$

$$
\omega_n - \frac{1}{\tau}
$$

Step responses of a second order underdamped system for various values of the damping coefficient are shown in the following figure.

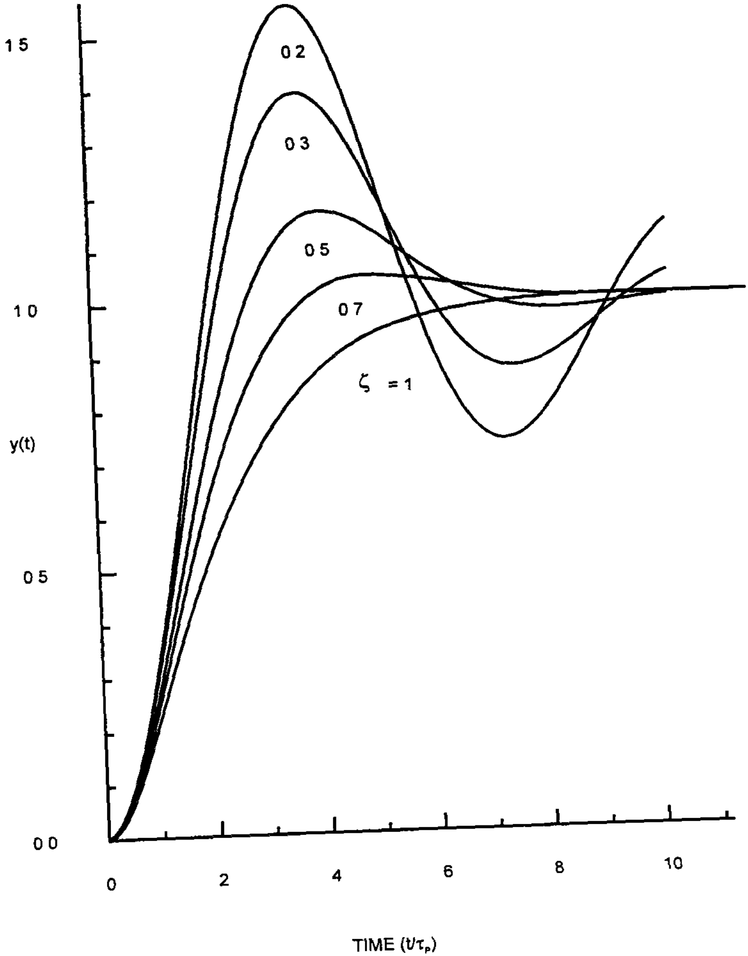


Fig 2-14. Second order underdamped step responses for various values of the damping coefficient.

Note that the overshoot O depends only on the damping coefficient ζ and not on the parameter τ. The percent overshoot as a function of ζ is shown below.

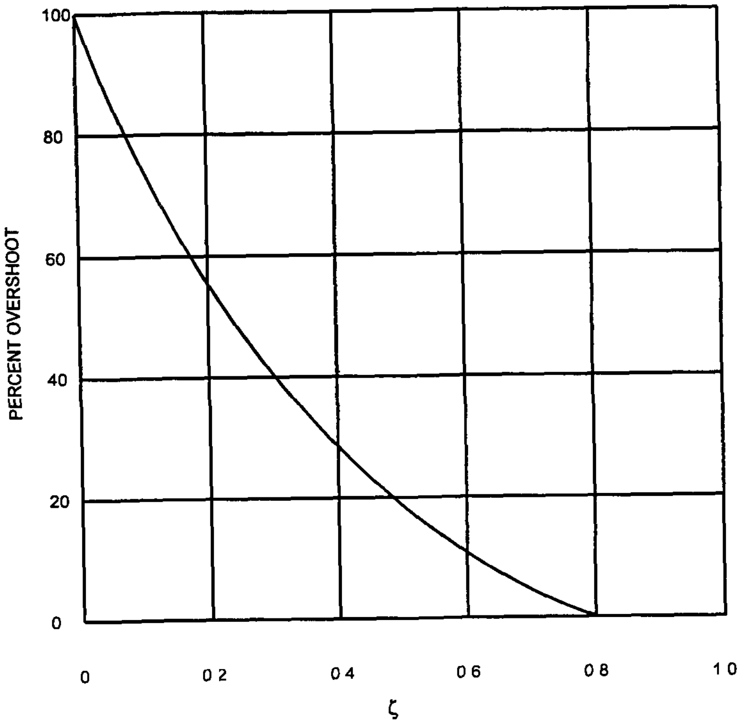


Fig 2-15 Plot of Percent Overshoot vs Damping Coefficient.

Solving for the damping coefficient ς in terms of the overshoot O yields

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

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.

<u>Example</u>: Tank level

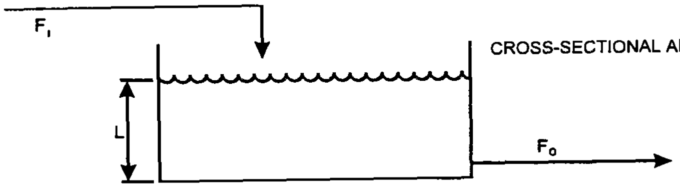


Fig. 2-16 Stock Chest Level.

An instantaneous material balance for this system is as follows.

$$\frac{dV}{dt} \quad F_{IN} - F_{OUT}$$

or

$$\frac{dL}{dt} - \frac{1}{A}(F_{IN} - F_{OUT})$$

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

$$L \quad \frac{1}{A}\int_0 (F_i \quad F_o)dt + L_o$$

Suppose the tank is initially empty $(L_o - 0)$ and there is no flow out of the tank $(F_o \quad 0)$. The leve will be

$$L \quad \frac{1}{A}\int F_i dt$$

If F_i is positive (flow going into tank), then the tank w l fil up

An integrating system can be described by

$$y(t_1) \quad K_P \int_0 u(t)dt$$

To keep a tank leve constant the difference between <u>all</u> in et 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:

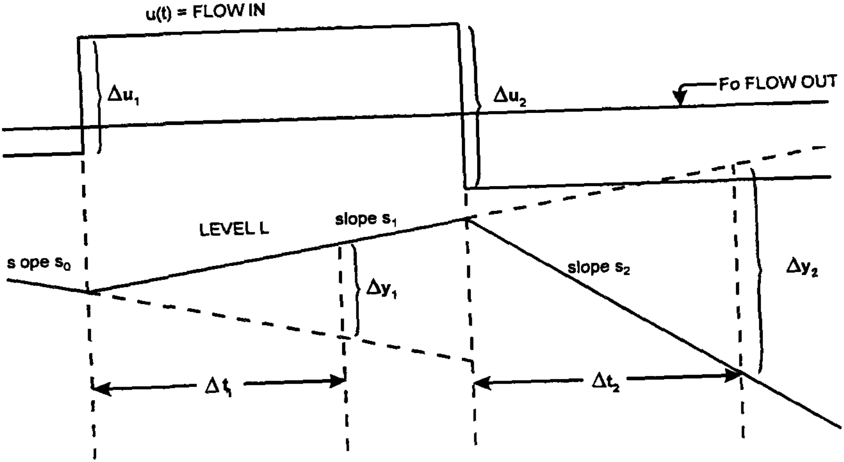


Fig 2-17 Bump test for a level process.

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 wil be covered in more detai 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 can not 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

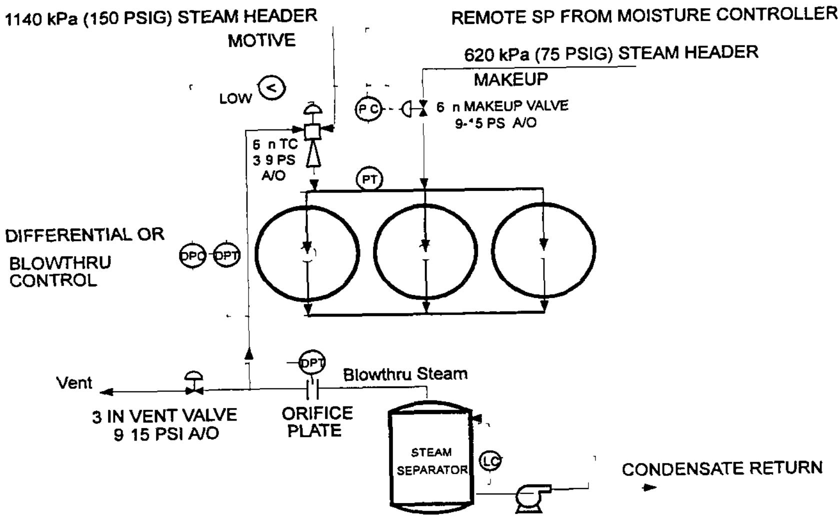


Fig 2-18 Typical Paper Machine Dryer Section

The Dryer pressure response to a bump test may look like either of those shown on the following two pages

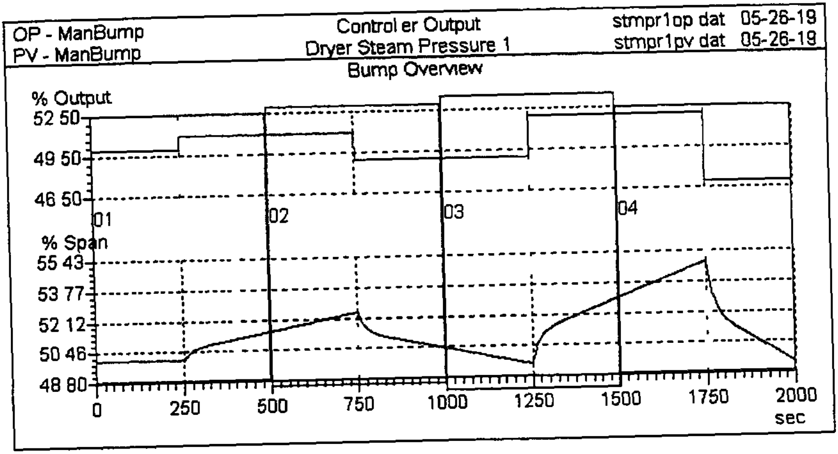


Fig. 2-19. Dryer Steam Pressure 1 bump test overview.

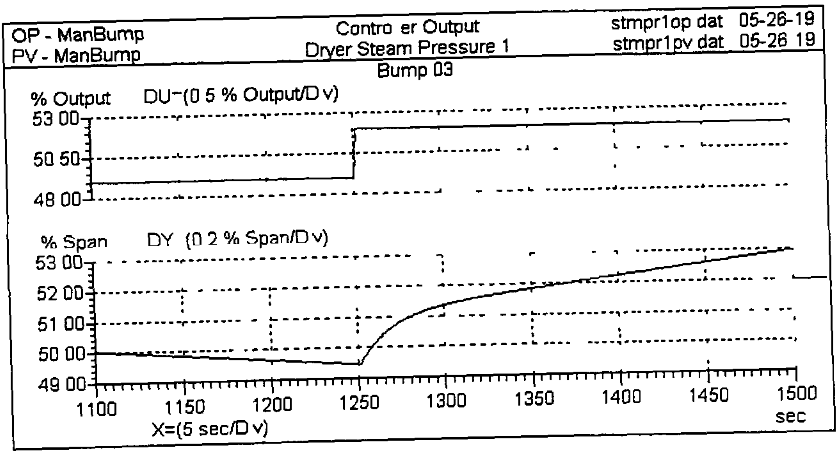


Fig 2-20 Dryer Steam Pressure 1 bump 3 close-up

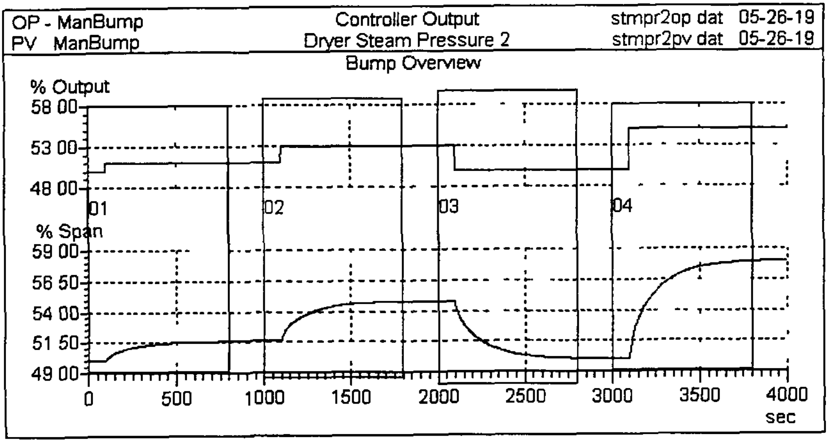


Fig 2-21 Dryer Steam Pressure 2 bump test overview

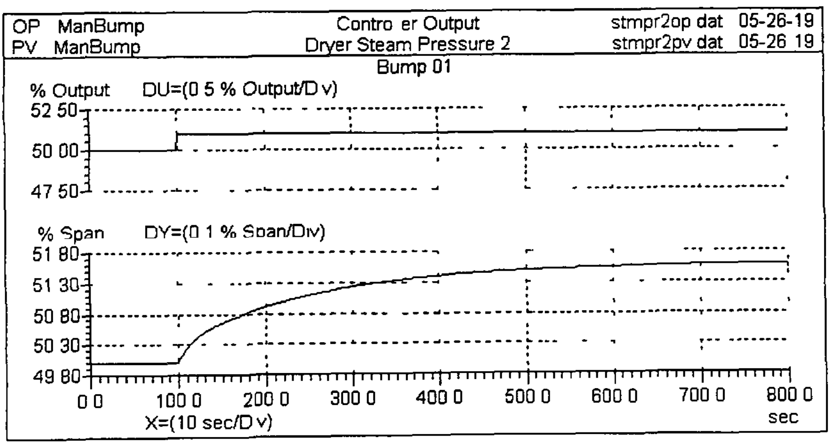


Fig 2-22 Dryer Steam Pressure 2 bump 1 close-up

Stock Chest Level With Two Stock Chests In Series

Consider the stock chest level control scheme shown below

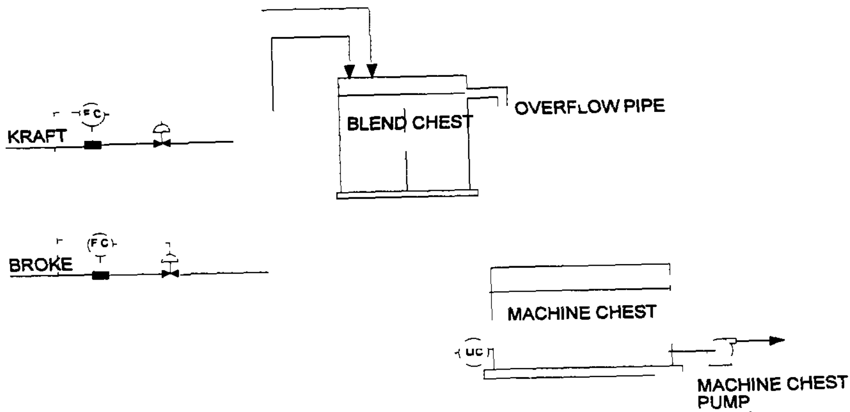


Fig 2-23 Stock blending system with gravity flow from Blend Chest to Machine Chest.

A sample bump test is shown on the following page

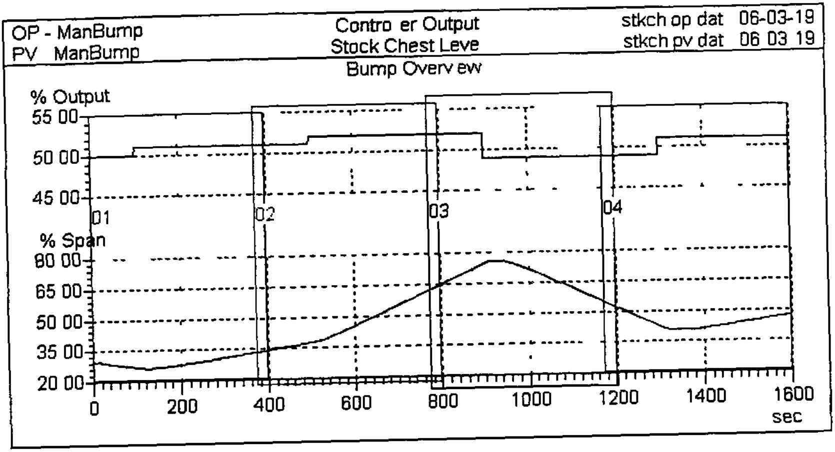


Fig 2-24. Machine Chest Level bump test overview

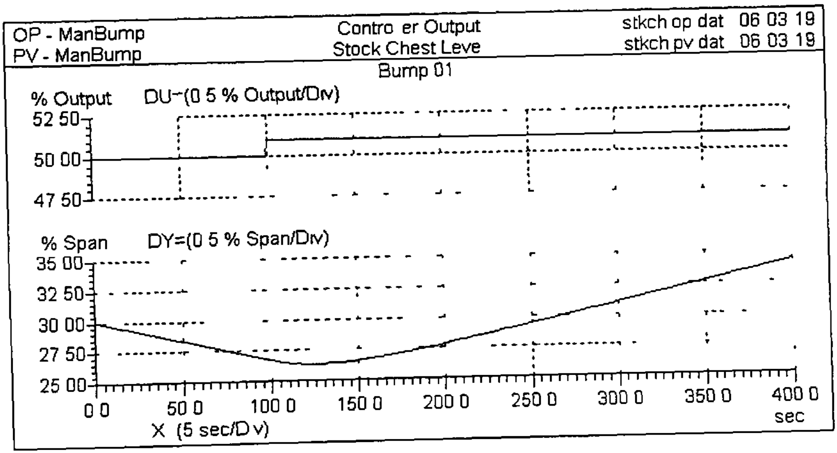


Fig 2-25. Machine Chest Level bump 1 close-up

Bleach Plant Steam Mixer Temperature

Consider the Bleach Plant Steam Mixer Temperature control loop shown below

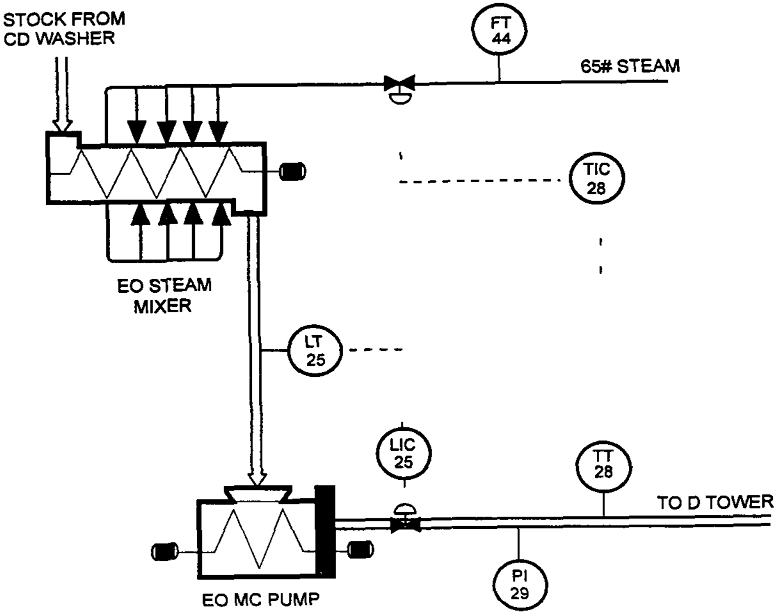


Fig 2-26 Bleach Plant Steam Mixer process area

A sample bump test is shown on the following page

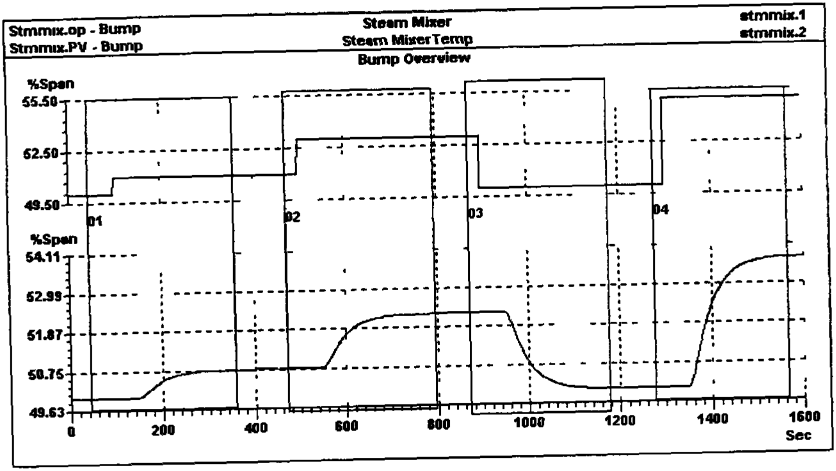


Fig 2-27. Steam Mixer Temperature bump test overview

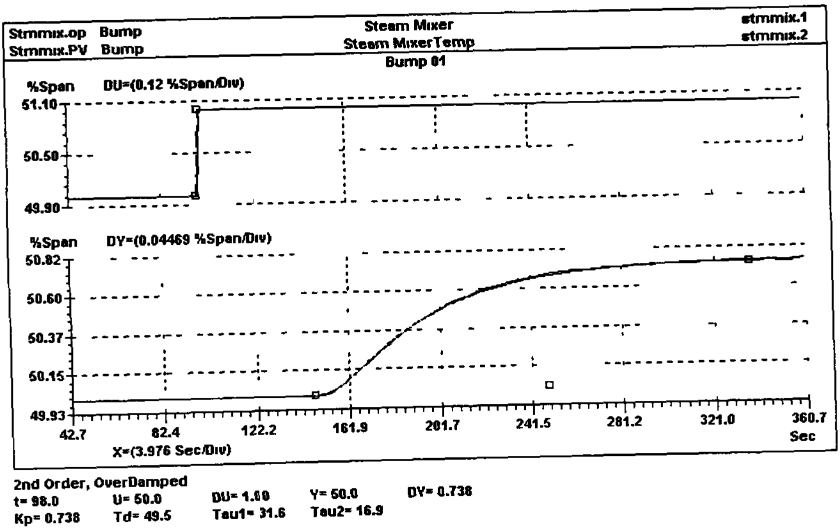


Fig 2-28 Steam Mixer Temperature bump 2 close-up

Boiler Steam Drum Level

Consider the Boiler Steam Drum Level control loop shown below.

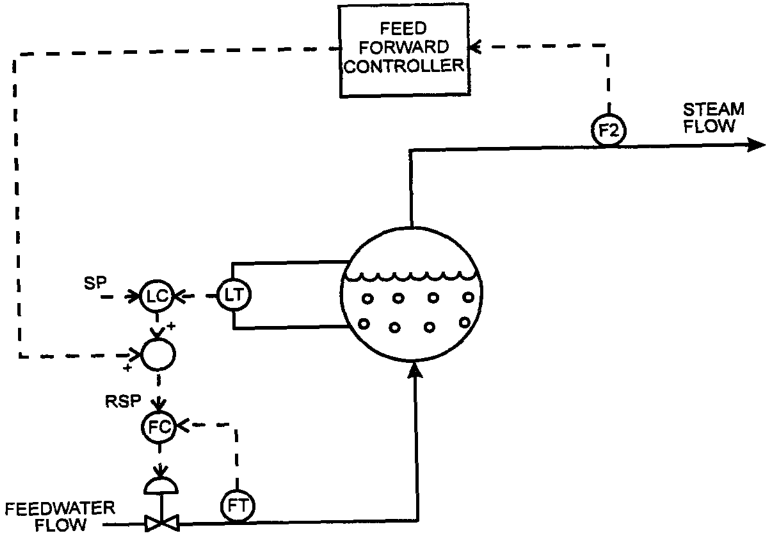


Fig. 2-29 Three-element Boiler Steam Drum Level

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

If the feedwater flow is increased, the l quid temperature w ll initially drop and many of the bubbles will collapse, with the result that the level will init al y drop (shrink-swel effect)

Eventually, the temperature wi l 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

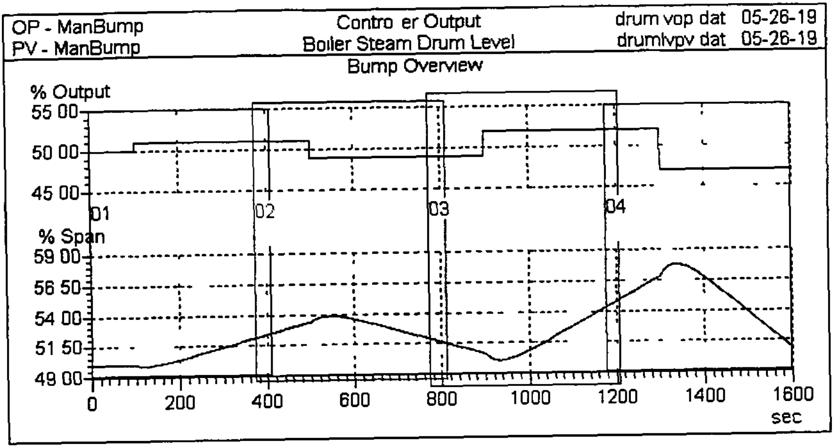


Fig. 2-30. Steam Drum Level bump test overview.

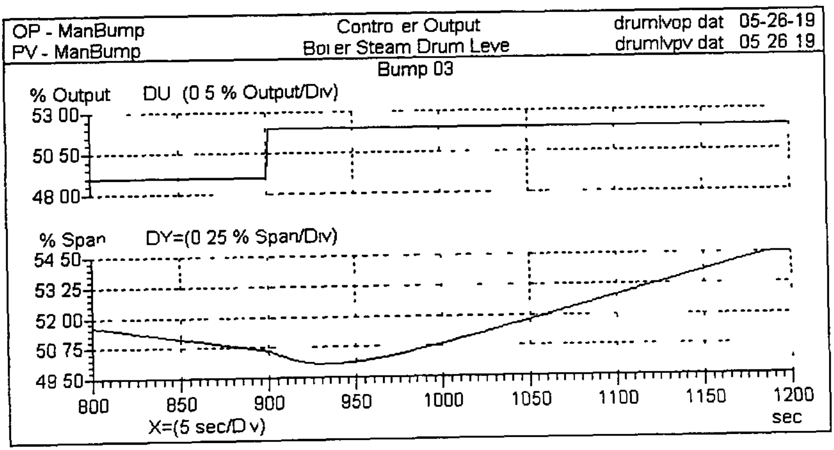


Fig. 2-31 Steam Drum Level bump 3 close-up

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

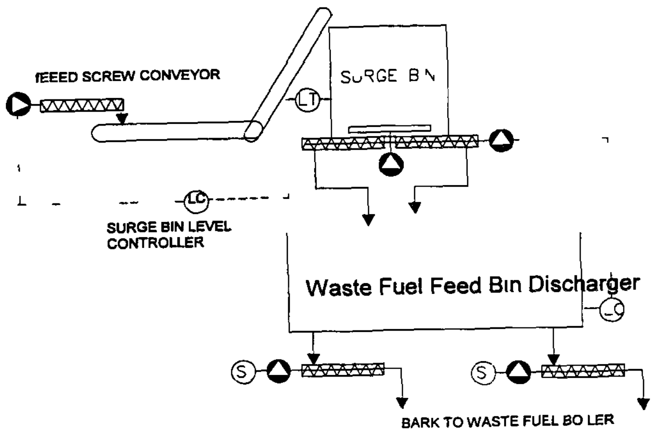


Fig 2-32 Waste Fuel Boiler Bark Fuel Transport System

A sample bump test is shown on the following page

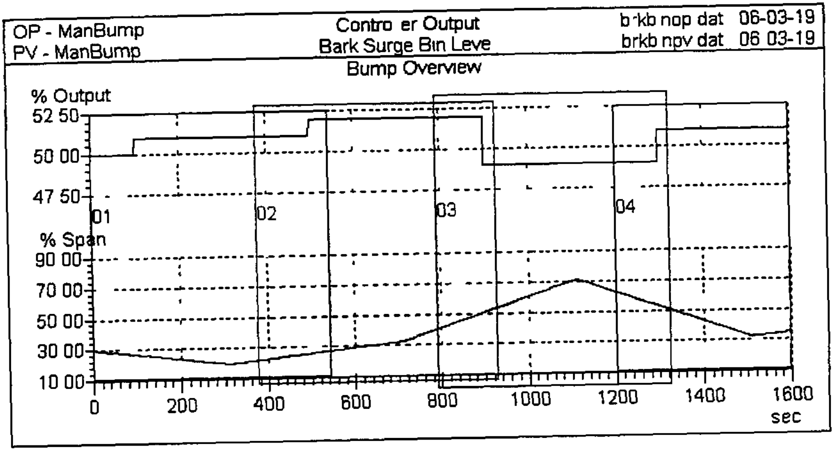


Fig 2-33 Bark Surge Bin Level bump test overview.

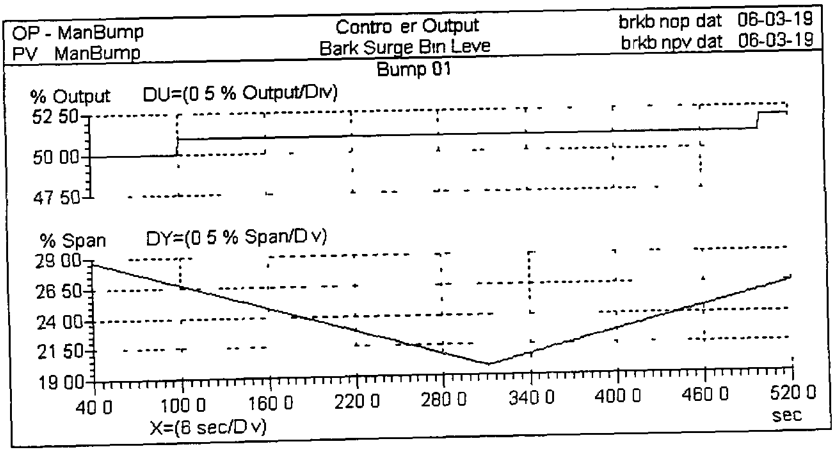
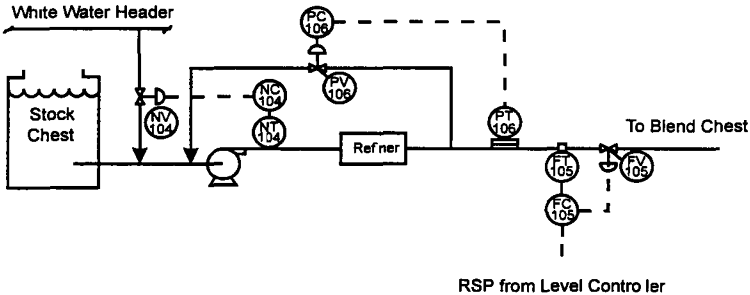


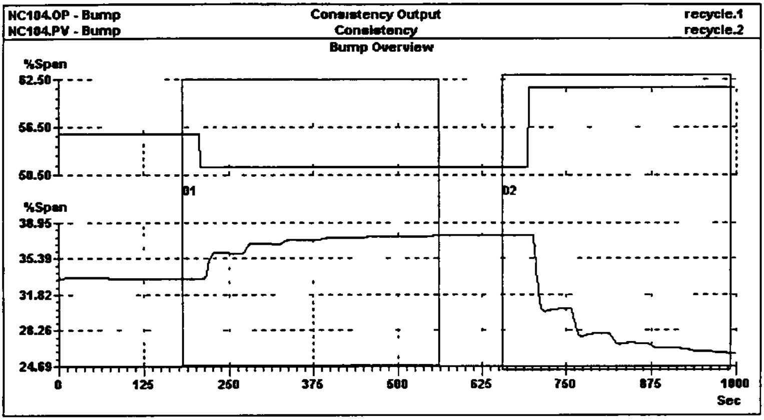
Fig 2-34 Bark Surge Bin Level bump 1 close-up

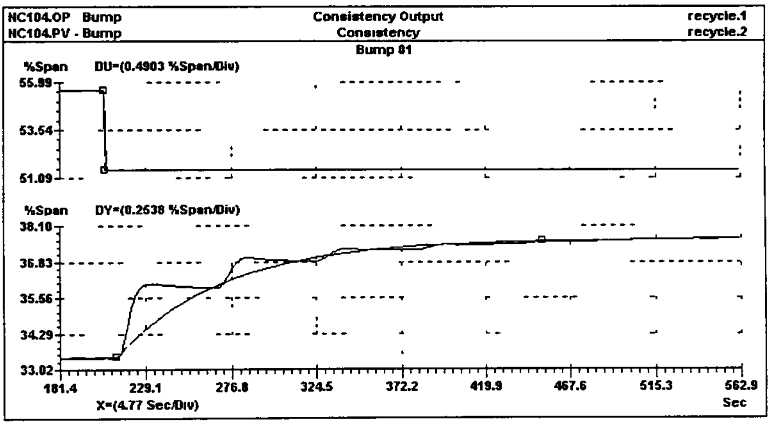
Consistency Loop with Recycle Dynamics

The prescence 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.



Sample bump test results are shown below





1st Order

t= 206	U= 55.6	DU= -4.09	Y= 33 4	DY= 4.16
Kp= 1.02	Td= 7.00	Tau1= 59.8		

APPENDIX 2A
SAMPLE BUMP TESTS

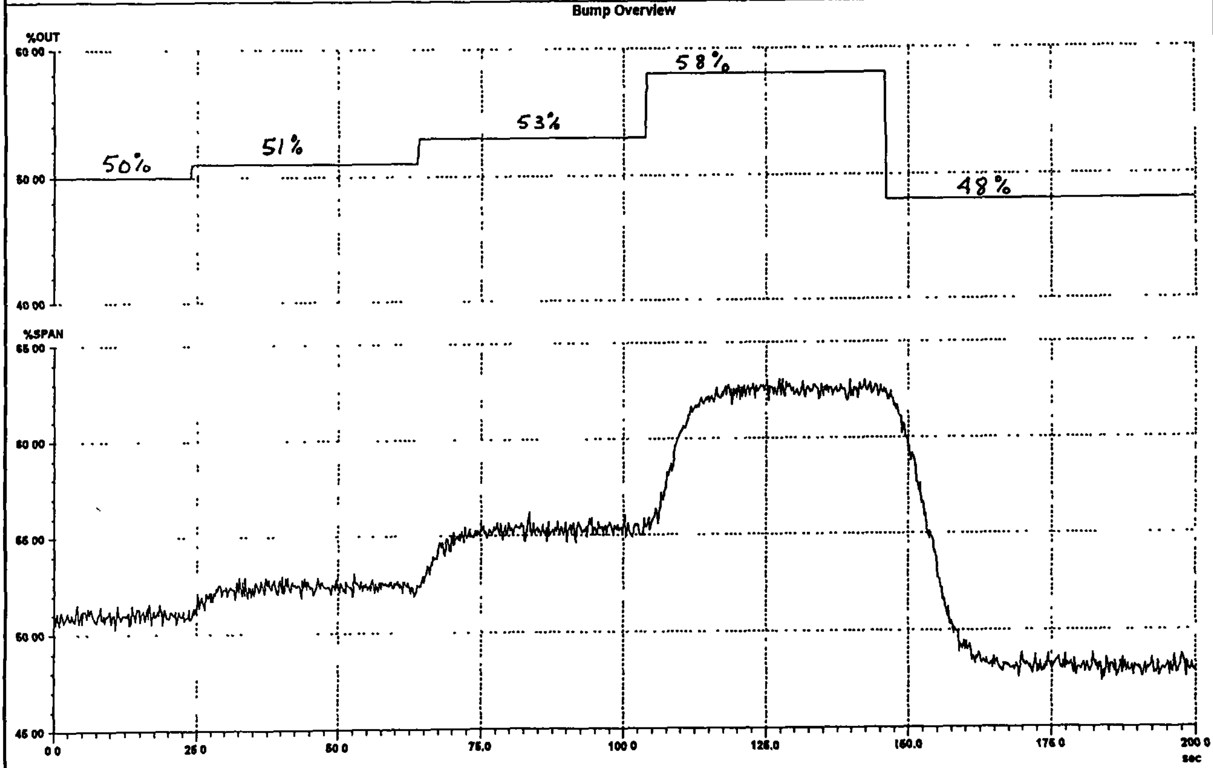
23 32 15 04 10 07

EnTech Control Engineering Inc.

ENTECH TORONTO

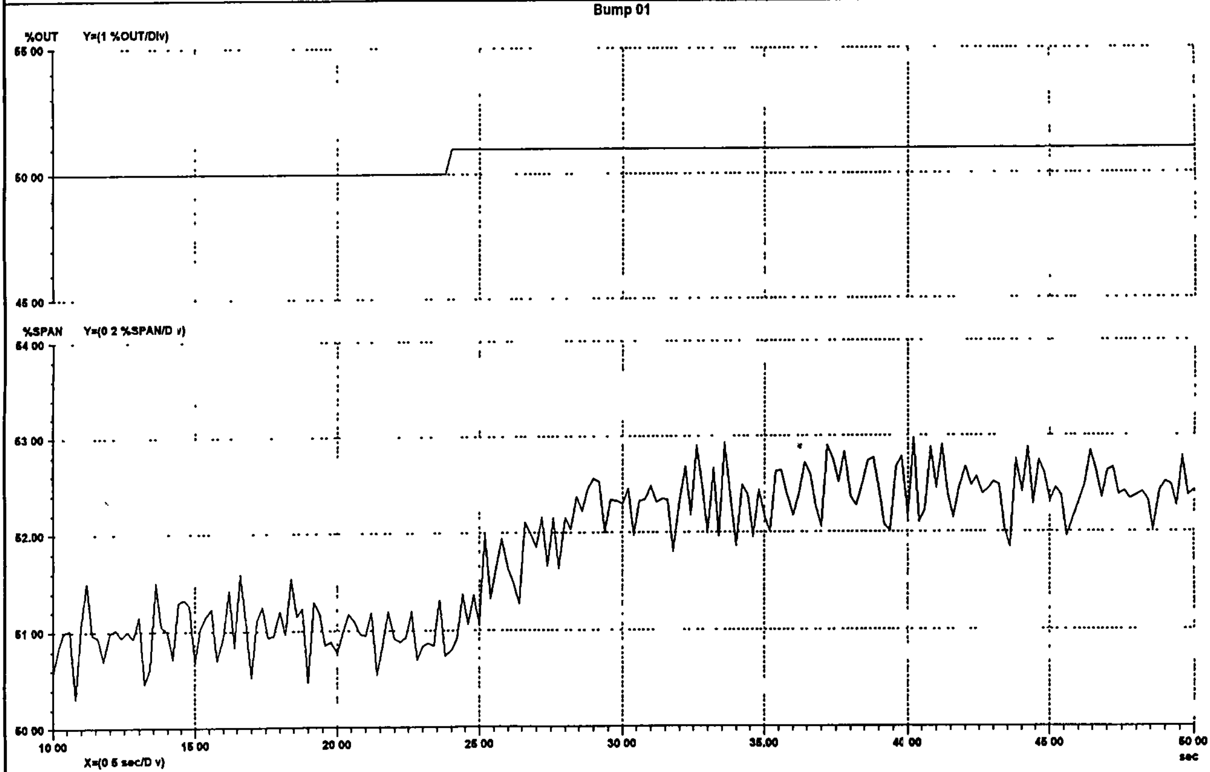
FC105 - MAN-BUMP
FT105 - MAN-BUMP

fc105bmp.dat 09-22-92
ft105bmp.dat 09-22-92



EnTech Control Engineering Inc. ENTECH TORONTO

FC105 - MAN-BUMP fc105bmp.dat 09-22-92
FT105 - MAN-BUMP ft105bmp.dat 09-22-92



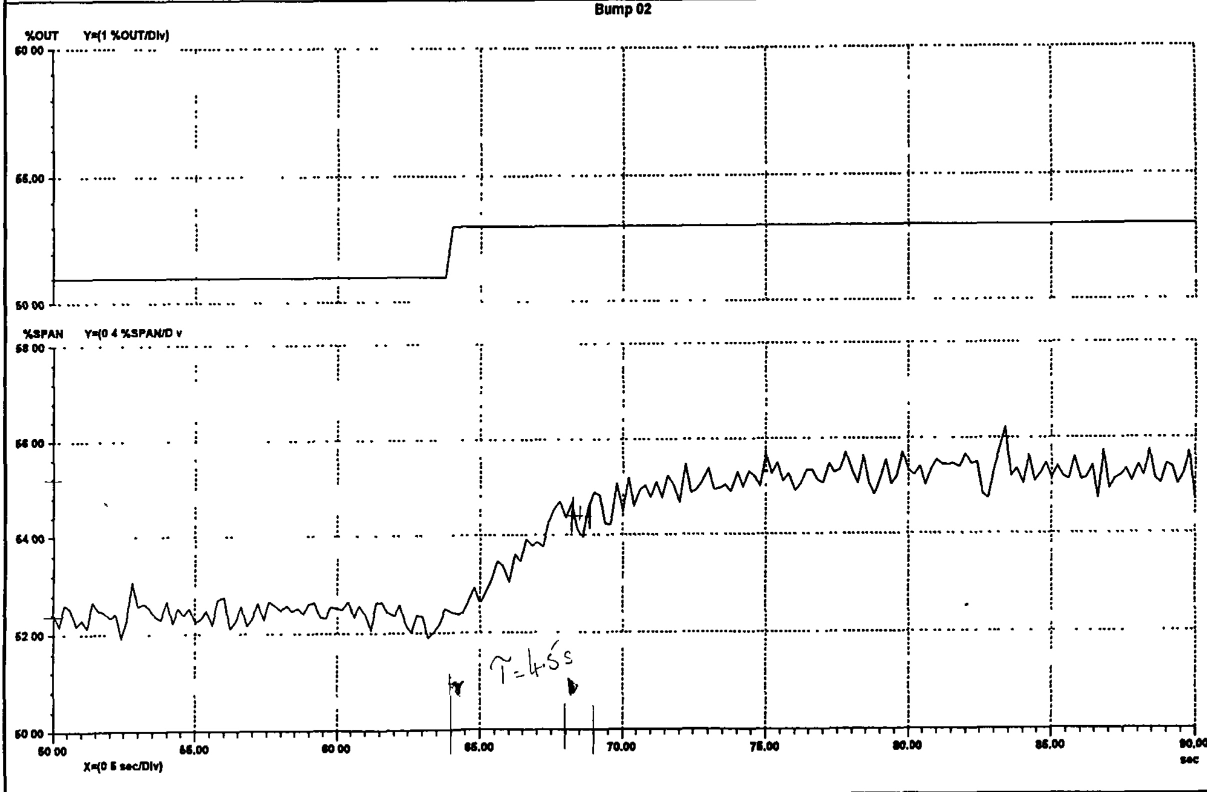
23 32 21 04 10 07

EnTech Control Engineering Inc.

ENTECH TORONTO

FC105 - MAN-BUMP
FT105 - MAN-BUMP

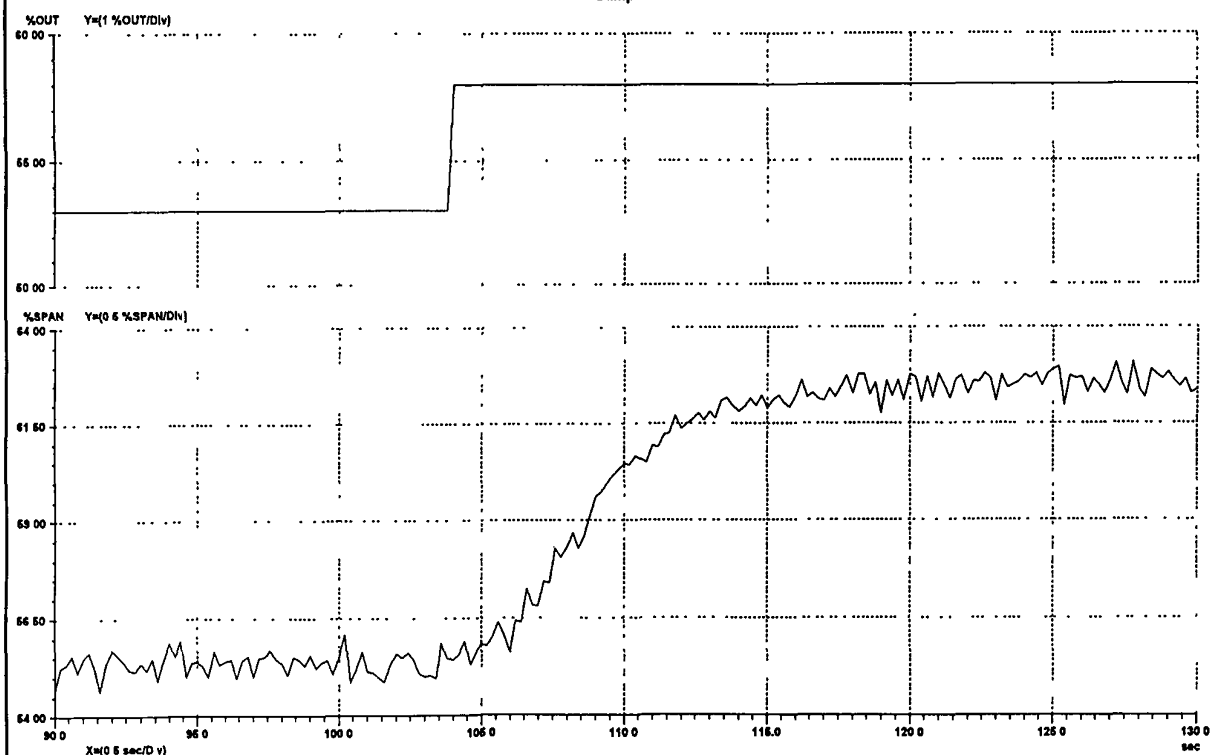
fc105bmp.dat 09-22-92
ft105bmp.dat 09-22-92



EnTech Control Engineering Inc.   ENTECH TORONTO

FC105 - MAN-BUMP   fc105bmp.dat 09-22-92
FT105 - MAN-BUMP   ft105bmp.dat 09-22-92

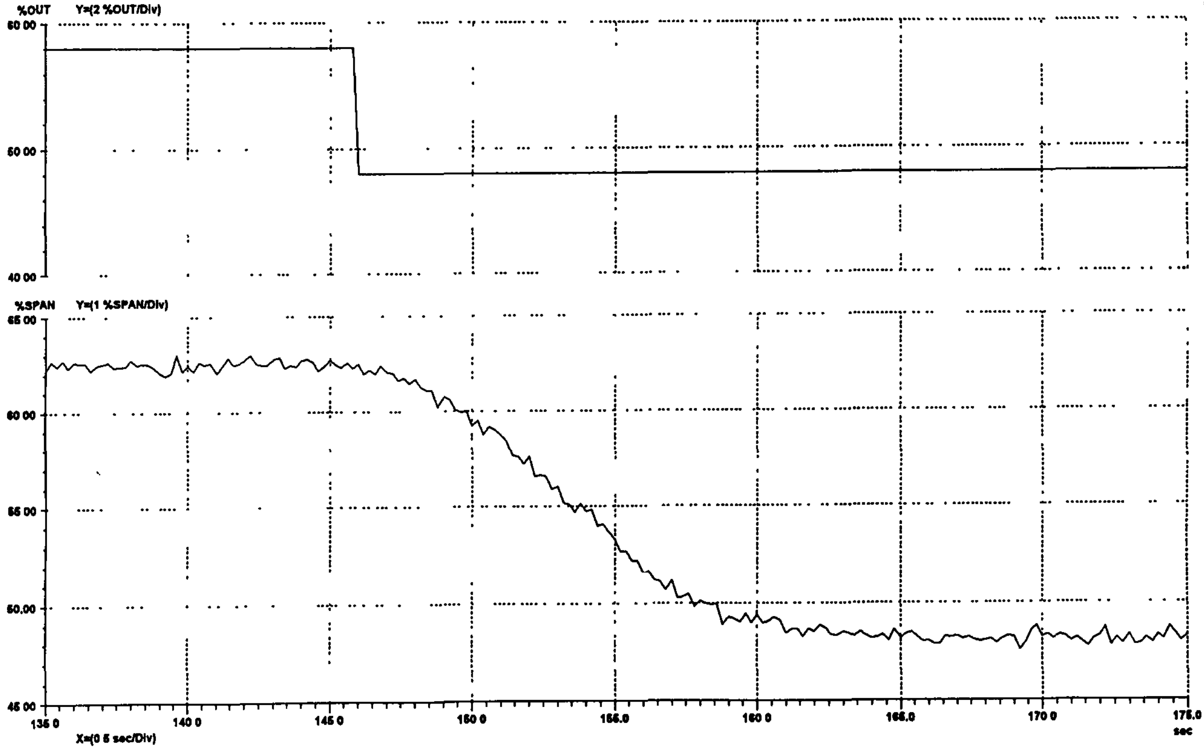
Bump 03



EnTech Control Engineering Inc. ENTECH TORONTO

FC105 - MAN-BUMP fc105bmp.dat 09-22-92
FT105 - MAN-BUMP ft105bmp.dat 09-22-92

Bump 04



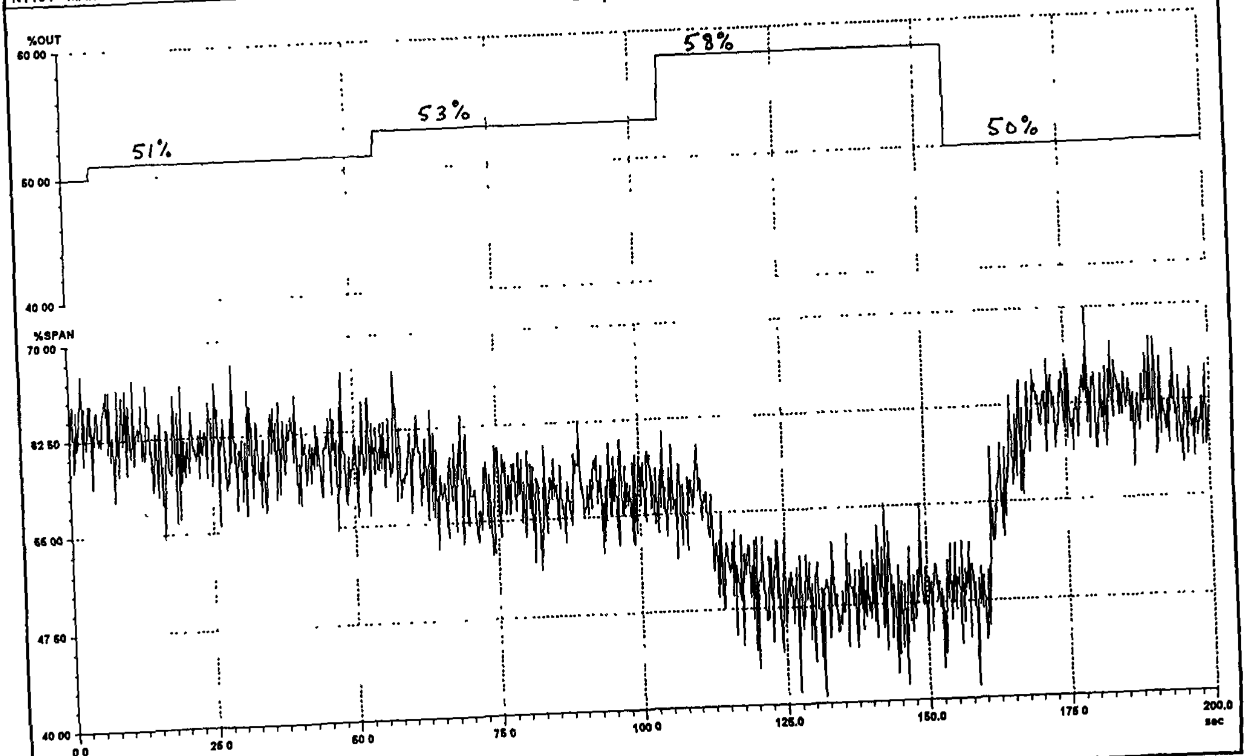
EnTech Control Engineering Inc.

ENTECH TORONTO

nc104bmp.dat 09-22-92
nt104bmp.dat 09-22-92

NC104 - MAN-BUMP
NT104 - MAN-BUMP

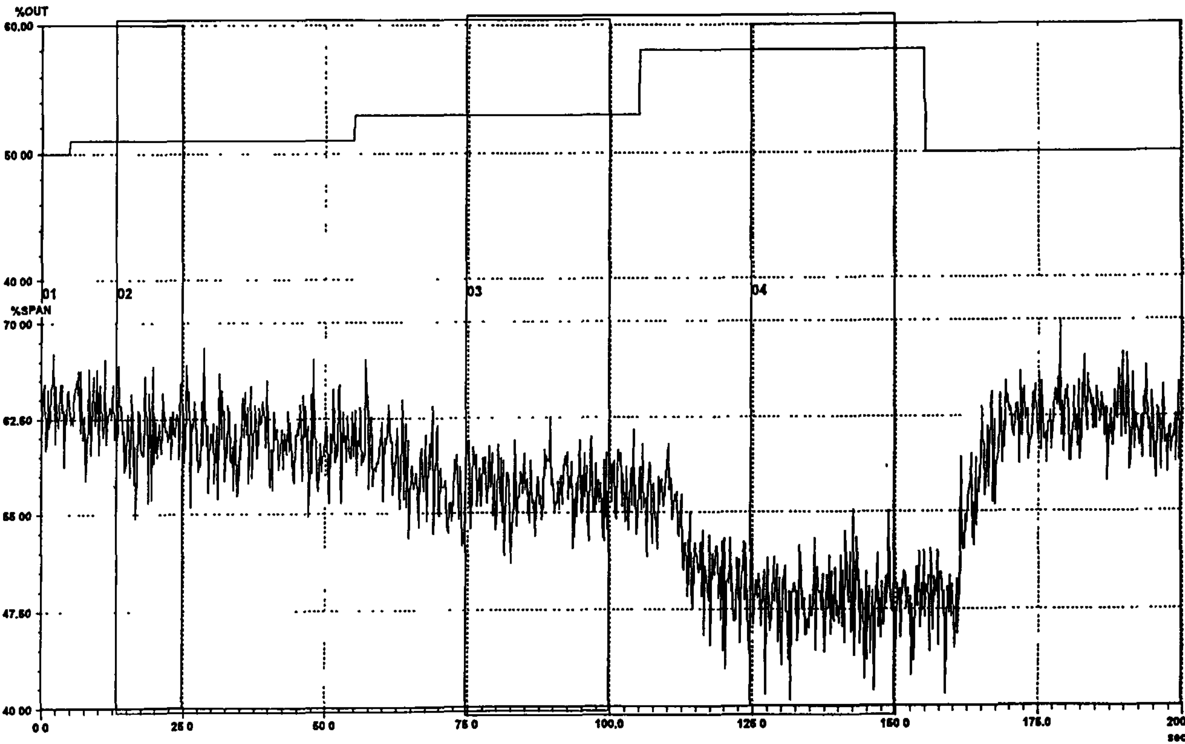
Bump Overview



EnTech Control Engineering Inc. ENTECH TORONTO

NC104 - MAN-BUMP nc104bmp.dat 09-22-92
NT104 - MAN-BUMP nt104bmp.dat 09-22-92

Bump Overview



23 32 26 04 10 07

EnTech Control Engineering Inc. ENTECH TORONTO

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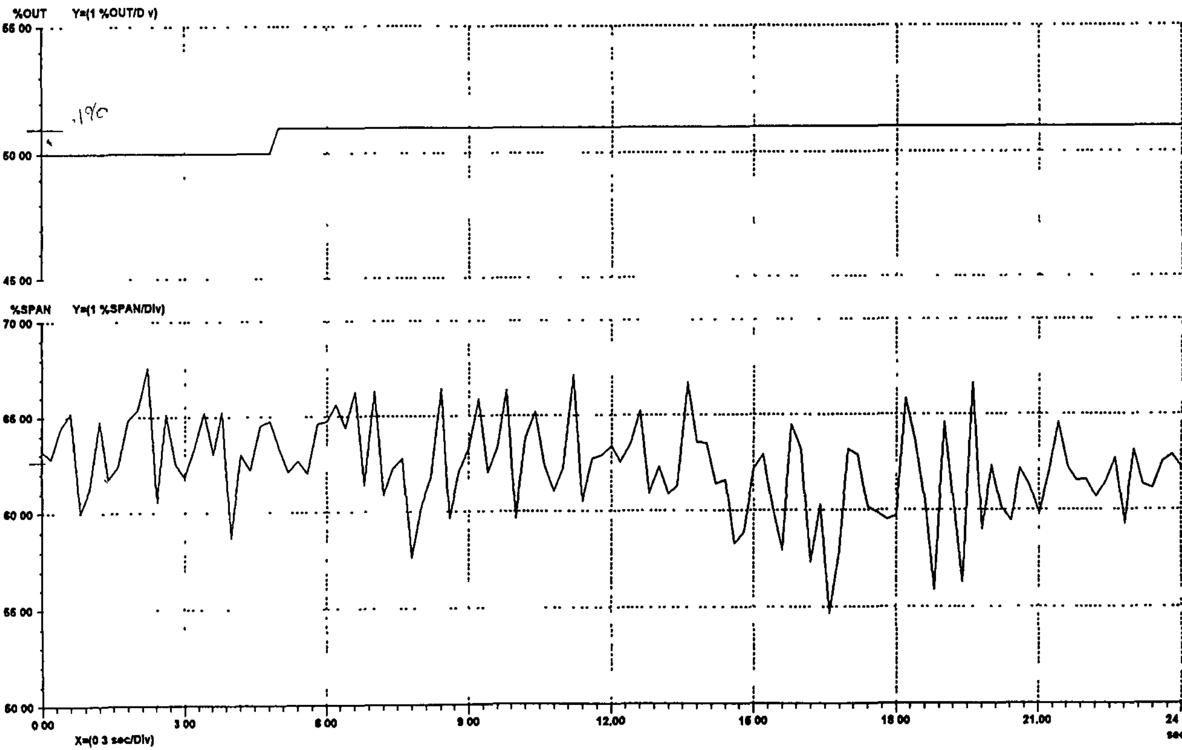
NT104 - MAN-BUMP nt104bmp.dat 09-22-92

Bump 01



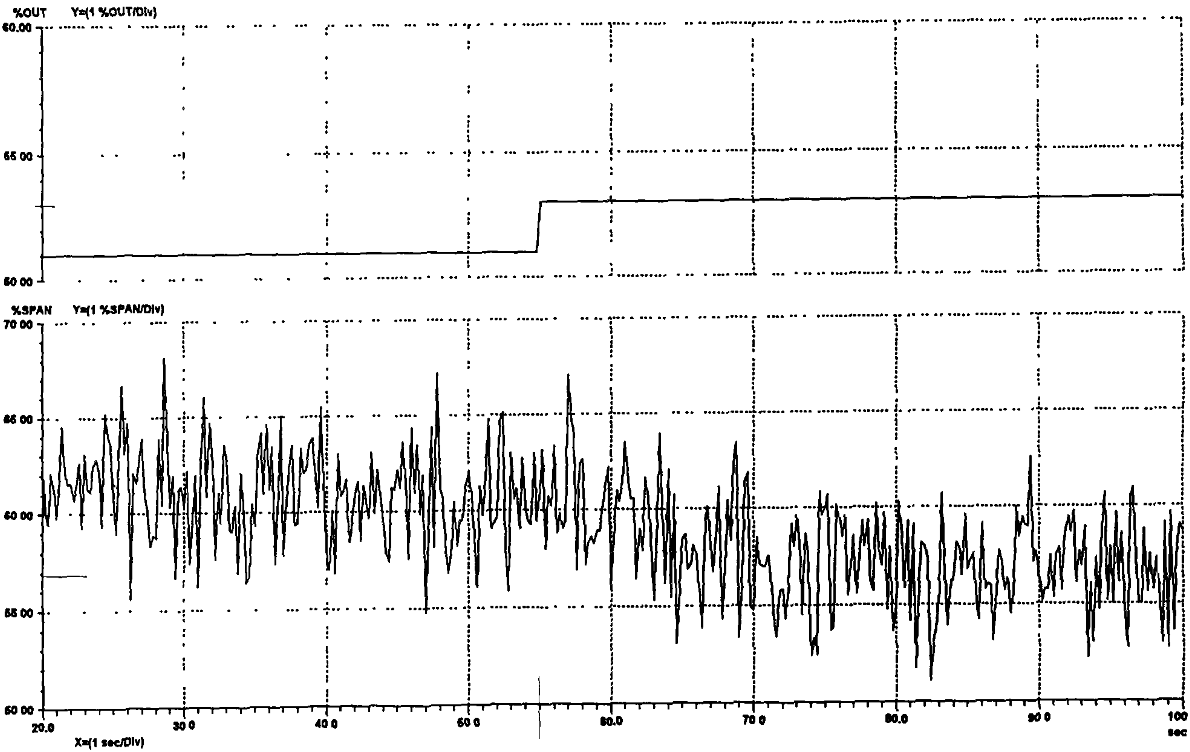
EnTech Control Engineering Inc.

ENTECH TORONTO

NC104 - MAN-BUMP
NT104 - MAN-BUMP

nc104bmp.dat 09-22-92
nt104bmp.dat 09-22-92

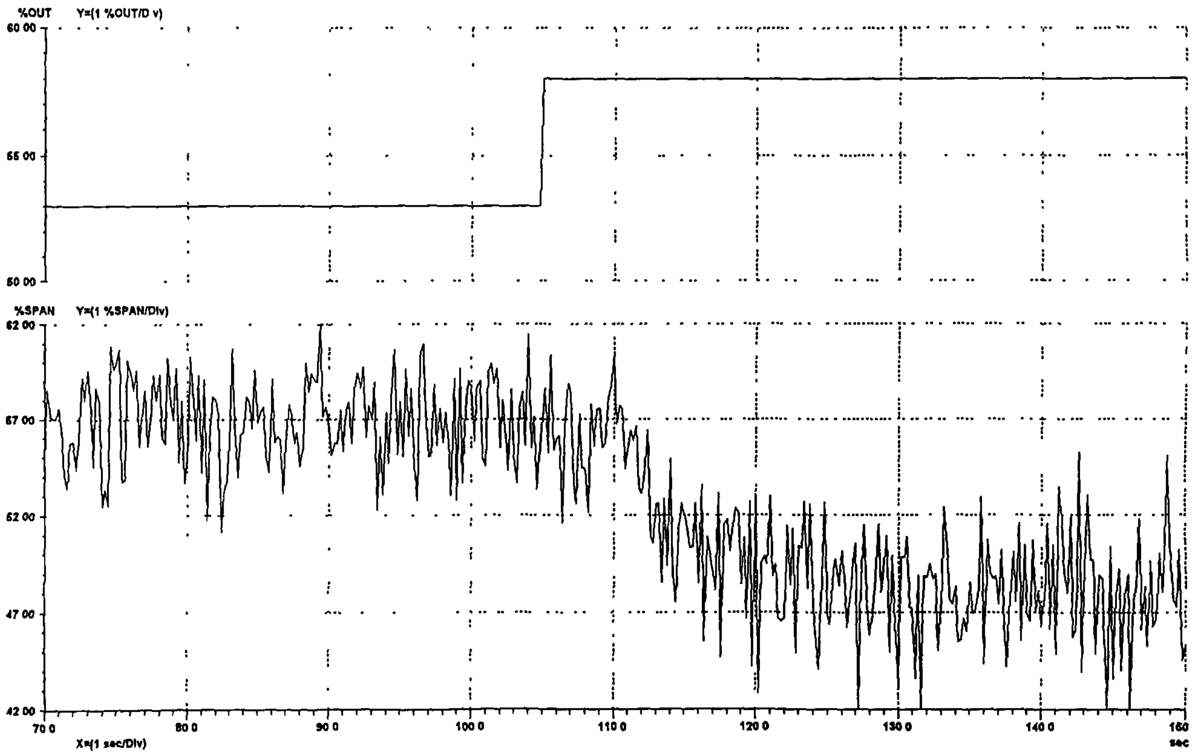
Bump 02



EnTech Control Engineering Inc. ENTECH TORONTO

NC104 - MAN-BUMP nc104bmp.dat 09-22-92
NT104 - MAN-BUMP nt104bmp.dat 09-22-92

Bump 03



EnTech Control Engineering Inc. | ENTECH TORONTO

NC104 - MAN-BUMP
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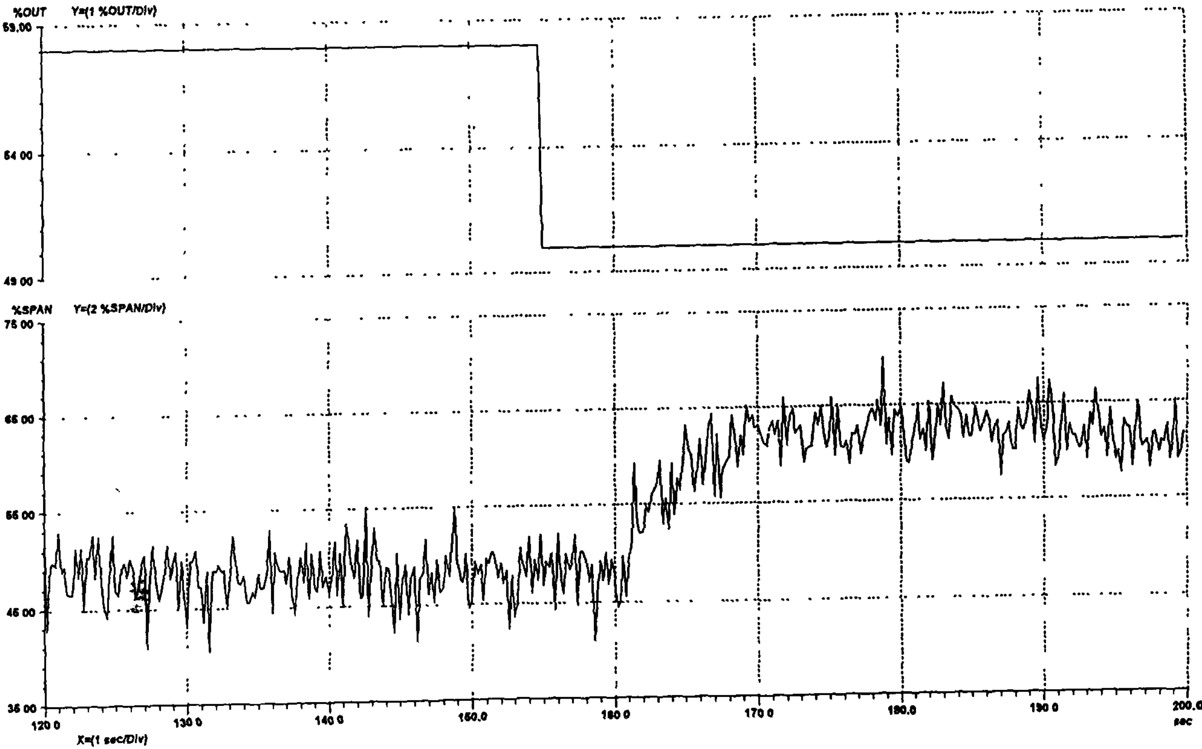

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3. NONLINEARITY IN PROCESS DYNAMICS

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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

<u>Example</u>

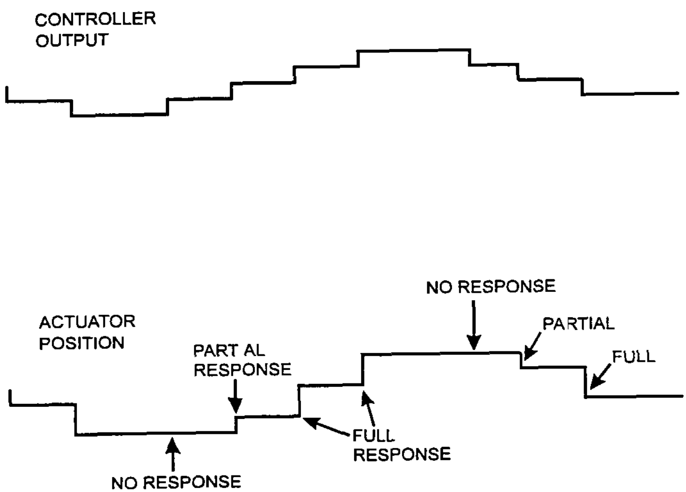


Fig 3 1

In this example the actuator did not move immed ately after the direction of motion was changed The second upwards step on y gave a part al 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 sim lar response on the downward steps is also due to back'ash

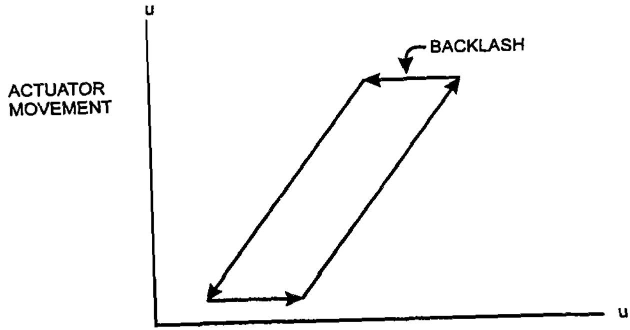


Fig. 3-2

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

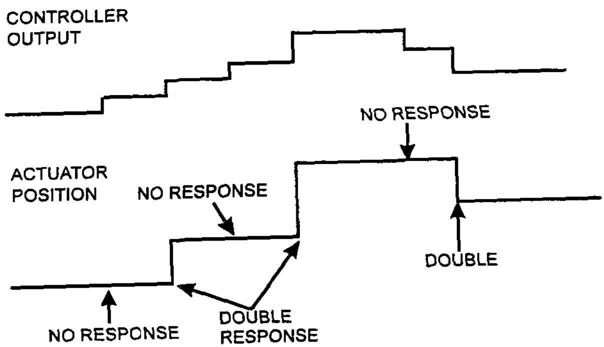


Fig 3-3

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 mot on

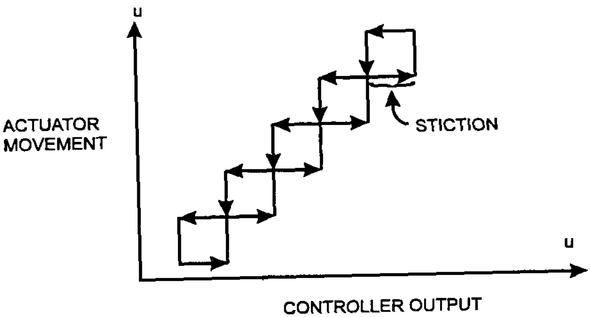


Fig 3-4

The EnTech specification for control valve performance allows a combined backlash and st ction of 1% as indicated in Figure 3 5 below

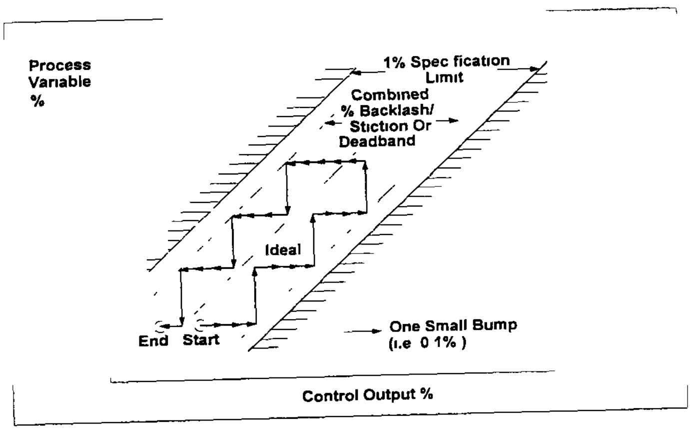


Figure 1

Figure 3-5 EnTech specification for Backlash and St ction

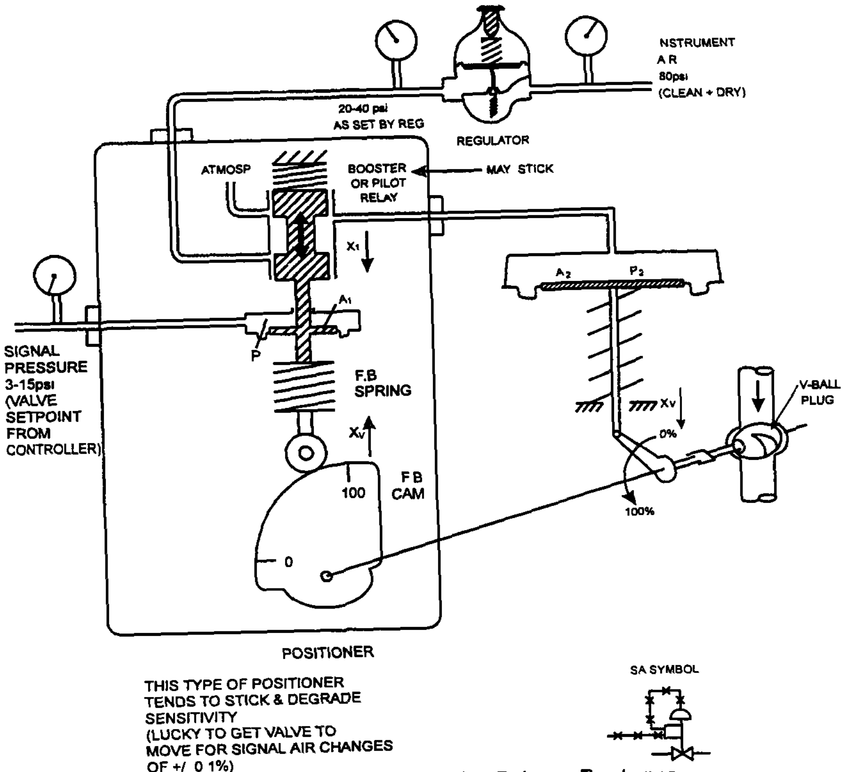


Figure 3-6 Spring Return Diaphraghm with Motion Balance Positioner

Comments on Backlash and Stiction

- These non-linearities may be non-linear with actuator position
- They may be time dependant
  (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 asymetric
  (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%

  $\Rightarrow \quad \Delta U = 1.0\% + 0.1\% = 1.1\%$

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 <u>600%</u> 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

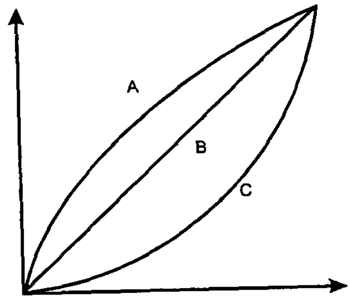


Fig 3-7

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 ΔP, the slopes give K_P

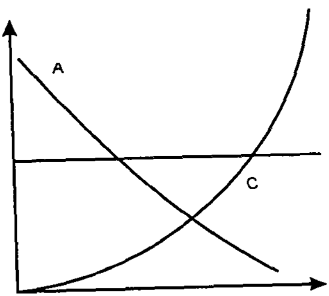


Fig 3-8

From this it would appear as though a linear valve is best. But ΔP usually changes as the valve opens or closes. We often find

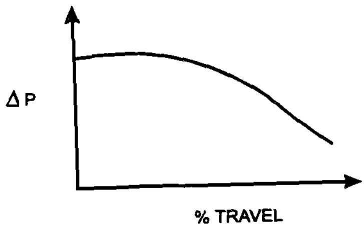


Fig 3-9

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 I 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%)

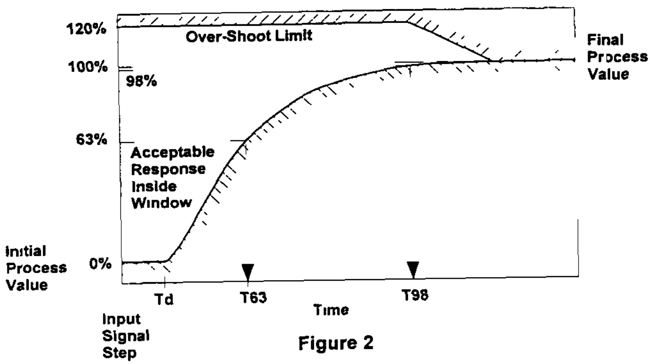


Figure 2

Figure 3-10 EnTech Specifcation for Speed of Response

Td = deadt me before any response is visible (seconds)
T63 = t me to reach 63% of a step change in input signal (seconds)
T98 = t me to reach 98% of a step change in input signal (seconds)
BW - bandwidth (Hz), the frequency at which the plug or trim position amplitude is attenuated by -6 dB Bandwidth has been inc uded in the specifcation for completeness, even though it is d'fficult to test for in most cases

Speed of Response

Table I

Valve Size (inches)	Valve Size (cm)	Td (sec)	T63 (sec)	T98 (sec)	BW (Hz)
0-2	0-5	0.1	0.3	0.7	1 6
>2-6	>5-15	0.2	0.6	1.4	0.8
>6-12	>15-30	0.4	1.2	2.8	0.4
>12-20	>30-50	0.6	1.8	4.2	0.27
>20+	>50+	0 8	2.4	5.6	0.2

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 re-sca ed appropriately.

Electric Actuators

High Torque Electric Actuators

High torque electric actuators have become ava lable for both rotary valves and damper drives as a replacement for pneumatic actuators These actuators a e 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) Wh le 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 w th the slew rate w ll determine the actua contro resolution Often this is wel bellow 1:1000 and may result in a nonlinear limit cycle.

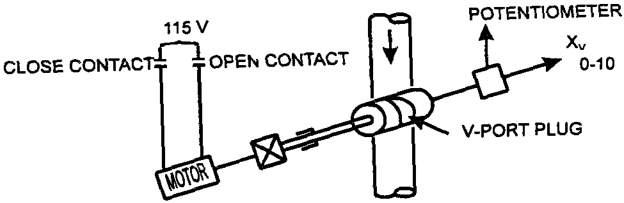
Another potentia problem is backlash, which typically occurs in the spline coupling between the gear-down output shaft and the valve plug For large valves with high torque requ rements this appears to be a b gger problem Back ash of as much as four seconds of actuation time has been reported

An associated problem, although not the fault of the valve s that in many basis weight control systems, the thick stock flow contro loop is be ng executed at only 5 seconds per sample This leads to prob ems with al asing and naccurate s ow contro

TYP CAL CONTROL PANEL

OPEN O O CLOSE 41 3 DIGITAL REPORT

COMP O O MANUAL



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

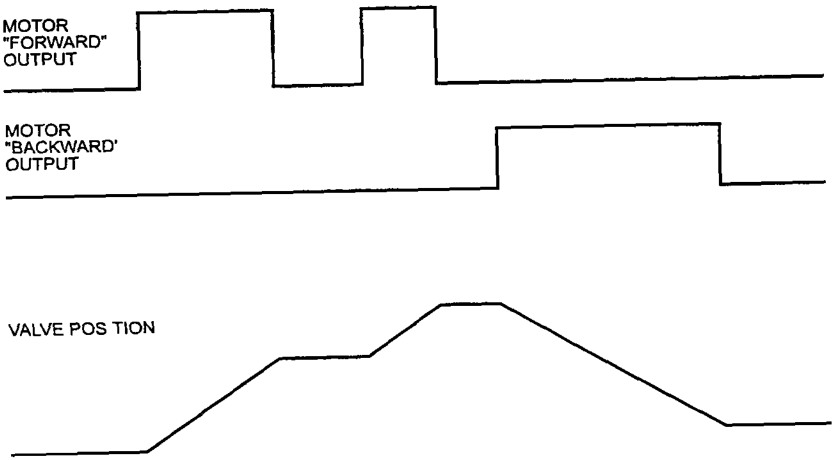


Fig 3-12

A specia form of st ction 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 pu se has no effect on the process output

In this instance, a deadband may be used to prevent unnecessary wear on the elements This deadband shou'd be app'ied 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 compensat on shou d be made equal to the minimum requ red actuation t me

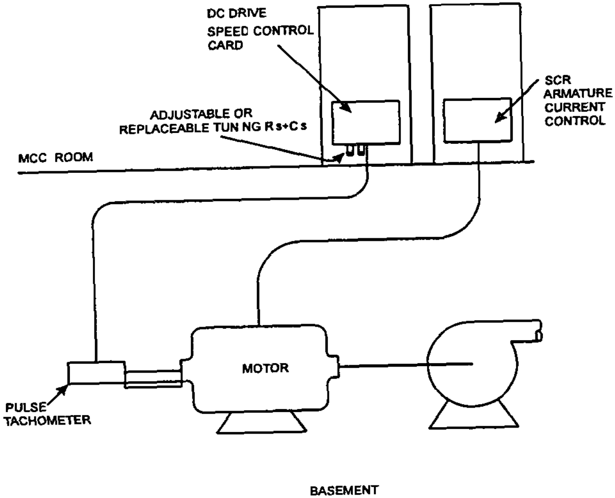
VARIABLE SPEED DRIVES FOR FANS AND PUMPS

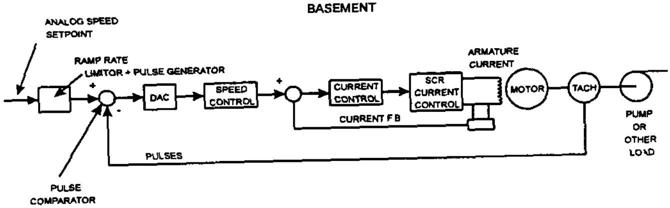
- o DC Drives - proven technology,
  - essentially linear dynamics,
  - time constant of 0.1 to 0.5 seconds.

- o 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'.

o 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 DR VE PEOP E (GE REL ANCE etc)
TYP CALLY TUNE AS FOLLOWS

CURRENT LOOP 30RAD/SEC eg $\lambda_{CURRENT} \approx$ 1/30 0 03SEC
SPEED LOOP 3RAD/SEC eg $\lambda_{SPEED} \approx$ 1/3 0 3SEC

NORMA LY OBJECT VE S TO GET A
SPEED STEP RESPONSE LIKE TH S

NOTE T IAT TH S S UNDERDAMPED
(ξ ~ 0 8) AND CAN BE THE CAUSE OF
OSC LLATORY BEHAV OUR

Figure 3-13 Variab e 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 cons der 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)

<u>Example</u> Variable Time Constant

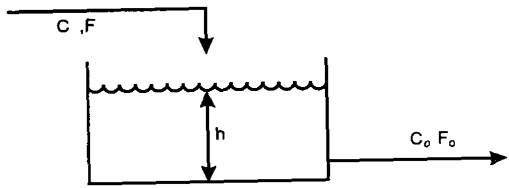


Fig 3-14

$$V \quad h * A_x \qquad (A_x \quad \text{cross sectional Area})$$

$$\tau \quad \frac{V}{F} \qquad (\text{assume } F_i \quad F_o)$$

The time constant wi'l vary with flow as wel as ıeve

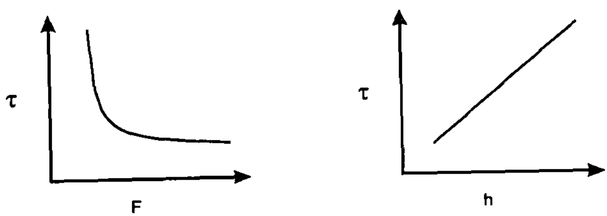


Fig 3-15

Example. Variable Gain with pH

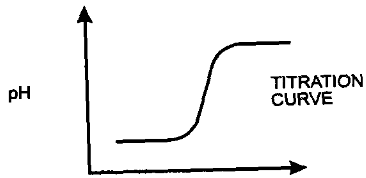


Fig 3-16

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

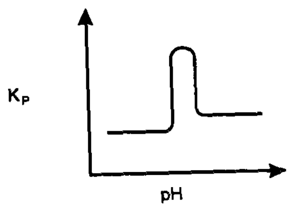


Fig 3-17

The process ga'n is also related to the flowrate of the 'controlled flow

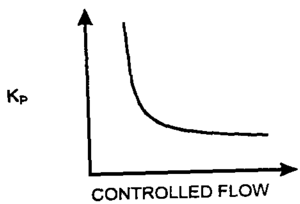


Fig 3-18

<u>Example</u> Variable 'Rate of Integration'

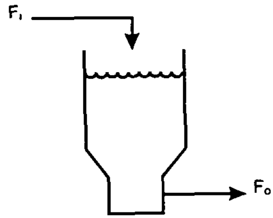


Fig 3-19

$$L = \int_0^T \frac{1}{A}(F_i - F_o)dt + L_o$$

Note $1/A$ cannot be taken outside since it may change with time

'Rate of Integration' = $1/A$

Fig 3-20

Example: Variable Time Delay

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

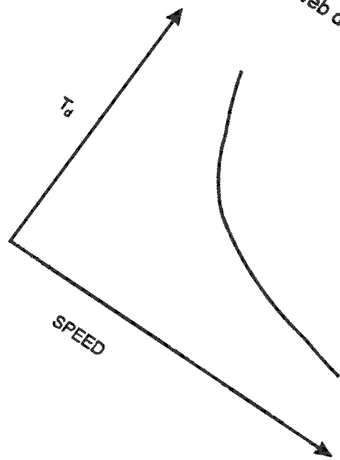
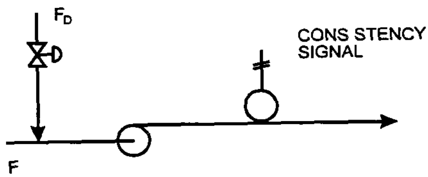
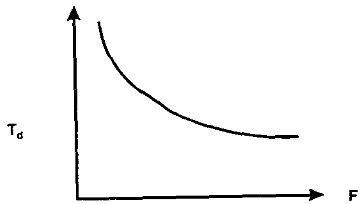


Fig.3-21

Example: Variable Gain, Time Delay, Time Constant

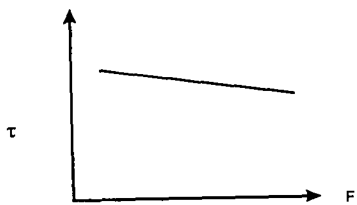
Consistency Control





Deadtime decreases with Stock flow

Process Gain decreases w th Stock Flow



Mixing time decreases slightly with Stock Flow

F gure 3-22 Non-Linearities in Consistency loop

<u>Example</u>: Variable Gain

Drying Rate on Paper Machine

Temperature differential is the driving force for drying

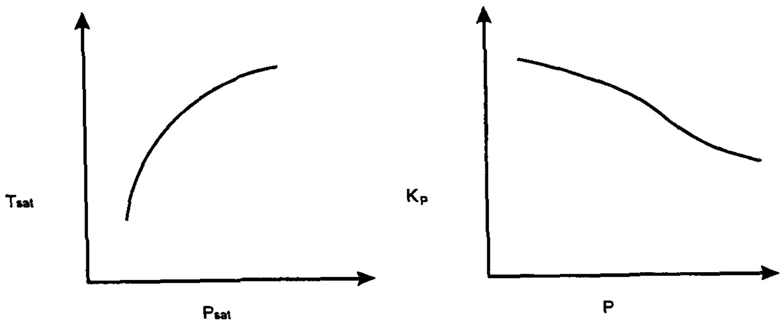


Fig 3-23 Non-Linear dynamics of Moisture Loop

Many other factors affect K_P

- machine speed
- 'rimming' characteristics
- press moisture

4. LAPLACE TRANSFORMS AND TRANSFER FUNCTIONS

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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

$$£(y(t)) = Y(s) = \int_0^\infty y(t)e^{-st}dt$$

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 al transforms of interest in process control are available.

Define the inverse Laplace Transform

$$£^{-1}(Y(s)) = y(t)$$

Common Laplace Transform Pairs

Time Function $f(t)$	*Laplace Transform* $F(s) - \pounds(f(t))$
$\delta(t)$ (unit impulse)	1
$u(t)$ (unit step)	$\frac{1}{s}$
e^{at}	$\frac{1}{s-a}$
t (unit ramp)	$\frac{1}{s^2}$
te^{at}	$\frac{1}{(s-a)^2}$
$e^{at}\frac{t^n}{n!}$	$\frac{1}{(s-a)^{n+1}}$
$\sin(\omega t)$	$\frac{\omega}{s^2+\omega^2}$
$\cos(\omega t)$	$\frac{s}{s^2+\omega^2}$
$e^{at}\sin(\omega t)$	$\frac{\omega}{(s-a)^2+\omega^2}$
$e^{at}\cos(\omega t)$	$\frac{s\quad a}{(s-a)^2+\omega^2}$

Table 4-1 Common Laplace Transforms

Properties of the Laplace Transform

1 Multiplication by a constant does not affect the transform.

$$\pounds(cy(t)) = c\pounds(y(t))$$
$$= cY(s)$$

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

$$\pounds\left(y_1(t) + y_2(t)\right) = \pounds\left(y_1(t)\right) + \pounds\left(y_2(t)\right)$$
$$= Y_1(s) + Y_2(s)$$

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

$$\pounds\left(\frac{dy}{dt}(t)\right) - s \cdot Y(s)$$

$$\pounds\left(\frac{d^2y(t)}{dt^2}\right) - s^2 \; Y(s)$$

etc.

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

$$\pounds\left(\int_0^t y(t)\,dt\right) = \frac{Y(s)}{s}$$

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

$$
\begin{aligned}
£\left(\tau\frac{dy(t)}{dt}+y(t)\right) &= £\left(Ku(t)\right)\\
£\left(\tau\frac{dy(t)}{dt}\right)+£\left(y(t)\right) &= K£\left(u(t)\right)\\
\tau£\left(\frac{dy(t)}{dt}\right)+Y(s) &= KU(s)\\
\tau sY(s)+Y(s) &= KU(s)\\
(\tau s+1)Y(s) &= KU(s)\\
Y(s) &= \frac{K}{\tau s+1}U(s)\\
Y(s) &= G(s)U(s)
\end{aligned}
$$

where

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

$G(s)$ is called the *transfer function* of the dynamic system

EnTech

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 prov des 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

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

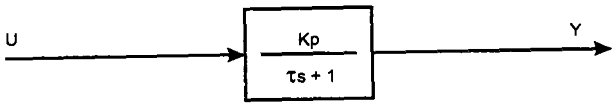
Note that the transfer function contains the same two parameters, K and τ, which characterize the dynamic system model. Also note that K can be characterized by

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

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*



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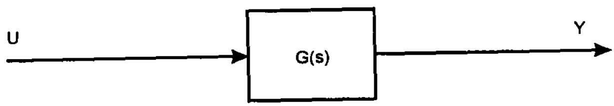
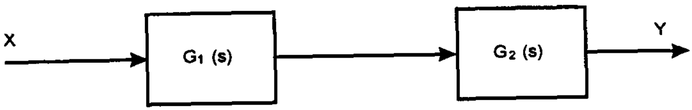


Fig 4-1 Block diagrams representing dynamics using transfer functions

Block Diagram Reduction

Rule 1: Blocks in series



IS EQUIVALENT TO

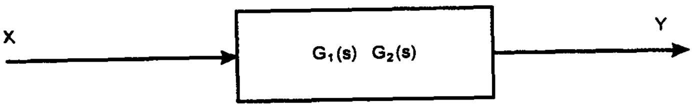
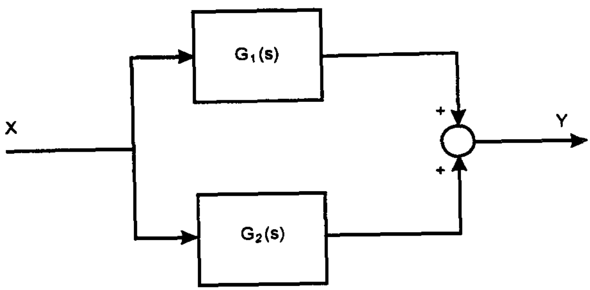


Fig. 4-2. Block diagram reduction of blocks in series.

Rule 2 Blocks in parallel



IS EQUIVALENT TO

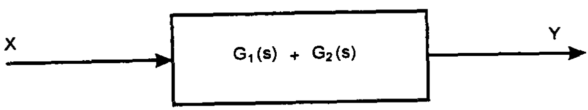


Fig. 4-3 Block diagram reduction of blocks in parallel.

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

Process Model	*Transfer Function*
1. Pure gain	$G_P(s) = K_P$
2. Pure delay	$G_P(s) = e^{T_d s}$
3 First order	$G_P(s) = \dfrac{K_P}{\tau s + 1}$
4 First order with deadtime	$G_P(s) = \dfrac{K_P e^{T_d s}}{\tau s + 1}$
5 Second order	$G_P(s) = \dfrac{K_P}{\tau^2 s^2 + 2\tau\zeta s + 1}$
6 Integrating	$G_P(s) = \dfrac{K_P}{s}$

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.

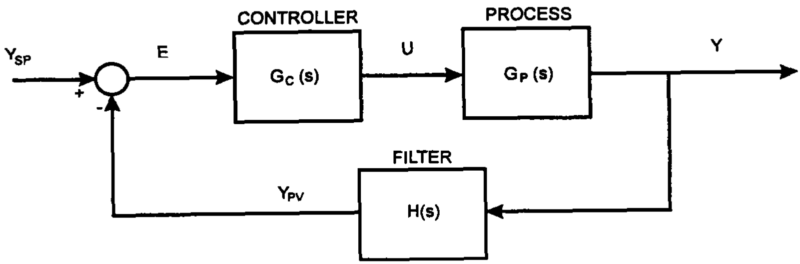


Fig. 4-4 Block Diagram of a control loop.

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 τ_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 wil be described by the following transfer function.

$$G_C(s) = K_P + K_I/s = \frac{K_P s + K_I}{s}$$

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 β (lead time constant) and τ (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

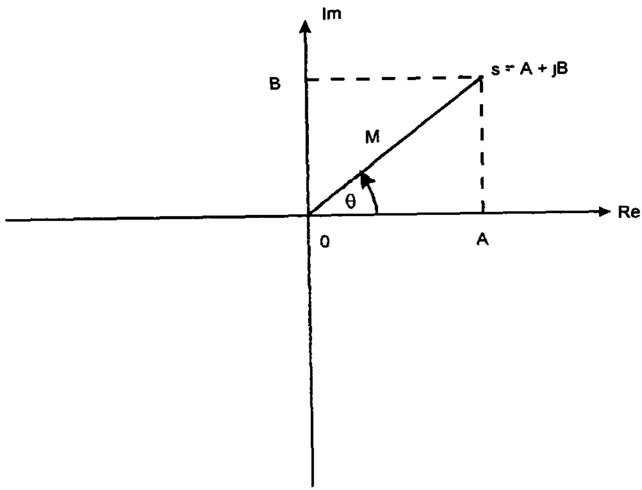


Fig 4-5 Representing a complex number as a point in the complex plane

The *polar representation* of s is given by s $Me^{j\theta}$ M s is called the *absolute value* s and θ 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 x s and the transfer function ~~poles~~ zeros represented by o s

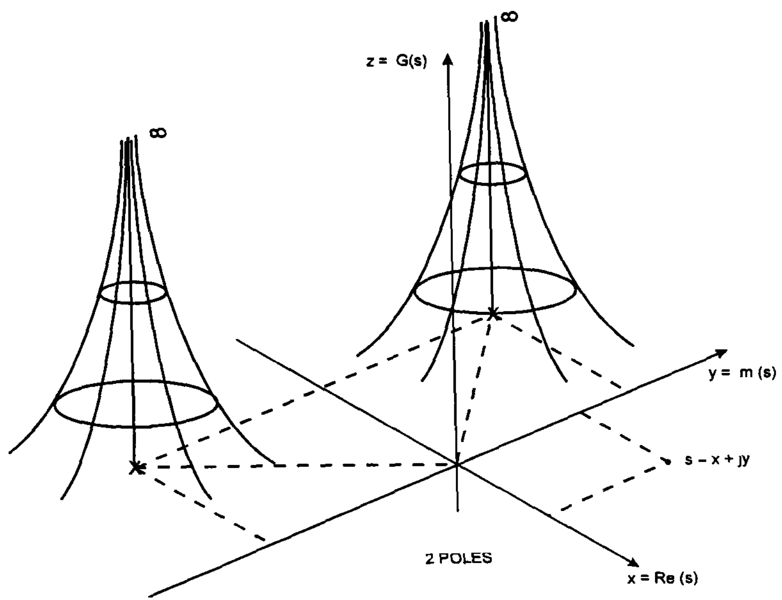


Fig 4-6 Geometric interpretation of transfer function poles and zeros

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 $s = \ 1/\tau$.

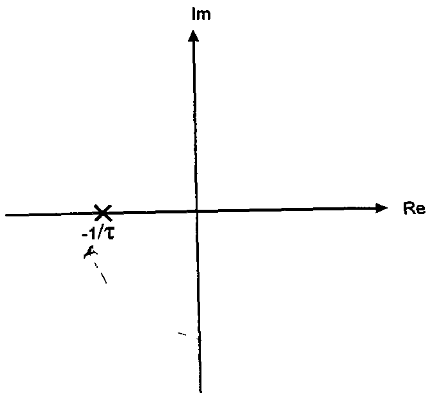


Fig 4-7 Pole-Zero Plot of a first order system

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*.

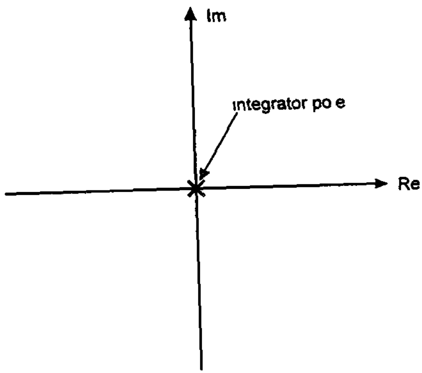


Fig 4-8 Pole-Zero Plot of an integrator

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, τ_1 and τ_2 are all positive parameters

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

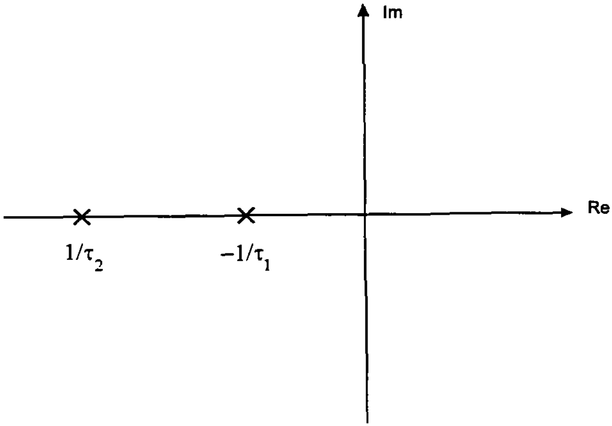


Fig 4-9 Pole-Zero Plot of an second order overdamped system

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 w th 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.

$$G(s) = \frac{K}{\tau^2 s^2 + 2\tau\zeta s + 1} \qquad \zeta < 1$$

where K, τ and ζ 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:

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

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

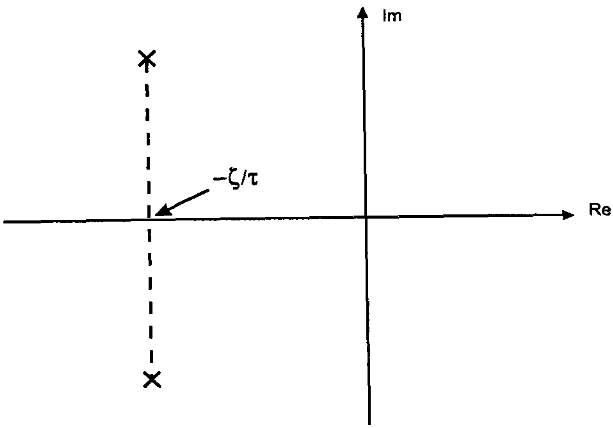


Fig 4-10 Pole-Zero Plot of an second order underdamped system

Recall that the step response will overshoot the final value and eventually approach the fina 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

<u>Example</u>

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

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

<u>Case 1</u> $\tau > 0$

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

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

Bounded output Stable

<u>Case 2</u> $\tau < 0$

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

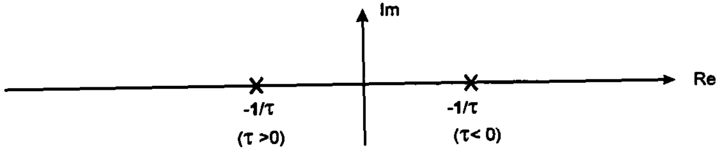


Fig 4-11 Pole locations for a stable system and unstable system.

$\tau > 0$ 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

<u>Example</u> First Order System

Transfer function	$G_P(s) = \dfrac{K_P}{\tau s + 1}$
Characteristic Equat on	$0 = \tau s + 1$
Root of Characteristic Equation	$s = \quad 1/\tau$

<u>Problem</u> The roots of the characteristic equation are not easily determined when the order of the equation is greater than 2

<u>Example</u> $As^3 + Bs^2 + Cs + D = 0$

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

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

where K is the gain of the transfer function and τ is the time constant There is a single pole at $s = -1/\tau - a$

A pole on the real axis is associated with a first order response. A pole on the negative real axis at $s = a$ is associated with a first order response with time constant $\tau = 1/a$ Step responses for various real pole positions are shown in the following figure

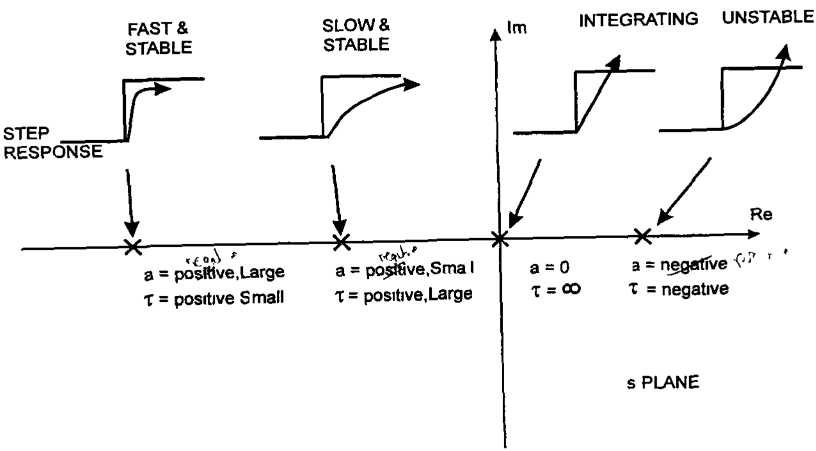


Fig 4-12 Step responses for various pole positions on the real axis

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 $s_1 = j\omega$ and $s_2 = -j\omega$. 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 ω (cyclic frequency $f = \omega/2\pi$ and period $T = 2\pi/\omega$) This is illustrated below

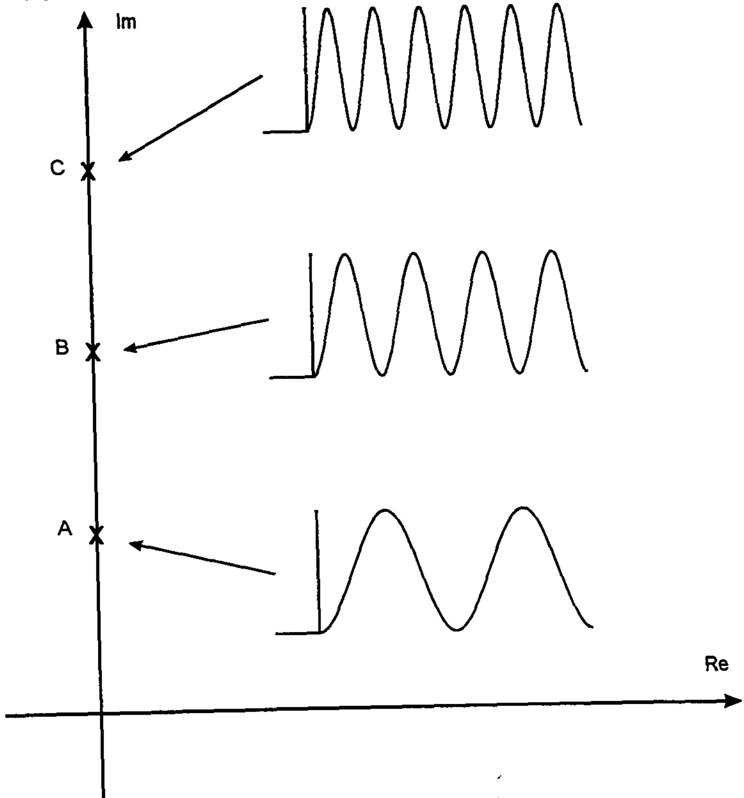


Fig. 4-13 Step responses for various pole positions on the imaginary axis.

Pole Position Interpretation of Complex Conjugate Pole Pairs

Consider the second order underdamped transfer function

$$G(s) = \frac{K_P}{\tau^2 s^2 + 2\tau\zeta s + 1} = \frac{K_P \omega_n^2}{s^2 + 2\zeta\omega_n s + \omega_n^2}$$

where τ is the second order time constant, $\omega_n = 1/\tau$ is the undamped natural frequency and ζ is the damping coefficient.

There are two complex conjugate poles located at:

$$s_{1,2} = -\omega_n\zeta \pm j\omega_n\sqrt{1-\zeta^2} = \sigma \pm j\omega$$

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 ω (imaginary part of the pole). Thus the cyclic frequency will be $f = \omega/2\pi$ 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 σ is the real part of the pole).

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

This is illustrated in the following figures.

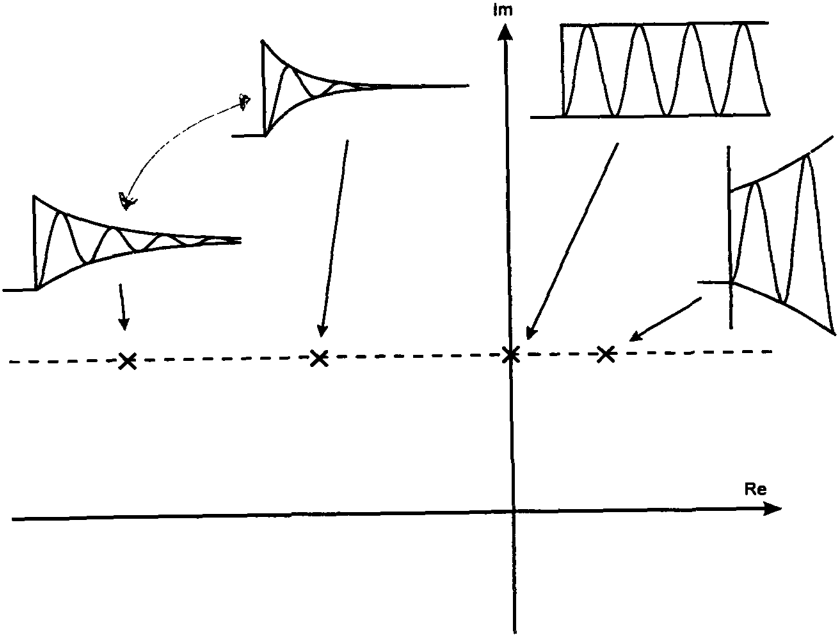


Fig. 4-14. Effect on step response of varying the real part of complex conjugate poles.

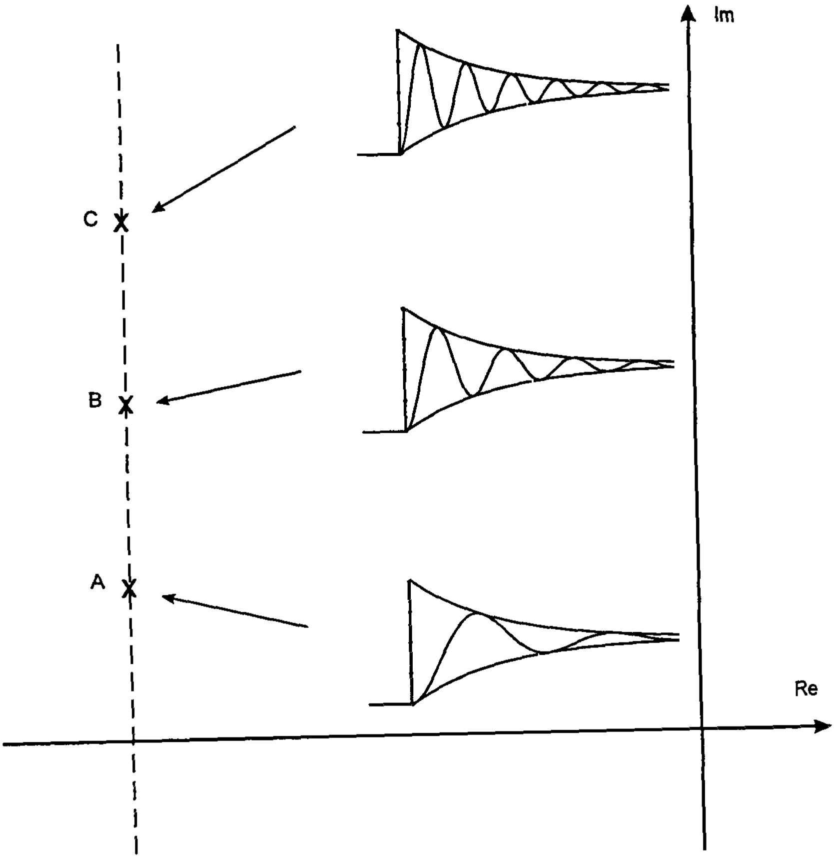


Fig 4-15 Effect on step response of varying the imaginary part of complex conjugate poles

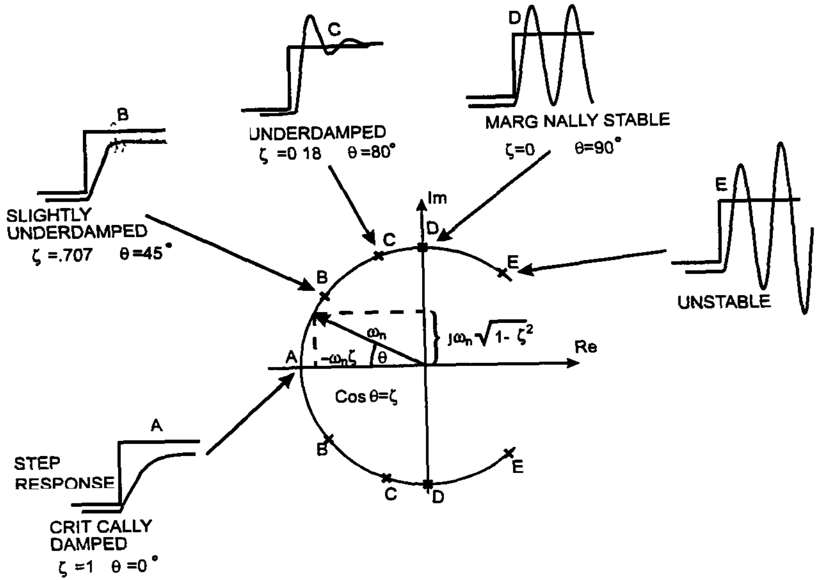


Fig 4-16. Effect on step response of varying the angle θ of complex conjugate poles

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

<u>Example</u>

<u>System 1</u>

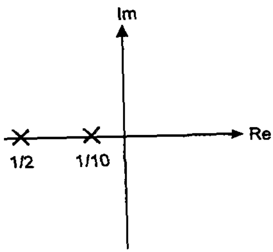
(Process model)

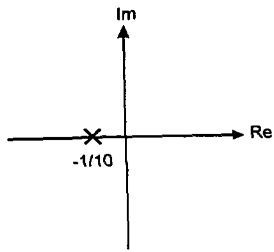
$$G_1(s) = \frac{K_p}{(10s+1)(2s+1)}$$

<u>System 2</u>

(Simplified model determined by dominant pole)

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





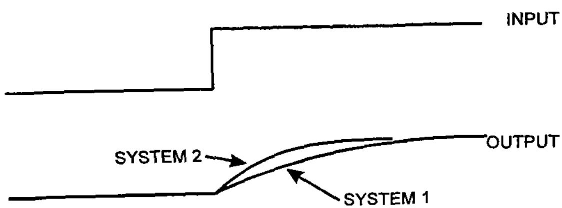


Fig 4-17 Step responses of a second order overdamped system and its dominant dynamics

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

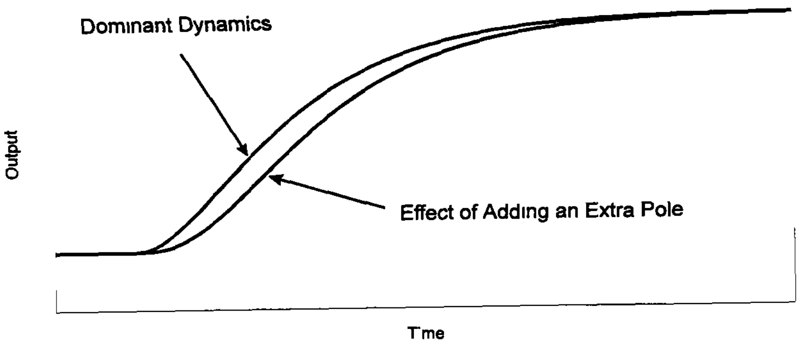


Fig 4-18 The effect of an additional real pole on the step response.

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

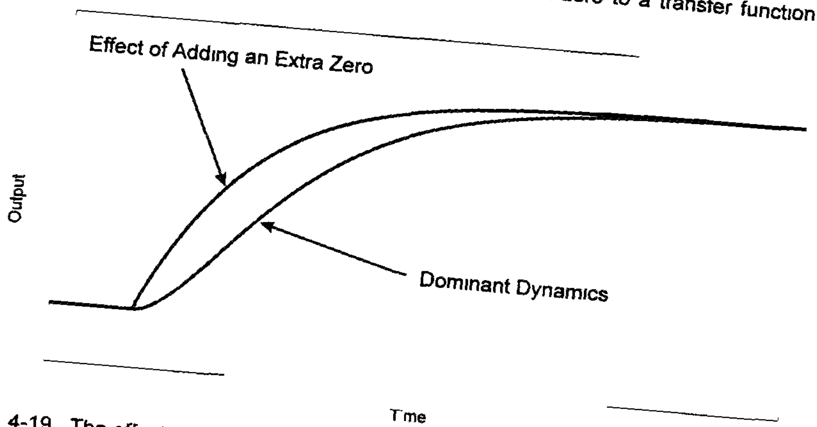


Fig 4-19 The effect of an additional real zero on the step response

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 β is the lead time constant and τ is the lag time constant Thus there is one pole at $s = -1/\tau$ and one zero at $s = -1/\beta$. 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 β/τ and a first order system with a gain of $1 \;\; \beta/\tau$ as shown below.

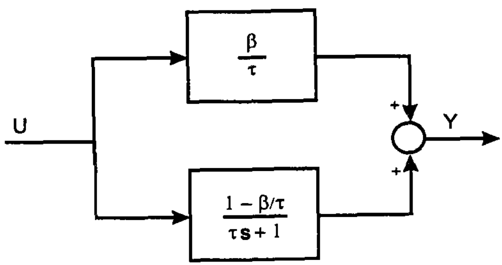


Fig 4-20 Representation of a lead-lag as a parallel combination.

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) wil be determined by the first order part namely the lag time constant τ The effect of the system zero will be to speed up the initial part of the response, the larger the ratio β/τ, the larger will be the effect on the speeding up effect on the initial part of the response

Step Response of the Lead-Lag Transfer Function

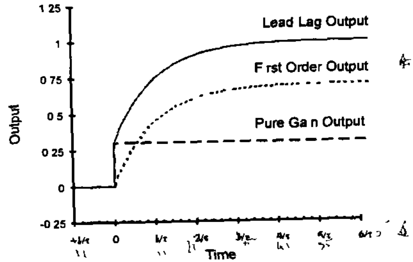


Fig 4-21. Step response of a lead-lag transfer function with $\beta < \tau$

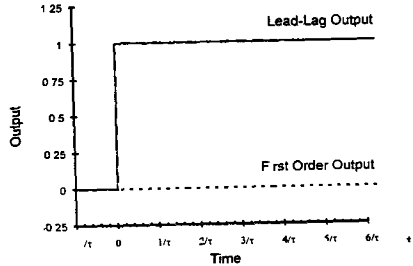


Fig 4-22 Step response of a lead-lag transfer function with β τ

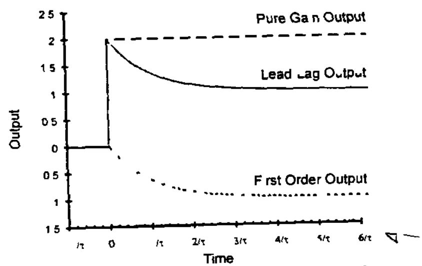


Fig 4-23 Step response of a ead lag transfer function with $\beta > \tau$

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.

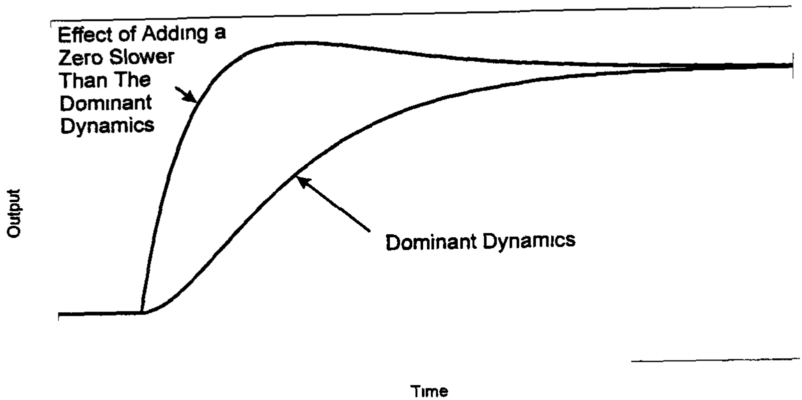


Fig. 4-24. Overshoot caused by a zero slower than the dominant pole(s)

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.

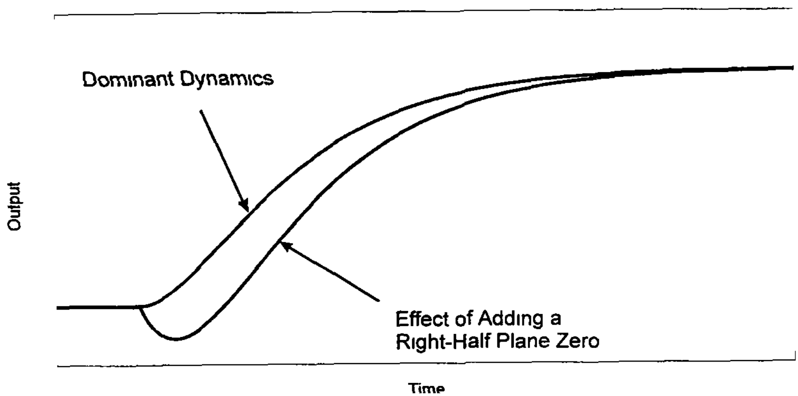


Fig 4-25 Wrong way response caused by a right half plane zero

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

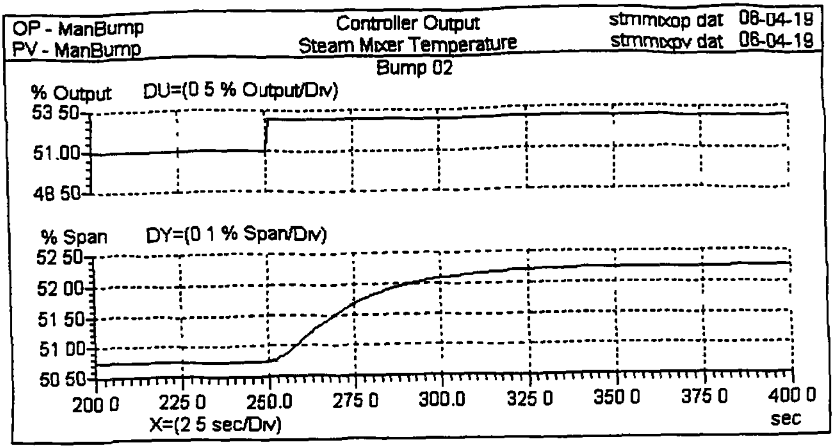


Fig. 4-26. Sample step response.

Dominant Dynamics Transfer Function: $G(s) = \frac{}{\tau s +}$

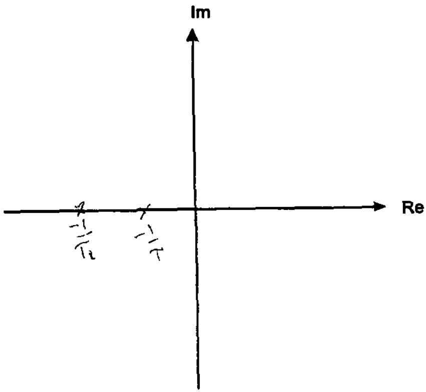


Fig. 4-27. Pole Zero Plot for sample dynamics

Model Transfer Function: $G(s) = \frac{k}{\tau s +} \; \frac{1}{\tau_2 s +}$

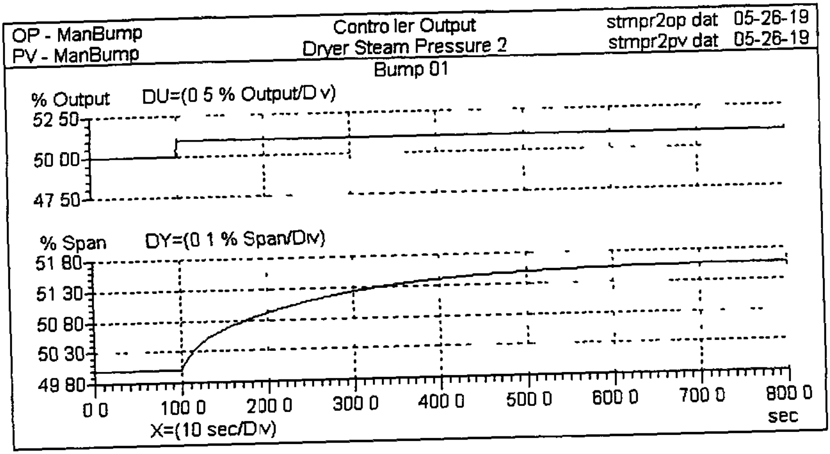


Fig 4-28 Dryer Steam Pressure step response

Dominant Dynamics Transfer Function $G(s) - \frac{k}{\tau s + 1}$

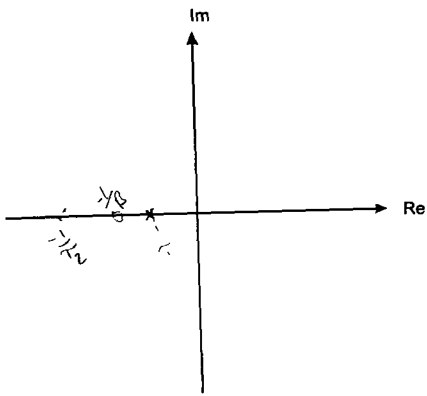


Fig 4 29 Pole Zero Plot for Dryer Steam Pressure dynamics

Model Transfer Function $G(s) - \frac{k((\ +1)}{(\tau s+1)(\tau_2 s+1)}$

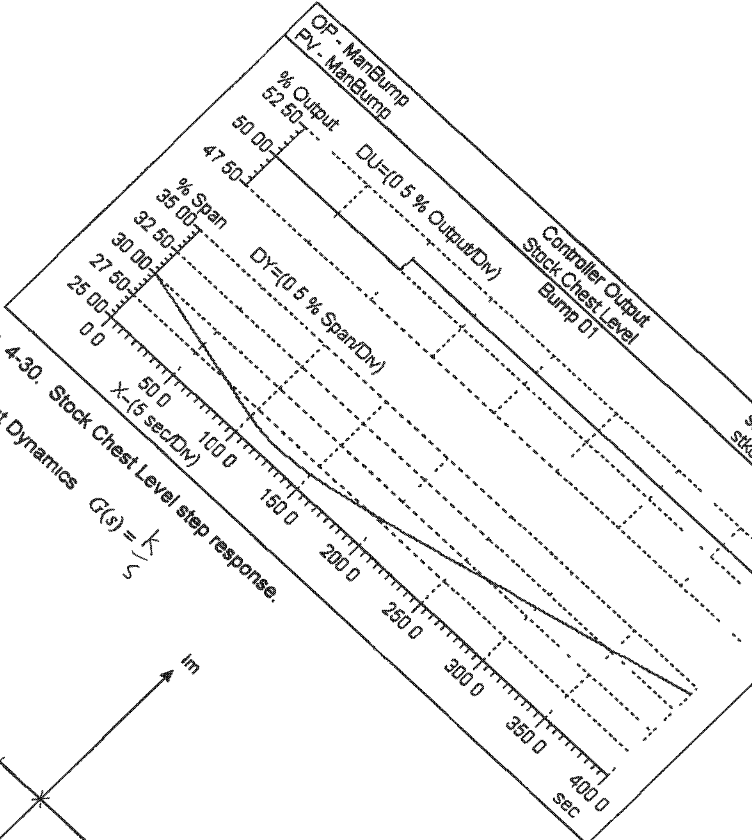
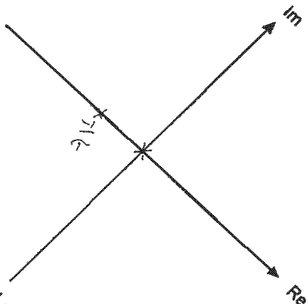


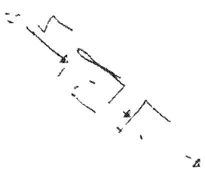
Fig 4-30. Stock Chest Level step response.

Dominant Dynamics $G(s) = \frac{K}{s(}$



Pole-Zero Plot for Stock Chest Level dynamics.

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



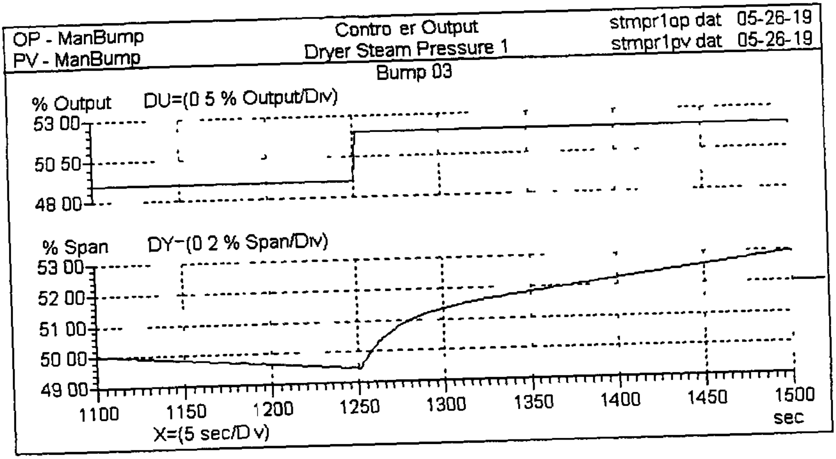


Fig 4-32 Dryer Steam Pressure step response

Dominant Dynamics Transfer Function: $G(s) =$ (Because of dominant tau f)

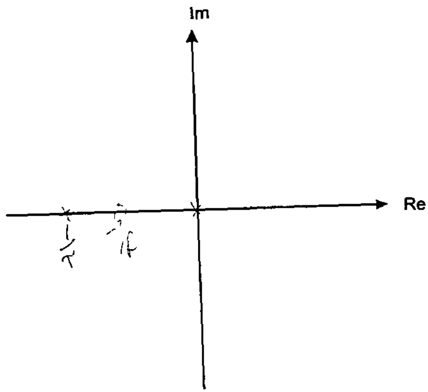


Fig 4-33 Pole-Zero Plot for Dryer Steam Pressure dynamics

Model Transfer Function $G(s) =$

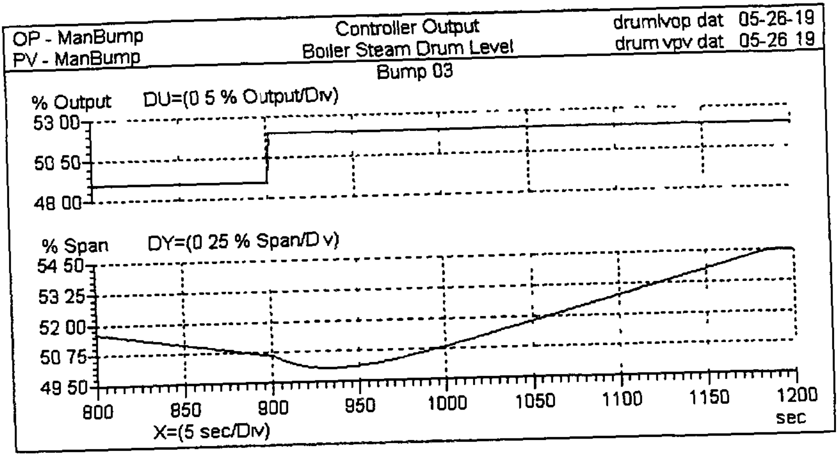


Fig 4-34 Boiler Steam Drum Level step response.

Dominant Dynamics Transfer Function $G(s) = \frac{k}{s}($ "dc gain ramp)

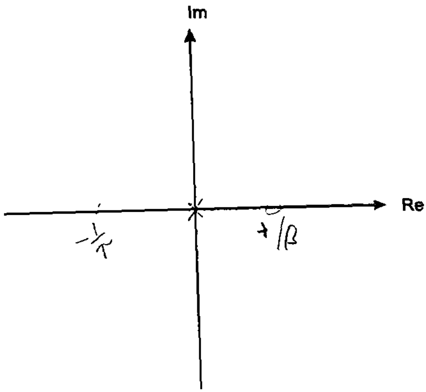


Fig 4-35 Pole-Zero Plot for Boiler Steam Drum Level dynamics

Model Transfer Function: $G(s) = \frac{k}{s} \cdot (1-\beta s) \cdot \left(\frac{1}{\tau s+1}\right)$

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5. FREQUENCY RESPONSE

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Sinusoidal Inputs

The frequency response of a system (eg. 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.

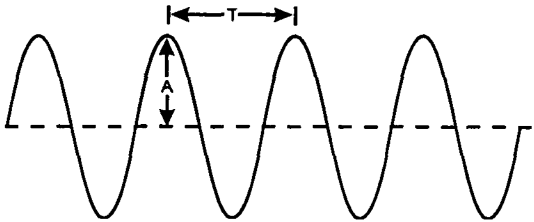


Fig 5-1

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

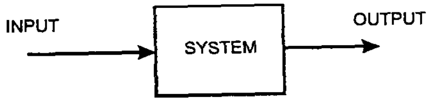
(i) the cyclic frequency

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

(ii) or, the angular frequency

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

Frequency Response



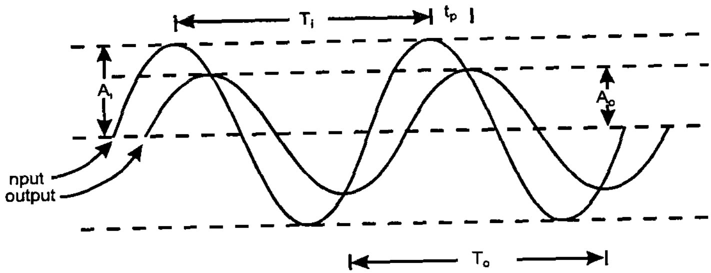


Fig 5-2

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

$$T_o = T_i$$

$$AR - \frac{A_o}{A_i}$$

$$\phi \quad \frac{t_P}{T_i} \text{ x } 360 \quad \text{(degrees)}$$

AR is called the <u>amplitude ratio</u> <u>(gain)</u>
ϕ is called the <u>phase shift</u>

Example 1 Chest consistency from inlet to outlet

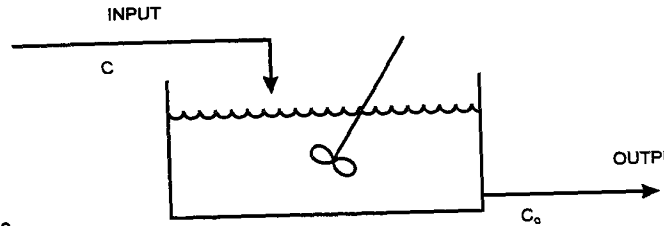


Fig 5-3

Step Response

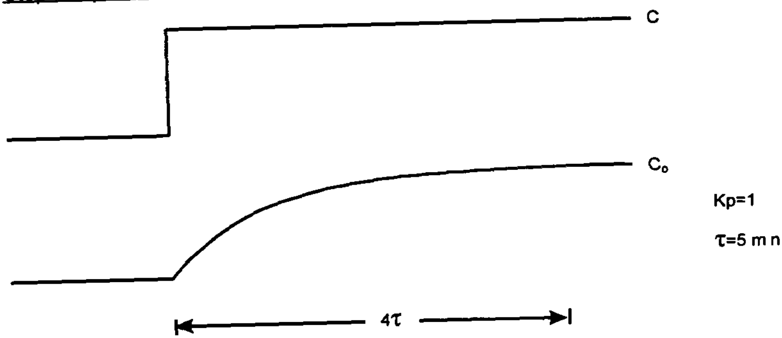


Fig 5-4

Frequency Response

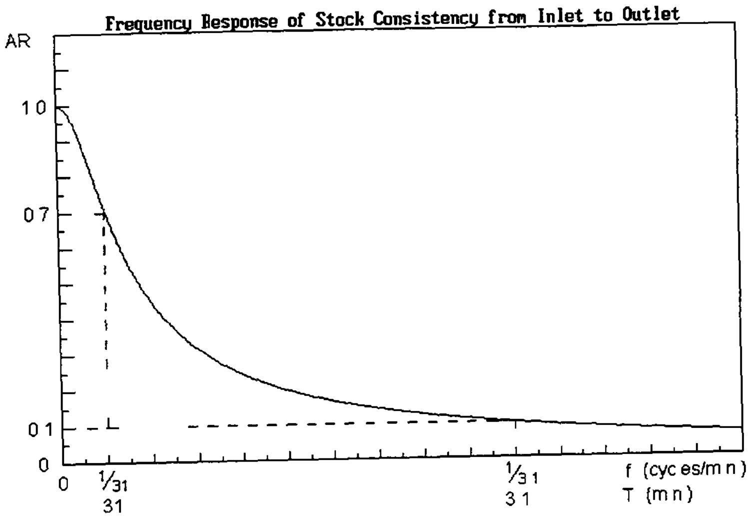


Fig 5-5

Example 2 F rst order filter, with filter time constant τ_f

Step response

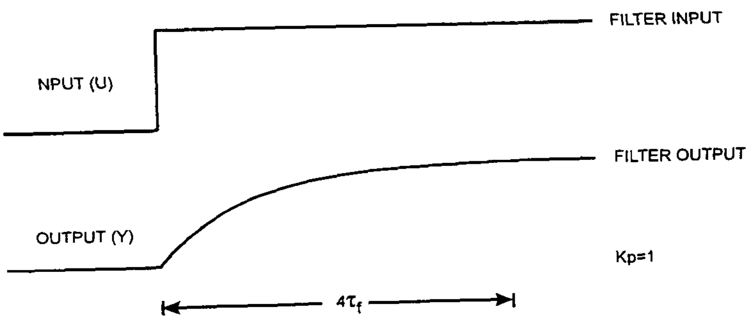


Fig 5-6

Frequency response

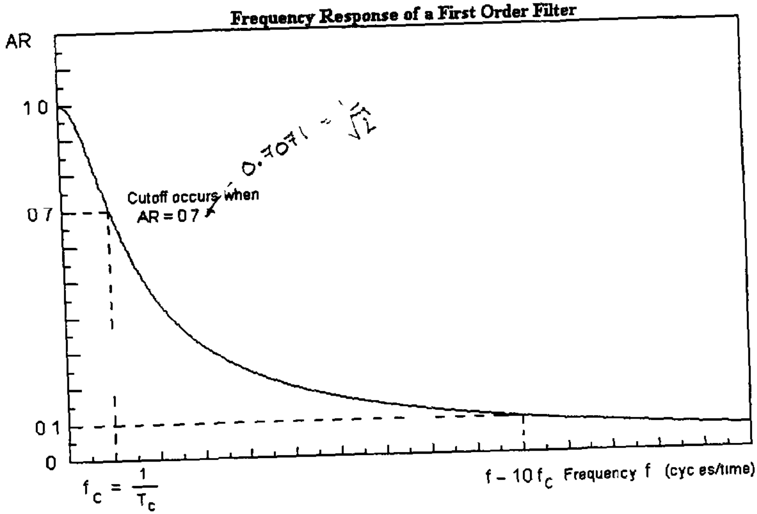


Fig 5-7

$$T_c = 2\pi\tau_f$$

$$f_c = \frac{1}{2\pi\tau_f}$$

$$\omega_c = \frac{1}{\tau_f}$$

First Order System

K_P - process gain
τ - process time constant

The amplitude ratio is given by.

$$AR = \frac{K_P}{\sqrt{\omega^2\tau^2 + 1}}$$

where ω is the input angular frequency

The cutoff frequency ω_c is

$$\omega_c = \frac{1}{\tau} \qquad \text{(rad/time unit)}$$

At the cut-off frequency

$$AR = \frac{K_P}{\sqrt{2}} = 0.707 K_P$$

Using cyclic frequency $f - \dfrac{\omega}{2\pi}$

$$AR - \frac{K_P}{\sqrt{\left(\frac{f}{f_c}\right)^2 + 1}}$$

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 $T \rangle\rangle T_c$ (low frequency)

$$AR \approx \frac{K_P}{\sqrt{1}} = K_P$$

For $T \langle\langle T_c$ (high frequency)

$$AR \approx \frac{K_P}{\sqrt{\left(\frac{T_c}{T}\right)^2}} = \left(\frac{T}{T_c}\right) K_P$$

<u>Example</u> 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

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

(a) when input period is 25 min

$$AR = \frac{K_P}{\sqrt{\left(\frac{T_c}{T}\right)^2 + 1}} = \frac{1}{\sqrt{\left(\frac{31.4}{25}\right)^2 + 1}} \approx 0\ 62$$

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

$$AR \quad \frac{K_P}{\sqrt{\left(\frac{T_c}{T}\right)^2 + 1}} - \frac{1}{\sqrt{\left(\frac{31\ 4}{5}\right)^2 + 1}} \approx 0\ 16$$

Note

$$AR \approx \left(\frac{T}{T_c}\right) K_P \quad \frac{5}{31\ 4} - 0.16$$

T (min) input period	AR amplitude ratio	
314	100%	slower than the cutoff
31 4	70%	at the cutoff
3 14	10%	10 times faster than the cutoff
0 314	1%	100 times faster than the cutoff

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

AR	AR (dB)
1	0
10	+20
0.1	-20
0 707	-3
2	+6
0.5	-6

Example Bode Plot of a first order system with process gain K_P and process time constant τ

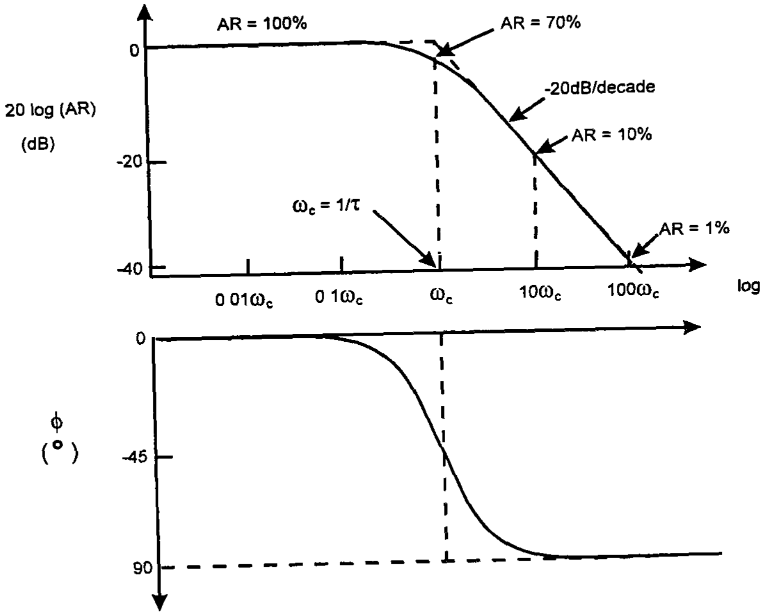


Fig.5-8

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

- the AR asymptotes intersect
- $AR = 0.707$
- $20 \log(AR) = -3\text{dB}$
- $\phi = -45^o$

The maximum phase lag is 90^o

Example. Second order system.

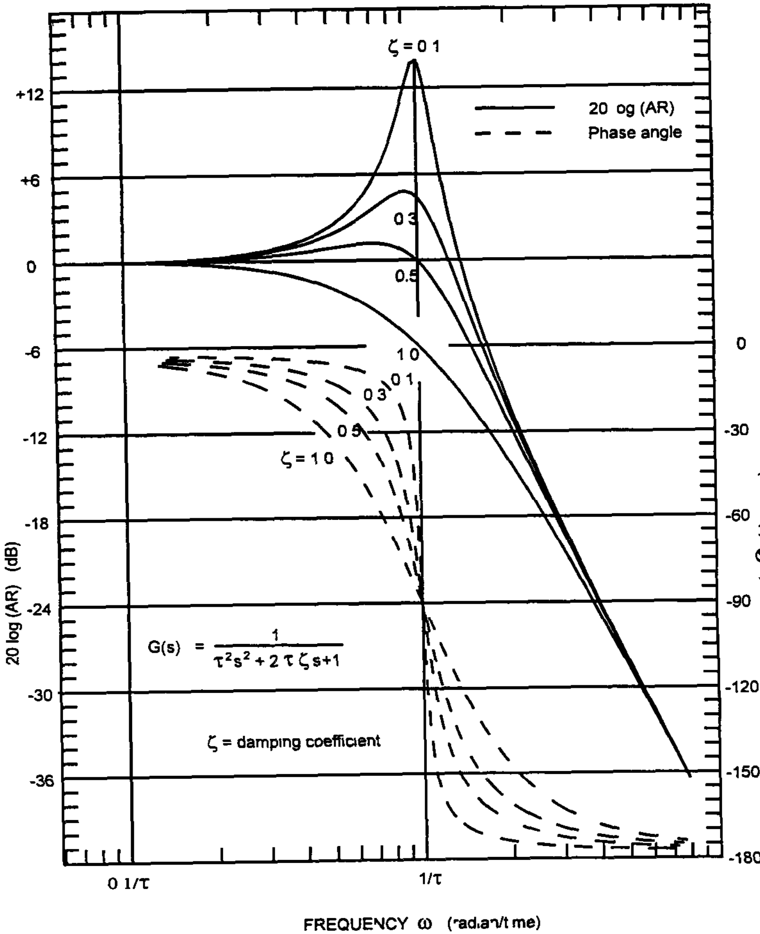


Fig.5-9

A second order underdamped system may exhibit a resonant peak This will occur for $\varsigma < 0.707$. 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)

<u>Example</u> PI Contro ler

6. PID CONTROLLERS

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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

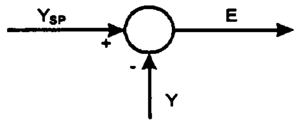


Fig 6-1

where: Y is the process measurement

Y_{SP} is the setpointor 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.

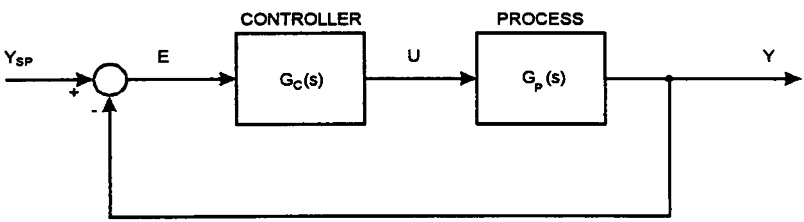


Fig 6-2

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

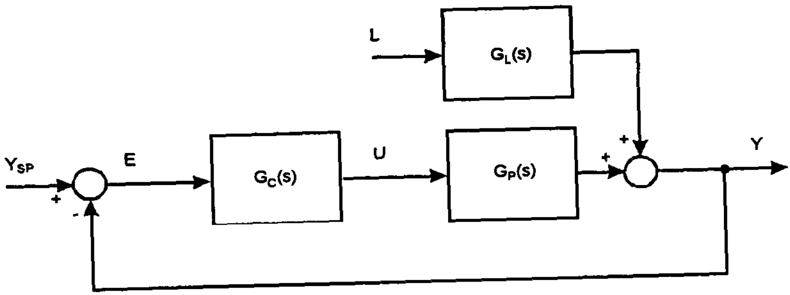


Fig 6-3

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

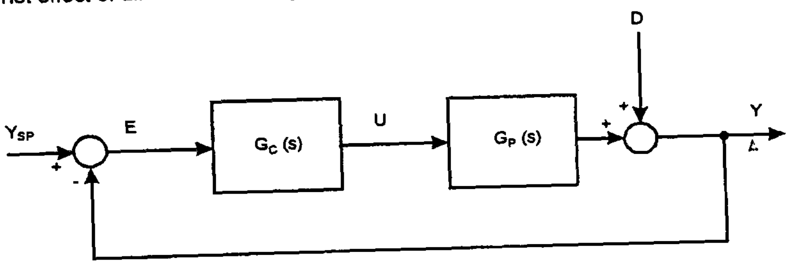


Fig 6-4

$Y - D = Y(G_C \times G_P)$

$\frac{Y}{D} =$

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

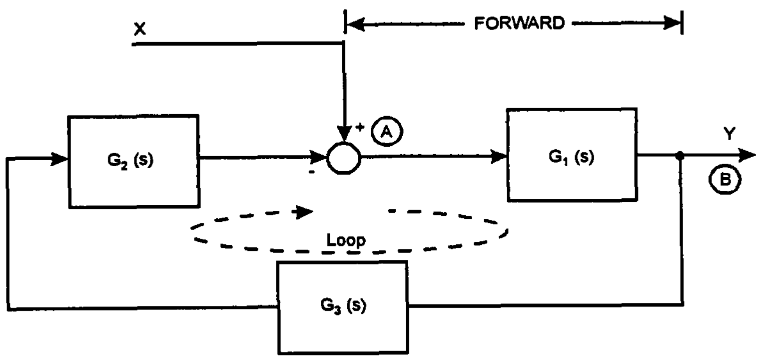
$$G_{LD}(s) = \frac{1}{1 + G_C(s)G_P(s)}$$

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:

$$G_{\text{CLOSED LOOP}_{AB}} = \frac{G_{AB}(\text{OPEN LOOP})}{1 + G_{LOOP}} = \frac{G(\text{FORWARD})}{1 + G_{LOOP}}$$



is equivalent to

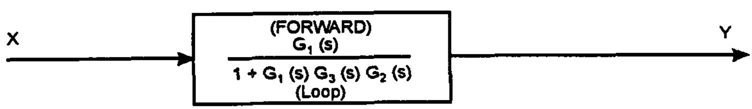


Fig. 6-5

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:

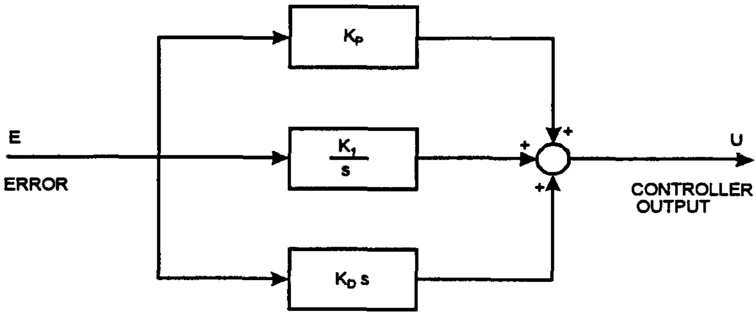


Fig. 6-6

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

The controller transfer function is:

$$G_C(s) = K_P + K_I \tfrac{1}{s} + K_D s$$
$$= \frac{K_D s^2 + K_P s + K_I}{s}$$

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:

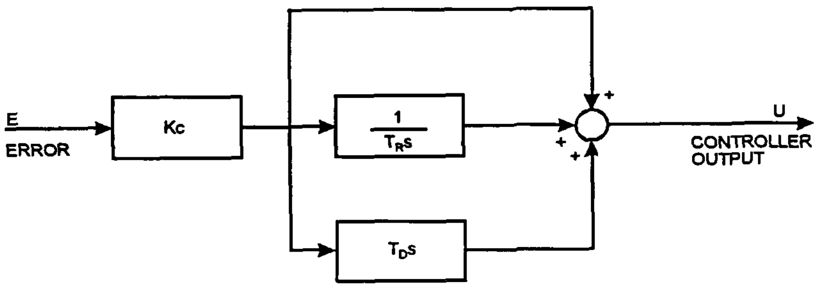


Fig 6-7

$$u = K_C\left(e + \frac{1}{T_R}\int e\,dt + T_D\frac{de}{dt}\right)$$

The controller transfer function is

$$G_C(s) = K_C\left(1 + \frac{1}{T_R s} + T_D s\right)$$
$$= \frac{K_C}{T_R}\left[\frac{T_R T_D s^2 + T_R s + 1}{s}\right]$$

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 equipment and in seconds in European and Japanese equipment

Classical Form of the PID Controller (Ser...

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

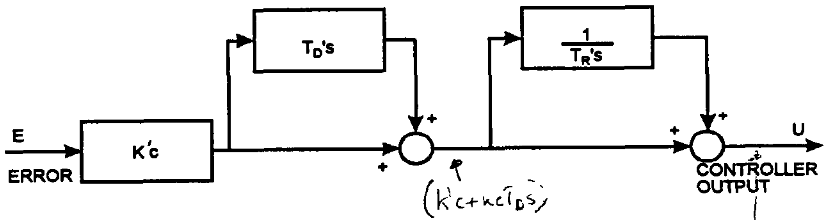


Fig. 6-8

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

The controller transfer function is.

K. change

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

The tuning parameters are

- K_C' called the <u>controller gain</u>,
- T_R' called the <u>reset time</u> and
- T_D' called the <u>derivative time</u>

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, <u>they are not identical</u>

PROPORTIONAL

The proportional term will move the output PROPORTIONAL to the error

This component is equal to the following,

P = Kc * error.

where Kc = proportional gain, normally expressed in units of $\frac{\%OUTPUT}{\%SPAN}$

error = (Setpoint - Process), normally expressed in units of % Span

Consider the setpoint bump shown below

Assume

Kc is equal to 0.5 $\frac{\%Output}{\%Span}$ 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 .

$$P = (Kc * error) = 0.5\ \frac{\%Output}{\%Span} * 2\%Span = \quad \%Output$$

This control action taken at the setpoint bump or in response to a disturbance is often referred to as the proportional k ck

PROPORTIONAL ONLY CONTROL

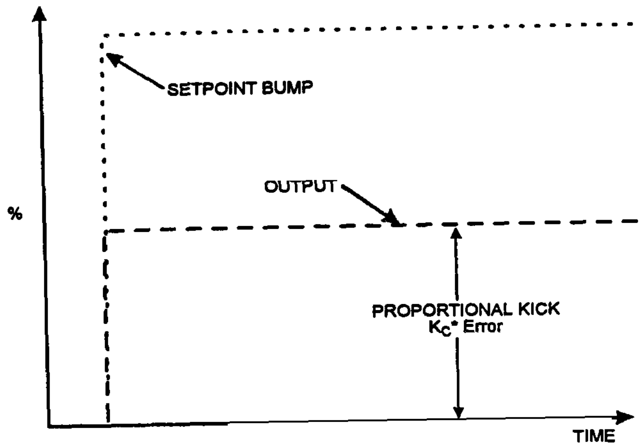


Figure 6-9 Proportional Only - Error is Constant

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 setpo nt 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

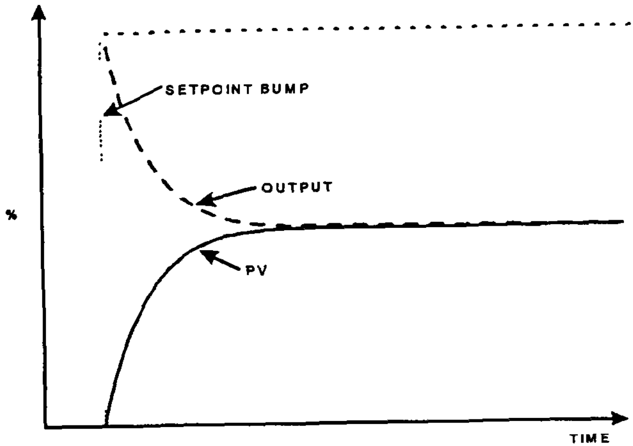


Figure 6-10 Proportional Only - Offset from Setpoint

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

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

INTEGRAL

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.

$$I = \frac{Kc}{T_R} * \int_0^t error(t)dt$$

where T_R = Reset Time, normally expressed in minutes,

and, $\int_0^t error(t)dt$ = 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 BUM
NO PROCESS CHANGE

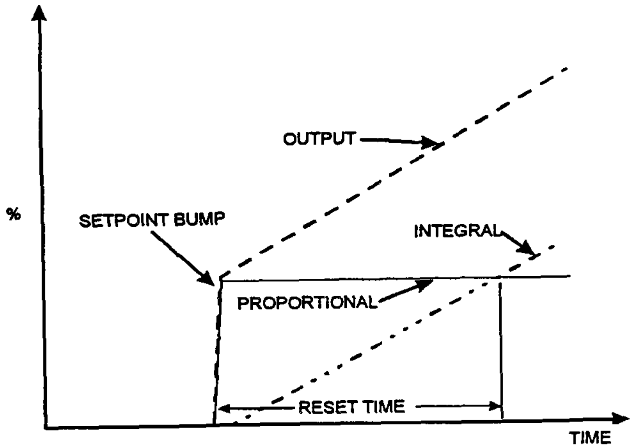


Figure 6-11 PI Controller Response to Setpoint Bump - Error is Fixed

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

PROPORTIONAL ONLY CONTROL

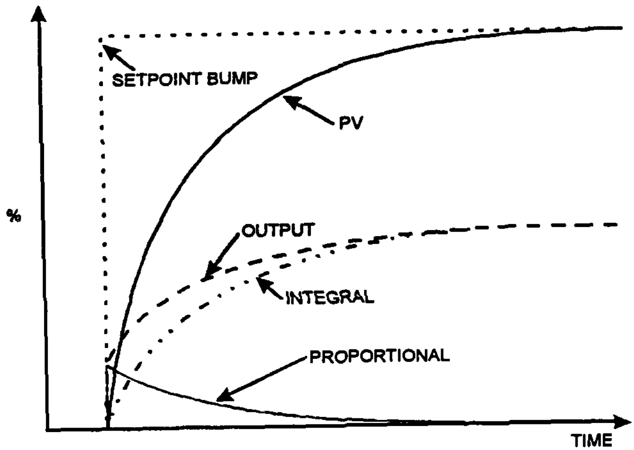


Figure 6-12 PI Control er Response to Setpoint Bump - Error not Fixed

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

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

DERIVATIVE

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

$$D = Kc * T_D * \frac{de}{dt}$$

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.

PI CONTROLLER
NO PROCESS FEEDBACK

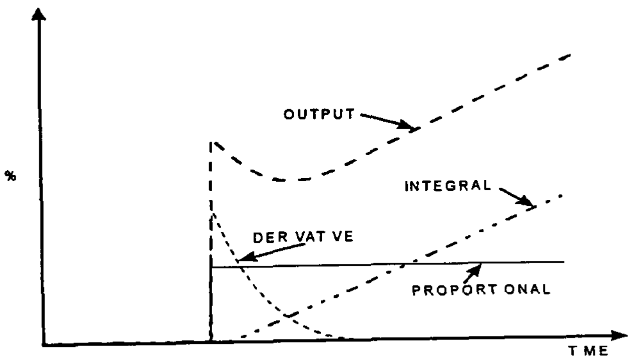


Figure 6-13 Sum Total of the Proportional, Integral and Derivative Control Terms

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.

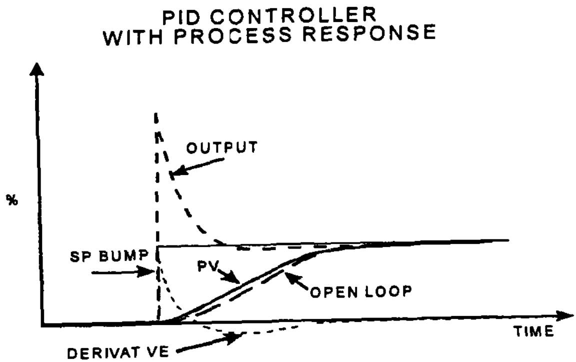


Figure 6-14 Open Loop Second Order Response Versus Closed Loop PID Tuned Response

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

$$K_{C\ \mathrm{Standard}} = K_{C\ \mathrm{Classical}} \frac{(T_{R\ \mathrm{Classical}} + T_{D\ \mathrm{Classical}})}{T_{R\ \mathrm{Classical}}}$$

$$T_{R\ \mathrm{Standard}} = T_{R\ \mathrm{Classical}} + T_{D\ \mathrm{Classical}}$$

$$T_{D\ \mathrm{Standard}} = \frac{T_{R\ \mathrm{Classical}} \cdot T_{D\ \mathrm{Classical}}}{(T_{R\ \mathrm{Classical}} + T_{D\ \mathrm{Classica}})}$$

Also NOTE the following

For Foxboro $K_C = \frac{100\%}{PB}$

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

Series)

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

$$K_{C\,\text{Parallel}} = K_{C\,\text{Classical}} \frac{(T_{R\,\text{Classical}} + T_{D\,\text{Classical}})}{T_{R\,\text{Classical}}}$$

$$K_{I\,\text{Parallel}} = \frac{K_{C\,\text{Classical}}}{T_{R\,\text{Classical}}}$$

$$K_{D\,\text{Parallel}} = K_{C\,\text{Classical}} T_{D\,\text{Classical}}$$

* Note that Bailey terminology for the proport onal gain is Kp, this should not be confused with the PROCESS GAIN Kp

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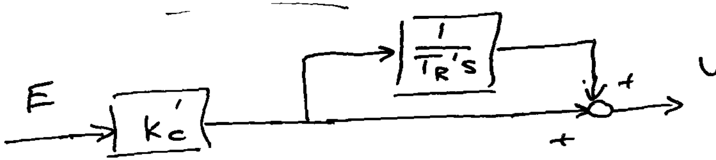
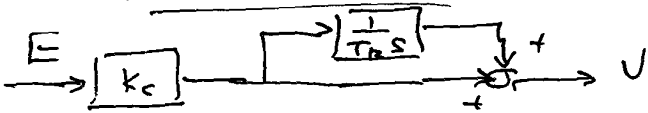
PI control

Classical Form



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

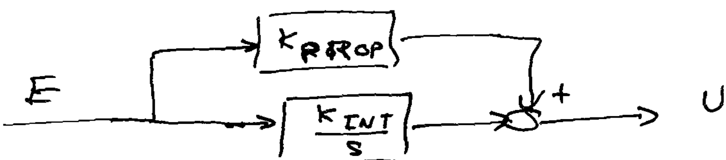
Standard Form



$$G_c(s) = K_c\left(1 + \frac{1}{T_R s}\right) = K_c\left(\frac{T_R s + 1}{T_R s}\right)$$

$$= K_c + \frac{K_c}{T_R}\left(\frac{1}{s}\right)$$

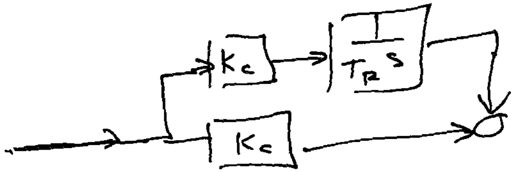
Parallel Form



$$G_c(s) = K_{PROP} + K_{INT}/s$$

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

23 34 13 04 10 07



PI only

To convert
classical / standard to Parallel.

$$K_{PROP} = K_c$$

$$K_{INT} = \frac{K_c}{T_R}$$

23 34 17 04 10 07

7. LAMBDA TUNING FOR FIRST ORDER PROCESSES

23 34 19 04 10 07

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 ca led the "COMFORT ZONE', shown below Unfortunately, it is hard to achieve exce lence in manufacturing through 'feel' alone

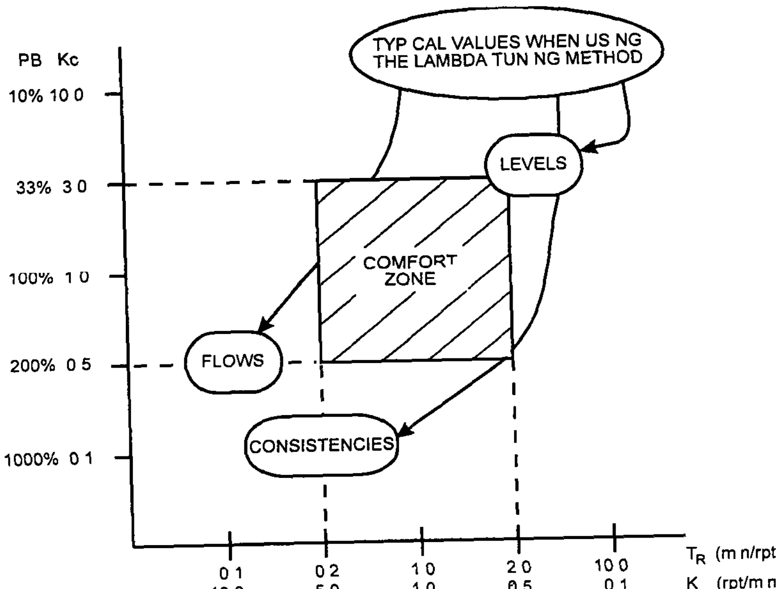


Figure 7-1 Typical Tuning Parameters vs Lambda Tuning Parameters

METHODS OF TUNING

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 1940's and 1950's 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

ZIEGLER-NICHOLS TUNING METHOD

BACKGROUND

The Ziegler-Nichols method was first developed in 1942. It is based on concepts first used for anti-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

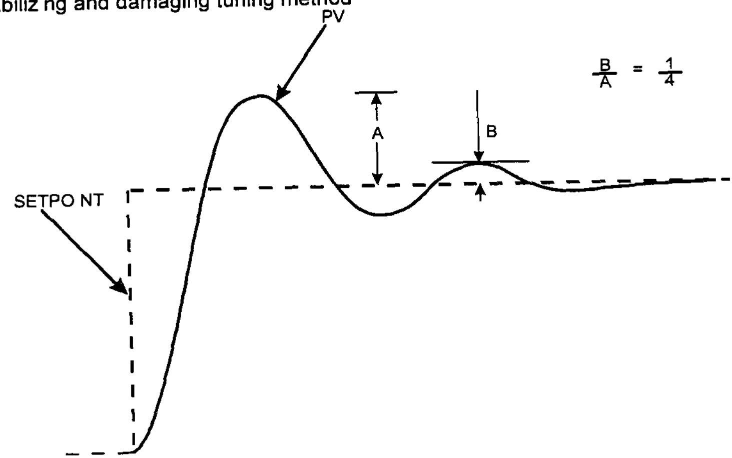


Figure 7 - 2 Quarter-Amplitude Damped Response Using Ziegler-Nichols Tuning Method

The tuning results are recommended only as a reasonable starting point

ZIEGLER NICHOLS TUNING METHOD

PROCEDURE

The tuning procedure s 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 contro er s gain K_C (or decrease PB) until a SUSTAINED OSCILLATION is recorded The ampl tude of the OSCILLATION must not be increas ng or decreas ng In order to cause the loop to cyc e, a sma l setpoint change may be requ red

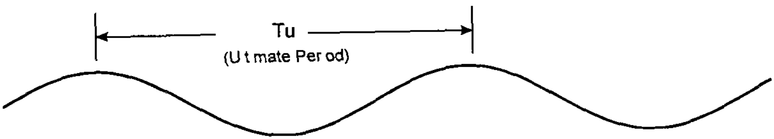


Figure 7 3 Measure the Ult mate Period From the Sustained Oscil ation

3 The fo owing parameters are then recorded

K_C or PB th s is ca ed the ULTIMATE GA N (the gain wh ch caused the osci ation)

T_U th s cal ed the ULT MATE PERIOD (the per od of the osc llat on)

4 Use these numbers to calculate the tun ng parameters accord ng to the Table 7-1

ZIEGLER-NICHOLS TUNING METHOD

TYPE OF CONTROLLER	PROPORTIONAL		RESET		DER
	K_C	PB(%)	T_R (min/rpt)	K_I (rpt/min)	TD (min)
P ONLY	0 5 K_U	2 PB_U			
PI	0 45 K_U	2.22 PB_U	T_U/1 2	1 2/T_U	
PID	0.6 K_U	1.67 PB_U	T_U/2	2/PB_U	T_U/8

Table 7-1 Formu a Table For Ziegler-Nichols Tuning Method

CONCLUSIONS

1 The method does not provide a knowledge of the important nonl near ties (i e backlash and st ction) in the loop

2 The method requires that the proportional gain be increased until the process osci lates which may be unacceptab e for operational reasons

3 It is d fficult to determine if the sustained oscil ation is a resu t of true process dynamics or nonlinear ties in the loop

4 The controller s speed of response is not defined in the procedure It is thus d fficult to tune two loops to have an ident'cal speed of response Th s is a common tuning objective

COHEN-COON TUNING METHOD

BACKGROUND

The Cohen-Coon Method, also known as the reaction curve method, was devised n the early 1950's Though the procedure is different, its objective is also to produce a quarter-amplitude DAMPED response as with Z eg er Nichols

Its development was due to the fact that it was not practical to make loops oscil ate during the tuning procedure Imag ne mak ng the boiler feed water flow cycle!

PROCEDURE

The procedure is s mp e It consists of perform ng a series of bump tests on the oop, while in MANUAL (as was discussed in Lecture 1) and recording the following process parameters

Deadtime	Td	=	
Time Constant	τ	=	___________
Process Ga n	Kp	=	

The tuning parameters are ca culated using the appropriate formu as in the Table 7 2

COHEN-COON TUNING METHOD

TYPE OF CONTROLLER	PROPORTIONAL		RESET		DERIVATIVE
	K_C	PB	T_R (min/rpt)	K_I (rpt/min)	T_D (min)
P ONLY	$\frac{Tau}{K_p T_d}\left(1 + \frac{T_d}{3Tau}\right)$	$\frac{300 K_p Td}{(3Tau + Td)}$			
PI	$\frac{Tau}{K_p T_d}\left(0\ 9 + \frac{T_d}{12Tau}\right)$	$\frac{1200 K_p Td}{(10\ 8Tau + Td)}$	$Td\frac{(30Tau + 3Td)}{(9T_{au} + 20Td)}$	$\frac{(9Tau + 20Td)}{Td(30Tau + 3Td)}$	
PD	$\frac{Tau}{K_p T_d}\left(1.25 + \frac{T_d}{6Tau}\right)$	$\frac{600 K_p Td}{(7\ 5Tau + Td)}$			$Td\frac{(6Tau - 2Td)}{(22Tau + 3Td)}$
PID	$\frac{Tau}{K_p T_d}\left(1\ 33 + \frac{T_d}{4Tau}\right)$	$\frac{400 K_p Td}{(5\ 32Tau + Td)}$	$Td\frac{(32Tau + 6Td)}{(13Tau + 8Td)}$	$\frac{(13Tau + 8Td)}{Td(32Tau + 6Td)}$	$\frac{Td}{(11Tau + 2Td)}$

Td = Deadtime
T_D - Derivative Time

Table 7 2

CONCLUSIONS

1 The formulas shown al heav ly rely upon deadtime They are not *appl cable f T_d is zero* Th s re iance on deadt me is mainly due to the fact that in the ear y 1950 s most controllers were pneumat c, and pneumatic lines introduce DEADTIME A general rule of thumb for est mating DEADTIME s that 100 feet of 1/4 pneumat c tub ng introduces about 1 5 seconds of deadt me

2 The testir g t me is usual y quicker than the Ziegler-N chols procedure t al ows the tuner to ident fy poss b e ma ntenance problems (i e BACKLASH STICTION, etc), and causes fewer process upsets

3 As w th the Z egler-Nichols method the tuning w l cause the process to cyc e on a setpo nt bump or a oad d sturbance

LAMBDA TUNING METHOD

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 Dahlin 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 CHOOSING LAMBDA, it does offer great FLEXIBILITY and REPEATABLE RESULTS

EnTech

LAMBDA TUNING METHOD

RESPONSE

By choosing LAMBDA, the tuner dec des how fast the PV wı reach the SP.

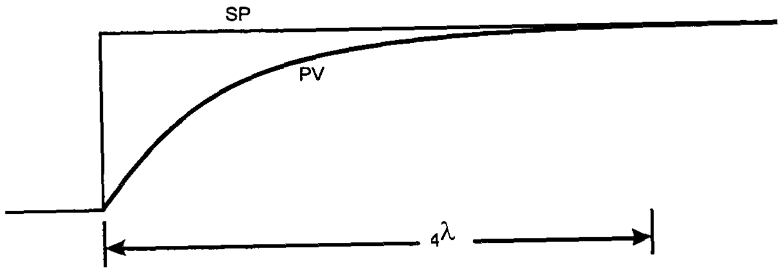


Figure 7- 4 Lambda Tuned Response

THE PV ESSENT ALLY 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

τ	CONTROLLER IN MANUAL
λ	CONTROLLER IN AUTO

λ – [illegible]
τ [illegible]

LAMBDA TUNING METHOD

PROCEDURE

The procedure is qu te simple.

1 Perform a series of bump test and obtain the following results

PROCESS DEADTIME	**Td**	=	________
PROCESS TIME CONSTANT	τ	=	________
PROCESS GAIN	**Kp**	=	________

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 τ.

2 Choose the LAMBDA value

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

3 Us ng τ, **Td**, **Kp** and **LAMBDA**, calculate the PI tuning parameters as fo lows,

T_R (min/rpt)	K_I (rpt/min)	K_C (%Out/%Span)	PB %PB
τ	$\frac{1}{\text{Tau}}$	$\frac{T_R}{Kp(\lambda + T_d)}$	$\frac{100 * Kp(\lambda + T_d)}{T_R}$

Table 7-3 Lambda Tuning Formulas

The Reset Time is *always* set equal to the open loop time constant τ The ca culation of the proport onal gain is a function of the process gain, the desired Lambda va ue, the open loop t me constant and the deadt me

LAMBDA TUNING METHOD

EXAMPLE:

The process dynamics of a flow oop are as follows

$K_P = 2 \ \dfrac{\%Span}{\%Output}$

T_d = 2 seconas

τ = 5 seconds

Assum ng the des red ambda were 15 seconds, K_C and T_R would be calculated as follows

$T_R = \tau$ 5 seconds = 0 083 mınutes

$$K_C = \frac{T_R}{K_p(\lambda + T_d)} = \frac{5}{2(15+2)} = 0\ 147\ \frac{\%Output}{\%Span}$$

Transfer Functions.

First order process. No deadtime

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

PI controller (classical Form)

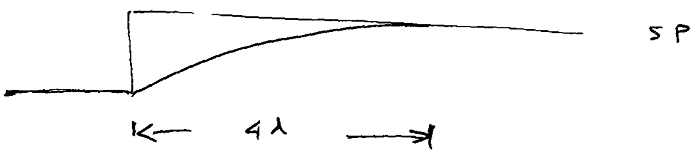
$$G_c(s) = K_c\left(\frac{T_R s + 1}{T_R s}\right) = K_c\left(1 + \frac{1}{T_R s}\right)$$

Select λ and λ-tune the controller.

$$T_R = \tau$$

$$K_c = \frac{1}{K_p} \cdot \frac{T_R}{\lambda}$$

$$G_c(s) = \frac{\tau}{K_p \lambda}\left(\frac{\tau s + 1}{\tau s}\right)$$



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closed - loop transfer function

$$G_{sp}(s) = \frac{G_c(s)\,G_p(s)}{1 + G_c(s)\,G_p(s)} \qquad \text{pg } 6\text{-}3$$

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

$$= \frac{1}{\lambda s} \qquad (\text{loop trans. fn})$$

Therefore

$$G_{sp}(s) = \frac{\frac{1}{\lambda s}}{1 + \frac{1}{\lambda s}}$$

$$= \frac{1}{\lambda s + 1}$$

First order with gain of 1.
and time const $= \lambda$.

23 34 35 04 10 07

CHOOSING 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 effectively 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
~~**Lambda** $= \tau/2$ **to**~~ **Lambda** $\leq 2 * \tau$

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.

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The following example shows the process and controller responses with **Lambda** is set equal to τ. The process dynamics are as follows,

Kp = 2 % Span/% Output
Td = 0 seconds

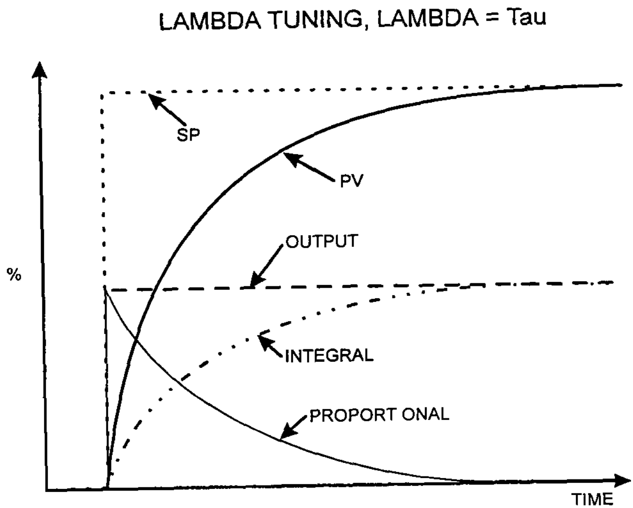


Figure 7- 5 Lambda Tuned Controller - Lambda = τ

In this case the controller output steps to it's 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 (**Lambda**) can be identical to the open loop time constant (τ)

The Proportional and the Integral terms are also shown Note that the proportional 'kick immediately brings the output to the 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 aggress ve 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

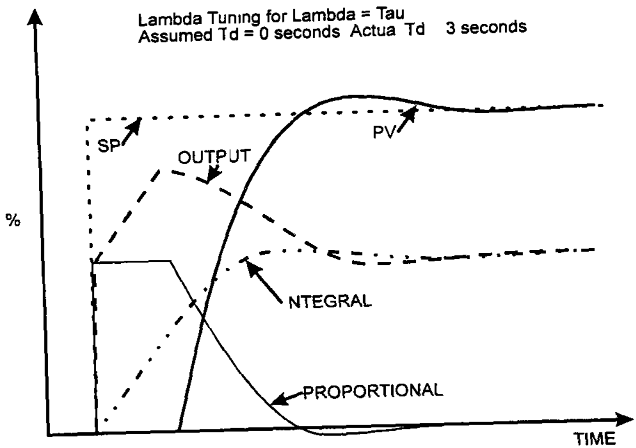


Figure 7- 6 Lambda = τ with 3 second deadtime

Because the PV does not change over the deadtime period, the controller output continues to ncrease 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, Lambda = Tau, Td = 3 Seconds, Kp = 4%, Span/%Ou
Note Tuning based on Kp = 2% Span/%Out, Td = 0 Seconds

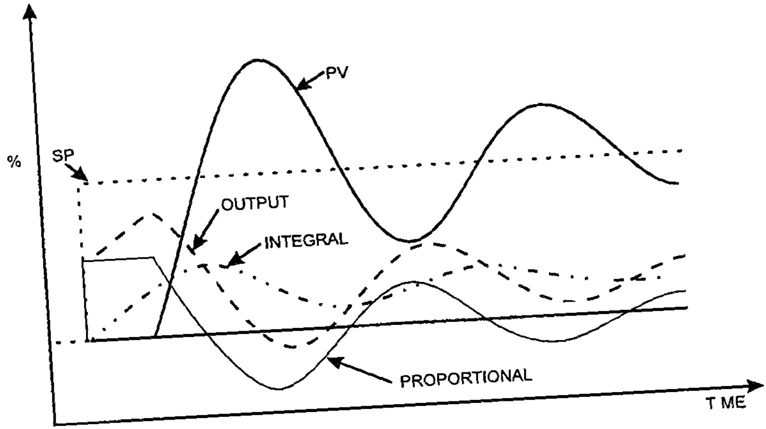


Figure 7- 7 Lambda = τ 3 seconds deadt me, Kp Increased

Under these cond t ons the response to a setpo nt bump is very oscillatory

RECOMMENDED TUNING
Lambda = 3 * τ or Slower

It is generally recommended that the Lambda value be equal to or *slower* than 3 * τ. 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 **Kp**, **Td** and τ 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 * τ** is shown. The process dynamics are the same as in the examples above i.e. Kp = 2 % Span/% Output, Td = 0 seconds. Note that the proportional 'kick' achieves one third of the total controller output change.

LAMBDA TUNING, LAMBDA = 3 * TAU

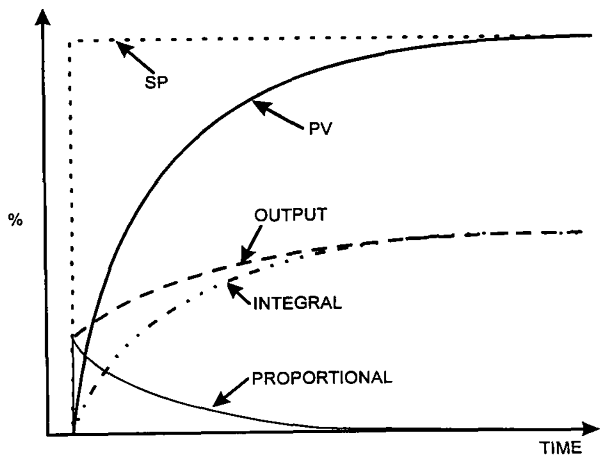


Figure 7- 8 Lambda = 3 * τ

The process dynamics are changed as shown in the example above. The deadtime is increased to 3 seconds and the process gain is increased to 4 %Span/%Output. When Lambda is equal to **3 * τ**,
the tuning is more forgiving to changing process dynamics as shown below.

LAMBDA TUNING, LAMBDA = 3*Tau Td ⁻ 3 SECONDS Kp 4% Span/ %Output
(TUNING BASED ON Kp = 2% Span/Output Td = 0 SECONDS)

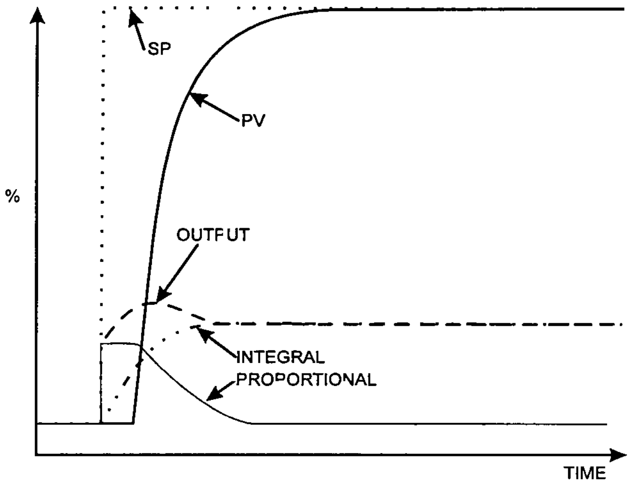


Fıgure 7- 9 Lambda = 3 * τ, Kp, Deadtıme Dynam cs Altered

In summary though ıt m ght be possıble to go faster,

LAMBDA = 3 * τ 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 τ, the deadt me 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 * Td OR SLOWER IS A SAFE STARTING POINT

EFFECTS OF FILTERING ON CONTROL PERFORMANCE

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.

$$T_{FILTER} = \frac{Tau}{2 \text{ to } 5}$$

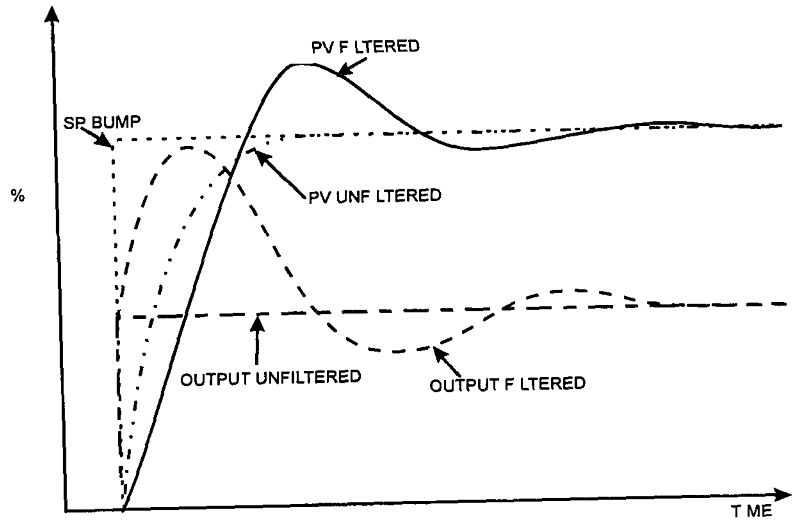
b) the FILTER time constant T_{FILTER} should be 5 to 10 times FASTER than the LAMBDA value chosen for the control loop. Excessive dampening will destabilize the loop and lead to cycling.

$$T_{FILTER} = \frac{LAMBDA}{5 \text{ to } 10}$$

Consider the following example. The controller has been tuned with a Lambda value of 10 seconds, which is equal to 1 * τ. Following retuning, a 10 second PV filter is installed at the controller. The unfiltered and filtered responses to a setpoint bump are compared in the figure below.

EFFECTS OF FILTERING ON CONTROL PERFORMANCE

PI CONTROLLER - 10 SEC PV FILTER VERSUS UNFILTERED



F gure 7-10 Oscillatory Setpoint Response due to excessive fltering

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 flter ng at only one locat on Since a prec se Filter t'me constant can be entered in a d gital controller, this s usua ly considered the best cho ce However, t is important to note that the process must be filtered at each point of sampling n a dig tal control system in order to avoid signal aliasing Ant -a iasing fi ters wil be discussed later

CONTROLLER TUNING EXAMPLES

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 % Span / % Output

τ = 6 6 seconds

T_d = 13 8 seconds

INITIAL TUNING

Kc = 3 7 % Output / % Span

T_R = 45 seconds

T_D = 30 seconds (Derivative)

T_{Filter} = 120 seconds

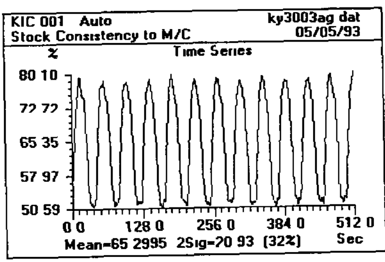
In the past, the operations group had complained about the amount of noise in the consistency signal and also to the slow response to setpoint bumps

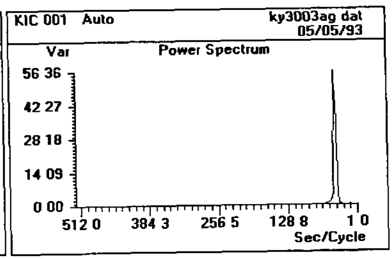
Questions:

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 * T_d = ________ seconds

3 * τ = ________ seconds

2 If Lambda is set equal to ________ seconds

- the total time to reach the new setpoint is ________ seconds

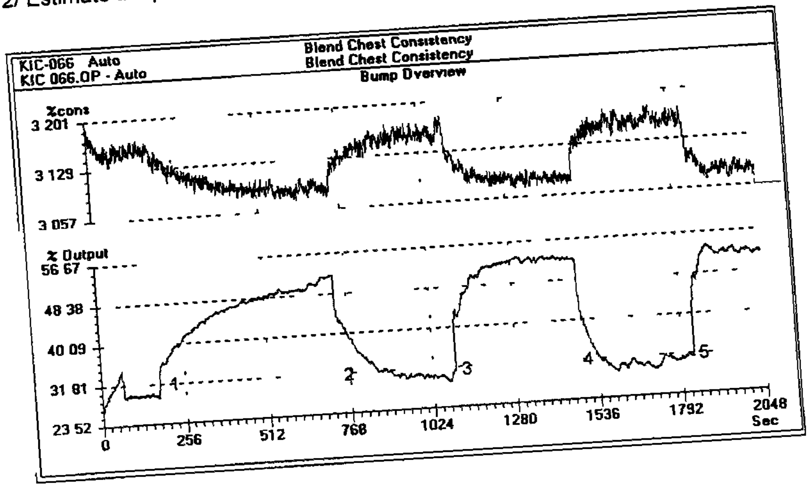
the total filter time in the loop should be less than ________ seconds

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 ncreasingly 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

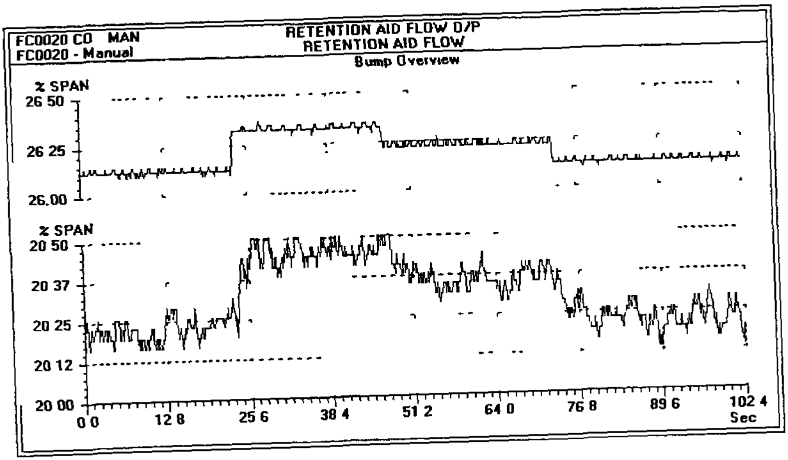
Retention aid is added to the machine whitewater in order to increase first pass fiber retention on the wire. The following time series plot shows an open loop bump and a closed loop setpoint bump

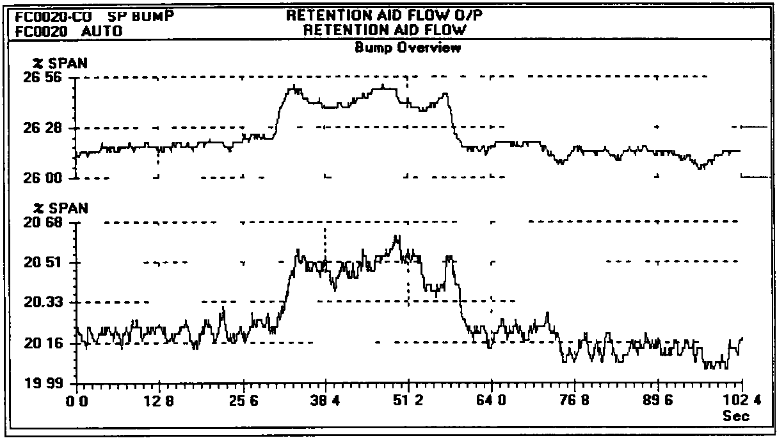
Questions:

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.



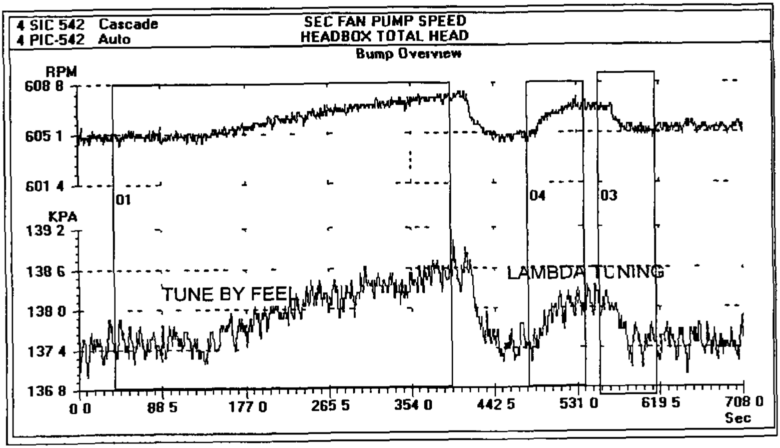

FC0020-CO SP BUMP
FC0020 AUTO
RETENTION AID FLOW O/P
RETENTION AID FLOW
Bump Overview
% SPAN
26 56
26 28
26 00
% SPAN
20 68
20 51
20.33
20 16
19 99
0 0
12 8
25 6
38 4
51 2
64 0
76 8
89 6
102 4
Sec

EXAMPLE 7-4

The fol owing figure shows successive setpoint bumps on a Paper Mach ne Headbox Tota Head controller The f rst is the Tune by Fee response, and the following are Lambda tuned responses.

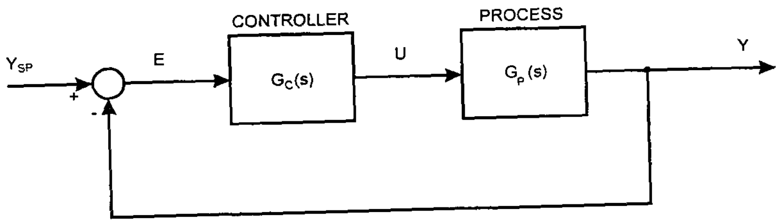
Questions:

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

Cons der the standard control loop configurat on



The closed loop transfer funct on (as always) is given by

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

regardless of $G_C(s)$!

Consider designing an arbitrary control er such that any particu ar des red closed loop transfer function $G_{SP}(s)$ can be achieved Spec fying the closed oop transfer function $G_{SP}(s)$ wil require the contro ler transfer funct on to be (by a gebraic man pu ation)

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

Suppose that we want the closed loop system to respond as a first order process to a setpo nt change Thus $G_{SP}(s)$ s chosen to be

$$G_{SP}(s) \quad \frac{1}{\lambda s+1}$$

where λ (ambda) is the des red closed oop t me constant Note that the gain is 1 Th s s to ensure that the change in the output w l match the setpo nt change in steady state while g v ng the required dynamics

To obta n a c osed oop system which s first order w th t me constant λ, wil requ re

$$
\begin{aligned}
G_C(s) &= \frac{1}{G_P(s)} \cdot \frac{G_{SP}(s)}{1 - G_{SP}(s)} \\
&= \frac{1}{G_P(s)} \cdot \frac{\left(\dfrac{1}{\lambda s + 1}\right)}{1 - \left(\dfrac{1}{\lambda s + 1}\right)} \\
&= \frac{1}{G_P(s)} \cdot \frac{1}{\lambda s}
\end{aligned}
$$

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

$$
\begin{aligned}
G_C(s) &= \frac{1}{G_P(s)} \cdot \frac{1}{\lambda s} \\
&= \frac{\tau s + 1}{K_P} \cdot \frac{1}{\lambda s} \\
&= \frac{\tau}{K_P \lambda} + \left(\frac{1}{K_P \lambda}\right) \frac{1}{s} \\
&= \frac{\tau}{K_P \lambda}\left[1 + \frac{1}{\tau s}\right]
\end{aligned}
$$

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

$$T_R \quad \tau$$

$$K_C \quad \frac{1}{K_P}\left(\frac{\tau}{\lambda}\right) \quad \frac{1}{K_P}\left(\frac{T_R}{\lambda}\right)$$

(reset tıme)

(contro er gaın)

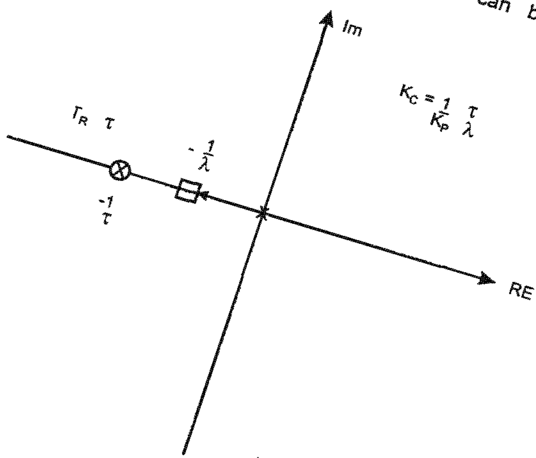
LAMBDA TUNING RULE

Note that the choice of the reset time is determined strictly from the open loop t me 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 or gin (always) and one zero whose location is determined by the reset time

A f rst 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 effect vely cancelling both the po e and zero We are now left with a sing e closed loop pole whose location we are free to choose by varying the proportional gain (see Root Locus example).

This method is meant pr mar ly for processes that can be reasonably approx mated as first order.



8. NONLINEARITY IN CONTROL LOOPS

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NON-LINEARITIES RESULTING FROM THE CONTROL STRATEGY

EXAMPLES

Split Ranging Strategy

A split ranged controller has one output but more than one man pu ated 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 Leve 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 be relatively high Control performance will suffer as a result It is part cularly important to understand the process requirements of the loop in order to make sound tuning decis ons

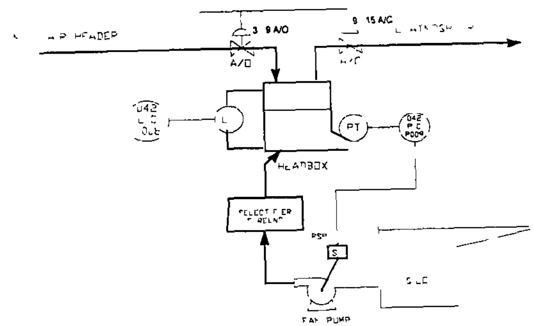


Figure 8-1 Air Padded Headbox Using Spl t Ranged strategy to control Leve

Low / High Select Strategy

The low / high select strategy configuration permits a manipulated variable to be shared by two or more controllers. This strategy is typically employed to extend the control range and/or to allow safe operation of a process system.

On a paper machine, a Dryer section which recirculates blowthrough steam with a Thermocompressor(TC) almost always uses a Low-select 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 / low 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 fall below 50% and non-linear 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.

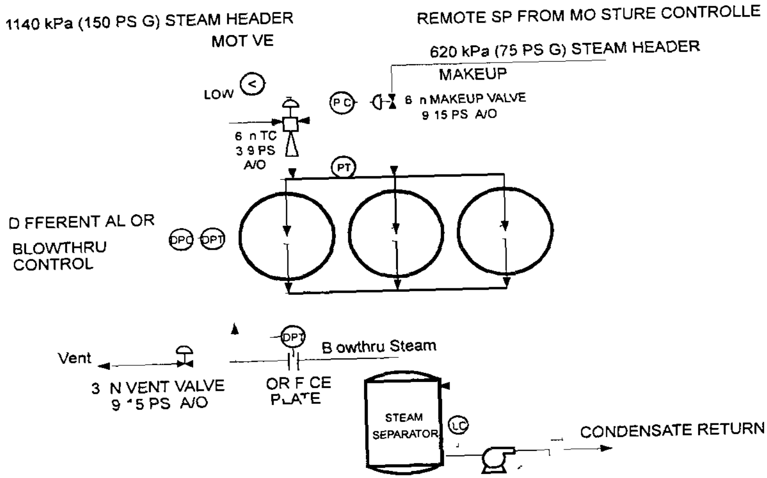


Figure 8.2 Dryer Section using Low Select strategy

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 ga n be ng dependent on stock consistency (e Blend Chest controller output scaled to Dry Tons target rather than a Wet flow target)
2/ Broke addition strategy effects the Blend Chest leve controller process gain (i.e Broke is ratioed to virg n fibre (rather than d rect y based on Level control er demand) making the Level process gain dependent on the Broke rat o)

Logic Controllers

Refiner Control

1/ Ref ner plates move at *fast* speed or *slow* speed depend ng on the magnitude of the error
2/ Hysterisis deadband employed for contro of Ref ner p ate pos tion

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'.

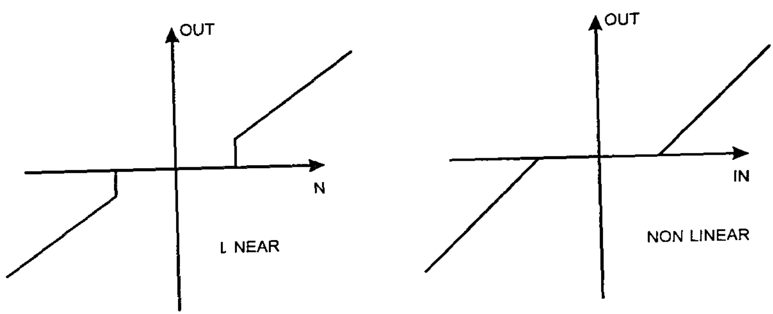


Fig 8-3 Linear and Non Linear Deadbands

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.

<u>A Special Case</u>

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

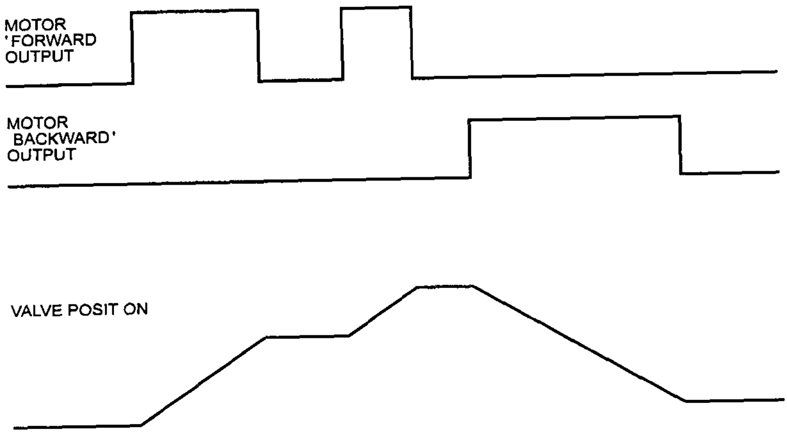


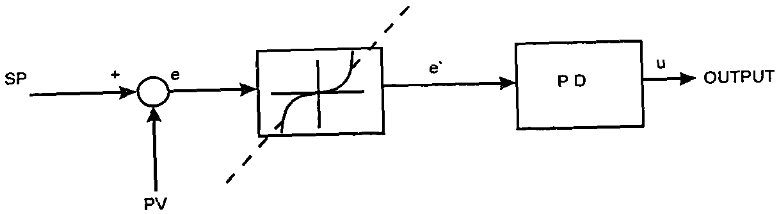
Fig 8-4 Velocity Mode Actuator

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 <u>output</u> and not the <u>error</u> 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



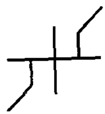


Fig 8 5 Error Characterization

Strategies for Dealing with Non-Linearities

1/ Fix them
This strategy is particularly applicable to contro valve design and maintenance The single biggest sources of short term variability are control valve backlash and stiction These can be minimized *first* by specify'ng the dynamic performance requireme1t prior to the valve purchase and *second* by employ ng a pro-active maintenance approach.

2/ Hide Them in An Inner Loop
Cascade control refers to a configuration where the control er output of one loop (outer or master control er) becomes the setpoint of another controller (inner or slave controller) This s shown in block diagram format below

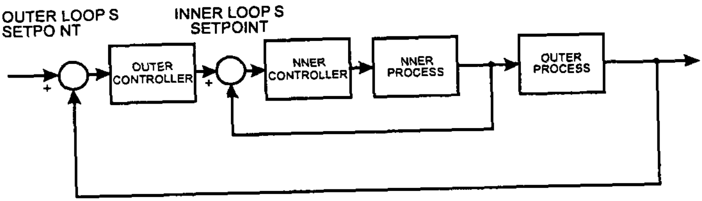


Fig 8-6 Block Diagram of Cascaded Loops

There are severa motivat ons for design ng a cascade configuration First, the inner flow control loop w l take control act on to attenuate an nner process disturbance *before* it has an effect on the oute- p-ocess Seconc y instal ng a- inner control loop wil linearize (to some degree) the dynam cs of the outer control loop This is because a unit step n the outer oop contro ler output will result in a constant change to the inner control loop setpoint The nonlinearit es in the control va ve will be effect vely passed on to the nner controller

In the examp e below, the Brightness controller cascades a setpoint to the CL_2 flow control er 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 resu ting from the control valve flow character stic and/or backlash and stiction w ll be largely eliminated

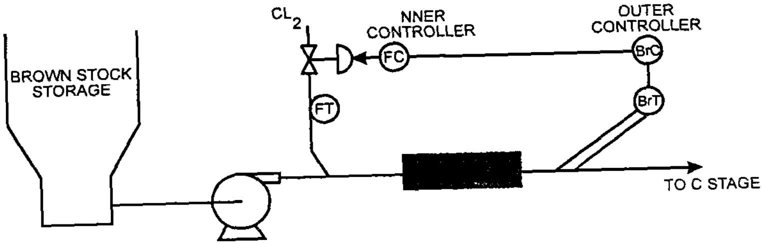


Fig 8 7 Brightness Control

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
- H ghest Resonant Peak

Example Cons stency Contro

Max (K_p) at M n (F)

Max (T_d) at Min (F)

Max (τ) at Min (F)

$\Rightarrow$ Tune loop at Minimum flow

Example pH Control

Max (K_p) Centre' pH

$\Rightarrow$ Tune loop at 'Centre' pH

Example Moisture Control

Max (K_p) at Min (P)

$\Rightarrow$ Tune loop at Minimum pressure

Another possibi ity which may be availab e on certain distr buted control systems is 'Grade' Dependent Tuning This would allow for d fferent tuning on different grades or at different operating points. This alternative may be cons dered 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 nonl nearities such as valve flow characteristics, backlash/stiction, the size of the bump and others These nonlinearit es 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 nonlinear ties The bump test results must be interpreted with respect to a tuning strategy to develop the 'best control er tun ng

Some typical results from a bump test series are shown below

K_P	T_d	τ
0 8	3 0	2 0
0 5	2 5	3 0
1 3	4 0	2 0
1 5	3 0	4 0
1 1	2 0	4 0
0 9	3 0	3 0

Us ng the most conservative process dynam cs w ensure c osed loop stabil ty under a l operat ng condit ons The concess on wh ch accompan es this tun ng strategy is slugg sh control performance under some operat ng cond tions Adaptive tuning may be required to ensure adequate contro performance under a conditions

Consequences of K_P Changing

$$\lambda = \frac{T_R}{K_C K_P} \qquad K_C \text{ ca culated based on h gh } K_P$$

As the process gain gets smal er, λ wi get longer and the loop wi become more sluggish Since K_C was probab y ca culated based on a high va ue of K_P λ wi l vary from the va ue chosen to s ower va ues Shou d the process ga n ncrease above that used to calculate K_C, the loop wi l speed up This may cause the fast dynamics to osci late

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

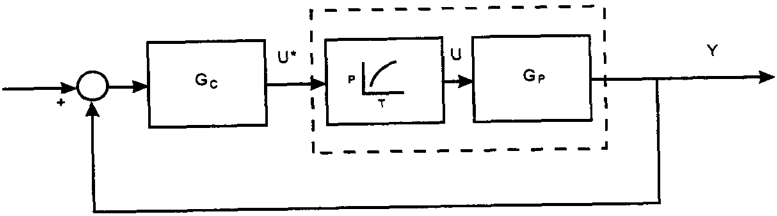


Fig 8-8 Non-Linear Moisture controller

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

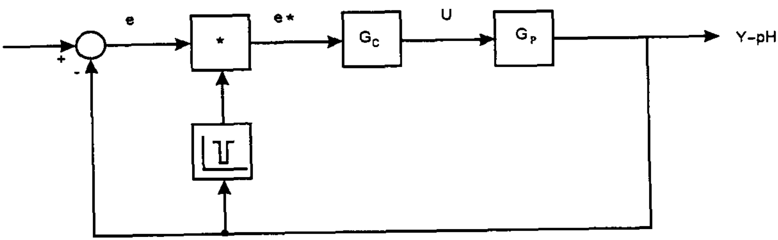


Fig 8-9 Non-Linear pH controller

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- inear control are

- vary estimated T_d as function of machine speed for weight control w th Dahl n or Smith
- character ze a valve for flow control Then compensate proposed U for valve behavior
- vary K_P & T_R according to the estimated τ on a Lambda Tuned PI control er.

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 contro is not covered in this course

9. SETPOINT RESPONSE AND LOAD RESPONSE

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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 Rejection
   - 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

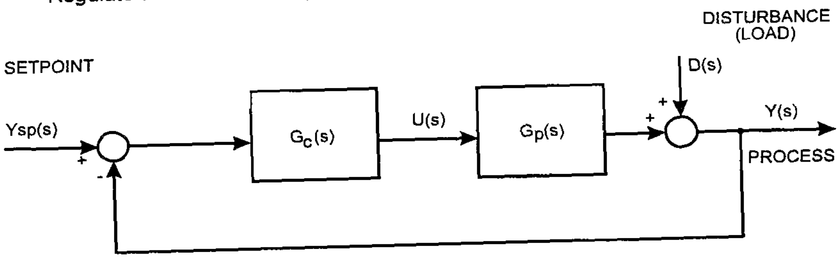


Fig 9-1 Block diagram of a feedback control loop

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

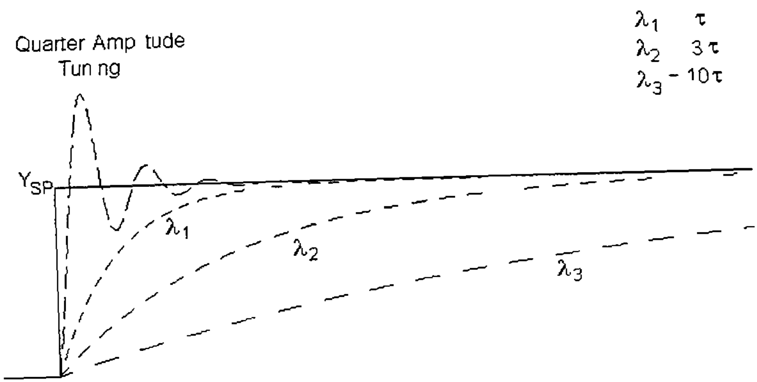


Fig 9 2 Setpoint responses for various controller tunings

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:

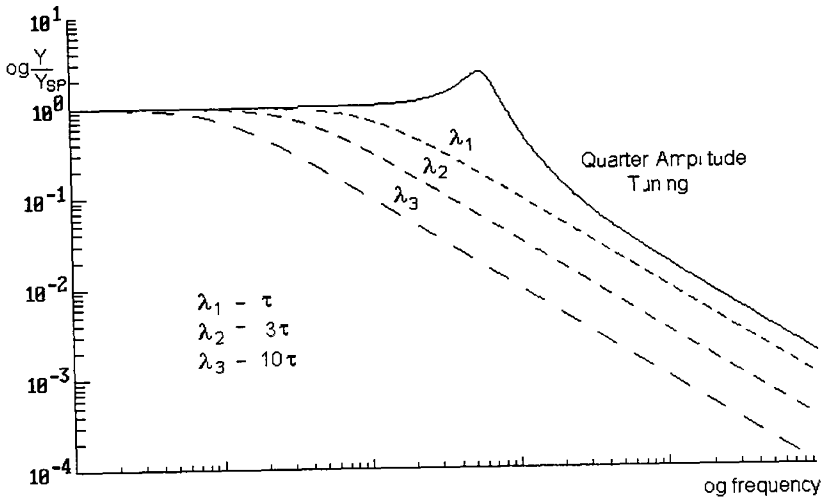


Fig 9-3 Setpoint tracking Bode plots for various controller tunings

Note The period at which frequency response starts to decay (cut-off frequency cut-off period) is given by the following formulas

cut off frequency $\quad f_c = \dfrac{1}{2\pi\lambda}$

cut-off period $\quad T_c = 2\pi\lambda$

Note that Ziegler Nichols and other oscillatory tuning will amplify certain frequencies (resonant frequency).

CASCADED LOOPS

In a cascaded loop s tuat on the outer loop (or master loop) updates the setpoint of an nner oop (or slave Loop) The nner loop then adjusts ts process to achieve the new setpoint This n turn affects the outer loop s process dynamics

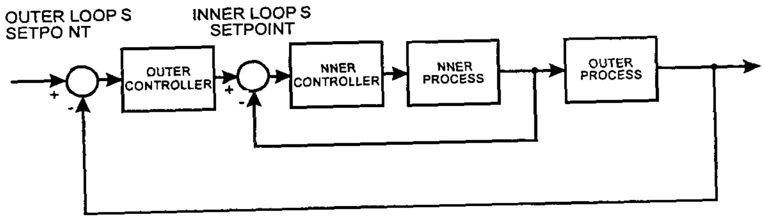


Figure 9-4 Block Diagram of Cascaded Loops

In the pulp and paper industry, there are many examp es of cascaded oops In the paper machine area for instance, the sheet moisture control er (OUTER LOOP) cascades a setpoint to the steam pressure control er (NNER LOOP)

It s mportant to tune the INNER controller so that it s fast enough to reasonably track the setpoint from the OUTER oop A good rule of thumb is that the OUTER loop shou d be tuned at least 5 times SLOWER than the INNER loop Th s means

$$\boxed{\lambda_{OUTER} > 5 \times \lambda_{INNER}}$$

If th s rule is not kept, then both the Inner and Outer processes may tend to osc late

BRIGHTNESS CONTROL EXAMPLE

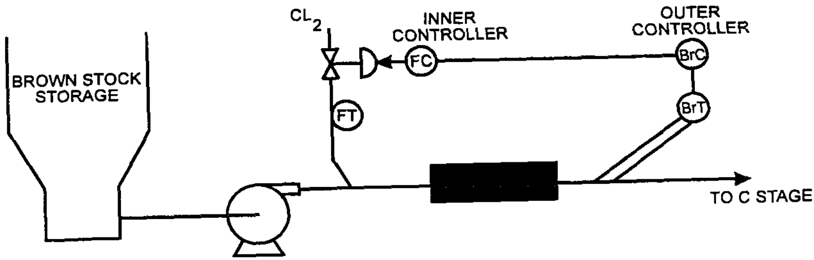


Figure 9.5 Typical C Stage Brightness Control

The dynamics of the Brightness and CL_2 loops are as follows

	Brightness	CL2 Flow
Kp	1 %Span/%Out	1 %Span/%Out
Tau	10 sec	5 sec
T_d	10 sec	0 sec

If the Cl_2 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

$$\text{LAMBDA}_{\text{FLOW}} = \text{15 SECONDS}$$

$$\text{LAMBDA}_{\text{BRIGHTNESS}} = \text{75 SECONDS OR SLOWER}$$

The brightness example described above is illustrated in the following figures

In the f rst figure, the Inner flow loop is tuned w th 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

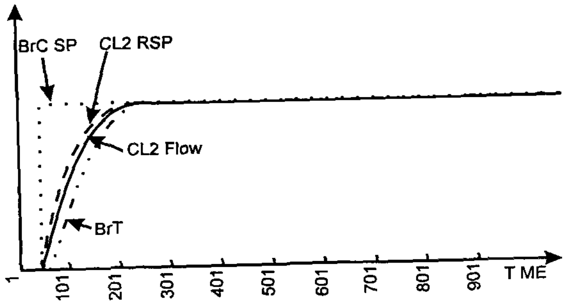
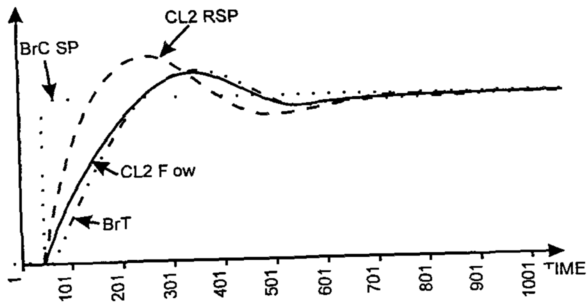


Fig 9 6 Brightness, Flow Responses to a Brightness SP Bump

In the second figure the Lambda value of the Outer loop are both equal to 75 seconds The Inner flow loop is unable to effect ve y track the Brightness control er setpo nt, and the response is oscillatory

CASCADE LOOPS - INNER & OUTER LAMBDA = 75 SECONDS



F g 9-7 Br ghtness 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.

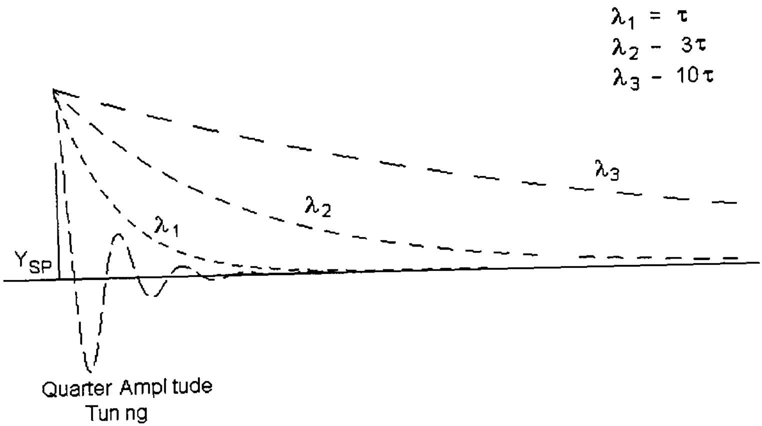


Fig 9-8 Load recovery for various controller tunings

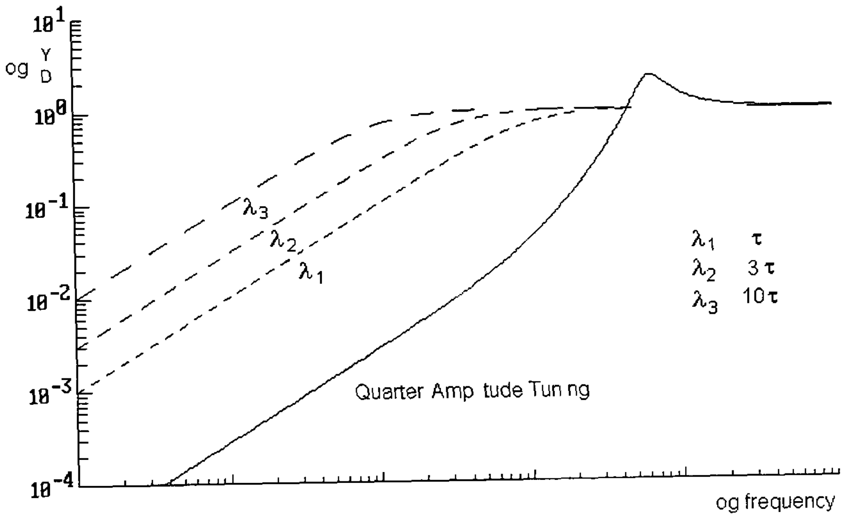
The transfer function from the disturbance D to the PV Y is

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

Examp e For a 1st order process with a PI contro ler that has been λ tuned

$$G_{LD}(s) \quad \frac{\lambda s}{\lambda s+1}$$

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 setpo nt value However, as D changes more and more rapidly, the loop wi l be less and ess able to bring the process back to setpo nt in time Hence, for very rapid d sturbances no effective correction wil be done and the gain wi l be un ty The frequency response for the regulator is shown in Fig 9-9



F g 9-9 Regulator Bode p ots for var ous contro ler tun ngs

Note The lower the frequency of the d sturbance, the more effect ve y w ll the loop regulate or reduce the amp itude of the fluctuat ons in D Attenuat on starts at the cut off frequency

$$\text{cut-off frequency} \qquad f_c \quad \frac{1}{2\pi\lambda}$$

$$\text{cut off per od} \qquad T_c \quad 2\pi\lambda$$

As the disturbance frequency is reduced attenuation will occur at 1 decade/decade (-20 dB/decade). This corresponds to

Disturbance Period	$1000T_c$	$100T_c$	$10T_c$	T_c	$0.1T_c$	$0.01T_c$
Variability Remaining	0.1%	1%	10%	70%	100%	100%

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 $T_c = 2\pi\lambda \simeq 59.7$ sec. $\cong 1.0$ min

Disturbance Period	100 min	10 min	1 min	6 sec
Variability Remaining	1.0%	10%	70%	100%

Trade Offs in Choosing Lambda

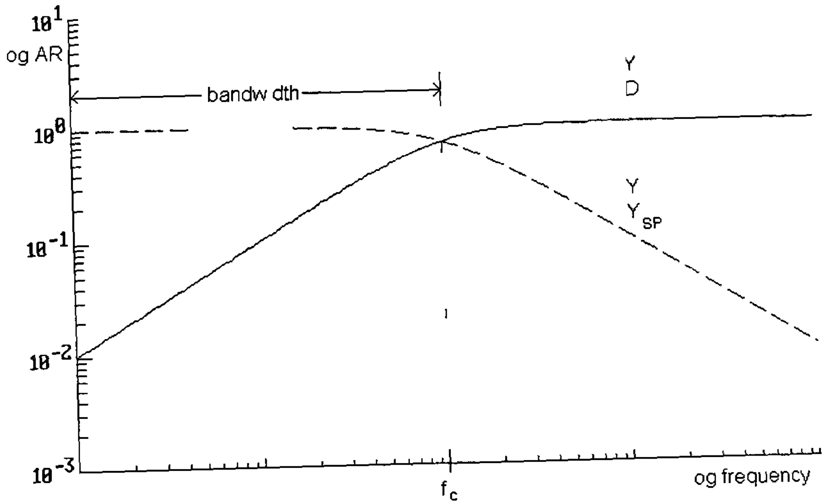


Fig 9 10 Setpo nt track ng and regu ator Bode p ots for a Lambda tuned first order oop

cut-off frequency f_C $\frac{1}{2\pi\lambda}$

cut off per od T_C $2\pi\lambda$

Trade offs in choosing λ

λ *smaller*

- h gher bandw dth
- faster setpo nt response
- better disturbance reject on
- less robust

λ *larger*

- sma er bandw dth
- slower setpoint response
- ess disturbance rejection
- more robust

EnTech

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

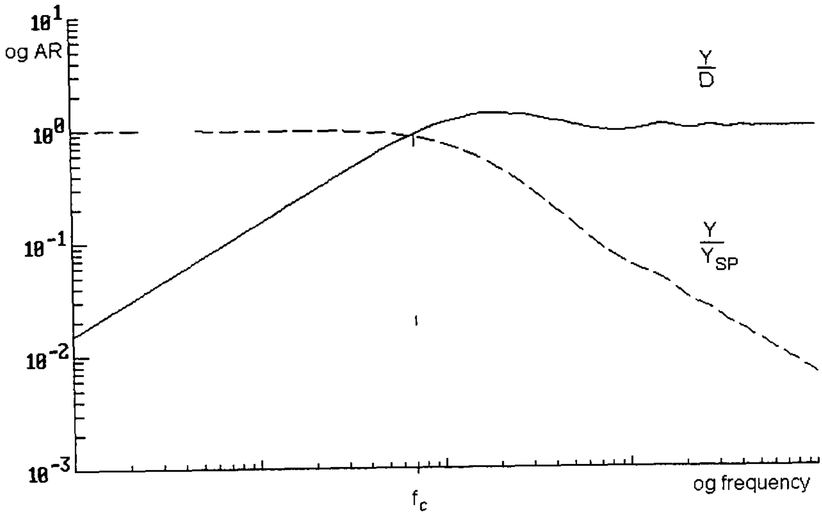


Fig 9 11 Setpoint tracking and regulator Bode plots with deadtime present

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

$$f_C = \frac{1}{2\pi(\lambda + T_d)}$$

$$T_C = 2\pi(\lambda + T_d)$$

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 is shown in the following figure

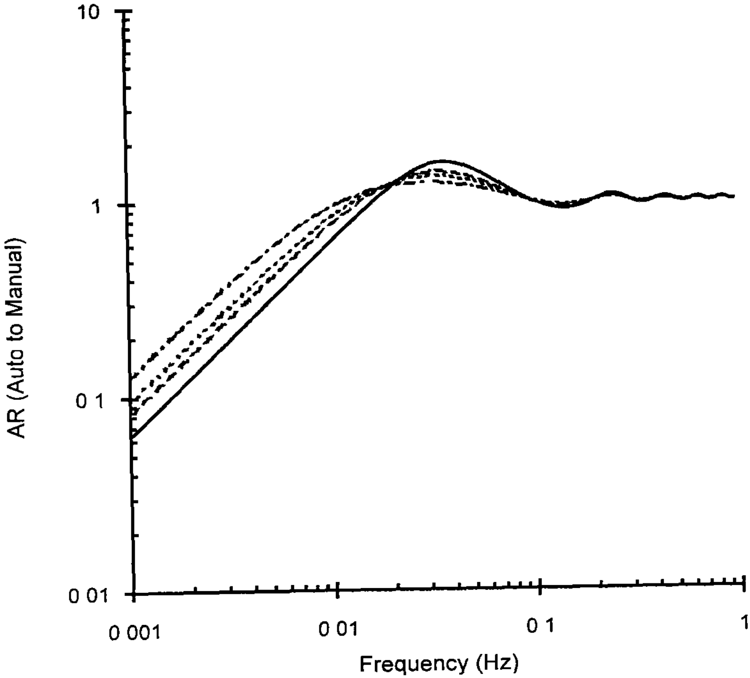


Fig 9 12 Regulator Bode plots for various values of Lambda

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

The dependency of the resonant peak M_P on the Lambda to deadtime ratio λ/T_d is shown in the following figure

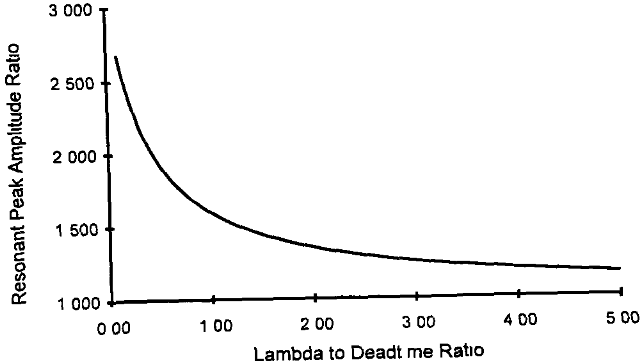


Fig 9-13 Regulator Resonance for a First Order with Deadtime Process

The dependency of the resonant period to deadtime ratio T_P/T_d on the Lambda to deadtime ratio λ/T_d is shown in the following figure

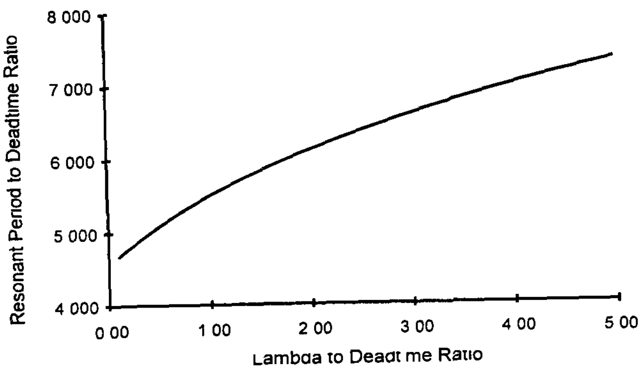


Fig 9-14 Regulator Resonant Period for a First Order with Deadtime Process

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 λ 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 rat o of deadt me to lambda decreases the contro oop wi exhibit more *control loop resonance*, result ng in increased amplification of d sturbances in a frequency range faster than the cutoff The following table gives the lower lim t for choosing λ re ative to the deadt me for a PI controller contro ling a first order w th deadt me process n order to cap the resonance at var ous va ues

Resonant Peak $AR = \frac{A_{AUTO}}{A_{MAN}}$	Resonant Peak in dB	Lower imit for λ to cap the Resonant Peak
1 26	2 0	$3 \times T_d$
1 35	2 6	$2 \times T_d$
1 41	3 0	$16 \times T_d$
1 59	4 0	T_d

An increased certainty in the mode parameters may just fy faster tun ng, however faster tuning wi ncrease 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 τ 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 2.5 seconds and various process deadtimes as shown in the following table

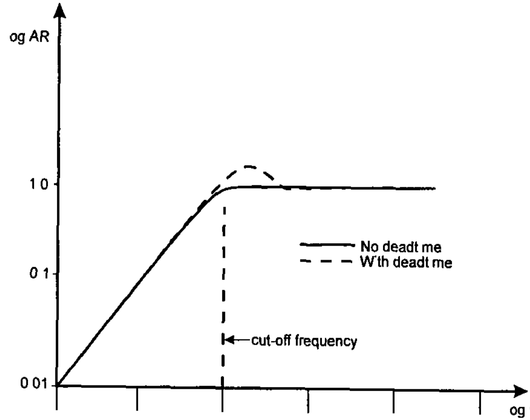
Process Deadtime T_d (sec)	Closed-Loop TC λ (sec)	Cutoff Period T_C (sec)	Cutoff Period T_C (min)
5	15	126	2.1
10	30	251	4.2
15	45	377	6.3
20	60	502	8.4
25	75	628	10.5
30	90	754	12.6

There is much benefit from keeping the process deadtime in a stock consistency control loop to a minimum by locating the consistency sensor as close as is 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

AR rat o of on contro to off control ampl tude



τ process t me constant

T_d process deadt me

λ c osed loop time constant

T_c cut off per od for d sturbance reject on

Process	τ	T_d	λ	T_c
HB Total Head	1 sec	0	3 sec	20 sec
Stock Flow	3 sec	0	10 sec	1 m n
Stock Cons stency	3 sec	5 sec	15 sec	2 0 min
Pu p Brightness	20 sec	20 sec	1 min	8 4 m n
PM Bas s We ght	45 sec	100 sec	4 min	36 min
PM Mo sture	2 min	2 min	6 min	50 min

Period of sine wave d sturbance

30 m n	3 min	20 sec	2 sec
100 m n	10 min	1 0 min	6 sec
200 m n	20 min	2 0 min	12 sec
14 hrs	84 m n	8 4 m n	50 sec
60 hrs	6 hrs	36 m n	3 6 m n
83 hrs	8 3 hrs	50 m n	5 min

F g 9-15 Cutoff Periods for typ cal pulp and paper control loops

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APPENDIX 9A

REGULATORY CONTROL EXAMPLE

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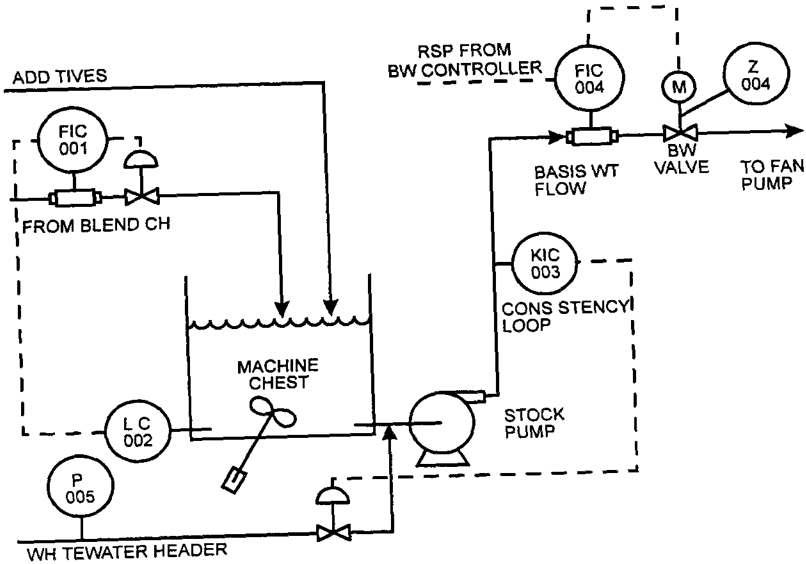
EXAMPLE:

The whitewater dilution header pressure is cycling When the consistency contro ler is in manual mode (no control act on), the machine chest consistency cyc es from 3 0 % to 3 2 % consistency. The Lambda value of the consistency controller is 10 seconds There is 5 seconds of process deadtime.



F g 9A 1 Machine Chest Consistency Loop

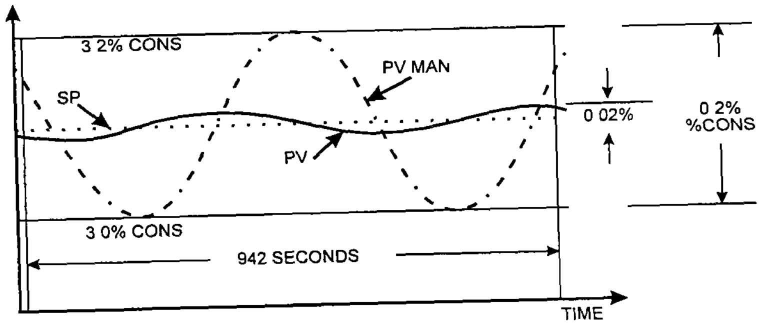
The CUTOFF period of the consistency controller is equal to

$$T_C \quad 2\pi(\lambda + T_d)$$
$$6.28 \times 15 \text{ seconds}$$
$$94.2 \text{ seconds}$$

Case 1:

The D lution Header cycles at a period of 942 seconds Th s s 10 t mes slower than the Cut-Off period of the cons stency control er

FREQUENCY RΓSPONSE REGULATORY CONTROL



F g 9A 2 Frequency Response - S ow D sturbance Cycle

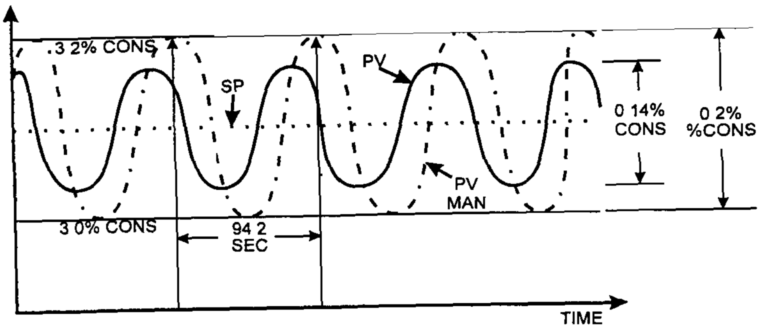
In this case the control er s successfu in reducing the impact of the d sturbance cycle by 90 % The ampl tude of the process cycle is only 0 01 % cons stency

$$AR = \frac{0.01\ \%\text{Consistency}}{0\,1\ \%\text{Consistency}} = 0\ 1$$

CASE 2

In th s case, the period of the disturbance cycle is equal to 94.2 seconds which equals the loop s cut off per od.

FREQUENCY RESPONSE REGULATORY CONTROL



F g 9A-3. Frequency Response - Disturbance Cyc e at Cutoff

The controller is only capable of reducing the d sturbance by only 30 % The amp itude of the consistency cycle is 0 07 % cons stency

$$AR = \frac{0.07}{01} = 0\ 7$$

CASE 3

In this case, the period of the disturbance cycle is 9 42 second or 10 times faster than the Cut Off period of the loop

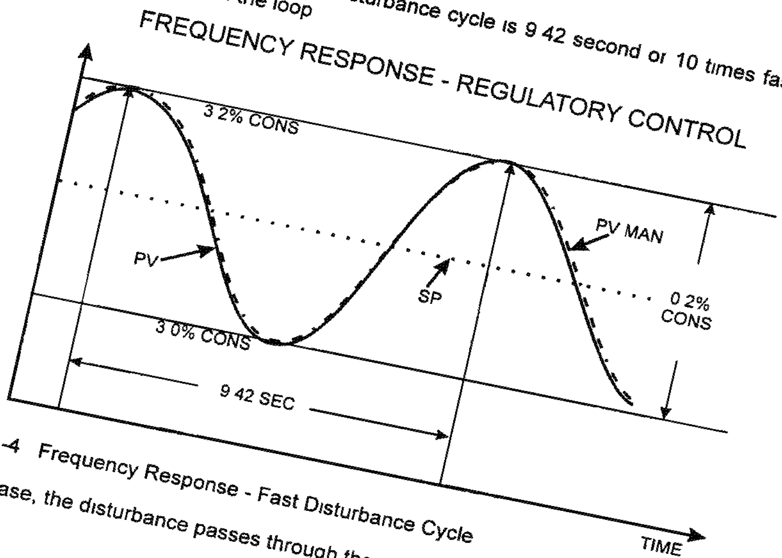


Fig 9A-4 Frequency Response - Fast Disturbance Cycle

In this case, the disturbance passes through the control er comp ete y unattenuated

$$AR \quad \frac{0.1}{0.1} = 1$$

10. TUNING INTERACTING CONTROL LOOPS

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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 examp e of interactive loops is the Total Head and Liquid Level loops in an air-padded headbox·

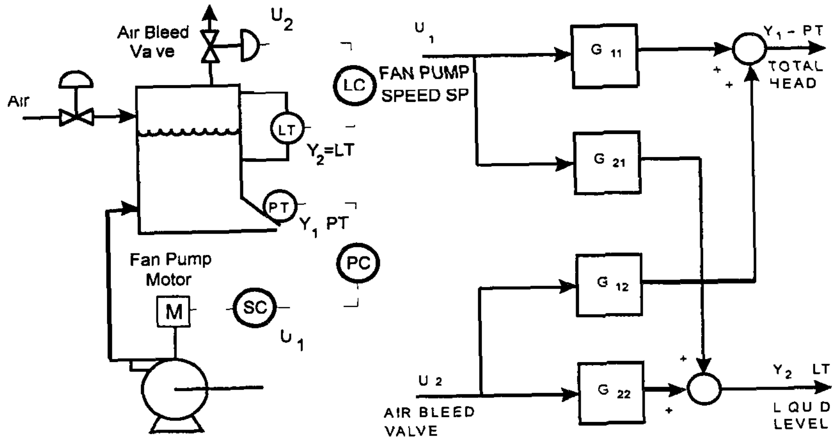


Fig 10-1 Headbox Total Head and L quid Level Process Diagram

In this example the evel s 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 wh ch the other loop w ll try to attenuate Th s carries the potent a 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 prev ous examp e, 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 tun ng method s that it allows us to quant fy "making one loop slower than the other by comparing the r Lambda values This a so allows for the systematic development of a tuning strategy for a group of several interact ng loops as we wi l see below

A b ock d agram representing control loop interaction s shown in Figure 10 2 Notice how the control er output of one controller creates a d sturbance for the other contro oop by driving through the coupl ng dynam cs

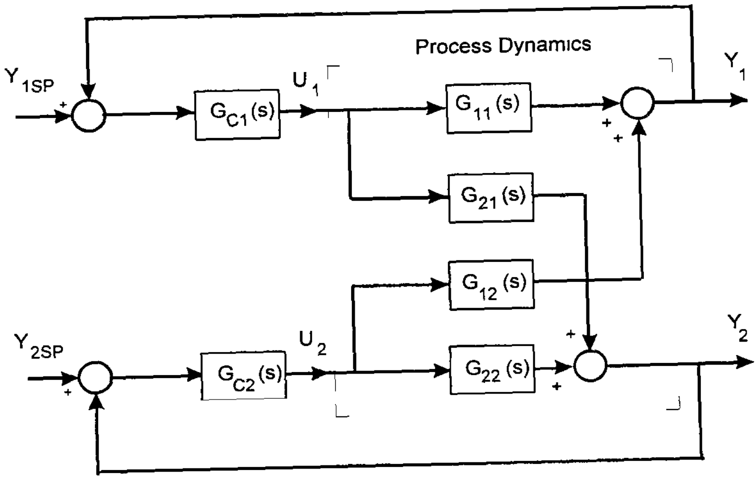


Fig 10-2 Block Diagram Il ustrat ng Control er Interaction

The general tuning procedure is as fo lows

1 Se ect one oop 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 g ves process dynamics for oop Y_1

3 Tune the more mportant loop Y_1 selecting the closed loop t me constant λ to g ve good performance (typically λ_1 max$(3\tau_1, 3T_{d1})$)

4 Put loop Y_1 in Auto mode (w th the above Lambda tun ng) and keeping loop Y_2 in Manua mode perform bump tests on loop Y_2

5 Using the process dynamics for Y_2, tune Y_2 with a closed loop t me constant λ_2 at least 5 t mes slower than λ_1, namely choose

$$\boxed{\lambda_2 > 5 \times \lambda_1}$$

n the Headbox examp e choose $\lambda_{LEVEL} > 5 \times \lambda_{TH}$

TUNING A SYSTEM OF INTERACTING LOOPS

Cons der the Stock Refining system shown ın Figure 10 3

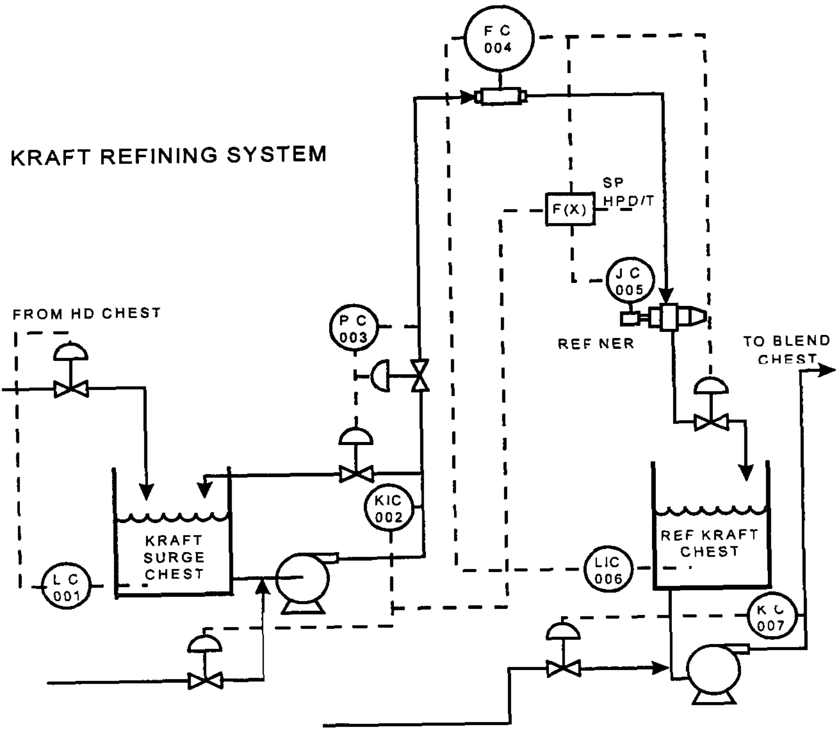


Fıg 10-3 Stock Refinıng System wıth Interact ng Loops

Process Objectives

The objective of this system is to safe y produce high y un form ref ned Kraft stock with the required strength characteristics The control strategy nvolved in ach ev ng these objectives inc udes the following

1 The Kraft Surge Chest consistency (KIC002) is control ed to an operator setpo nt by man pu ating the flow of d lution wh tewater to the pump suction

2 The Refiner inlet pressure (PIC003) is controlled to setpoint by manipulating the ma n stock 'ine pressure valve (0 to 50% A/O) and the rec rcu'ated flow contro va ve (50 to 100% A/C). During norma operation the controller output is above 50% and is manipulat ng the recirculation flow va ve

3 The stock flow through the refiner (FIC004) is contro led to a setpoint cascaded from the Ref ned Chest level control er The flow control valve s pos tioned downstream of the refiner

4 A specific energy control er (HPD/Ton) manipulates Refiner p ate position in order to maintain the specific energy at an operator entered setpoint The control er PV is ca culated from the Ref ner KW load. the Stock Flow and Stock Consistency us ng the following formula

$$HPD/Ton = (k * KW)/(Flow * Cons)$$

5 The Refined Stock consistency (KIC007) s contro led to setpo nt by man pu ating the d lut on valve to the Ref ned Kraft pump suction In this example the di ution whitewater for both cons stency oops s supp ed 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 avai able pressure drop across the flow control valve This wil also disturb the Refiner Inlet pressure. Adjusting the Pressure controller output wil 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 oad disturbance to consistency

* Increasing the stock flow controller output will increase the stock flow from the Chest and ncrease the cons stency
* Increasing the consistency controller output will decrease the consistency and decrease fr ctional losses, acting to ncrease both the flow and pressure It is common that the flow to consistency coup ing is much stronger than the cons stency to flow/pressure coupling
* Increasing the Pressure Valve position will increase the stock flow through the pump, which wil increase the consistency over the short term In this process design the recircu ated stock flow change w l reduce the chest consistency, bringing the discharge consistency back to or ginal value

4 *Kraft Surge Chest (KIC002) and Kraft Refined Chest (KIC007) Controllers*

These are coupled via a common whitewater header An increase n the Refined Chest consistency va ve will decrease the Header Pressure, wh ch decreases the dilut on flow to the Surge Chest consistency thereby decreas ng the Surge Chest Consistency

Step 1 Bump Tests to Determine Coupling Dynamics

The first phase of tuning must focus on developing an understanding of the process dynam cs and coupling dynamics in the system The follow ng tasks should be performed in the fo lowing order

* Al interactive controllers are placed into manual mode
* Bump tests are performed sequential y on interact ve controllers
* Process Dynamics and Coupling Dynam cs are defined The coupling ga ns are particularly important s nce this wil assist in defining the need for decoupling via tuning The Step 1 results provide guidance regarding the relative need for decoupling, constraints on Lambda se ection condition of the contro equipment etc

Summarizing the results into a spreadsheet of the form shown be ow s a useful method of appreciating the nature of system interact ons In the Refining example, this wou d be a 4 by 4 matrix

NPUTS / OUTPUTS	**1. Refined Stock Flow Cont. Output (FC004)**	**2. Refiner Inlet Pr. Cont. Output (PC003)**	**3. Specific Energy Cont. Output (HPDT)**	**4. Surge Chest Cons. Cont. Output (KC002)**
1. Refined Stock Flow (FC004)	FC004 K_{11} τ_{11} T_{11}	PC003/FC004 K_{12} τ_{12} T_{12}	HPDT/FC004 K_{13} τ_{13} T_{13}	KC002/FC004 K_{14} τ_{14} T_{14}
2. Refiner Inlet Pr. (PC003)	FC004/PC003 K_{21} τ_{21} T_{21} –	PC003 K_{22} τ_{22} T_{22}	HPDT/PC003 K_{23} τ_{23} T_{23}	KC002/PC003 K_{24} τ_{24} T_{24}
3. Specific Energy (HPDT)	FC004/HPDT K_{31} τ_{31} T_{31}	PC003/HPDT K_{32} – τ_{32} T_{32}	HPDT K_{33} τ_{33} T_{33}	KC002/HPDT K_{34} τ_{34} T_{34}
4. Surge Chest Cons. (KC002)	FC004/KC002 K_{41} τ_{41} T_{41} –	PC003/KC002 K_{42} τ_{42} T_{42} –	HPDT/KIC002 K_{43} τ_{43} = T_{43}	KC002 K_{44} τ_{44} T_{44}

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

1 Minimize freeness variability in the Refined Kraft Chest

2 Minimize consistency variability in the Refined Kraft Stock.

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 it's 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 smal and decoupl ng is therefore not a requ rement

The interaction between the two consistency loops may be significant however S nce the disturbances from the upstream chest are likely to be more severe, and we want to m n mize 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 fol owing 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 regard ng process time constant and process deadt me

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 Spec fic Energy control er at least 5 times s ower than the flow oop.

Refined Kraft Chest consistency loop - Tune the level loop to be at least 5 t mes 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 oop will be the fastest in the system Using the bump test results from Step 1, tune th s controller as fast as possible, consistent with normal Lambda select on constraints Tune the controller and place into Auto mode

2 Tune the Kraft Surge Chest consistency controller as fast as possible, consistent with the normal Lambda selection constraints Place the contro ler 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 w th the Pressure controller in Auto mode. The Flow wi l reach steady state after 4 Pressure controller Lambda values (the time required to bring the pressure back to setpo nt fo lowing the Flow controller output bump The process gain of the Flow loop wi l 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 control er 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 contr l er to be at least 5 times slower than the Flow controller.

6 Tune the Refined Kraft Chest Leve controller to be *as slow as possible* consistent with maintaining the level within an acceptable operating range The leve controller Lambda should be at least 5 times s ower than the Refined Stock Flow controller

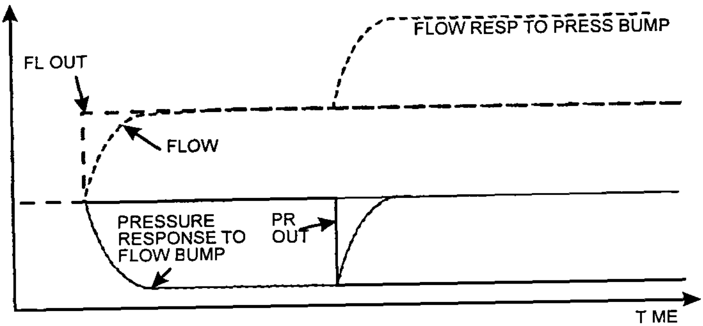
Step 4 Validate the Tuning

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 Ref ned Kraft Level should be made. The upstream controlled variables should maintain setpoint reasonab y well during the transitional period

EXAMPLE FAST PRESSURE

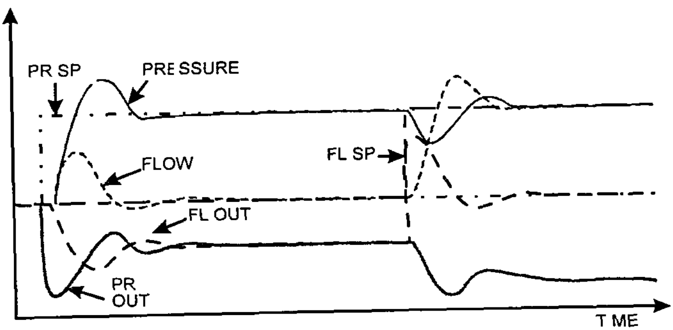
REFINER FLOW AND PRESSURE INTERACTION

INTERACTIVE LOOPS - FLOW AND PRESSURE IN MANUAL



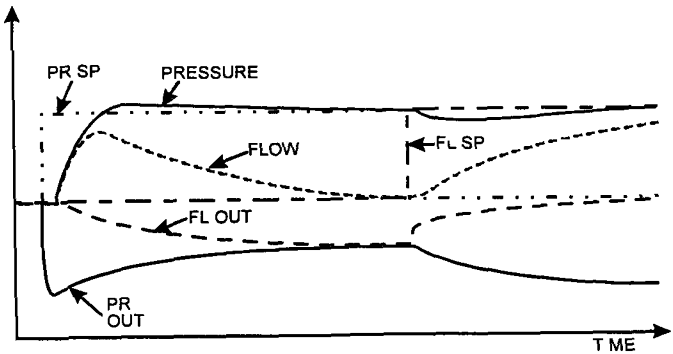
INTERACTIVE LOOPS - FLOW AND PRESSURE

LAMBDA OF PRESSURE CONTROLLER = LAMBDA OF FLOW CONTROLLE



INTERACTIVE LOOPS - FLOW AND PRESSURE

PRESSURE CONTROLLER 10 TIMES FASTER THAN FLOW CONTROLLE



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11. TUNING LEVEL CONTROLLERS FOR LOAD RECOVERY

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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 se f regulating processes

The most common integrating processes are LEVEL controllers Consider the basic LEVEL CONTROL LOOP

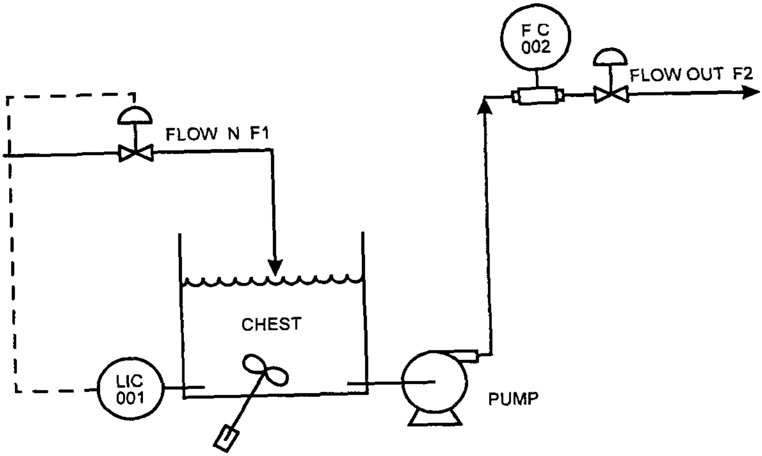
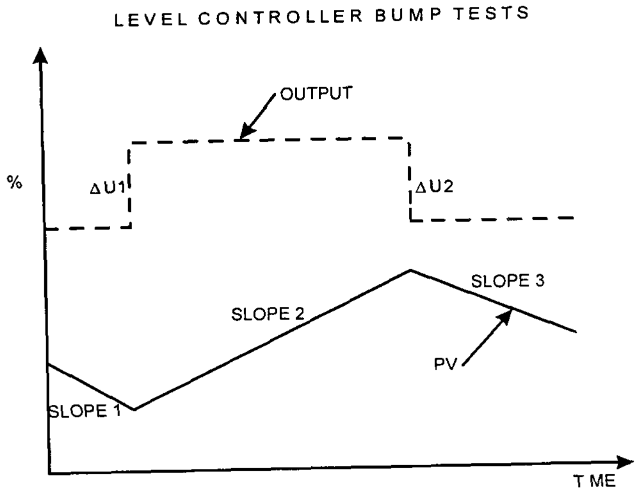


Fig 11-1 Basic Level Contro Loop

Bump Tests

Open Loop Bump tests on a Level controller look like the fol owing Two different methods can be used to ca culate the process ga n the slope method (F g 11-2) and the graph cal method (Fig 11-3) These two methods w ll g ve the same answer



F g 11 2 Leve Contro Bump Tests (Slope Method for Kp)

SLOPE METHOD

$$Kp_1 = \frac{SLOPE2\ SLOPE1}{\Delta U1}$$

$$Kp_2 = \frac{SLOPE3 - SLOPE2}{\Delta U2}$$

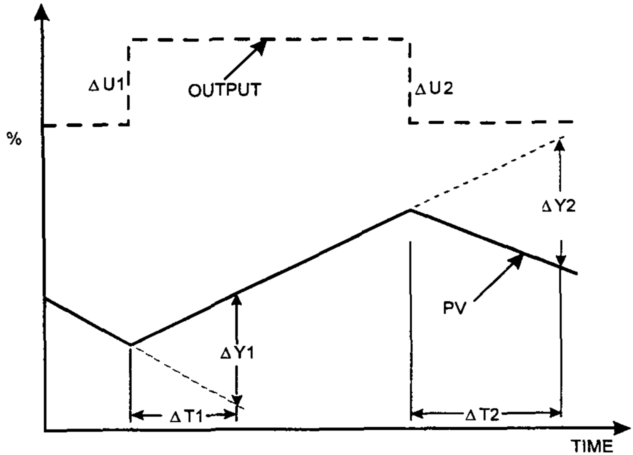


Fig. 11-3 Level Control Bump Tests (Graphical Method for Kp)

GRAPHICAL METHOD

$$Kp_1 = \frac{\Delta Y1 / \Delta T1}{\Delta U1}$$

and,

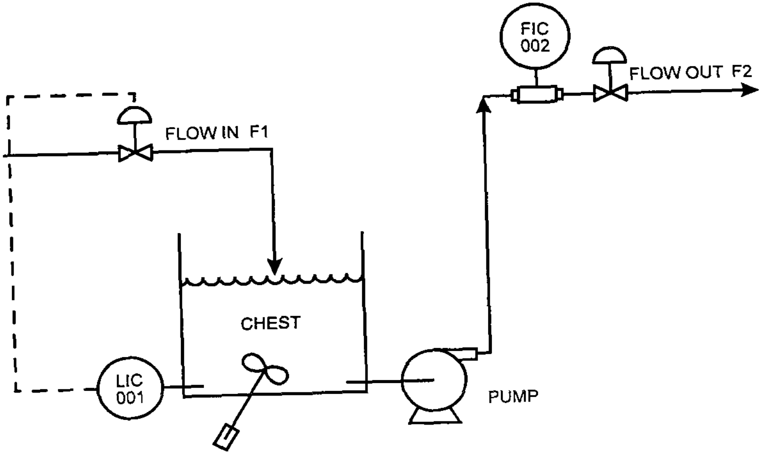
$$Kp_2 = \frac{\Delta Y2 / \Delta T2}{\Delta U2}$$

Differences in the values of Kp1 and Kp2 (or any other subsequent bump tests) might indicate some valve problems or other nonlinearities

Proportional Only Controller

Since the process be ng controlled s an integrat ng process one could argue that no integra act on (the "I part in a PI controller) s requ red

A proport ona only controller moves the output by an amount proport onal to the nput error s gna Consider the fo lowing proport onal on y level control loop be ow



F g 11-4 Leve Contro Tun ng

Proportional Controller Response To A Setpoint Change

A Assuming a steady state condition (i e. flow-in F1 is equal to flow-out F2) 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 n (F1) to ncrease, and F1 exceeds F2 The level wi l ncrease

C As the level increases, the error decreases, and the proport ona term is reduced The output to the contro ler moves back towards ts initial value, reducing the flow-in When the leve reaches the new setpoint, the flow-in is equa to the flow out

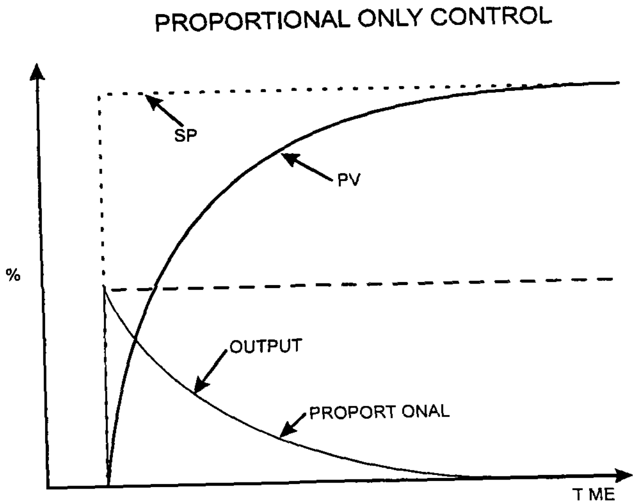


Fig 11 5 Level Controller - Proportional Only Response to Setpoint Bump

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 λ determine K_P
- $K_C \quad \frac{1}{\lambda K_P}$

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 (λ much faster than τ_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 in, the level will drop As the level decreases the error increases, and the controller output moves in proportion to the error

A block diagram representing the load disturbance is shown below

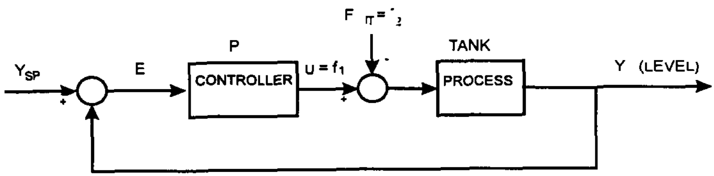


Fig 11-6 Block diagram for level control

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.

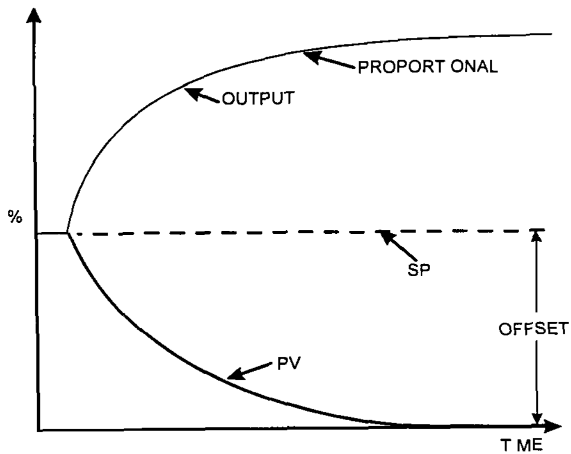


Fig 11-7 Level Controller - Proportional Only Response to Load Disturbance

CONCLUSION	RESET IS NEEDED TO REMOVE THE OFFSET

Tuning For Load Recovery

A block diagram representing level control using a PI controller is shown below

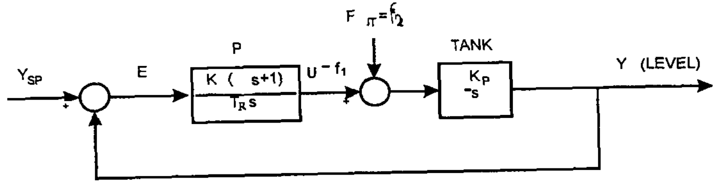


Fig 11-8 Block diagram for level control

The closed loop transfer function from the disturbance $D = F_{out} = f_2$ to the level Y using a PI controller is

$$G_{LD}(s) = \frac{G_P(s)}{1 + G_C(s)\,G_P(s)} = \frac{\dfrac{K_P}{s}}{1 + \dfrac{K_C K_P (T_R s + 1)}{T_R s^2}} = \frac{\dfrac{T_R}{K_C}\,s}{\dfrac{T_R}{K_C K_P}\,s^2 + T_R s + 1}$$

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

$$\boxed{\begin{aligned} T_R &= 2\lambda \\ K_C &= \frac{1}{K_P}\,\frac{2}{\lambda} \end{aligned}}$$

CRITICALLY DAMPED LEVEL TUNING RULE

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

$$\frac{T_R}{K_C K_P} = \frac{2\lambda}{\left(\frac{2}{K_P\lambda}\right)\cdot K_P} = \lambda^2$$

Using the above expressions, the closed loop transfer function from the d sturbance $D - F_{\text{out}} - f_2$ to the level Y becomes.

$$G_{LD}(s) - \frac{-\frac{T_R}{K_C}s}{\frac{T_R}{K_C K_P}s^2 + T_R s + 1}$$

$$- \frac{-K_P\lambda^2 s}{\lambda^2 + 2\lambda + 1}$$

$$- \frac{-K_P\lambda^2 s}{(\lambda s + 1)^2}$$

Thus the above tun'ng rule does in fact resu t n a closed loop response wh ch s critically damped with time constant λ, wh ch we are free to select

By substitut ng $\lambda - T_R/2$ in the equation for the control er ga n, the tun ng rule can be written alternatively as

$$\boxed{\begin{array}{l} T_R - 2\lambda \\[1ex] K_C - \frac{4}{K_P T_R} \end{array}}$$

CRITICALLY DAMPED LEVEL TUNING RULE

Lambda Interpretation in Level Controller Tuning

The question that remains is "What is the meaning of λ in the above tuning rule and how should we choose λ?"

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 level controller load recovery is shown in the following figure

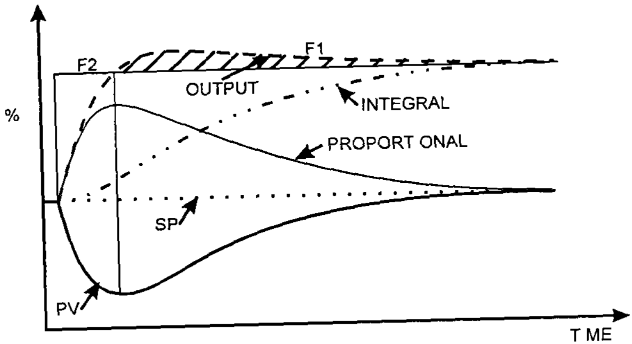


Fig 11-9 Level Control Response When Tuned For Load Recovery

In λ, the level change will be arrested and the flow in f_1, will equal the flow out f_2. In 6λ, 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 it 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 (rising) after a load step change as the arrest time T_{ARR}. λ is identical to T_{ARR}, and thus

$$\lambda = T_{ARR} = \frac{T_{REC}}{6}$$

Thus the tuning rule can be expressed as

$$\boxed{\begin{array}{l} T_R = 2T_{ARR} = T_{REC}/3 \\ K_C = \dfrac{4}{K_P \, T_R} \end{array}}$$

Level Controller Setpoint Response

The level response to a setpoint change is as follows:

PI CONTROL - SETPOINT RESPONSE

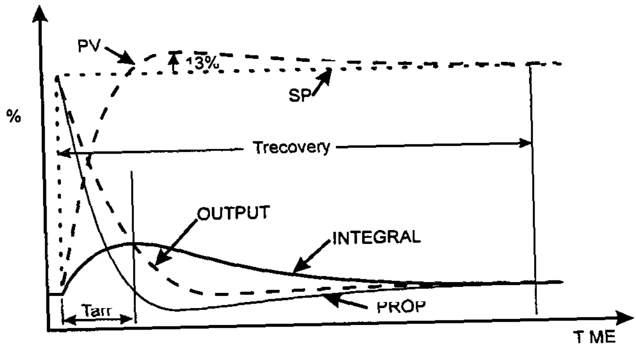


Fig 11-10 PI Controller Response to a Setpoint Bump

The important points to note are

1. The level first reaches the new setpoint at the Arrest Time (~ λ)
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 (6λ)

Tuning Rules For Integrating Plus Deadtime Processes

To tune an integrating plus deadtime process, the tuning rules are modified as shown below

$$T_R - 2\lambda + T_d$$

$$K_C - \frac{T_R}{K_P(\lambda + T_d)^2}$$

CRITICALLY DAMPED LEVEL WITH DEADTIME TUNING RULE

EXAMPLE:

The level process gain, Kp from a series of bump tests was 0 2 $\frac{\%Span/Min}{\%Output}$ The deadtime was 10 seconds or 0 167 minutes

The desired LAMBDA value (ARREST) time for this level control er is 5 minutes

The *Reset Time* is ca culated as follows

$$T_R = 2\lambda + T_d - \; 2\times 5 + 0167 \quad 1016 \text{ minutes}$$

The *Controller gain* is calculated as follows

$$K_C \quad \frac{T_R}{K_P(\lambda + T_d)^2} - \frac{1016}{0.2(5+016)^2} - \frac{1016}{0.2(5.16)(5.16)} \quad 19\ \%Output/\%Span$$

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.

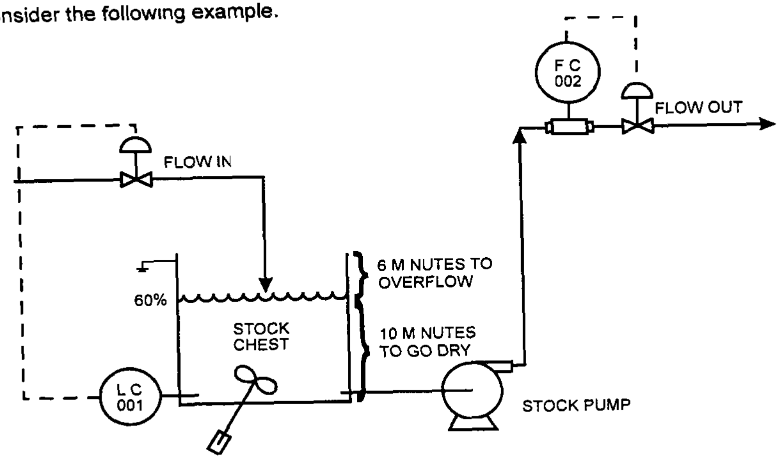


Fig 11-11 Operating conditions of a level loop

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

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.

λ	=	6 Minutes
	OR	
T_{Rec}	=	36 Minutes

The process gain (K_P) is determined from the bump tests. Knowing Kp and λ 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) wou d not always be chosen This would ensure that the controller will respond very aggressively to d sturbances, thereby maintaining the evel 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

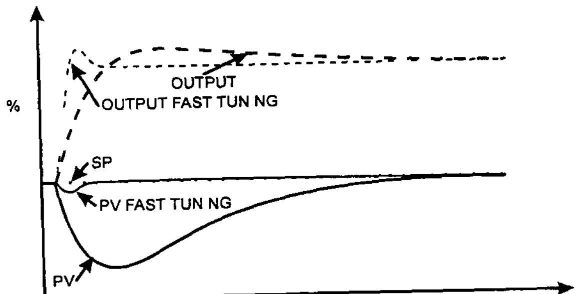


Fig 11-12 Fast Versus Slow Arrest Time

There are two important po nts to observe

1 With the fast tun ng, the level control er output increases very quickly, wh ch means that the flow rate into the chest increases very rapidly With the relative y slow tuning, the flow rate changes much more gradually The impact of the fast flow change on upstream process conditions (consistency freeness etc) wil be much more severe than with the more gradual change The upstream control ers wi l obviously have a greater abi ity to eliminate the impact of a gradual flow change versus a fast flow change

2 W.th the fast tuning the load disturbance s arrested quickly and the level decrease is small The fast tuning w ll hold a much tighter setpo nt 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

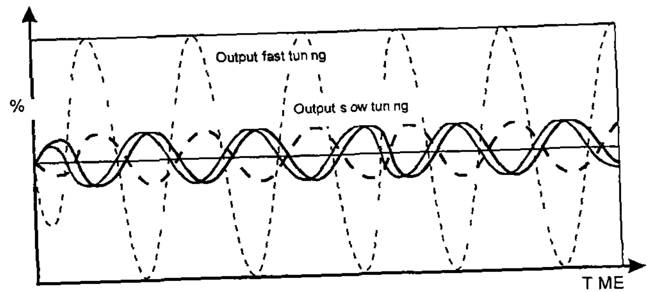


Fig 11-13 Impact of Fast Variability on Level Controller Output

Dampening can be used to attenuate the noise present while not significantly affecting the controller's performance The Filter Time Constant τ_f should be at least 10 times smaller than λ.

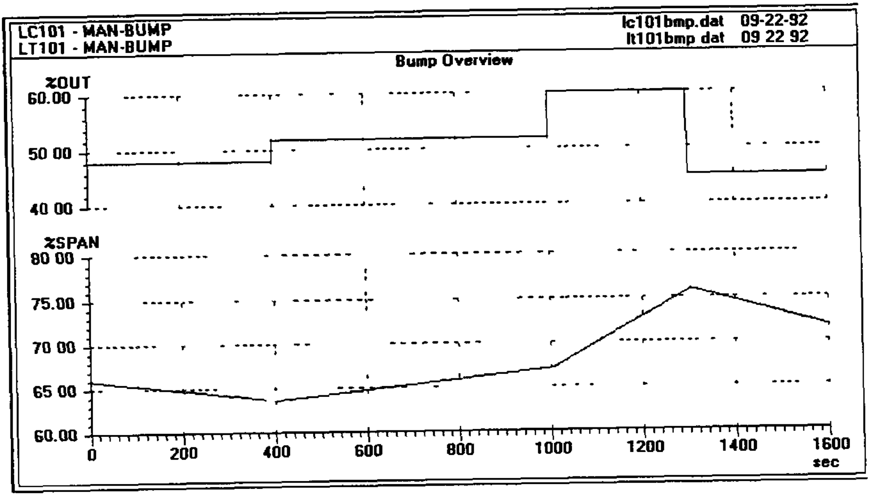
$$\tau_f < \lambda/10$$

Note that the faster the LAMBDA the faster the filter should be Excessive use of filtering can result in level cycling

Examples / Questionnaires
LEVEL CONTROLLER TUNING

Example 11-1

Calcu ate the process gain for LC101 in units of $\frac{\%Span / Min}{\%Output}$ Please note that th s is s mulated data and there are no nonlinearities.



Checklist

1 The Process Ga n for a Level control er can be ca culated us ng the ------------- method or the ------------ method

2 Using the ---------- method, the process gain is $\frac{\Delta Y \ / \Delta T}{\Delta U}$

3 Us ng the ----------- method, the process gain is $\frac{SLOPE2\text{-}SLOPE1}{\Delta U1}$

Example 11-2

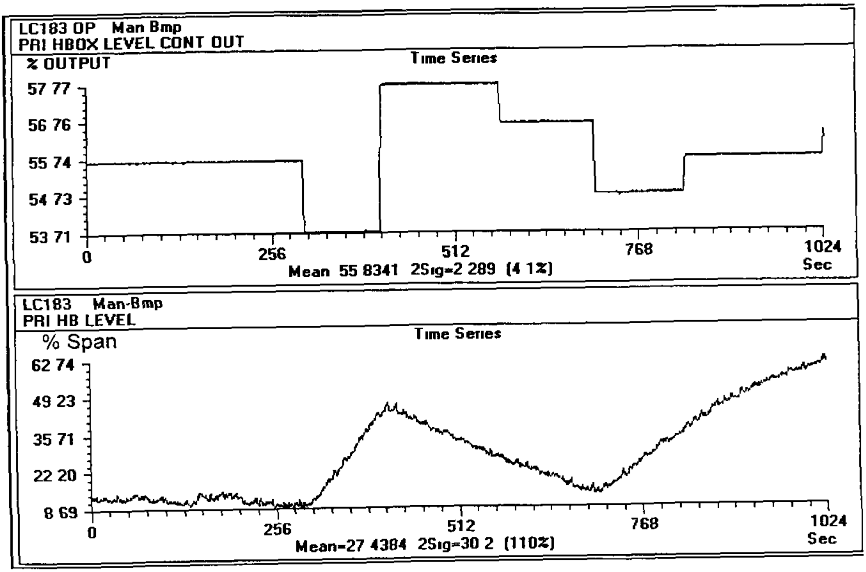
The bump tests shown below were taken from an a r padded headbox The level controller man pulates the aır supply valve to the aır pad n order to ma ntaın the level at setpoınt

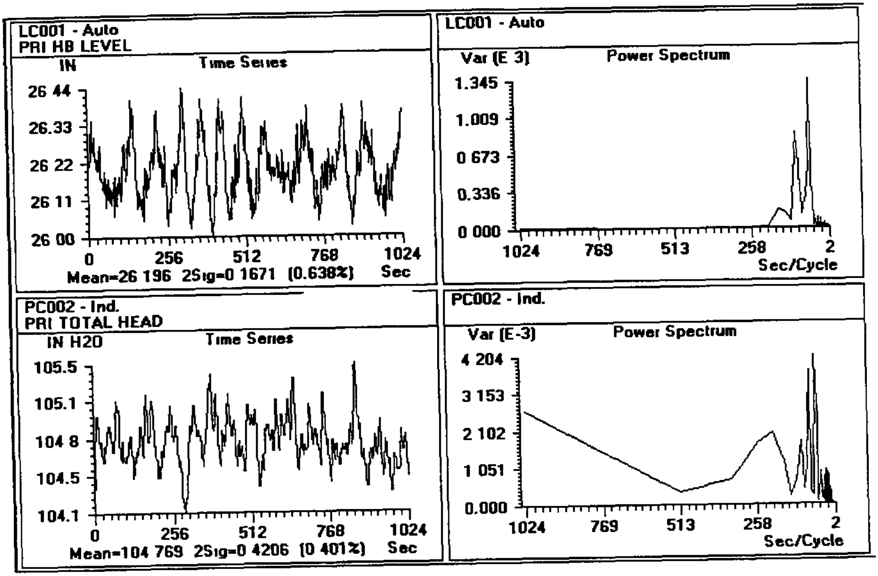
Questions

Q1 Calculate the process gaın for thıs evel controller

Q2. Can you ıdent fy any nonlınearıtıes?

Q3 Can you exp a n the common cycle n the Headbox Level and Total Head?





Checklist

1. Have enough bump tests been performed to properly evaluate backlash and stiction in the control valve

2. For a level 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

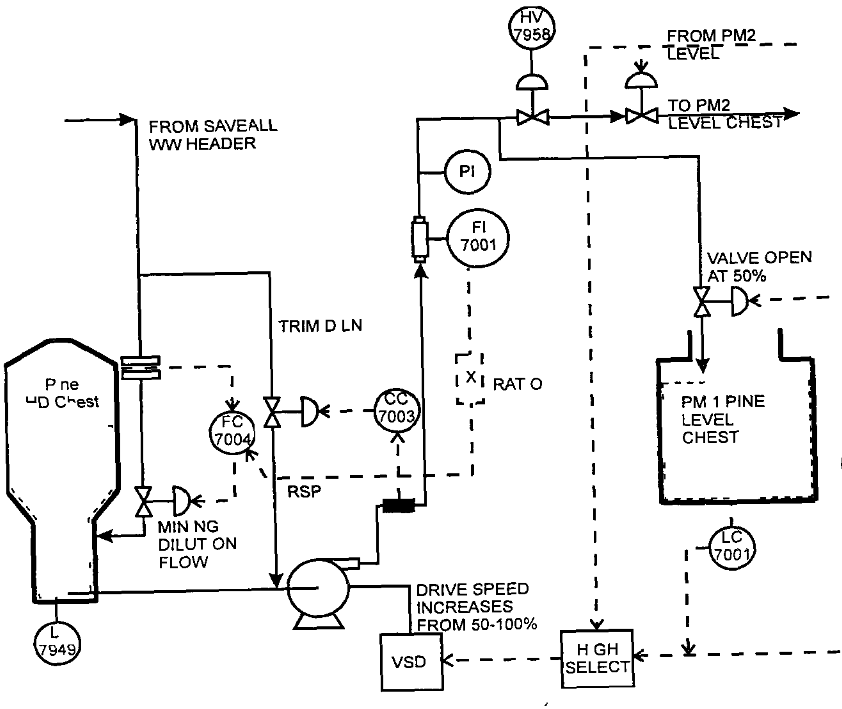
The simplified P&ID shows a pine High Density (HD) and level chest process and control system The time series data shows the pine chest level, HD flow, HD consistency and coarse dilution flow

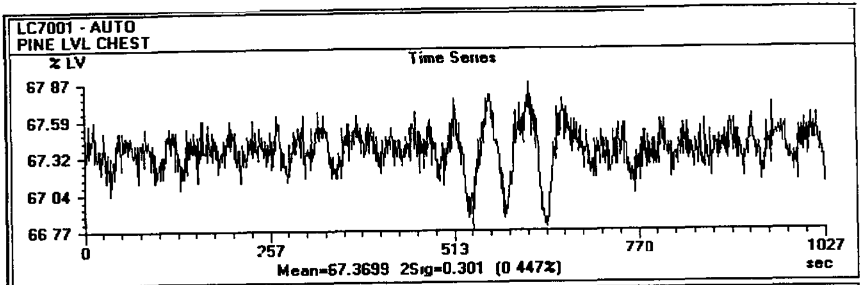
Questions:

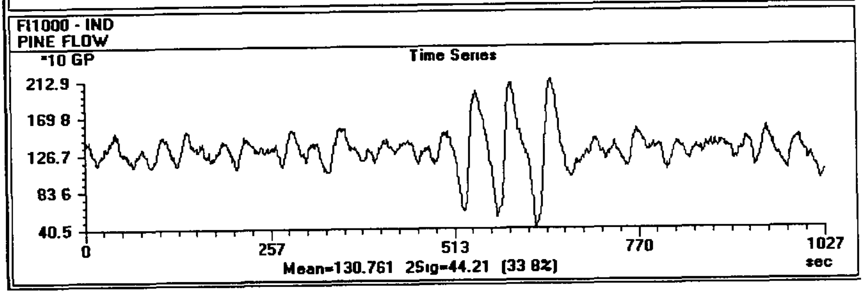
1 How important is the variability in the pine chest level

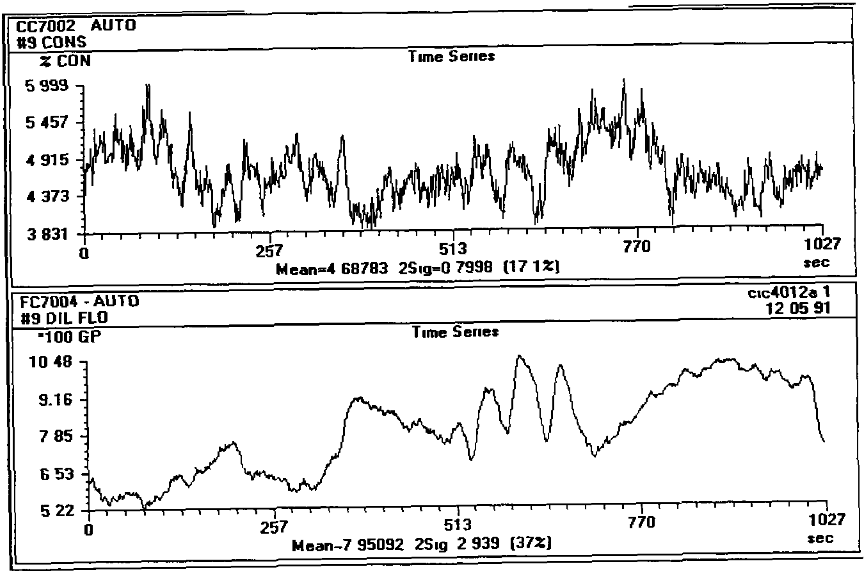
2 What might be causing the 60 second cycle in the pine chest level ?

3 Comment on the control strategy Can you suggest any improvements?




LC7001 - AUTO
PINE LVL CHEST
% LV
Time Series
67.87
67.59
67.32
67.04
66.77
0
257
513
770
1027
sec
Mean=67.3699 2Sig=0.301 (0.447%)


FI1000 - IND
PINE FLOW
*10 GP
Time Series
212.9
169.8
126.7
83.6
40.5
0
257
513
770
1027
sec
Mean=130.761 2Sig=44.21 (33.8%)


CC7002 AUTO
#9 CONS
% CON
Time Series
5 999
5 457
4 915
4 373
3 831
0
257
513
770
1027
sec
Mean=4 68783 2Sig=0 7998 (17 1%)
FC7004 - AUTO
#9 DIL FLO
cic4012a 1
12 05 91
*100 GP
Time Series
10 48
9.16
7 85
6 53
5 22
0
257
513
770
1027
sec
Mean-7 95092 2Sig 2 939 (37%)

APPENDIX 11A

SAMPLE BUMP TESTS

23 36 05 04 10 07

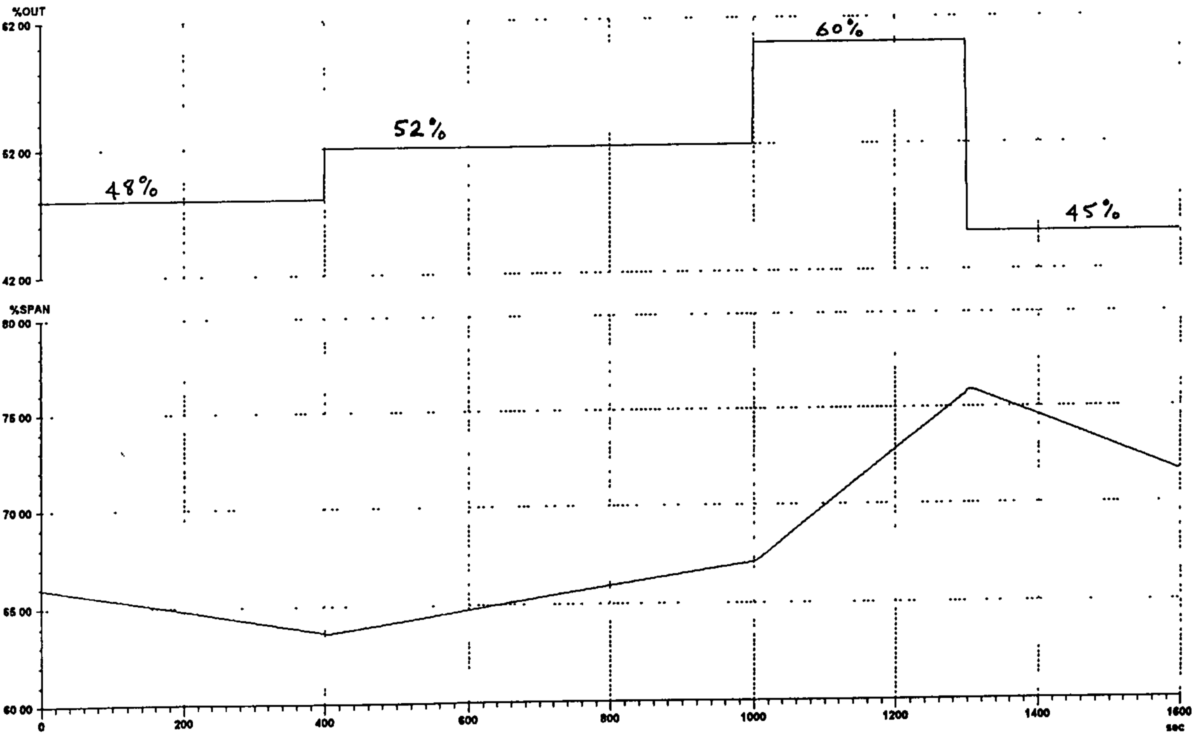
23 36 07 04 10 07

EnTech Control Engineering Inc. | ENTECH TORONTO

LC101 - MAN-BUMP
LT101 - MAN-BUMP

lc101bmp dat 09-22-92
lt101bmp dat 09-22-92

Bump Overview



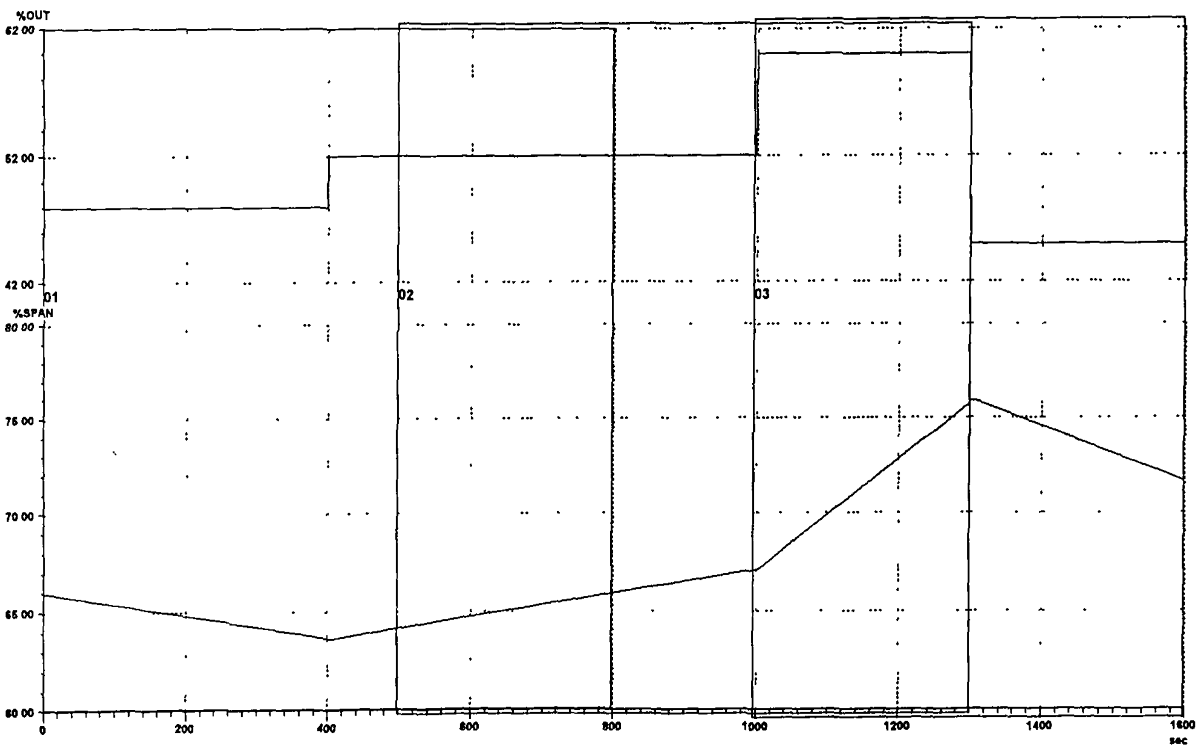
EnTech Control Engineering Inc.

ENTECH TORONTO

LC101 - MAN-BUMP
LT101 - MAN-BUMP

lc101bmp dat 09-22 92
lt101bmp.dat 09-22-92

Bump Overview

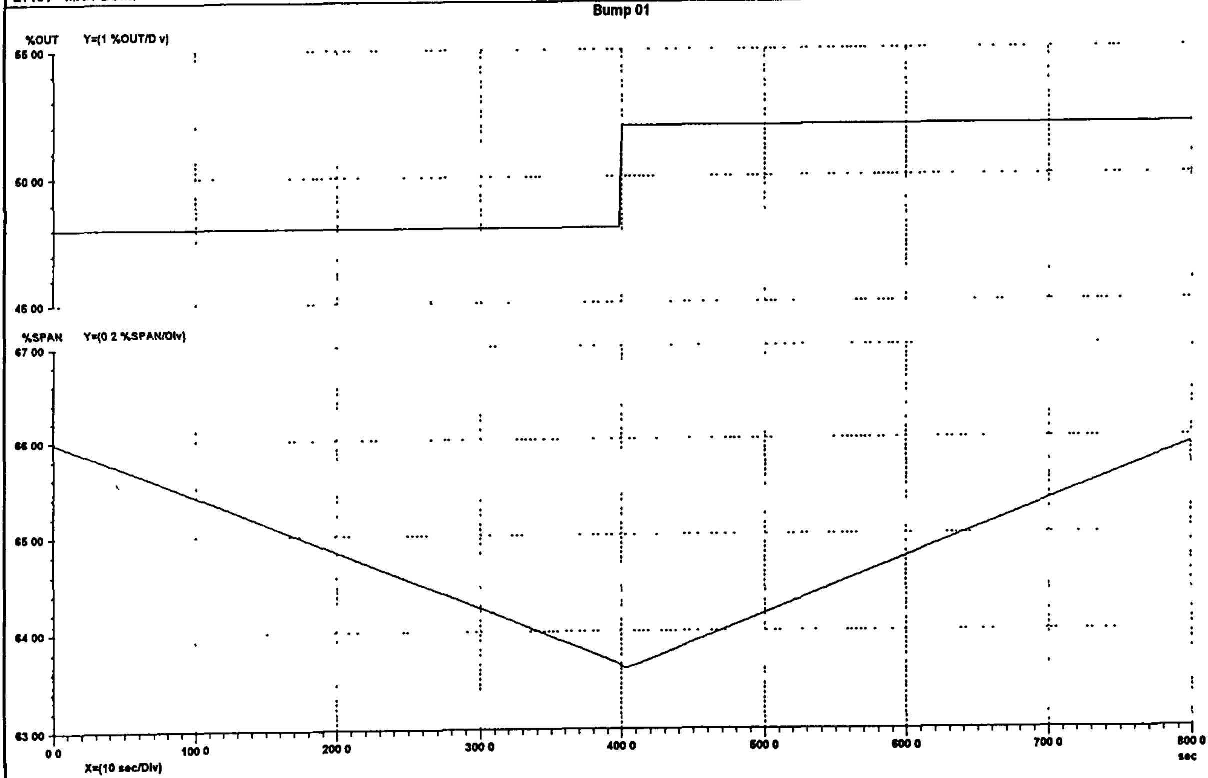


EnTech Control Engineering Inc.

ENTECH TORONTO

LC101 - MAN-BUMP
LT101 - MAN-BUMP

lc101bmp.dat 09-22-92
lt101bmp.dat 09-22-92

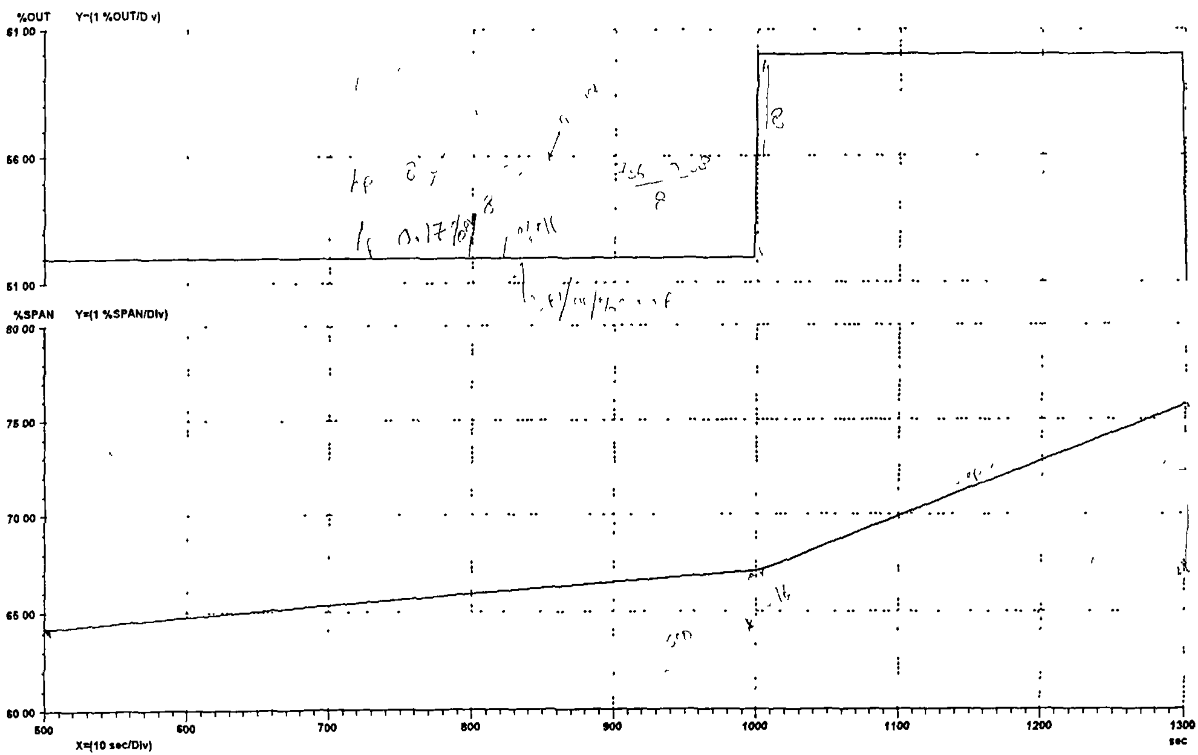


23 36 09 04 10 07

EnTech Control Engineering Inc. ENTECH TORONTO

LC101 - MAN BUMP lc101bmp.dat 09-22-92
LT101 - MAN BUMP lt101bmp.dat 09-22-92

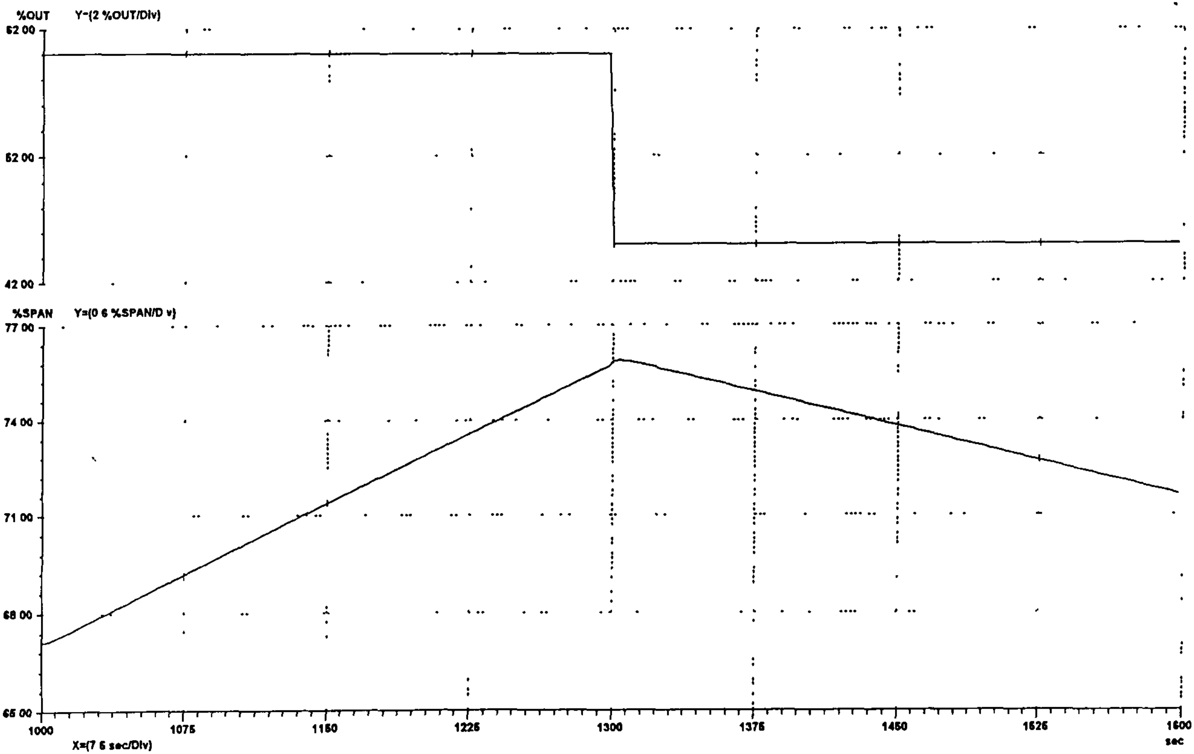
Bump 02



EnTech Control Engineering Inc. ENTECH TORONTO

LC101 - MAN-BUMP lc101bmp.dat 09-22-92
LT101 - MAN-BUMP lt101bmp.dat 09-22-92

Bump 03



23 36 10 04 10 07

12. TIME SERIES ANALYSIS

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 it's 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 σ^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 σ, is probably the most widely used statistic and is just the square root of the variance, namely

$$\sigma = \sqrt{\mathrm{var}[x_k]}$$

If the data series is normally distributed (seldom the case in continuous process industries) approximately 95 % of all the data points will fall within +/- 2 σ of the Mean

4 The variability is equal to 2 * σ 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

$$\% \text{ Variability} = \frac{2 * \sigma * 100}{\text{MEAN}}$$

Power Spectrum

The power spectrum describes the frequency content n a data series by decompos ng the s gnal nto many sine waves, known as harmon cs It is part cular y useful because t ass sts in ident fying any sign ficant cycles in the process This s important for a number of reasons

1 It wil assist in diagnosing the source of product cyc es A product cycle at a g ven frequency is caused by a process cycle at the same frequency Thus t assists in linking process disturbances with product disturbances Ultimately, this is an important tool in reduc ng product var abil ty

2 It will ass st in analyzing control loop performance

3 It w l prov de ns ght nto control or process design modifications which m ght aid n reducing process var abi ity

The plot of variance versus frequency is known as the 'Power Spectrum plot The power spectrum nd cates how much variance ex sts at a part cular frequency or per od The sum of the indiv dual var ances at each frequency is equa to the total var ance in the data ser.es

The power spectrum analysis has a number of properties which are very useful in character zing a data series These are described n 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. $f = \dfrac{1}{T_1}$ 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 $f_N = \dfrac{1}{2T_s}$

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

$$f_m = m \cdot f_1 = \frac{m}{NT_s}$$

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:

Cons der Tımes Serıes data co lected with $T_s \quad 0.2$ sec and $N = 1024$ samp es

Fundamental Penod

$$T_1 \quad 1024 \times 0.2 \quad 204\ 8 \text{ sec / cycle}$$

Fundamental Frequency

$$f_1 \quad \frac{1}{204.8} \simeq 4\ 88 \ \times\ 10^{\ 3} \text{ Hz}$$

Nyquıst Penod

$$T_N \quad 2T_s - 0.4 \text{ sec / cycle}$$

Nyquıst Frequency

$$f_N \quad \frac{1}{2}(f_s) \quad \frac{1}{2}\left(\frac{1}{0.2}\right) \quad 2.5 \text{ Hz}$$

Harmonıcs present

There are $N/2 \quad 1024/2 \quad 512$ harmonıcs present g ven by

$$f_k \quad kf_1 \quad \frac{k}{204\ 8} \text{ cycles / sec}$$

$$T_k \quad T_1/k \quad 204.8/k \text{ sec}$$

The perıod and frequency of the f rst few harmon cs are shown ın the follow ng table

Harmon c k	Perıod T_k sec	Frequency f_k Hz
1	204 8	0 00488
2	102 4	0 00977
3	68 3	0 01465
4	51 2	0 01953
512	0 4	2 5

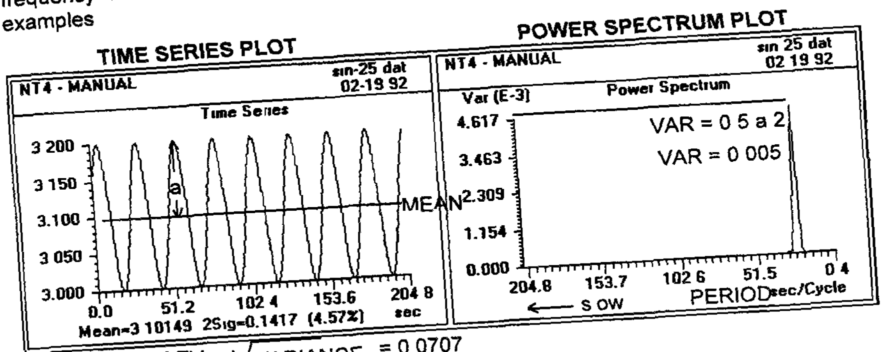
Notıce that the frequencıes are equa ly spaced

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



$$\text{STD DEV} = \sqrt{\text{VARIANCE}} = 0.0707$$

$$\text{VARIABILITY} = 2 * \text{STD DEV} = 0.141\ \%$$

Fig 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

$$\text{Variability \% of mean} = \frac{0.141}{3.1} * 100 = 4.57\ \%\ \text{of mean}$$

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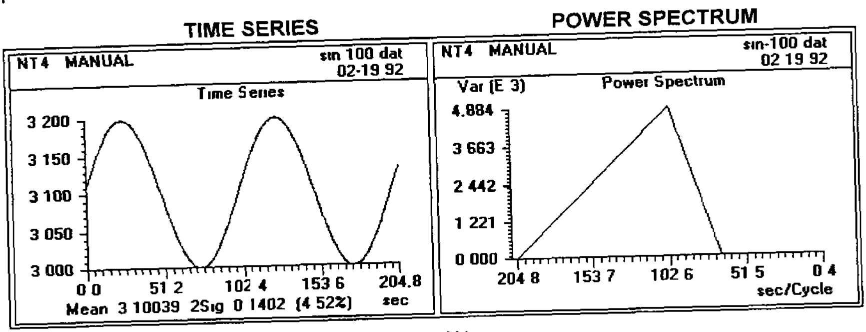
The fo lowing example again shows a cons stency s gnal containirg a s ng e cycle The per od of the cycle in th s example is 100 seconds



F g 12-2 Power Spectrum Plot of Single Sine Wave

This example illustrates that the variability is independent of the period of the cyc e In this example, the amp itude of the cycle is again 0 1 % consistency. The power spectrum ndicates that there is a sing e cycle at a per od of 100 seconds/cyc e and that the variance contributed by this cycle is approximate y 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.

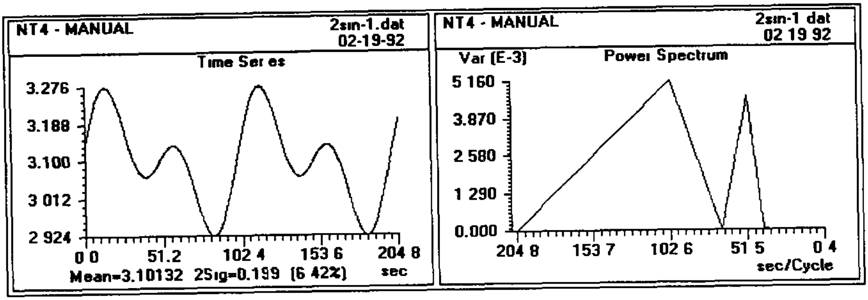


Fig. 12-3 Sum of two Sine Waves - Equal Magnitude

It is still possible to 'eyeball' the period of the two cycles, although more difficult to judge their 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 * σ or 0.2 % Consistency. Expressed as a percentage of the mean the variability is equal to $\frac{0.2}{3.1}$*100 or 6.47 % of mean. These expected values are in fact indicated on the time series plot above.

The usefulness of the power spectrum becomes increasingly evident as the complexity of the data ser es increases Consider the following process s gnal composed of three s ne wave cycles The amplitude and period of these cyc es are as fo lows

	PERIOD	AMPLITUDE	VARIANCE	VARIANCE,%
CYCLE 1	100	0 05	0 0013	7 1
CYCLE	50	0 10	0 0050	28 6
CYCLE 3	25	0 15	0 0113	64 3
TOTAL			0 0175	100

In this case each cyc e dces not contr bute equa ly to the overal variance The unequal peaks in the power spectrum indicate the relative contributions and are consistent with our expectations

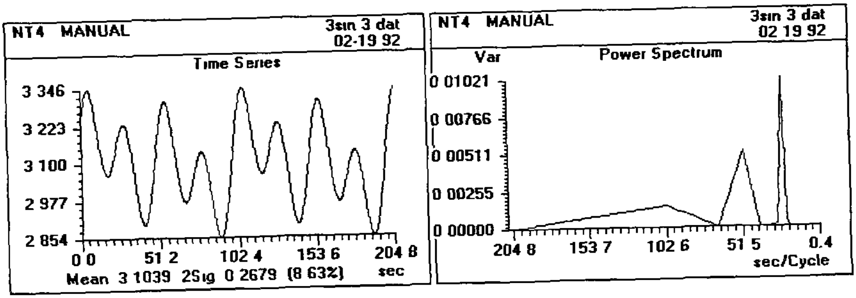


Fig 12-4 Power Spectrum - Sum of 3 Sine Waves Unequa Ampl tudes

It would clearly be d ff cu t to accurate y judge the per od and the ampl tude of the three cycles in the process signal in th s examp e Yet, th s is a re ative y simple signal Most process measurements are far more comp ex and are increasing y d ff cu t 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 n the process signal to the total variance.

For example, in the previous example the percentage contr bution of the 100 50 and 25 second cycles were 7 2, 28 6 and 64 3 % respective y The cumulative spectrum plot for this process signa , is shown below·

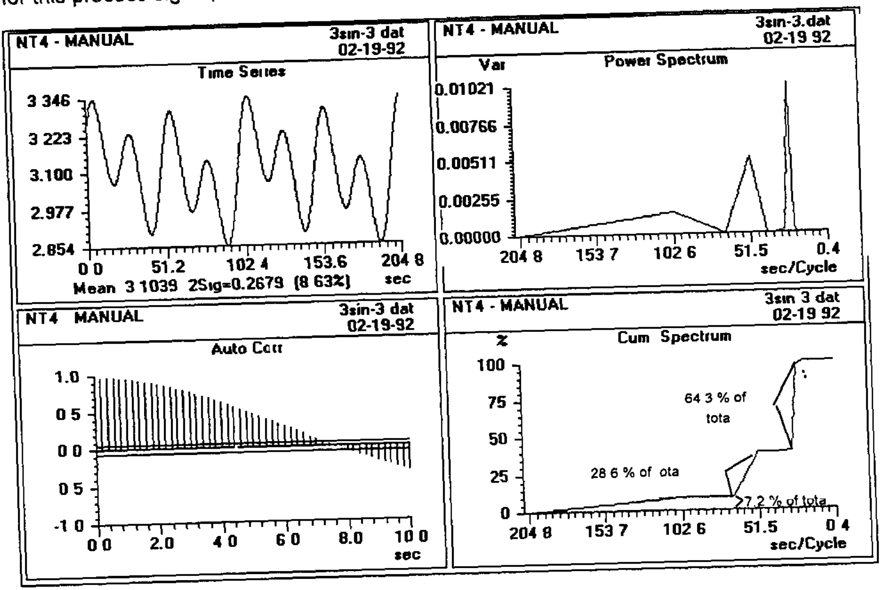


Fig 12-5 Cumulative Spectrum

The cumu at ve spectrum plot ind cates 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 var ance in the process signal is slower than 25 seconds/cycle

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White Noise

A process signal that contains purely random variabil ty s termed WHITE NOISE Such a process signal is shown below

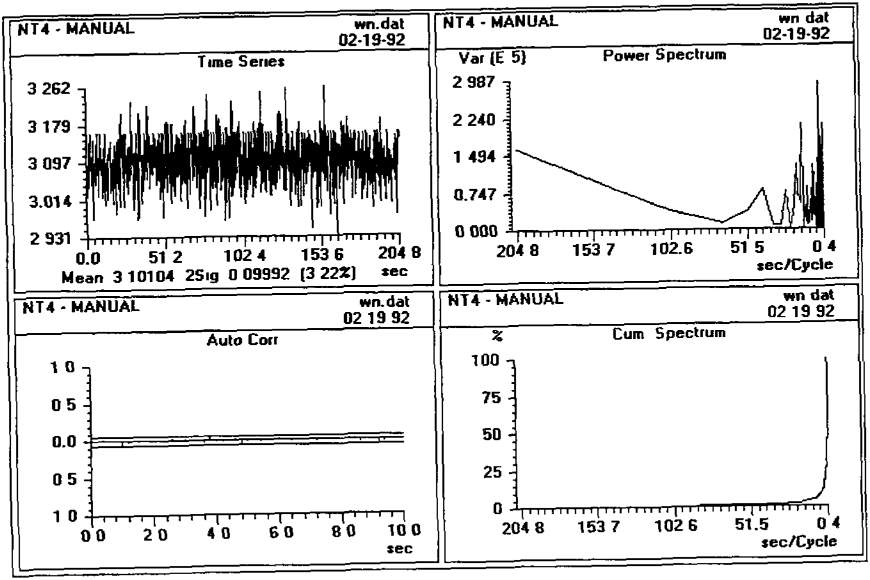


Fig 12 6 Time Ser es Plot of White No se

Since the white noise signa is pure y random, the variance is d v ded equa ly among a l frequencies included in the power spectrum Thus, the white no se s gnal conta ns slow variability as we l as fast variab lity (equa variance at all ana yzed frequencies) Note that the magnitude of the average var ance present in each harmonic s very smal (about 0.5×10^{5}) since the total variance is d vided equa ly n th s case nto 512 frequency components

Autocorrelation

The Autocorrelation plot ıs considered the best indicator of the randomness ın a t me series If the data in a process sıgnal is completely random (whıte no se) then all of the vertıcal lag ınes ın the autocorrelatıon plot wıll fall withın the horızontal confidence limıts

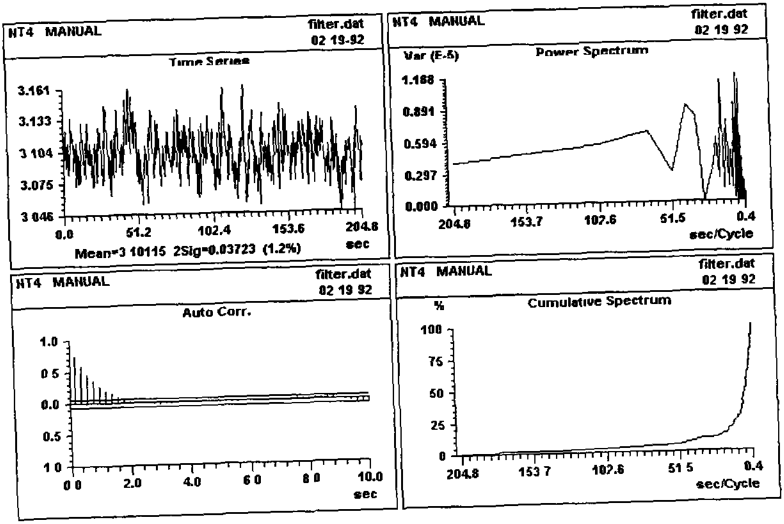
White Noise ıs often associated with measurement noıse The process sıgnal can be thought of as the combınation of the true process value and the measurement noise This combıned signal ıs the only process measurement ava lable for control purposes However it ıs clear that taking corrective action to compensate for measurement noise wı l ıncrease the true process varıabı ity

The concept of mınımum varıance contro states that the controlle should be tuned to el mınate only true process varıabıl ty Once a whıte noıse sıgna ıs ach eved, no further control improvement can be made The autocorrelatıon funct on can be used to assess whether mınımum varıar ce control ıs beıng approached Subject to certa n assumptıons M nımum Varıance Control wıl be achıeved when the autocorrelat on functıon decays to zero (falls wıthın the confidence lımıts) wıth n one plus the number of lags representing to the process deadtıme

Filtered White Noise

An ıdeal wh te no se s gnal contaıns equa varıance at a l frequenc es It ıs of ınterest to d scover how the wh te noıse s gnal s shaped' by a f rst order fi te

The whıte noıse sıgnal shown ın Fıg 12-6 s passed through a f rst order f lter The tıme serıes p ot of the resu tıng sıgna ıs shown n Fıg 12-7 below



F g 12 7 Fıltered Whıte Noıse

Note that the varıabı ty n thıs s gnal ıs much ower than ın the orıg nal whıte noıse s gna

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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.

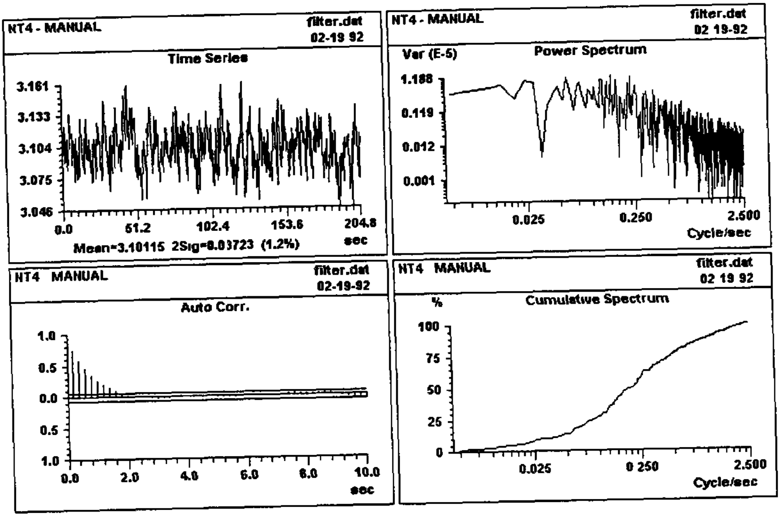


Fig 12-8 Filtered White Noise using log-log Power Spectrum

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 $2*\pi*\lambda$) 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

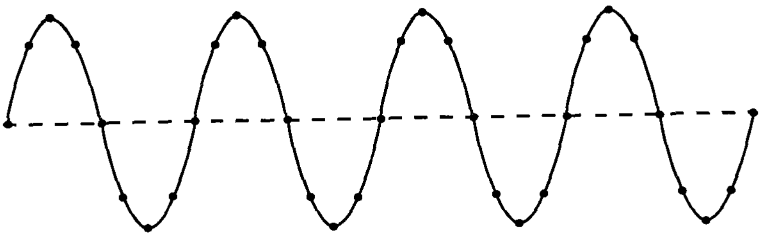


Fig 12 9 Effect of Sampling a Signal

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 will attempt to control out this false variability This is called ALIASING

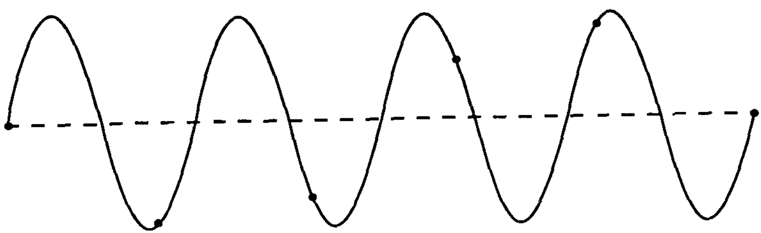


Fig 12-10 Effect of Aliasing

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

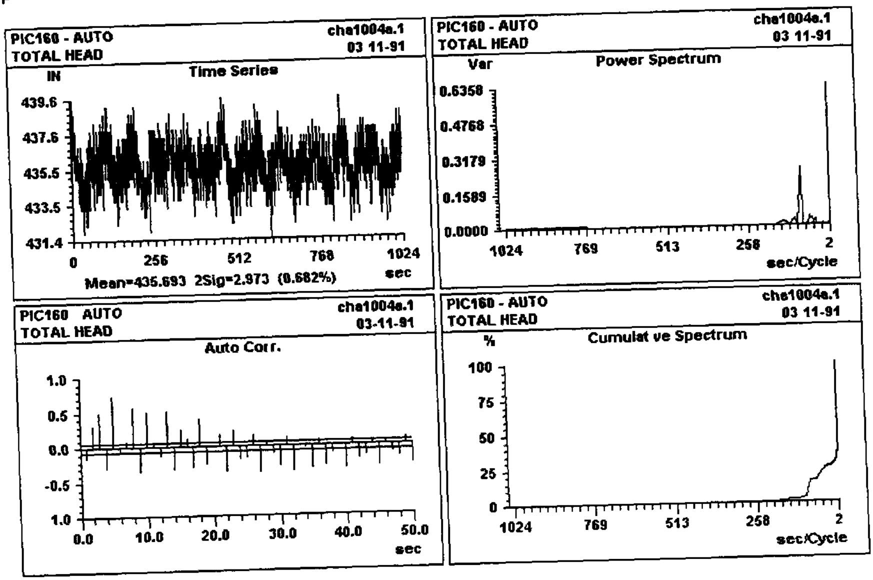


Fig 12-11

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 samp e the Tota Head at 20 m l iseconds The 2 second cycle disappears Now, a very strong cycle at 0 4 seconds/cycle is obv ous, accounting for approximately 90% of the total variance

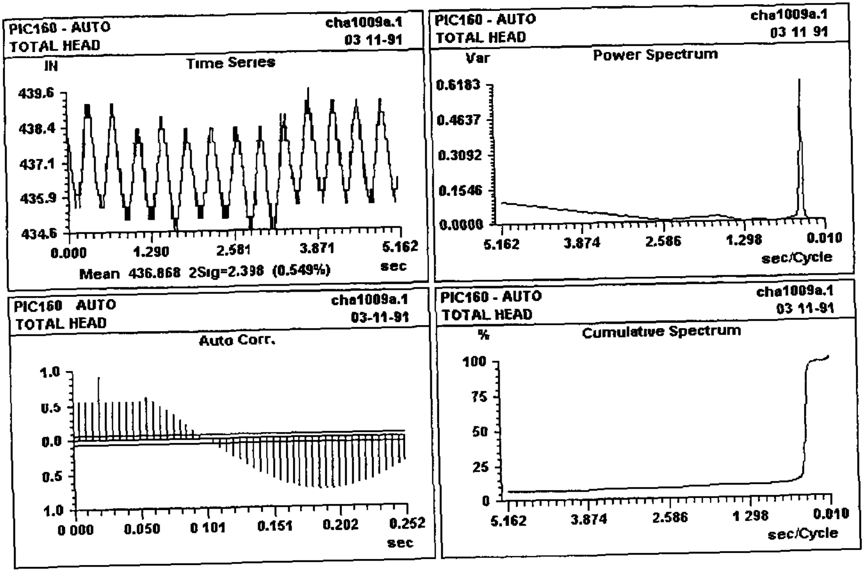


Fig 12 12 Sampl ng the Tota Head every 20 m ll seconds

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 (i e 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. Th s is i lustrated in the following example The process s gnal cons sts of a 17 second cycle with an amplitude of 0.025% cons stency, and white noise with a variance of 0 0025

	AMPLITUDE	VARIANCE
WHITE NOISE		0 0025
17 SEC/CYCLE SINE WAVE	0 025 %Cons	0 0003
TOTAL VARIANCE		0 0028

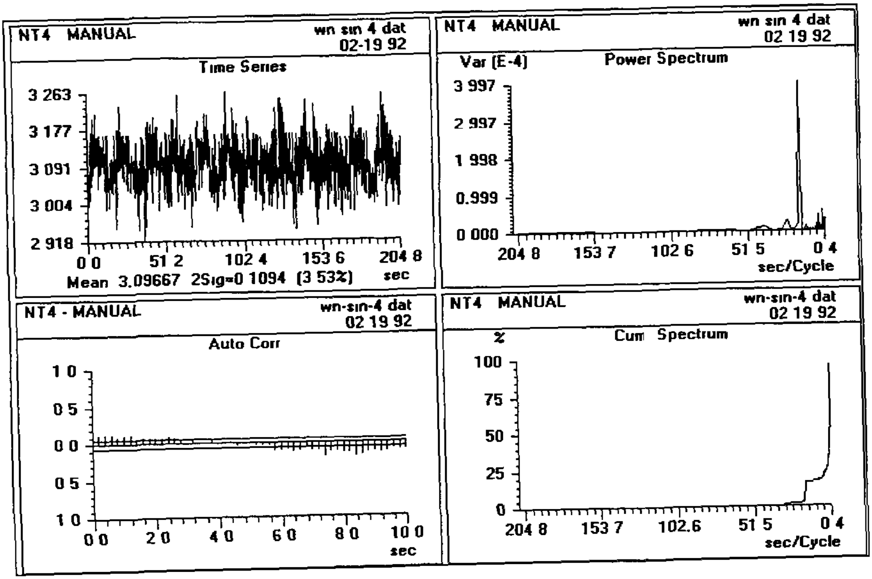


Fig 12 13 White Noise Plus Sing e Sine Wave

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 a one Note that whi e the 17 second cyc e dom nates the power spectrum plot, t on y 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 mach ne would be a likely source of the BW variabil ty
3 Why might the variability be higher on the heavy grade

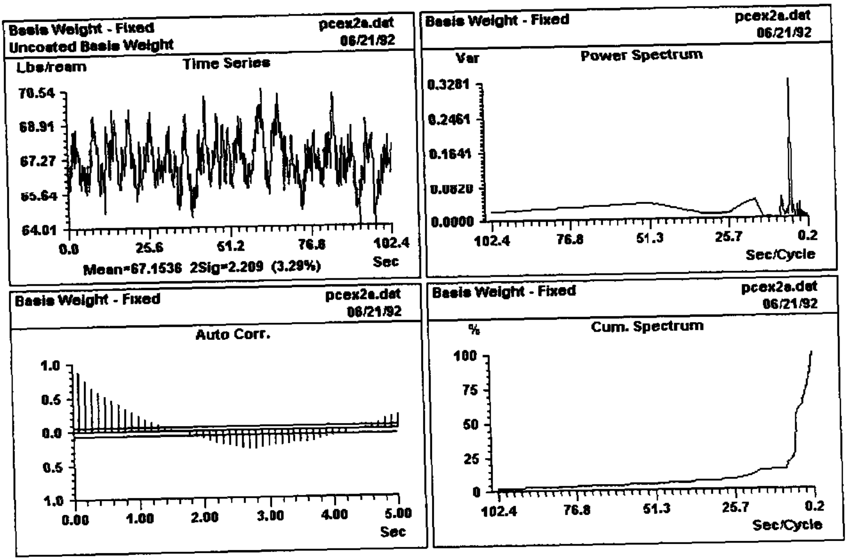


Figure 1.1 - BW Variability - Sampling at 0.1 second intervals, 1024 samples.- 68 lb grade

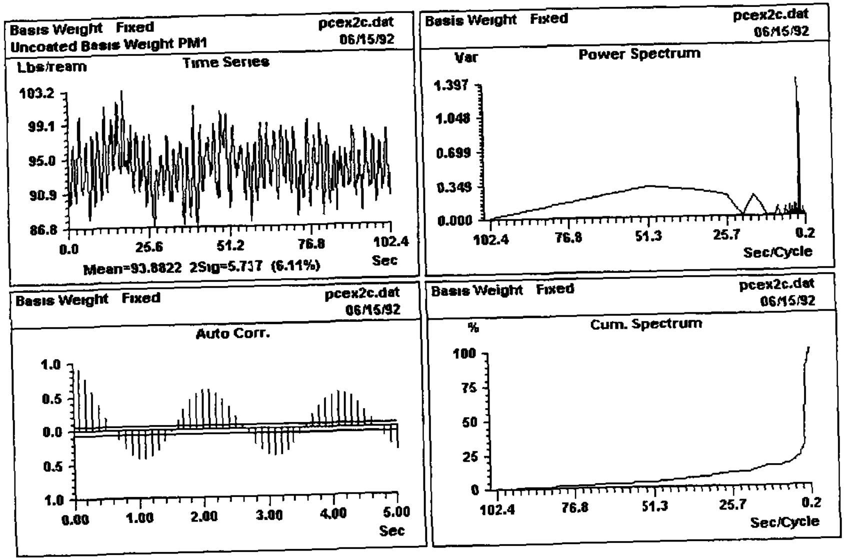
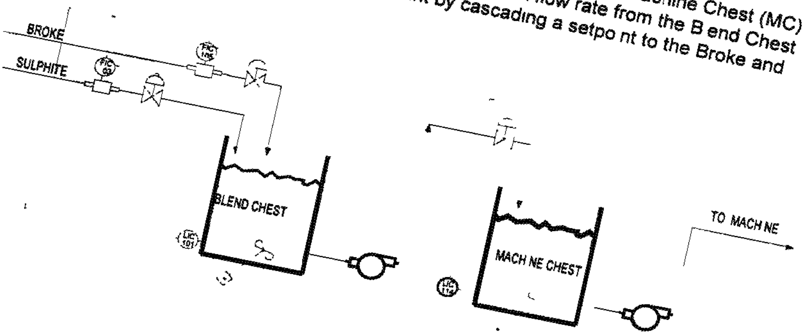


Figure 1.2 - BW Variability - Sampling at 0.1 second intervals, 1024 samples. 96 lb grade

EXAMPLE 2
Stock Chest Level Variability

The stock preparation area of a tissue machine is shown be ow The Machine Chest (MC) leve control er ma ntains level at setpoint by adjusting the stock flow rate from the B end Chest The B end Chest Leve control er mainta ns setpoint by cascading a setpo nt to the Broke and Su ph te flow controllers



Questions

1 What is the dominant cyc e in the process data
2 Can you explain why a l of the process variab e in this system contain the same cyc e
3 What tests would you conduct to determine the ultimate source of the process cyc es

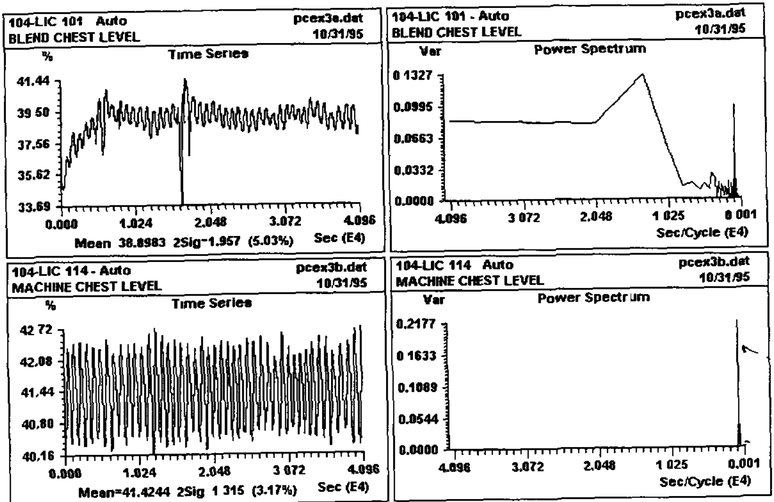


Figure 2.1 Variability in the Blend Chest and Machine Chest Levels

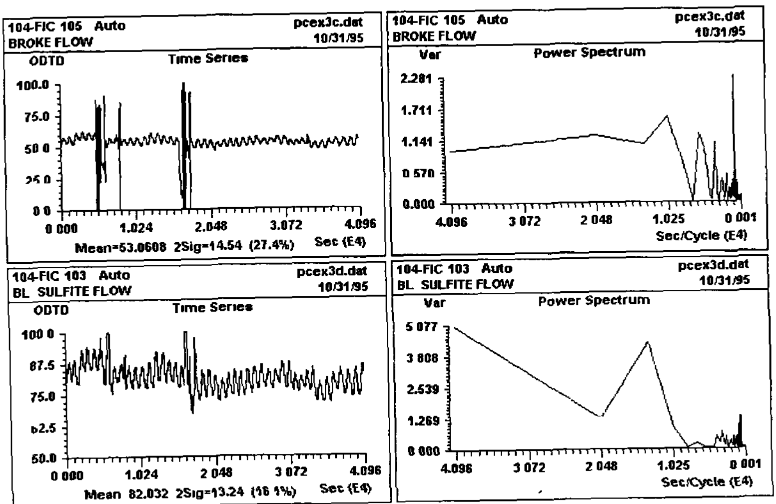


Figure 2.2 Variability in the Stock Flows to the Blend Chest

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 1.3 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.

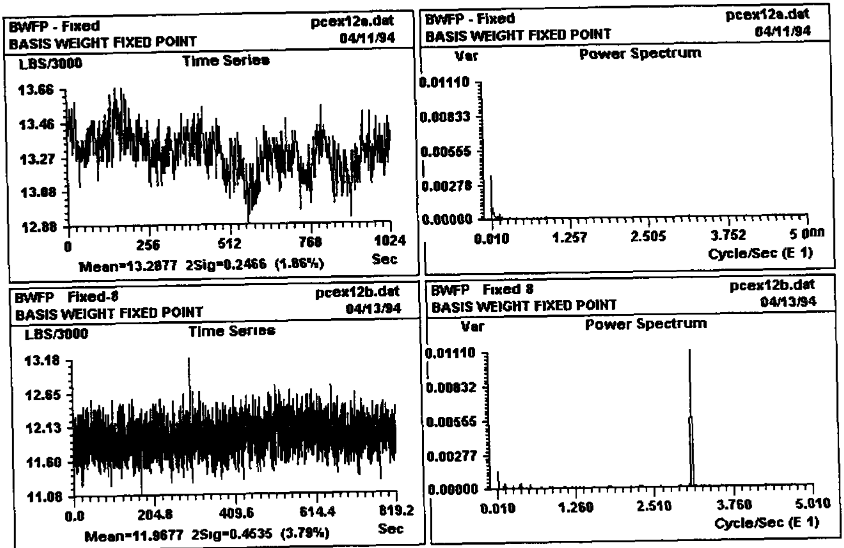


Figure 3.1 BW Variability under different test conditions.

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-lineant es are very often the most sign ficant source of process variabı ity Backlash, Stiction and Posit oned Overshoot are examples of hard non- inear ties which wil 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 F ow loop in the Lean Stock system The f rst plot shows the process and controller output with the loop in Auto mode The fo owing plot shows a ser es of open oop bump tests

Questions

1 Identify the nature of the control valve non-l nearity
2 D scuss the impact of mod fying the Lambda value on the per od and ampl tude of the flow cycle

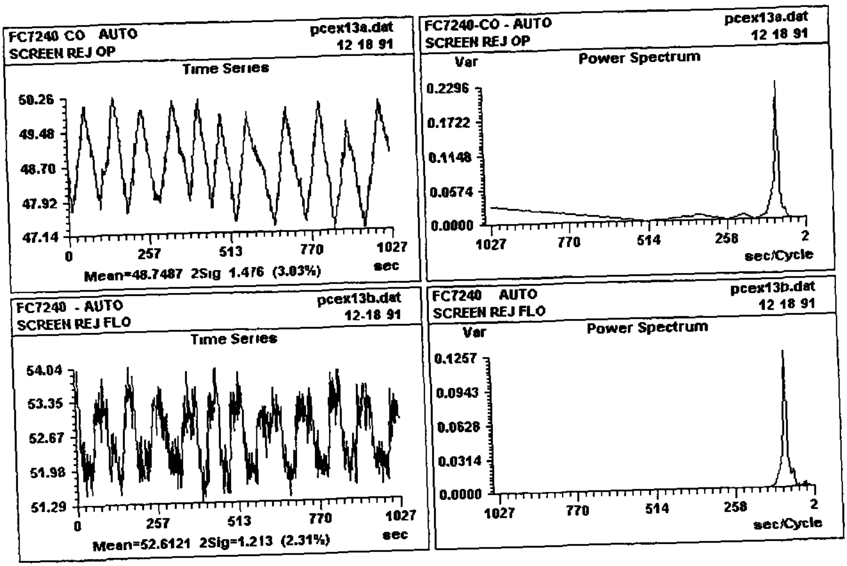


Figure 4.1.1 - Variability caused by Control Valve Non-linearity

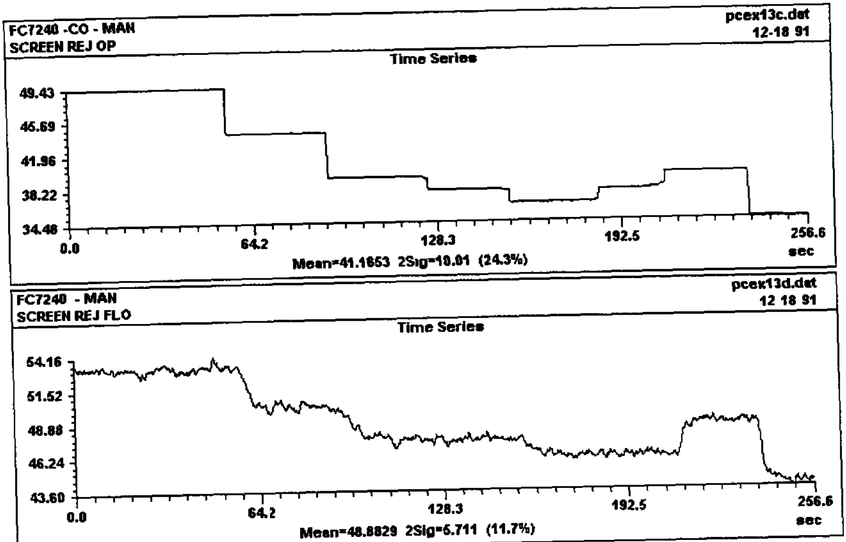


Figure 4.1.2 Open Loop Bump Tests on Screen Rejects Flow Loop

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 part cularly n the case of integrat ng processes or cascaded loops

Figure 4 2 1 shows a series of open loop bumps on a Thick Stock consistency loop F gure 4 2 2 shows a series of setpo nt bumps on a Refined Kraft flow oop

Questions

1 Quant fy the backlash shown in Figure 4 2 1
2 Discuss whether retun ng this contro loop is a good dea?
3 Quantify the backlash ind cated in the Figure 4 2 2 test ng
4 Estimate the response deadt me resulting from the back ash?

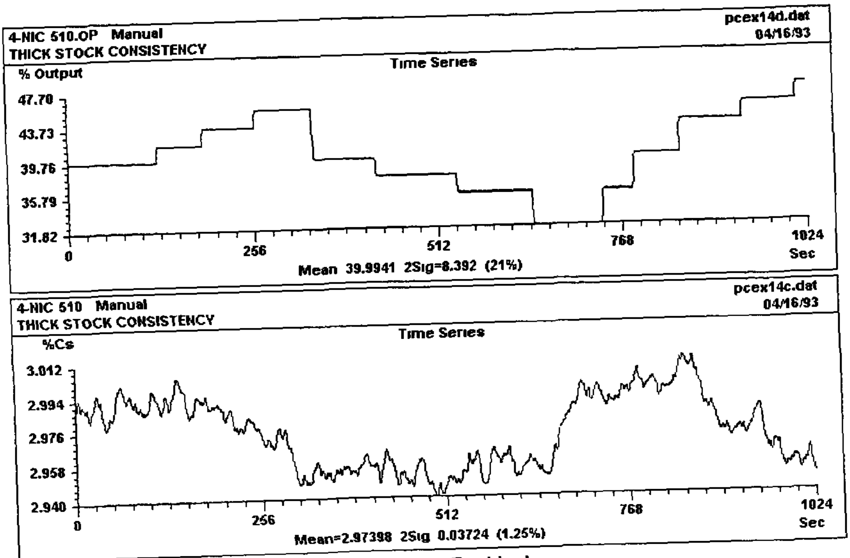


Figure 4.2.1 Open Loop Bump tests - showing Backlash.

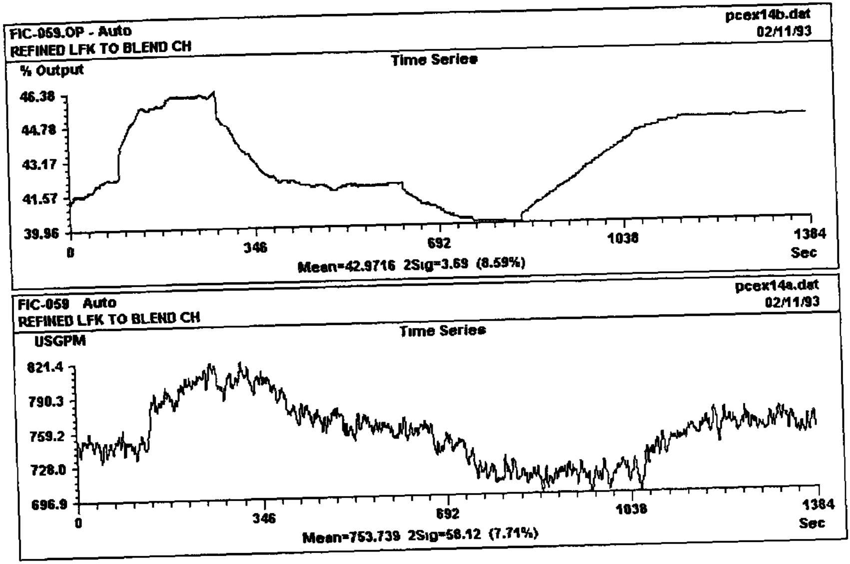
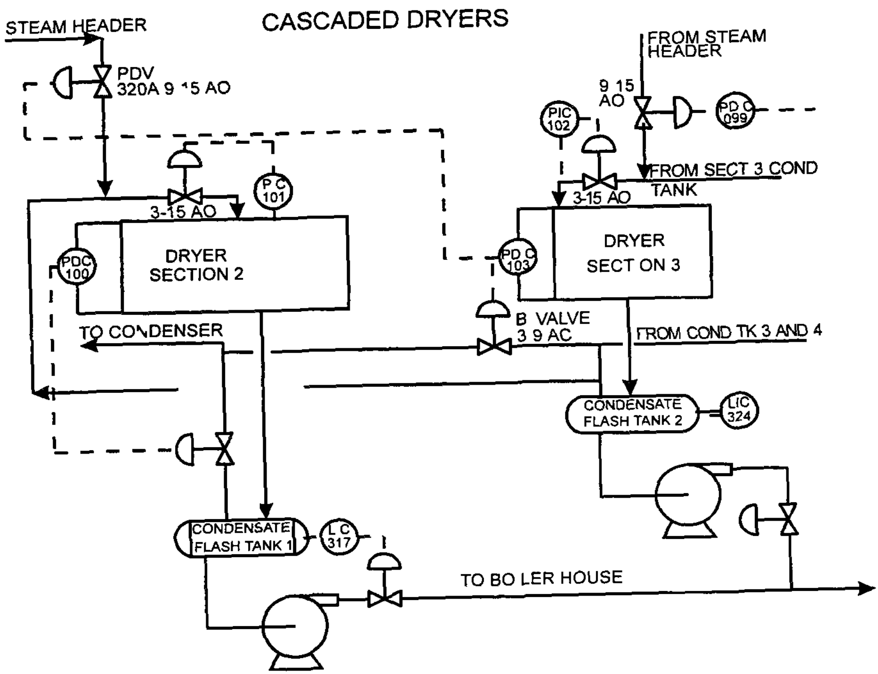


Figure 4.2.2 Setpoint Bumps in the presence of Backlash

EXAMPLE 5
Time Series Analysis and Process Upsets

The fol ow ng data was taken from Section 3 of the Dryer Sect on Pressure and Different a loops shown in the sketch be ow The data was collected immediate y fol ow ng startup



Questions

1 Character ze the pressure and d fferent al variab l ty during this startup period
2 D scuss the possible benefits of the power spectrum p ot n ana yz ng th s startup data
3 Why does the d fferent al decrease so dramat ca ly fol ow ng the sheet break How does the d fferent a controller compensate for th s disturbance

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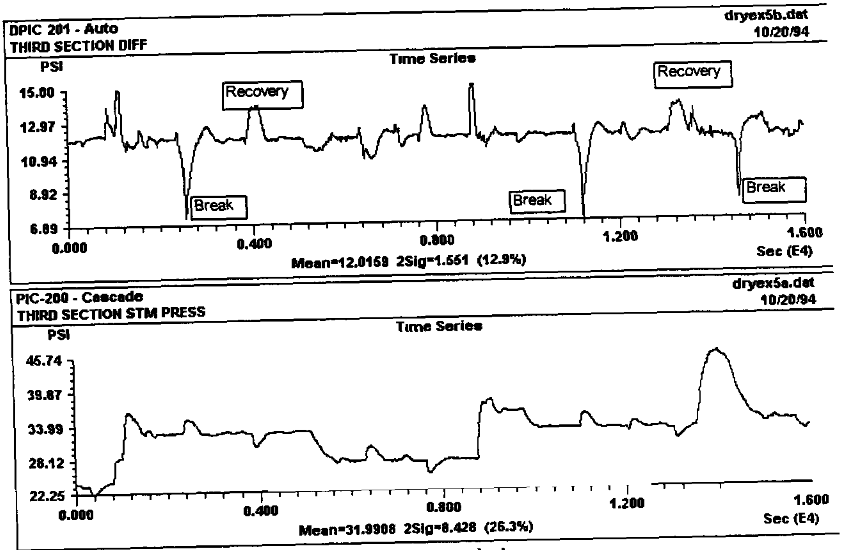
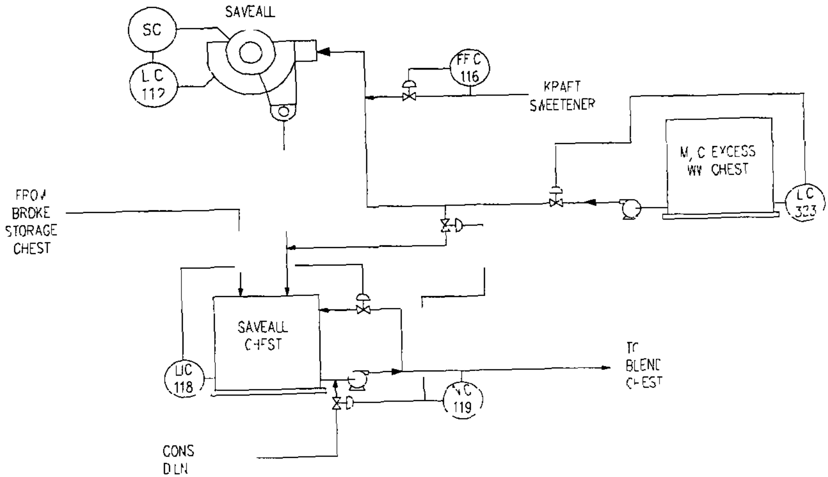


Figure 5.1 Dryer Data gathered during startup period.

EXAMPLE 6
Process Inherent Cycle

In a typica Savea l the mesh covered disks rotate through a Vat f led w th med um cons stency stock (0 5 - 2 0 %) The Vat level is controlled to setpoint (typically 50% level) by adjusting the disk speed As the d sk speed increases the f ltrate flow rate ncreases As the disk rotates through the Vat the f ltrate rate varies causing a level cycle at a period equa to the d sk revo ution time A time series plot of a typ ca Savea l Vat leve is shown below



Questions

1 Shou d the process nherent cyc e be a cons deration when deve op ng a tun ng strategy for th s oop

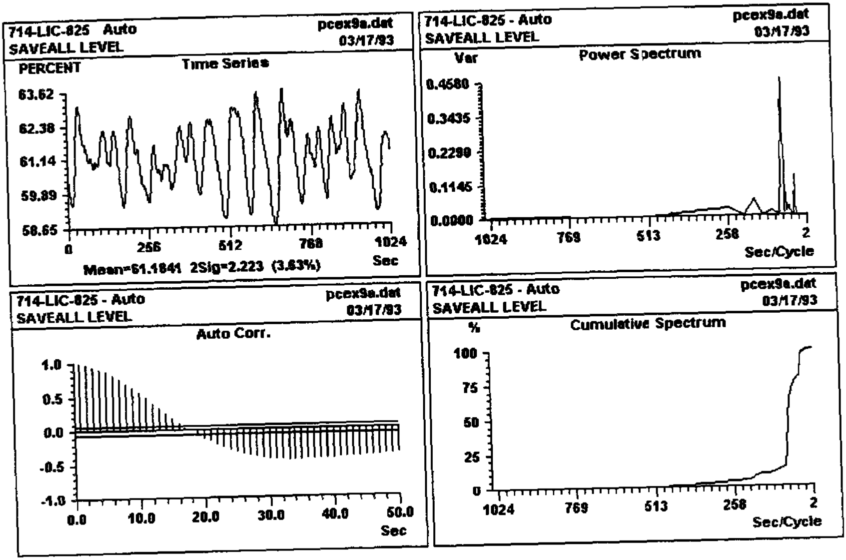


Figure 6.1 Saveall Level Variability

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 collection periods

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

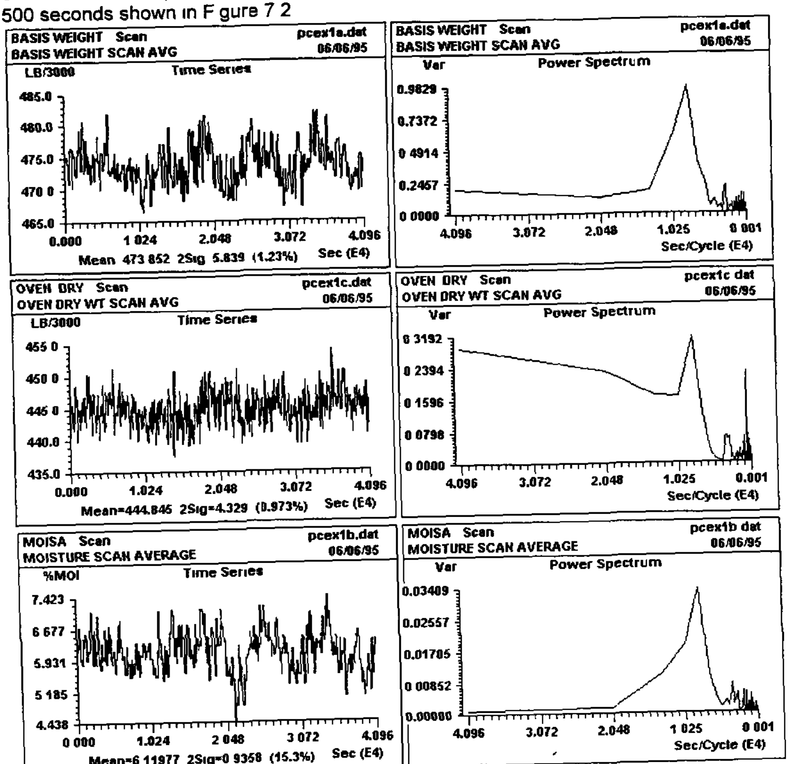


Figure 7.1 Scan Average BW, Conditioned Weight and Moisture Variability.- 40000 second run

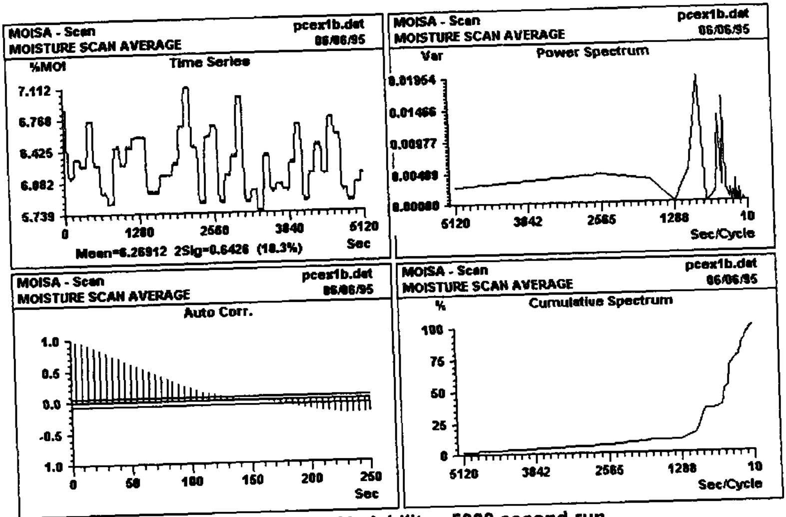


Figure 7.2 Scan Average Moisture Variability - 5000 second run

EXAMPLE 8
Signal Integrity

W th the advent of d gita transmitters m croprocessors have found there way out to the f eld devices This has been beneficia in terms of perform ng tasks such as remote re spann ng and ca brat on However because the s gnal s sampled in the field there is the possibı ty of aliasing occurring at the fie d device Furthermore the signa often undergoes various forms of digita signa processing such as sp ke suppress on or mov ng average ca culat ons The reso ution of the dig tal transmitter may be inadequate because of A/D resolut on The mpact of these non-linearit es 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 examp es of digital sensor data

Questions

1 Descr be the non-linearity shown n F gure 8 1 Is this acceptab e for th s appl cat on?
2 Estimate the A/D reso ut on n the data s gnal shown in Figure 8 2 The eve transmitter is spanned from 0 to 100 % level
3 A Fan Pump speed s gnal s compared with simultaneous y co ected Tota Head s gna in F gure 8 3 A bump test showing the total head response to a fan pump speed bump s shown n Figure 8 3a Does the fast cycle in the Fan Pump speed signa appear to be real?
4 Exp a n the d fferences in the power spectrums shown n Figure 8 4 The top plot shows an ana og transmitter signal The simu taneously co ected Smart transm tter data s shown on the bottom plot The sensors are both flush mounted to the Headbox

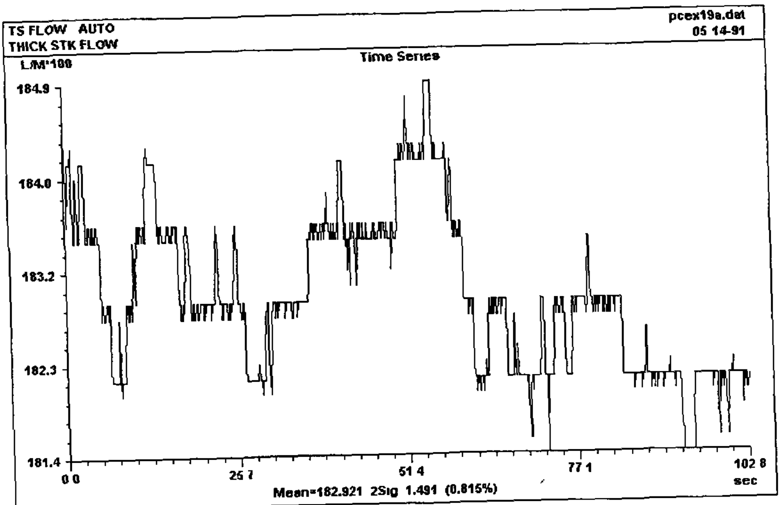


Figure 8.1 Signal from Mag Flow Sensor measuring Thick Stock Flow

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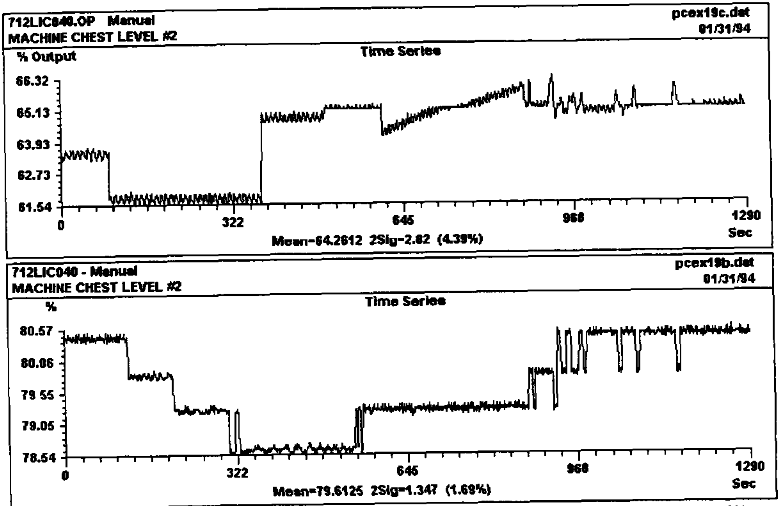
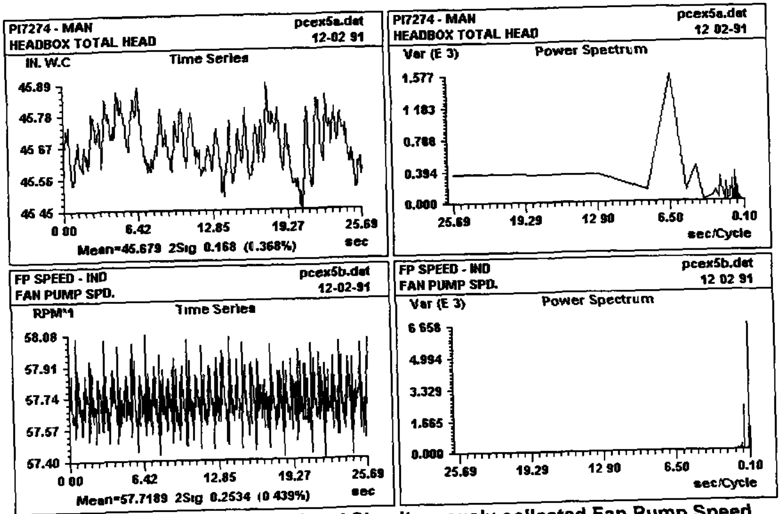


Figure 8.2 Level response to Open Loop Bumps - using Smart Level Transmitter



8.3 Variability in the Total Head and Simultaneously collected Fan Pump Speed

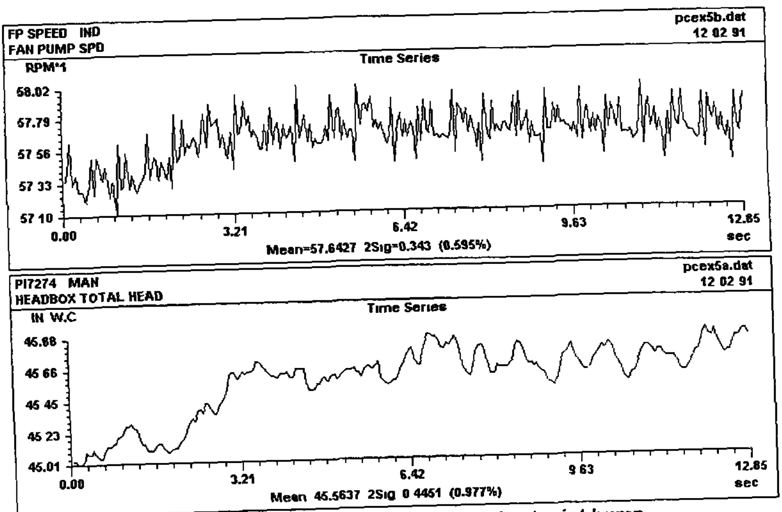


Figure 8.3a Total head response to Fan Pump Speed setpoint bump.

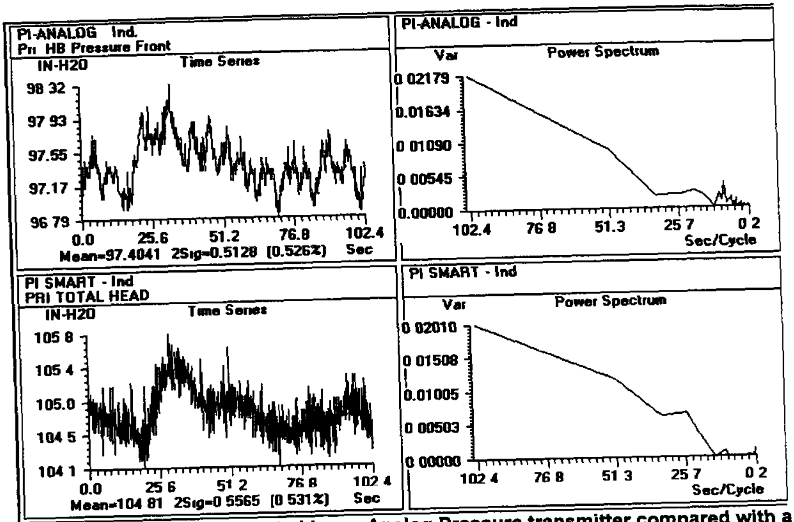
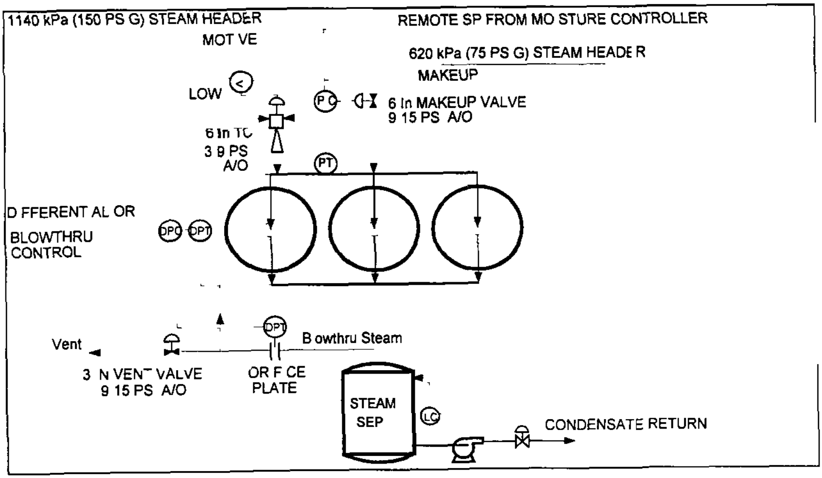


Figure 8.4 Variability Reported by an Analog Pressure transmitter compared with a Smart transmitter.

EXAMPLE 9
Paper Machine Dryer Control

In the case of Paper Machine Dryers it is often difficult to ach eve acceptable control performance because the process dynamics of the Pressure and D fferentia control loops vary widely depend ng on operating regime Thus, wh le contro performance appears to be acceptab e at the grade where the tuning was performed, the processes wi l be h gh y variabi ty on a different grade

Questions
1 Describe how the dynamics of the B owthru flow controller vary with thermocompressor (TC) sp nd e position
2 Exp a n why the Blowthru variable begins to cyc e so vigorously
3 What this tell us the power spectrum plot



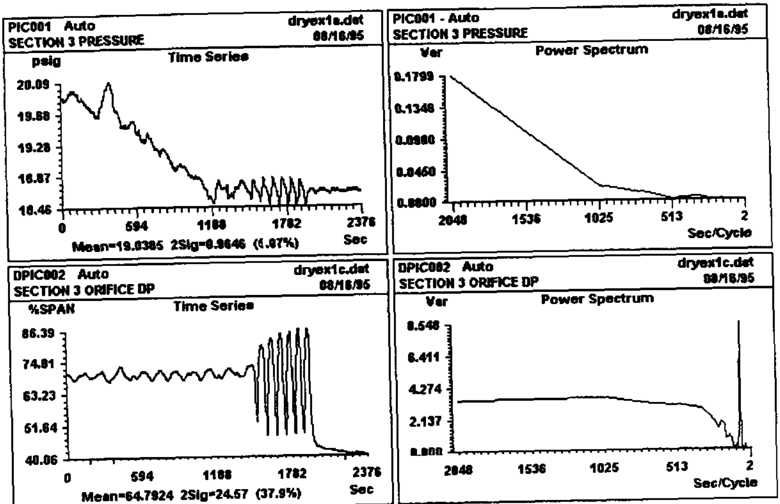


Figure 9.1 Variability Results from the Pressure and Blowthru loop of Dryer Pressure and Blowthru Loops

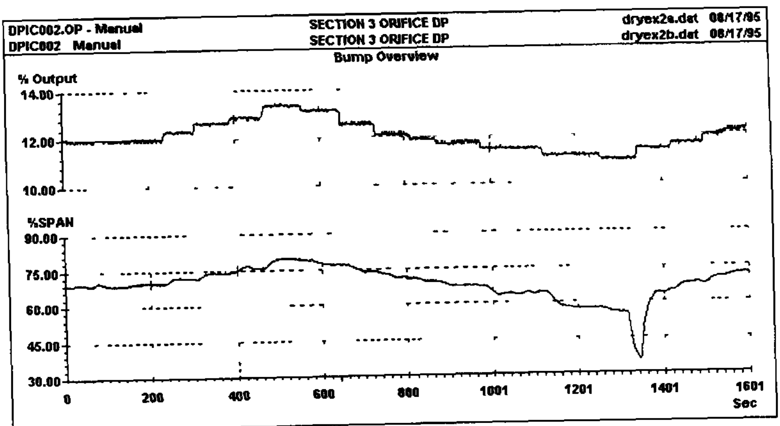
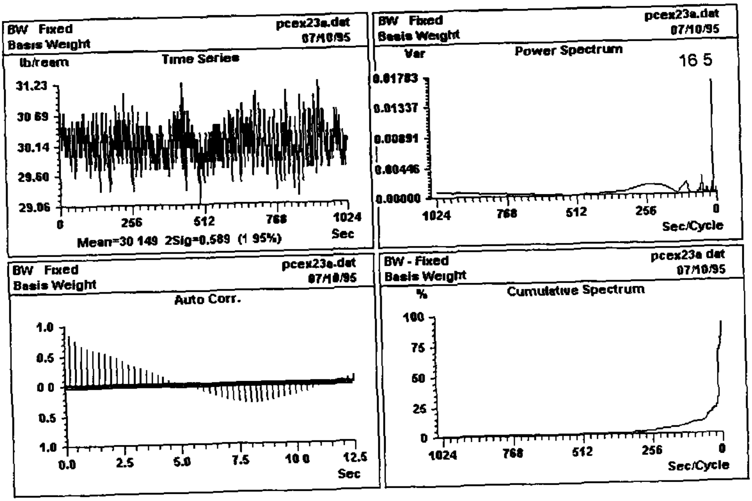


Figure 9.2 Open Loop Bump test results from Blowthru Control Loop.

EXAMPLE 10
Fast Basis Weight Variability

The Basıs We ght data shown ın Fıgure 10 1 was collected from a Newspr nt machıne runn ng at 4000 fpm As can be seen from the power spectrum plot there s a domınant cycle ın the data at 16 5 seconds The cumulatıve spectrum plot represents more than 50% of the varıance



Questions

1 What process area would be a l ke y cand date as the source of th s cyc e?
2 What methods could be emp oyed to determ ne the source of th s cyc e?

The time series plots shown in Figure 10 2 show the Basis Weight data compared to Tota Head data As can be seen from the power spectra, the 16 5 second cyc e is dominant in both sets of data This is an indication that the cycle in BW may be caused by the Tota Head var ability.

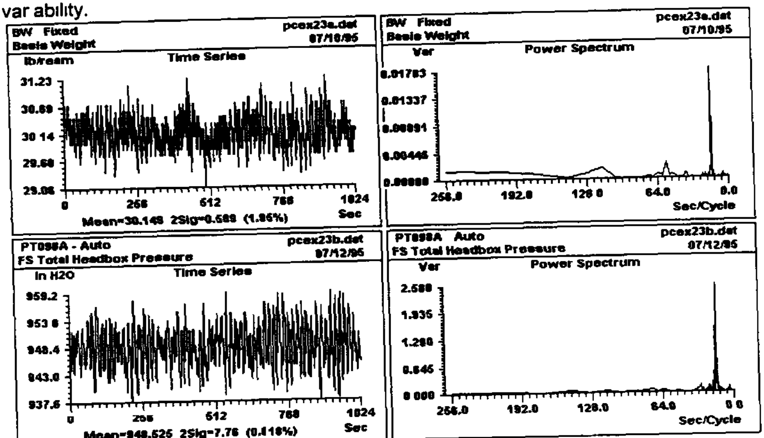


Figure 10.2 Basis Weight and Total Head variability

In order to determine if this is the defin te source of the BW var ab l ty the Tota Head contro loop was placed nto manua mode and data was co lected on both the Total Head and Bas s We ght As can be seen from the time series p ots shown in Figure 10 3 the 16 5 second cyc e s no onger visible

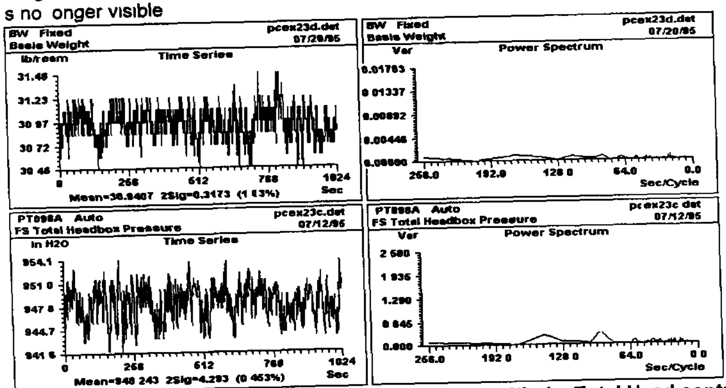


Figure 10.3 Basis Weight and Total head variability with the Total Head controller in Manual mode.

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14. INTEGRATED PROCESS AND CONTROL DESIGN

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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 w'll 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 opt mizing the process objectives - usually involv ng 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 variabil ty 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 Lambda = Corner Period/2π

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 wil still be variability remaining .

2 *Make Effective Use of Process Mixing*

- Improving the agitation in Mixing Chests 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

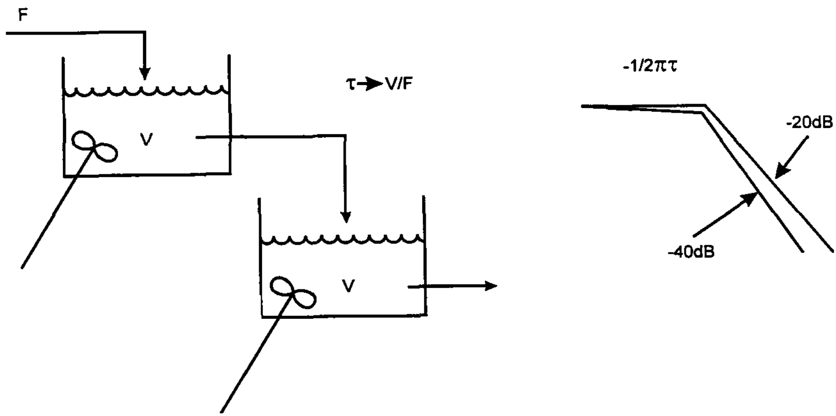


Fig. 14-1

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

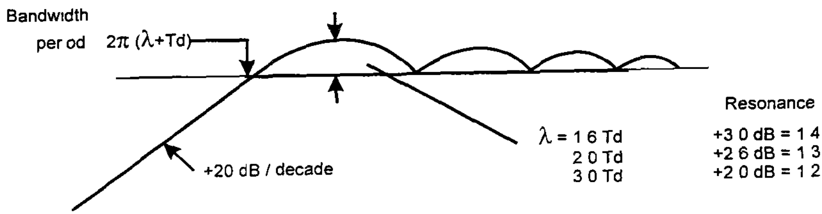
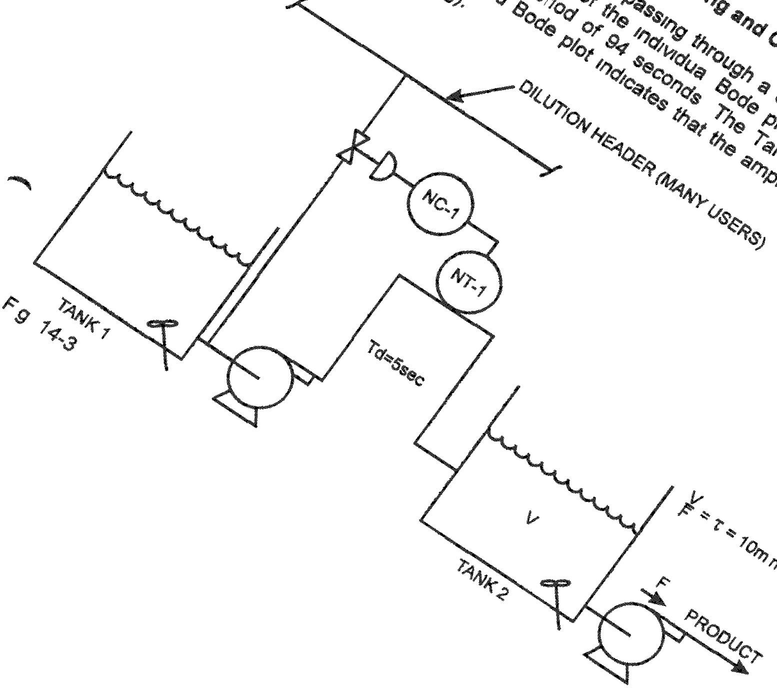


Fig 14-2

Attenuation of Mixing and Control in Series

The Bode plot for a disturbance passing through a consistency control loop
with tank mixing is the sum of the individua Bode plots In the examp e be
controller has a Cutoff period of 94 seconds The Tank Mixing Cutoff perio
m nutes. The combined Bode plot indicates that the amplitude ratio does not ex
0 1 (10% remaining).



F g 14-3

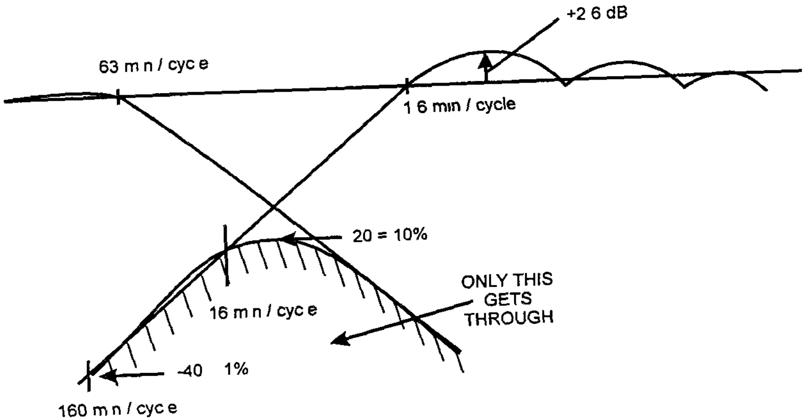

+2 6 dB
63 m n / cyc e
1 6 mın / cycle
20 = 10%
ONLY THIS
GETS
THROUGH
16 m n / cyc e
-40 1%
160 m n / cyc e

Fıg 14-4

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 redirected 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.

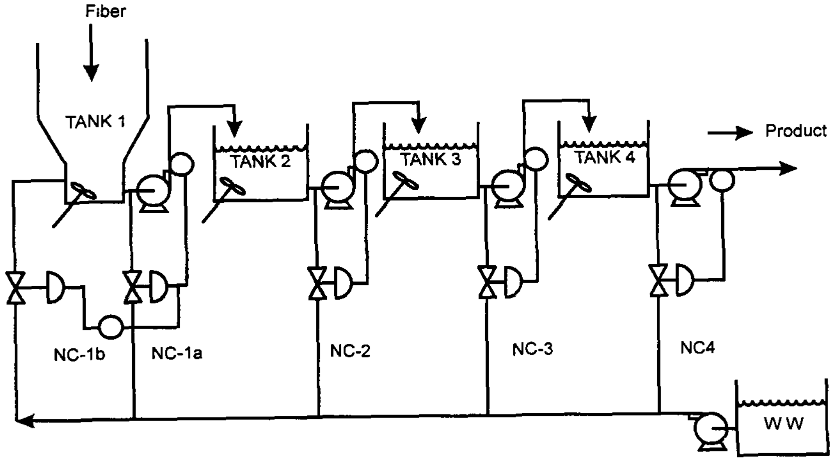


Fig 14-5

This design does not effectively redirect variability away from key process stream (i e 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 mechanısm for varıabılıty pathways ıs shown ın the followıng block dıagram

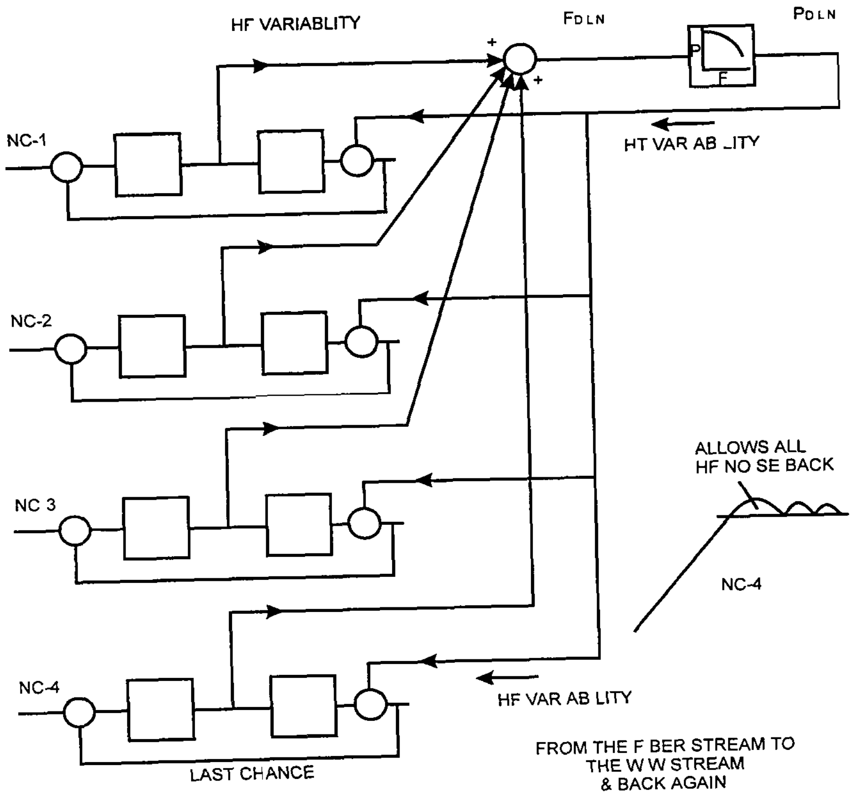


Fıg 14 6

Eliminating Variability Pathways

Eliminating the variability pathway from upstream consistency loops to the Tank
consistency can be achieved by installing a whitewater pump which s dedicated
Tank 4 consistency loop Disturbances to NC4 will now be l mited by the Tank 4
Pass filter

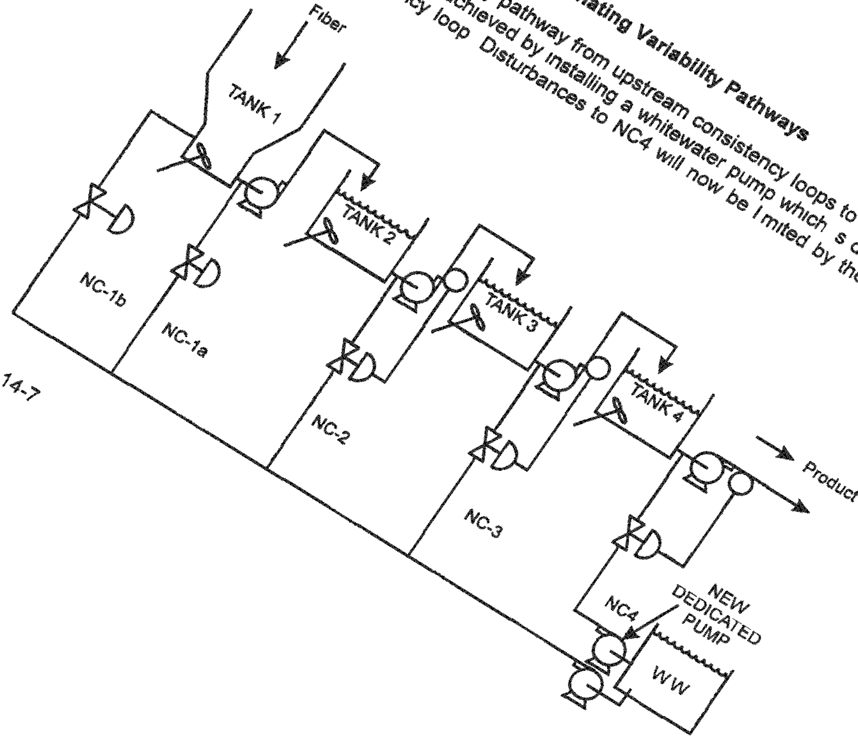
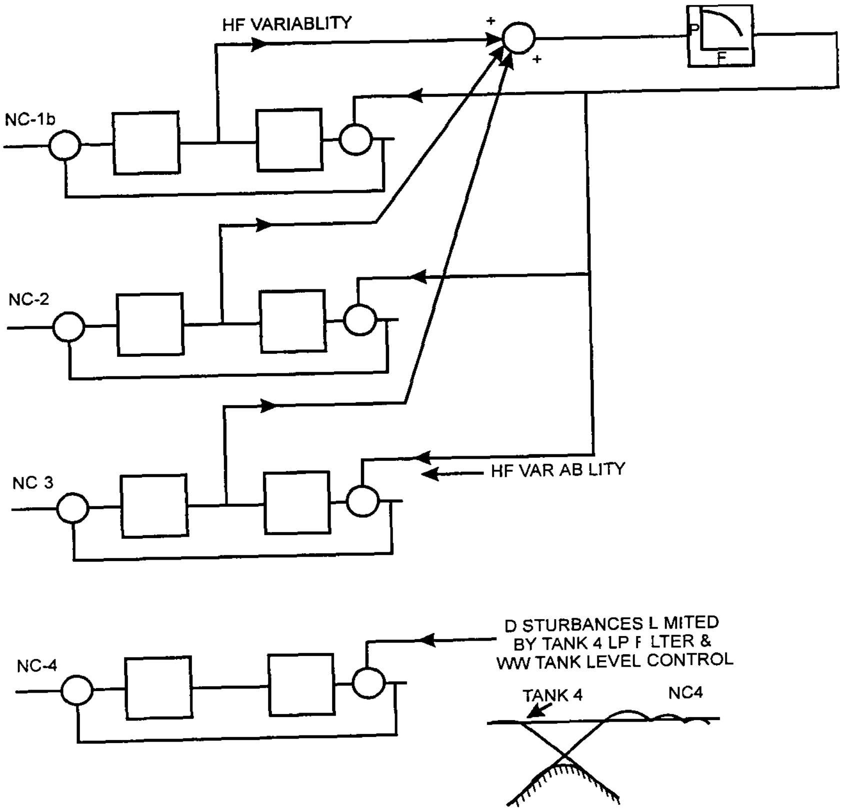


Fig 14-7

Dynamic Representation of Elimination of Variability Pathways

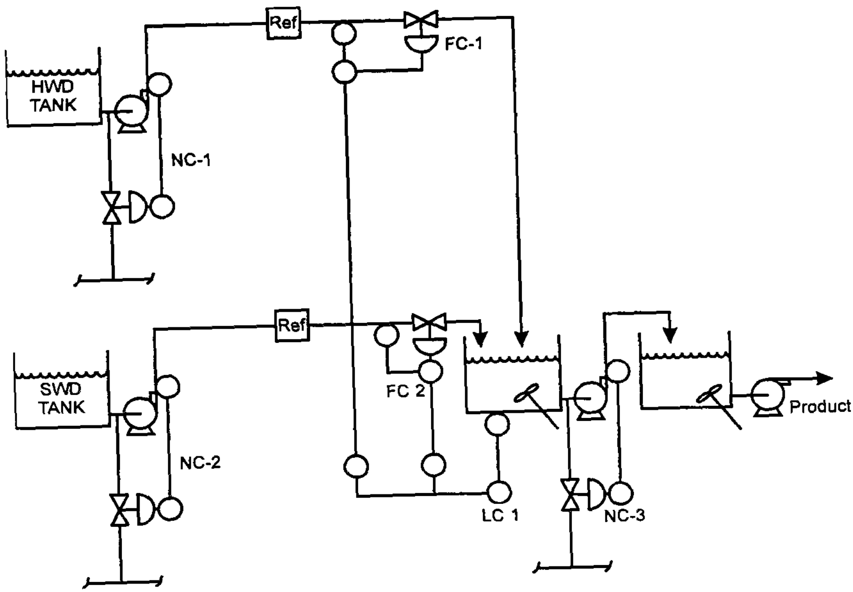


F g 14 8

Tuning Strategy - How To Select Lambda

The process design and control strategy design determine the variabi ity pathways The Lambda selection defines how much of the process var ab lity w ll 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



F g 14-9

NC-1, NC-2

These are very important loops. Severe disturbances are expected, but there is much downstream m xing which will attenuate higher frequencies The best approach is to tune these contro lers as fast as possible while limiting resonance to less than an AR of 1 41 (3 dB)

FC-1, FC-2

These stock flows part cipate in the Stock Blending strategy Since the primary objective of the Blending system is to produce a uniform furn sh, both flow loops should have the *same* Lambda value The flows will track the cascaded setpoint (from the level controller) ident cally producing a constant ratio In addition, s nce 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

Th s control er shou d be tuned as slow as poss ble, whi e 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 variabi ity on upstream processes Tight level control is not nearly as important to product variability and machine effic ency as is tight stock consistency and ref ner control

NC-3

This is the final consistency control point Hopefully the variabil ty is reasonably ow and random n nature The minimum variance cr ter a can therefore be considered on this loop

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EnTech Seminars

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Toronto Ontario M9B 6E5
Tel 416 626-6908 Fax 416 626-7691
email entech@entechcontrol com

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 deadtime 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**.

PCE I	January 27 - 31	March 10 - 14	June 2 - 6	September 15 - 19	November 3 - 7
PCE II	February 3 - 7	March 17 - 21	June 9 - 13	September 22 - 26	November 10 - 14

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 ongoing 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.

PDE I	January 20 - 24	April 7 - 11	September 8 - 12	November 17 - 21
PDE II	April 14 - 18	December 1 - 5		

MODERN LOOP TUNING

EnTech Seminars

*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 held in EnTech s training center or a mill chosen site, uses computer based dynamic simulation in place of the actual mill processes for the hands-on exercises described above

E & I Technicians - Instrument Specialists - Instrumentation Engineers

MLT I	On-site scheduled to suit mill requirements	**MLT+** adds one week of hands-on training with the EnTech Toolkit	
MLT II	February 18 - 21	April 15 - 18	October 7 - 10

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, savealls 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

Control Engineers - Process Control Engineers - Instrument Engineers - Systems Engineers

PCA II (Paper Machine) June 16 - 20 PCA III (Bleach Plant) April 28 - May 2 **PCA IV (Boiler)** May 12 - 16

Request Form

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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 general set of recommendations aimed at helping pulp and paper mills manage process and product variability. It suggests a minimum set of standards for human awareness, knowledge and skills in the various aspects of the process control engineering discipline that is needed at each stage of a pulp and paper mill design, start-up and on going operation/optimization.

1.0 Competitive Marketplace Background

The global market for manufactured products continues to focus on product quality and uniformity. Today, product uniformity specifications are approaching 1% and paper product can be rejected when it deviates outside specified limits – or when the variability characteristics of the products adversely affect the operation of the customer's secondary manufacturing – such as a printing press. There is now a need to pay more attention to the qualifications of personnel who design process equipment, process piping, process control installations, control strategies, configure DCS's, tune loops and optimize pulp and paper mill variability. EnTech™ Control Engineering Inc. has specialized in the analysis and optimization of pulp and paper manufacturing through mill variability audits and the training of mill staff for a decade. Mill variability audits have shown [1] that product variability is caused mainly by the combined behaviour of the many upstream process variables. In many cases, it is possible to trace these causes to individual control loops. Of equal importance is the impact of process variability on manufacturing efficiency in which small and subtle improvements in the process variability of key variables can have a large impact. Audits have identified that automatic control loops which cycle are a major cause of destabilizing product or in the pulp and paper industry.

1.1 Mill Audit Variability Results

To date, only about 20% of all control loops surveyed in mill audits have been found to actually reduce process variability in automatic mode over the short term [1]. The remaining 80% of loops were found to increase variability. Of these some were found to oscillate due to control valve non linearities [2] (about 30%), others due to controller tuning and controller equipment design limitations [3], (about 30%), some due to deficiencies in control strategy design (about 15%) while yet others were due to process design (about 5%). These findings are an indication of a massive "knowledge gap" in current design and maintenance practice which fails to even recognize that dynamic behaviour is a potential source of concern, let alone to provide effective designs of equipment, processes and procedures which optimize the impact of process dynamics on pulp and paper operations. To a great extent, process equipment design, mill design and mill operation today is thought of only in steady state terms – "so many tons of raw materials plus additives and energy will produce so many tons of product." When process variability becomes evident on instrument signals there is a strong tendency by many people to "smooth-it-out" through signal filtering – a form of denial.

1.2 The Knowledge Gap and The Control Engineer

Mill audit findings are an indication of a massive "knowledge gap" in current design and maintenance practice. This has occurred in spite of the fact that most engineers study automatic control theory in university then quickly forget the mathematics of control after graduation, thus leaving no foundation for future design procedures and practice. A new graduate is not a control engineer. A 'true' control engineer is one whose practical experience is leavened by, and based on, a sound foundation in dynamics and control theory and is a fully trained professional. There are very few such people in the pulp and paper industry today. As a result, decisions affecting control are often made without adequate awareness of the fundamental principles which govern dynamic behaviour. Mill variability audit results illustrate the impact of this knowledge gap.

1.3 The Fully Qualified, Competent Control Engineer – A Role Model

To become a "Fully-Qualified" control engineer is in fact a formidable task requiring years of experience, much determination and a very strong focus. It is a serious *'calling' which can be achieved over a* career coupled with continued learning. Even after graduation with a Masters or Ph.D. in control engineering, one has potential academic qualifications in the theory only where the "Fully-Qualified" control engineer is master of both theory and practice. There are only a handful of such people in the pulp and paper industry, however there are several hundred people who aspire to, and are working towards this goal today. If the control engineer is key to the variability reduction efforts of a pulp and paper mill, and if these efforts truly do impact on the competitive position of the mill, then it follows that the potential for these efforts to succeed are a direct function of the competence of this engineer. Competitiveness in manufacturing uniform product is both knowledge-based and control knowledge intensive.

The "Fully-Qualified" control engineer is experienced with each pulp and paper process area of interest [8], has a good command of control theory (e.g. [9,10,11,12,13,14,15,16]) and is skilled at *translating operational manufacturing* objectives into control objectives. Technical skills should also include process dynamics, transfer function models, dynamic simulation, algorithm selection, loop tuning, process and instrumentation troubleshooting with time series analysis, spotting process variability, and minimizing variability through control. Finally, the control engineer should be a good communicator, teacher, trainer and coach who is equally at home with a process operator, a technician or a mill manager.

1.4 Impact on Industry Thinking

Not only has the industry missed the technical skills of the control engineer but the leadership that they should have exerted has also been missing. What of establishing effective programs for loop *tuning, E/I technician training* (still being taught the highly oscillatory [17] Quarter-Amplitude-Damping loop tuning method!) and supporting effective management decision making? In short, with the absence (in effect) of the control engineer from the ranks of the pulp and paper industry, industry thinking has been allowed to 'slide' to the point where the concepts of process variability and dynamics are effectively absent. Absent at a time when variability, uniformity and quality are increasingly defining competitive position and company survival.

1.5 Advances in Process Control – Background

In the last decade modern model-based methods such as 'Lambda tuning' [6] and Internal Model Control (IMC) and Robust Control [7] and *Stochastic Control [16] have been used extensively in pulp and paper* mills with considerable success in variability reduction. These concepts are well established in the technical literature and collectively capture the essence of the variability and control problem in the pulp and paper industry. However, to be able to take advantage of these ideas, the right academic training, practical experience, organizational mandate and personal motivation are required – in short, to be a qualified, competent, control engineer.

2.0 Summary of Control and Dynamic Problems Encountered

Based on mill audit experience, the types of problems encountered include:

- excessive deadtime in transmitter installations
- excessive deadtime in conveyor designs
- inadequate process agitation and mixing
- air entrainment in stock lines
- oversized control valves
- excessive backlash and/or stiction in control valves
- excessive interaction in control strategies
- inappropriate control algorithms and tuning approaches
- *inappropriate controllers*
- excessively slow sampling rates
- lack of anti-aliasing filters
- digital transmitters with excessive deadtime (latency time)
- excessive DCS display ranges
- excessive use of report-by-exception
- control loops with default tuning
- faulty control strategy design and coding
- inability to locate the source of variability
- inability to organize the coordinated tuning of many loops
- inability to tune for given disturbances

Most of the above problems were caused by a lack of awareness and knowledge. It is important to note that the most serious problems identified in mill audits are ones which relate to process layout, equipment selection, piping and equipment design, as changing tile tanks, agitators and stainless steel pipes is both difficult and costly. Post audit optimization programs can more easily replace bad valves, re-tune loops, or change DCS control strategies and algorithms. However, this underscores the need to improve initial designs and hence the need to improve the awareness of equipment and process designers regarding the impact of their designs on the resulting process dynamics and product uniformity.

2.1 Purpose

This document sets out to define the elements of the control engineering discipline which are needed in the various phases of a pulp and paper mill design, start-up, operation and on-going optimization, together with the people who need to perform these tasks so that process variability, operating efficiency and product uniformity can be optimized. Each element is defined including the background and educational requirements that need to be mastered. An attempt is *made to identify the level of skill that each member of the 'manufacturing team'* (in the broadest sense) must have in order to allow success to be achieved in the end. The team is considered to comprise process equipment designers, piping designers, control system selection team, instrumentation selection and installation designers, control strategy designers, control loop designers, initial control loop tuning and start-up team, E/I technicians, process operators, production supervision, process engineers, process control engineers, testers, quality control staff and managers.

2.2 The Manufacturing Team

In the broadest sense, the manufacturing team consists of many people from the designers of the initial concept through to the operators. The list usually includes general consultants and equipment vendors, all of whom have played an increasingly important role in mill design in recent years. In order to simplify the categories, the following definitions have been made:

1) process designer (major process equipment design [equipment vendors], process designer, process flow sheet design, piping design, may include both mill engineering staff, but also includes major equipment vendor and general engineering consulting staff)
2) control system selection team (control system selection team responsible for control system selection, instrumentation equipment selection team, responsible for selection of transmitters, valves and field equipment, may include general engineering consulting staff)
3) instrumentation design (transmitter location and installation design, control valve selection and sizing, includes general engineering consulting staff)
4) control design (definition of control objectives based process operating and manufacturing objectives, design of Process and Instrumentation Drawings [P&ID's], control strategy design, control logic design [SAMA diagrams], control loop objectives, *control loop design, loop sheet drawings, operator interface* design, console configuration, display configuration, may include general engineering consulting staff)
5) loop tuning (setting of loop tuning constants during initial system configuration, start-up, on-going operation by start-up team,

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comm ss on ng team E/I techn c ans genera eng neer ng consulting or equipment vendor staff)

6) operations staff (process operators production s ipervisors, testers, qual ty contro staff)
7) process engineers (process eng neers product on eng neers, ass stants to product on super ntendents process trouble shoot ng staff)
8) process control eng neers, (process cor tro eng neers, contro eng neers var ab ty optim zation team)
9) managers (m managers, product on managers, eng neer ng managers technical managers, a so nc udes managers n equ pment vendor and genera consu t ng compan es)

The head ngs as def ned above are used throughout the rema n der of the document as each e ement of the subject s exam ned For each subject e ement each category w be rated n terms of the depth of sk requ red (e g A awareness S sk , K – know edge)

3.0 Awareness, Skills and Knowledge by Manufacturing Team Task

Hav ng def ned the contro engineer as the person where "the buck stops" on the subject of pu p and paper process var ab ty and contro , what fo ows s a systematic breakdcwn of th s core contro eng neer ng mater al into mportant e ements wh ch re ate to the typ ca prob ems that have been uncovered dur ng m aud ts These in turn are c ass f ed by the eve of sk or knowledge needed by each person or task w th n pu p and paper manufactur ng framework

Definitions – Skill Levels: Awareness, Skill, Knowledge
Three gross sk eve s of know edge or sk are arb trar y def ned be ow and used n th s document n order to s mp fy the concept

Awareness: Th s s mp y means that the nd v dua (usua y one n authority) needs to be genera y aware of the poss ble need for a type of act on or cons derat on Awareness mp es the respons b ity of invo v ng sk ed and know edgeab e peop e to ensure that correct des gn or operat on s ach eved Awareness does not requ re e ther sk or know edge

Skill: The nd v dua should be ab e to do a spec f c task proper y by fo ow ng a set of ru es or procedures Deep knowedge s not a prerequ site as ong as the procedure s correctly fo owed Sk mp es awareness but not necessar y know edge n wh ch case the need ex sts for a rev ew of the work done by someone w th know edge to ensure correct des gn or operat on

Knowledge: The nd v dua must have a f rm grasp of the fundamenta s of the subject both theory and pract ce and be capab e of app y ng this know edge to evo ve a successfu des gn or set of ru es to be fo owed by sk ed peop e t fo ows that the theory s understood and s used on a da y bas s to so ve rea prob ems Know edge mp es both awareness and sk

3.1 The Bump Test

Anyone working d rect y w th a contro oop the process operator the E/I techn c an, or the process eng neer shou d be sk ed n perform ng bump tests to check the "hea th" of the oop As an examp e for a given f ow oop a 1% bump test shou d resu t n about a 1 5% response n the process var ab e n ess than ten seconds otherw se someth ng has gone wrong

3.2 Deadtime

Deadt me s the most de-stab z ng dynam c phenomenon wh ch occurs n process contro Deadt me is usua y the resu t of transport de ay caused by conveyors, p p ng react on towers etc and severe y l m ts the ab ity of a feedback contro er to perform Everyone shou d be aware of th s and those nvolved in design or operation shou d be try ng to m n m ze it

3.3 Process Mixing

Process var ab ty can be reduced by two methods process contro , wh ch can on y reduce ow frequency or s ow var ab ty, and process m xing, wh ch can reduce h gh frequency or fast var ab ty The two are complementary However m audits have often dentif ed that the actua m x ng character st cs of tanks are often far ess effective than pred cted by the des gned res dence t me The mathemat cs of process mode ng, transfer funct ons and Bode plots app es equa y we to the m x ng character st cs of tanks Th s approach offers a conc se numer ca method of predict ng the effect veness of a m x ng operation and ts potent a mpact to reduce var ab ty at a g ven frequency and everyone shou d be aware of th s Those nvo ved n process des gn or troub eshoot ng shou d be ab e to ana yze s mulate and accurate y pred ct the effectiveness of process m x ng on product un form ty

3.4 Air Entrainment

A r entra ned n stock nes or other process streams not on y mpacts paper mak ng and other react ons but t a so adverse y affects nstrumentat on s gnals such as f ow meters cons stency etc Th s n turn strong y detracts from contro performance Process des gners and troub eshooters shou d know how to e m nate the entry of a r nto hydrau c systems

3.5 Control Valve Selection and Sizing

M aud t results have dent f ed poor contro va ve performance as the b ggest s ng e source of process var ab ity (see EnTech™ Contro Va ve Dynam c Spec f cat on [2]) n the pu p and paper ndustry There are two separate causes of prob ems

- the contro valve des gn se ect on of va ve assemb y components (e g actuator or pos t oner) or state of ma ntenance wh ch have resu ted n excess ve back ash or st ct on,
- the valve has been severe y over s zed The net result s that the contro oop w m t cyc e (a non near system mode of behav our) cont nuous y and de-stab ze process var ab ty There s a strong tendency for va ves to be over s zed due to the natura des re to ensure adequate capac ty and at m n mum everyone shou d know that a va ve s severe y overs zed f t s 15% open under norma operat on and that th s w ncrease var ab ty

3.6 Control Loop Interaction

t often happens that two contro oops mutua y de-stab ze each other for nstance, the contro of pressure and f ow n the same ne n process des gn t s mportant to m n m ze the occurrence of such process nteract ons wherever poss b e A des gners shou d know how to m n m ze these effects wh e sti meet ng the operat ona objectives Contro des gners shou d know how to des gn decoup ng a gor thms to overcome nteractions wh ch cou d not be avo ded

3.7 Control Algorithm Selection and Tuning

Contro a gorithm se ect on and tun ng requ res know edge wh ch es at the heart of the contro eng neer ng d sc p ne and nvo ves extensive tra n ng [9 10 11 12 14] Anyone nvo ved n contro or nstrumentat on shou d have at east bas c tra n ng n modern oop tun ng concepts Once the dynam cs become comp ex a forma contro eng neer ng background s needed

3.8 Control System Selection

Contro system se ect on requ res know edge of the spec f c process dynam cs n quest on and how these trans ate nto contro er requ re ments such as samp ng rates a gor thm structure, f ter ng needs etc [3], (For deta s see EnTech™ Contro er Dynamic Spec f cat on (Vers on 1 0 11/93) Those invo ved n contro system techno ogy shou d be sk ed n these cr t ca ssues

3.9 Digital System Sampling Rates

With the advent of the DCS there has been a tendency to s ow down contro oop execut on n order to save comput ng t me n the contro processors. Th s tendency s dangerous and very damag ng to contro performance. Samp ng rates shou d be determ ned based on process needs [3] and everyone nvo ved n contro or nstrumentat on shou d know how to determ ne these

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3.10 Anti-Aliasing Filters

Anti-a as ng f lters are necessary on each process measurement to prevent contro from nduc ng process var ab ty due to a ack of s gna ntegrity The amount of filtering needed s a funct on of the sampling rate [3] and everyone nvolved n contro or nstrumentation shou d know how to determ ne these

3.11 Digital Transmitters

The advent of d g ta transmitters has moved the ssue of samp ng rates, anti-a as ng and d g ta s gna process ng nto the transm tter tself Transm tter selection now nvo ves know edge of these ssues [3] and everyone nvo ved n contro or nstrumentation shou d know the criter a for se ect on

3.12 DCS Display Ranges

Often process operators are unaware of process var ab ty prob ems because the var ab e in question s be ng d sp ayed on a w de sca e such as 0% to 100% Process var ab ty of on y 1% or ess, wh ch may ndeed be cr t ca y mportant w now appear as a stra ght ne Clear y the se ect on of operator d sp ay ranges s cr t ca f the operator s to 'see s gn f cant process var ab ty Anyone see ng 'stra ght nes' on a DCS screen shou d f rst suspect poor y chosen d sp ay ranges

3.13 DCS Report-by-Exception

The advent of the DCS has brought report-by except on as a method of s gna transm ss on n order to m n m ze data h way traff c Report by-except on shou d never be used ns de a contro oop, as th s s extremely de-stab z ng Everyone nvo ved n contro or nstrumenta t on shou d know th s

3.14 Faulty Control Strategy Design and Coding

W th the advent of d g ta contro systems many peop e have become invo ved n des gn ng and cod ng' contro strateg es n contro system software In most cases this s seen as noth ng more than an exerc se n computer programm ng without any realiza t on of how the dynamic behav our of the process s be ng affected To des gn and code contro strateg es n computer software, a background n d g ta contro s needed [12]

3.15 Process Trouble-Shooting Using Time Series Analysis

Troub e shoot ng process prob ems us ng t me-ser es ana ys s too s s the heart of the process var ab ty aud t ng sk Know edge of t me ser es analys s techn ques and process dynam cs s requ red [15]

3.16 Spotting/Reducing Variability to Optimize Pulp and Paper Manufacturing

Mak ng un form product eff c ent y n a pu p and paper m l n the face of randomly vary ng raw mater a s and add t ves es at the heart of becom ng more competitive Th s ab ty starts w th the process operator be ng ab e to spot var ab ty when t s seen t does not stop here but cont nues through to the contro eng neer's sk to m n m ze var ab ty with stochast c contro [16]

4.0 Summary – Skills Matrix

n conc us on a l of the recommendations are presented n summary n Tab e as a sk s matrix

Table I - The Skills Matrix

Category	Proc Des n	Cont SysSel	Instr Des n	Cont Des'n	Loop Tune	Ops Staff	Proc Eng	Cont Eng	Manager
Level of Skill									
Bump Test	A	A	A	A	S	S	S	K	A
Dead T me	S	A	S	S	S	A	S	K	A
Process M x ng	K	A	A	A	A	A	K	S	A
A r Entra nment	K	A	A	A	A	A	K	A	A
Va ve Se ect on/S z ng	A	K	K	A	S	S	S	S	A
Contro Loop nteraction	S	S	S	K	S	A	S	K	A
A gor m/Se ect on/Tun ng	A	S	S	S	S	A	A	K	A
Cont Sys Selection	A	S	A	K	S	A	A	K	A
Samp ng Rates	A	S	S	K	S	A	A	K	A
Ant -Al as ng	A	S	S	S	S	A	A	K	A
D g ta Transm tters	A	S	S	S	S	A	A	K	A
DCS D sp ay Ranges	A	S	S	K	S	S	K	K	A
Report by-Exception	A	S	S	S	S	A	A	K	A
Con Strat Des gn/Cod ng	A	S	S	K	S	A	A	K	A
Process Troub e-Shooting	A	A	A	A	S	A	K	K	A
Spott ng/Reduc ng Var ab'ty	A	A	A	A	S	S	K	K	A

References and Bibliography

1 B a kowsk W L *Dreams Versus Real ty A V ew From Both S des of the Gap* Keynote Address Contro Systems 92 Wh stler British Columbia 1992 Pub shed Pulp and Paper Canada 94 11 (1993)

2) EnTech™ Contro Va ve Dynamic Spec f cation (Ver s on 2 1 3/94)

3) EnTech™ Automatic Contro er Dynam c Specif cation (Vers on 1 0 11/93)

4) EnTech™ Competency n Process Control ndustry Recommendations (Vers on 1 0 3/94)

5) Taylor G *The Role of Control Va ves n Process Performance* 80th Annua Meeting Techn ca Sec tion Canad an Pulp and Paper Assoc ation Montrea (1994)

6) Dah n E B *Designing and Tuning Digital Controllers* nstrum and Contro Systems 41 (6) 77 1968

7) Moran M and Zaf r ou E *Robust Process Contro* Prentice Ha 1989

8 Kocurek M J Series Ed tor *Pulp and Paper Manufacture* Jo nt Textbook Committee of the Pulp and Paper ndustry Vol to 10

9) Stephanopou os G *Chemical Process Control An Introduction to Theory and Practice* Prentice Hal 1984

10) Coughanowr D R and Koppel L B *Process Systems Analys s and Control* McGraw-H 1965

11) Seborg D E Edgar T F and Mellichamp D A *Process Dynamics and Control* John W ey & Sons 1989

12) Frank n C F Powe J D and Workman M L *D g tal Control o Dynamic Systems* Add son-Wes ey 1990

13 Doy e J C *Analysis of Contro Systems W th Struc tured Uncerta nty* EE Proc Part D 129 242 1982

14) G bson J E *Nonl near Automatic Control* McGraw-H Book Company 1963

15) Box G E P and Jenk ns G M *T me Series Analys s Forecasting and Control* Ho den-Day 1976

16) Astrom K J *Introduction to Stochastic Contro Theory* Academ c Press NY 1970

17) Z eg er J G and N cho s N B *Optimum sett ngs for automatic controllers* Transact ons ASME pp 759-768 1942

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EnTech™

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 optimizing pulp and paper mill process variability. The full document is available upon request.

1.0 Competitive Marketplace – Background

The global market for manufactured products continues to focus on product quality and uniformity and more and more attention is being paid to process control equipment and its condition. EnTech™ Control Engineering Inc. has specialized 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 outside specified limits, or when the variability characteristics of the products adversely affect the operation of the customer's secondary manufacturing such as a printing press. Mill audits have shown[1] that product variability is caused mainly by the combined behaviour of the many upstream process variables. In many cases it is possible to trace these causes to individual control loops. Of equal importance is the impact of process variability on manufacturing efficiency in which small and subtle improvements in the process variability of key variables can have a large impact. Audits have identified that the cyclic behaviour of automatic control loops is a major cause for destabilizing production in the pulp and paper industry.

To date, only about 20% of all control loops surveyed were found to actually reduce process variability in automatic mode[1] while the remaining 80% of loops were found to increase variability. Of these, some 30% were found to oscillate directly due to control valve nonlinearities[2] and another 30% were found to oscillate due to controller tuning. It is recognized that controller tuning is to a great extent a human training issue (a separate issue from that of controller design). Most people are only familiar with the original controller tuning method known as Quarter Amplitude-Damping (published by Ziegler and Nichols in 1942[3]) or, they just "tune-by feel." These methods often produce loops which tend to cycle and destabilize product uniformity. By nature, trial and error methods accept the capabilities of existing controllers without question. For this reason controllers have not advanced significantly in dynamic performance terms over the years despite the major advances in control system implementation technology from pneumatic to electronic to digital with 64 bit microprocessors. In fact, the tendency has been for each new generation of controllers to mimic the functional capabilities of the previous generation notwithstanding the availability of the additional computational power of the new technology. For this reason most digital controllers are "analog like", in that they do only what analog controllers could do in the past. What of deadtime compensation, velocity mode control, feedforward control? On the other hand, as digital controllers have become prevalent, sampling rates have been relaxed and in some cases, little or no attention has been paid to signal aliasing. The result has been badly implemented digital control which has destabilized product uniformity for reasons of slow sampling and signal aliasing.

In the last decade, modern model based methods such as Lambda Tuning[4] and Internal Model Control (IMC)[5] have been used extensively in pulp and paper mills with considerable success in variability reduction. The use of these methods has "pushed" existing controller designs to the limit. For example it has been very disappointing to discover that the recommended tuning parameters based on these methods could not be

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entered into some commercially available controllers because they were outside the controller limits or that a particular feature of the modern control solution could not be implemented because it was not part of the available control algorithm

Summary of Controller Problems Encountered

The types of problems encountered included

- slow digital controller sampling rates
- lack of anti aliasing filters,
- lack of controller 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

Control in the pulp and paper industry is provided via a wide range of equipment including distributed control systems *(DCS's) single loop controllers, programmable logic controllers* (PLC's), variable speed drives, computer control systems and specialized, application oriented systems (e g bleach plant brightness control systems, turbine governors) This specification is aimed at providing a guide for controller design that is suitable to meet the needs of the majority of pulp and paper mill process control problems in which modern model based control solutions could be used to enhance competitiveness The specification is intended for any application where 'general-purpose process control' is attempted in which an "automatic feedback controller" is connected to a transmitter or sensor on the one hand and a final control element on the other (it excludes variable speed drive internal dynamics such as speed and armature current) The specification has two intended purposes as a tool to be used by pulp and paper companies when specifying and assessing the capabilities of control equipment and as a design guide for control equipment suppliers

2.0 The Pulp and Paper Control Problem

Virtually all pulp and paper mills use PID controllers for "analog type" or continuous control (this includes digital control with *adequately fast sampling rates) with 4 to 20 mA outputs to* control valves and other actuators Derivative action has been turned off in 98% of all cases Most commercial controllers have controller gain adjustments ranging from about 0 1 to 10 A small number of important control loops deal with motor driven or "velocity mode" type actuators such as electric valves refiner plates and headbox vertical slices These loops by nature use discrete or velocity mode algorithms with digital output logic to handle increase/decrease motor action. In addition, some loops have extensive deadtime and require deadtime compensation algorithms Other loops require feedforward control in order to achieve acceptable dynamic performance

3.0 Specification

3.1 Bump Tests

The controller will have a 'manual' mode in which feedback control will be suspended. Should safety considerations preclude this under normal operation an equivalent 'test mode' will be available 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

It will be possible to record both the controller output and the process variable as seen by the controller at the controller update frequency This could be done either via a file save software utility or via assignable digital to analog converters having at least 12 bit resolution

3.2 Gain Scheduling as a Function of Error

Whereas gain scheduling is a useful feature which controllers should have available, a means of scheduling the controller gain ("Kc", for ISA standard form and classical form PID) as a function of controller error will be provided This will allow a preselected gain suitable for minimum variance control to be used as long as the absolute error is below a low value. Above this value, a different gain suitable for more aggressive control can be selected *Care should be taken to avoid the possibility of a* limit cycle between these values For a parallel path PID implementation the same method of gain scheduling must be arranged for all three parameters

3.3 Control Structure – Controller Filter

To permit IMC type control structures which may require the cancellation of a process zero (β value) with a controller pole, there will be a first order filter implemented in series with the PID algorithm The filter can be on either the controller error (before the PID) *or after PID The filter must be properly initialized* (bumpless transfer) The transfer function of the filter will be

$$G_F(s) = \frac{1}{(\tau_F s + 1)}$$

The time constant τ_F is to be user selectable between 0 seconds and 60 minutes

3.4 Control Structure

The available control structures will include the following combinations

P, I, PI, PID PID (both real and complex zeros)

Each of these will have a series filter available (see 3 3 above)

The derivative filter τ_D will be nominally one tenth and no more than one eighth of T_D

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3.5 Alternative PID Structures

f a ternat ve structures such as "D" on PV and 'P" on PV are offered they must on y be as opt ons There must be a fu P D a ternat ve dr ven by contro er error

3.6 PID Parameter Ranges

Parameters (K_C, T_R, T_D K_I K_D, τ_F)
These must be adjustab e cont nuous y (not n steps) and ca brated to at east 2 s gn f cant f gures

K_C	0.001 to 20 (% output)/(% span), (ISA stan dard or c ass ca PID)
K_C	0 0 to 20 (% output) (% span) (Para e PID)
T_R	0 1 seconds/repeat to 10 hours/repeat
T_D	0 seconds to 60 m nutes,
K_I	0 to 50%/%/sec,
K_D	0 to 1500 %/% m nutes
τ_F	0 seconds to 60 m nutes

3.7 Process Variable Filter

The process var ab e w l have a f ter pr or to the error summ ng junct on of the form

$$G_{F_{PV}}(s) - \frac{1}{(\tau_{PV}s+1)}$$

The t me constant τ_{PV} should be user adjustable from 0 to 1 0 m nute

3.8 Feedforward Control

A dynam c feedforward contro opt on w be prov ded as per

$$G_{C_{FF}}(s) - K_{C_{FF}}\left(\frac{\tau_{Lead}s+1}{\tau_{Lag}s+1}\right).e^{-sT_{FF}}$$

The feedforward contro er output w be added to the output of the feedback contro ler w th due regard for bump ess transfer, m ts and reset w nd up The user se ectab e adjust-ments w be·

$K_{C_{FF}}$ 0 to +/ 100, τ_{Lead}, τ_{Lag} 0 to 10 m nutes
T_{FF} 0 to 10 m nutes

3.9 Setpoint Filter

There w be a prov s on to f lter the setpo nt through a f ter of the form

$$G_{F_{SP}}(s) \quad \frac{1}{(\tau_{SP}s+1)}$$

w th a user se ectab e t me constant from 0 to 100 m nutes

3.10 Deadtime Compensation

There w be a deadt me compensation a gor thm ava ab e of the form

$$G_C(s) = \frac{(\tau s+1)}{K_P(\lambda s+1-e^{-sT_d})}$$

Parameter values w l nc ude·
τ 1 second to 10 m nutes, K_p 0 01 to 50,
λ 2 seconds to 30 m nutes T_d· 0 to 10 minutes

3.11 Controller Output Integrity

The contro er output shou d not be a tered dynam ca y by rate m ters or other means

3.12 Sampling and Controller Execution Rates (T_S)

Analog signal samp ng and contro er execution shou d be 10 times faster, and no slower than three times faster, than the process time constants Process time constants faster than 0 5 seconds need not be cons dered in the "general" process control problem. Samp ng can be s owed down to one samp e per second when time constants are equal to three seconds or slower.

Ana og s gna samp ng and contro er execut on shou d be at 0 1 seconds/samp e However t can be s owed down to between 0 3 seconds for t me constants of 1 0 second (T_S τ/3) to as s ow as 1 0 second/samp e for t me constants of 3 seconds and onger

3.13 Anti-Aliasing

Ant a as ng f ters sha be present at every po nt of s gna samp ng, from raw ana og nput to the po nt of use of the feedback signal at the contro ler At each po nt of samp ing the ant a as ng f ter sha prov de a m n mum of 12 dB of attenuat on at the Nyqu st frequency of the next samp er

3.14 Bit Resolution

The m n mum acceptab e b t reso ut on for both nputs and outputs s 12 b ts Report by exception report ng shou d never be used nside a control oop

3.15 Motor Driven Actuators – Velocity Mode Control

A ve ocity mode P a gor thm shou d be ava ab e of the form

$$\Delta u_k - K_C(e_k \quad e_{k\,1}) + K_I e_k$$
$$- K_C(e_k \quad e_{k\,1}) + \left(\frac{K_C T_S}{T_R}\right) e_k$$

The "t me durat on" d g ta output hand er shou d execute at a m n mum update t me of 0 05 seconds Preferab y, t shou d execute at such a rate wh ch wou d a ow the f na actuator to ach eve a quant zation equ valent to about 12 b ts of reso u t on or 1 4000 Parameter ranges w l be as defined for P D parameter ranges above Output im t ng and w nd up ssues must be cons dered.

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3.16 Controller Documentation

.ontro er documentat on w l nclude the fo owing
- contro ler transfer function for each opt on and s gna path,
- parameter un ts and,
- parameter ranges ava ab e

Nomenclature and Symbols

a 0 or 1 exponent on ntegrator (0–none, 1 present n transfer funct on),
β – ead t me constant (+ve or ve),
e_k current error (setpo nt PV),
e_{k-1} ast error
$G_C(s)$ – contro ler transfer funct on,
$G_F(s)$ f ter transfer funct on,
$G_{FF}(s)$ feedforward transfer funct on,
$G_{F_{PV}}(s)$ process variable fi ter transfer funct on
$G_P(s)$ – process transfer funct on,
$G_{FSP}(s)$ setpo nt fi ter,
I Integral a gor thm,
k current t me ndex for d screte t me d fference equat on
K_C contro er ga n
K_{CFF} – feedforward ga n,
K_D control er der vat ve gain (para e P D),
K_I – control er ntegra ga n (para e P D)
K_P process ga n,
P Proport ona a gorithm,
PD Proportional-Der vat ve a gor thm,
PI – Proportiona ntegra a gorithm,
PID – Proport ona ntegral-Der vative a gor thm
P D F – Proportiona ntegral-Der vat ve a gor thm w th ser es f lter,
τ – process t me co stant,
τ_1, τ_2 – process t me constants
τ_F f ter time constant,
T_{FF} – feedforward deadt me compensation
τ_{PV} process var ab e f ter t me constant
τ_{Lead}, τ_{Lag} – feedforward ead/lag t me constants,
T_d – deadt me,
T_D der vat ve t me (standard and c ass ca P D),
τ_D der vat ve term f ter (paras t c po e) norma y set to $T_D/10$ or $K_D/10$,
τ_{PV} process var ab e f ter t me constant,
T_R reset or ntegra t me (standard and c ass ca P D)
T_S contro er samp ng t me
τ_{SP} – setpoint f ter t me constant,
Δu_k – current des red change n control er output,
ζ damp ng coeffic ent

References

1) B a kowsk , W L , *Keynote Address, Control Systems 92* Conference, Wh st er, Br t sh Co umb a, Sept 1992 to be pub shed n Pulp and Paper Canada
2) EnTech™ Co tro Va ve Dynam c Spec ficatio (Vers o 2.0 7/93)
3) Ze g er J G and N cho s, N B., *Optimum settings for au tomatic controllers*, Transact ons ASME pp 759 768 1942.
4) Dah n, E B *Designing and Tuning Digital Controllers* n strumentation and Contro Systems 41 (6), 77, 1968
5) Moran M and Zaf r ou E *Robust Process Control*, Prent ce Ha 1989

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Digital Measurement Dynamics – Industry Guidelines

(Version 1.0, 8/94)

Keywords: sensor, transmitter, sampling, sampling rate, aliasing, filtering, anti-aliasing, delay

Note: In this document the words sensor and transmitter are used interchangeably to mean a measurement device which makes a calibrated signal available.

1.0 Competitive Marketplace Background

The global market for manufactured products continues to focus on product quality and uniformity and more attention is being paid to process control equipment and its condition. EnTech Control Engineering Inc. has specialized 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 outside specified limits or when the variability characteristics of the products adversely affect the operation of the customer's secondary manufacturing such as a printing press. Mill audits have shown [1] that product variability is caused mainly by the combined behaviour of the many upstream process variables. Audits have identified that the cyclic behaviour of automatic control loops is a major cause for destabilizing production in the pulp and paper industry and that sometimes this is caused by the measurement equipment.

Digitally implemented sensors and transmitters together with digital signal transmission, in which inappropriately designed signal sampling and signal conditioning especially have resulted in signal aliasing and a loss of signal integrity. If measurement data is to be useful for feedback control and trouble shooting [2] there are fundamental principles of dynamics to which the measurement process must conform.

As the process industries including pulp and paper, prepare for the advent of Fieldbus [3] technology where all field devices transmitters and valves will be digital, it is even more important to ensure a high level of basic measurement and signal integrity.

Summary of Digital Measurement Problems Encountered in Mill Audits

- slow digital sampling rates,
- inability to change sampling rates,
- signal aliasing
- no anti-aliasing filters,
- inability to change anti-aliasing filter settings
- low bit resolution
- non linear filtering & signal conditioning,
- excessively slow digital signal transmission
- digital transmitter time delays
- inability to access data via proprietary digital protocol
- inability to access the measured signal ahead of signal filtering
- inadequate signal conditioning documentation

Digital Measurement Dynamics – Recommendations – Purpose

Control in the pulp and paper industry is accomplished via a wide range of digital equipment including distributed control systems (DCS's), single loop controllers, programmable logic controllers (PLC's), variable speed drives, computer control systems and mill-wide historical data archiving systems. All of these systems communicate digitally. In the near future, most transmitters and control valves will also communicate digitally with all of the above equipment via the Fieldbus protocol [3]. Thus there is an even greater need to ensure that any problems related to digital signal processing are avoided.

These guidelines define the minimum requirements needed to provide the pulp and paper industry with measurements which are indeed useful as mills strive towards more efficient manufacture of higher uniformity product. This document is aimed at providing a design guide for sensor, transmitter and digital signal transmission equipment in order to ensure that the designs are capable of meeting the needs of process control, trouble-shooting and data acquisition, especially at a time where there are increasingly urgent needs in the pulp and paper industry for greater product uniformity and operating efficiency. This document has two intended purposes: as a guideline for pulp and paper companies when specifying or buying process measurement and data transmission equipment and as a design guide for equipment suppliers to the pulp and paper industry.

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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 Newtonian laws governing physical dynamic systems and can be modeled as a linear system with low-pass filter dynamics and additive noise. Most transmitters and sensors produce an analog output signal, typically 4 to 20 mA. This signal is essentially linear. Its bandwidth is limited by the low pass filtering of the measurement device, plus any internal filtering which in most cases is adjustable. This signal is suitable for both control and trouble shooting.

Digital Signals

There is no doubt that digital systems are flexible, powerful and here to stay. However, there are a number of important properties of digital signals which are sometimes overlooked, yet which are important for both process control and process trouble shooting. These relate to signal quantization, sampling rates and data transmission delays. If any of these are badly chosen, the resulting signal may be useless for trouble-shooting and even potentially hazardous for control. In addition, there are possibilities for altering the properties and character of the signal by aliasing, nonlinear filtering or report-by-exception. These may also render the signal useless for both control and trouble-shooting.

Signal Requirements for Control

Control algorithms such as PID are essentially linear. Control in the presence of nonlinear elements becomes very difficult and imposes very real limitations on control performance. The best approach is to eliminate all possible nonlinearities. Measurement nonlinearities should be avoided if possible.

Digital sampling rates for control should be a function of the process dynamics. For the pulp and paper industry, sampling rates should typically be 0.3 seconds per sample or faster for most common variables [4]. Digital sampling and control loop execution must be coincident (control must use the latest sample) and must also be evenly spaced in time.

Signal Requirements for Process Trouble-Shooting

Process trouble shooting requirements are more demanding than those for control, as the nature of the problem cannot be pre-defined in advance. Trouble-shooting involves being able to find the cause of some process or control related behaviour – say a cycle, an intermittent spike, or a particular type of noise. It may be necessary to examine a particular sensor output at sampling rates much faster than those used for control (typically 10 to 100 times faster). Also it may be necessary to examine a sensor output without any filtering at all in order to see if there is any evidence of a particular type of disturbance characteristic present. Both of these requirements are very different from those of normal control, yet it may be necessary to make these special measurements without disturbing the normal operation of the control system.

3.0 Digital Signal Properties

Digital signals have two important properties – quantization and periodic sampling. If either of these are badly chosen, they can make the signal of little use.

Quantization

All digital signals are digitized by some form of analog-to-digital conversion (A/D). In all cases 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 also limits the usefulness of the signal for both control and diagnostics in the pulp and paper industry. 12 bit resolution is usually adequate.

Sampling Rate

The act of sampling at some rate T_s limits the bandwidth of the signal to an upper frequency known as the Nyquist frequency f_N.

Where

$$f_N = \frac{1}{2T_s}$$

The Nyquist frequency is based on Shannon's Sampling Theorem and requires at least two samples per sine wave period in order to be able to reconstruct the sine wave frequency from the digital samples (the ability to reconstruct the amplitude accurately requires many more samples).

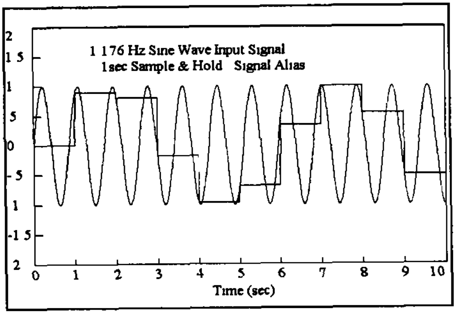

FIGURE 1 – EXAMPLE OF SIGNAL ALIASING

Signal Aliasing

If the original analog signal contains frequency content beyond the bandwidth of the sampler (beyond f_N), then signal aliasing will occur. Aliasing is the creation of a fictitious signal at slower frequency. **A signal alias does not actually exist in the original data. An aliased signal is absolutely accurate at each sample, but the mistake made is that the samples are not joined by straight lines.** Figure 1 shows an example of signal aliasing in which a 1.176 Hz sine wave is sampled via a sample-and-hold device at 1.0 seconds per sample. Shown are 12.5 cycles of the 1.176 Hz signal. The sampled signal alias is seen as another sine wave with a period of about 6 seconds. In this example T_s is 1.0 seconds. Thus the Nyquist frequency for this sampler is 0.5 Hz. As the input signal frequency is 1.176 Hz, which is faster than the Nyquist frequency, aliasing results.

Whereas the Nyquist frequency defines the limit at which aliasing occurs (Shannon's Sampling Theorem), some signal distortion and loss of integrity starts to occur at sampling rates which are much faster. Distortion starts to be visibly evident when the sampling rate is ten times faster than the signal content.

The ultimate danger of allowing aliasing to occur is that the control system will attempt to control out the alias signal components. If it does so, it will de-stabilize process and product uniformity as a result. Equally important is the fact that no one will be aware that this is happening.

Figure 2 illustrates the importance of providing proper anti-aliasing for control purposes. This is a dynamic simulation example which illustrates a unit step change in a process variable which has a first order time constant of 3 seconds. The step change is initiated after 4 seconds. The full change can be observed after four time constants or another 12 seconds, at second 16 of the illustration. The raw signal however is corrupted by a sine wave of 4.04 Hz and an amplitude of 1.0. There are four signals shown: 1) the raw signal, 2) the actual process change, 3) the raw signal sampled at 0.25 seconds with no anti-aliasing, and 4) the raw signal sampled at 0.25 seconds after a first order anti-aliasing filter of time constant 0.325 seconds (t_f = 1.3 T_s). The signals have been displaced vertically for clarity. It is evident that if signal 3) the raw signal sampled without an anti-aliasing filter, were used for control, that the resulting control action would destabilize the process severely for a completely fictitious reason.

Once a signal has been aliased, there is no way to retrieve the true signal, no matter how many times the aliased signal is filtered or

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sampled. **The only solution is to prevent signal aliasing from occurring.**

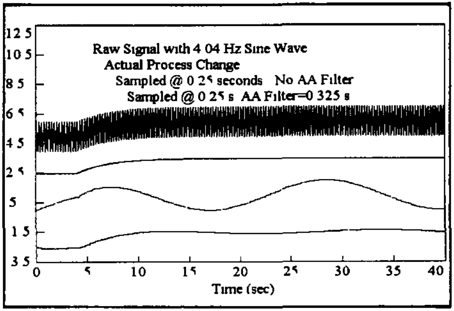

FIGURE 2 – THE IMPACT OF SIGNAL ALIASING ON CONTROL

Random Variability
The previous examples of signal aliasing used single sine waves to illustrate the point. In practice, process signals are usually corrupted by noise which tends to be random in nature (white noise). Mill trouble-shooting and auditing experience makes heavy use of time series analysis [1] in which Fourier analysis and power spectra are common tools. These techniques are key to measuring and understanding variability and its impact. The underlying principle of these techniques is that any random signal can be decomposed into its harmonic content, hence many sine waves existing concurrently and each at a different frequency. The impact of aliasing can be readily understood. Each harmonic which is faster than the Nyquist frequency will produce its own unique signal alias. This phenomenon is known as "fold-over" and involves the corruption of the power spectrum as a result of aliasing.

Anti-Aliasing
There are two ways to prevent aliasing. One is to sample faster than the fastest frequency components of the input signal, if these are known — usually they are not. The other is to install an anti-aliasing filter before the sampling device thereby filtering out all of the signal content which is faster than the Nyquist frequency of the sampler at the chosen sampling rate, hence preventing aliasing from occurring. The anti-aliasing filter should provide a minimum of 12 dB of attenuation at the Nyquist frequency. This can be provided by a first order low pass filter of time constant T_F = 1.3T_S. It can also be provided by an arithmetic averager with T_A = 4T_S (T_A = averaging time). The 12 dB recommendation represents a bare minimum requirement. Where possible, more effective anti-aliasing should be considered by employing a filter with a sharper cut-off and more attenuation at the Nyquist frequency.

Sub-Sampling
In digital control systems, not only are analog signals sampled to produce digital versions of the input signals, but the digital signals are usually sub-sampled many times later on. For instance, in modern DCS's the input sub-system which does the A/D conversion may run at say 100 ms while the controller sub-system may have variable sampling rates depending on the individual needs of each controller. Typical controller execution rates may be 1 second. Aliasing can hence occur between the input sub-system and the controller even if aliasing was prevented at the input sub-system. **It is the final sampling rate which governs aliasing.**

In DCS's there are many other uses for the data. For instance, the process values from the controller will be sampled again to update the operator's console. Typically this is done at a slower rate than the controller update frequency. The mill-wide archiving system will also sample the same data at a much slower rate, typically once per minute. Other control loops may sample the data at slower sampling rates for feedforward compensation. All of these operations are subject to data aliasing. **Anti-aliasing filters are required before each sampling operation** and must be matched to the sampling rate. **This is usually neglected in most systems today.**

4.0 Digital Signal Transmission

Signals can be transmitted digitally via data highways, data links, RS-232, Fieldbus, Ethernet, LAN's, protocol converters, gateways and other communications links and devices. All of this communication processing should be completely transparent to the control and trouble-shooting applications by providing 100% reliable on-time data transfer. This means that each data sample is transferred at the agreed-upon sampling rate without the introduction of noticeable time delays (say less than 5% of sample period).

Report-by-Exception
To prevent data highway overload, many digital systems employ report-by-exception whereby a signal is only updated via the data highway if its value has changed by at least the value of a selectable deadband. Often such deadbands are set at 1%. **This seriously limits the signal quantization and should never be used for control or trouble-shooting.**

5.0 Digital Sensors and Transmitters

Digital sensors and transmitters are becoming increasingly prevalent and are flexible, useful and here to stay. These are primarily micro-processor based devices, often referred to as "smart transmitters" which provide many additional useful functions such as onboard diagnostics and remote re-spanning and re-calibration features.

Digital sensors and transmitters sample their primary measurement elements internally prior to processing this data into a calibrated and filtered signal. Many sensors and transmitters retransmit this information as an analog signal (typically 4 to 20 mA). It is important to recognize that this "analog" signal is a retransmission of a digital signal, hence is capable of containing aliased information. Most transmitters in the future will transmit their output digitally on Fieldbus.

Some digital sensors and transmitters use primary measurement principles which are inherently analog in nature such as a force balance system. In such sensors, the analog measurement system itself can be considered as the anti-aliasing filter ahead of the A/D converter. The sampling rate of the A/D should be chosen in such a way that these measurement dynamics are considered as part of the anti-aliasing filter. Other sensors are inherently digital in nature. For instance, a pulsed-DC magnetic flow meter takes readings at 30 Hz typically. These readings are digital samples of the flow through the meter. There is no inherent anti-aliasing mechanism possible except for the fluid residence time in the flow meter volume which provides a natural averaging of the fluid velocity over the measurement volume. It is important to understand under what conditions this type of physical process might provide adequate anti-aliasing. Another example of a digital sensor is a paper machine infrared moisture sensor with a spinning optical filter wheel which measures light samples every rotation of the wheel. Clearly, such a sensor has no inherent mechanism by which aliasing can be prevented since each sample represents a separate "spot" on the paper, each one potentially at a different moisture value. Questions such as these need to be posed when such sensors are used for feedback control.

The micro-processor execution rate often determines the sampling rate of a digital sensor. Choice of sampling rate has serious implications for feedback control and even greater consequences for trouble-shooting. Some digital transmitters also introduce measurement "latency" or time delay. This introduces a measurement time delay or deadtime and may have very serious consequences for control. This should be limited to less than 5% of sample period.

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6.0 Fieldbus Implications for Digital Sensor/ Transmitter Design

Once F e dbus has become a rea ty process control system arch tec ture w undergo a major change A users w have access to the source of measurement information at the transm tter However each user has d fferent needs with regards to sampling rate and data ntegrity Contro requ res anti-a ased data every 0 3 seconds typi ca y The operator's conso e may requ re anti-a ased data every second, but on y when th s measurement s se ected on the screen The trend package may want anti-a ased data every 10 seconds The m -w de arch v ng package may want th s data together w th a data po nts n the m every m nute Meanwh e the troub e-shooter may want the anti-a ased s gna every 10 m seconds but on y for a short burst extend ng over a few m nutes dur ng wh ch time a mu ti-var ab e tr a s be ng run Alternatively the troub e-shooter may a so want access to the raw sensor data, shou d s gna a as ng be suspected

7.0 Sensor/Transmitter Requirements – Fieldbus Multi-Channel Design

Cons der ng F e dbus sensor/transm tter des gn and a of the needs and requ rements stated prev ous y the sensor/transm tter des gn for the future shou d nc ude the fo ow ng features

Field Measurement Verification Capability

The ab ty to nterrogate and ver fy the ntegr ty of a measurement s of paramount mportance t often happens dur ng process troub e-shoot ng that the ntegr ty of a measurement s brought nto quest on and steps must be taken to verify the measurement by compar ng the s gna to s gna s gathered from other sensors Such nvest gat ons re y on hav ng ana og s gna s ava ab e n order to have access to the fu sensor bandw dth Sensors wh ch are nherent y d g ta can on y prov de the s gna at the fastest samp ng rate ava lab e To prov de th s capab ty the sensor shou d have the fo ow ng term na s ava ab e n the f e d w th the fo ow ng s gna s access b e

1) An ana og output s gna (4 to 20 mA or 1 5 V) wh ch has not been f tered or d g t zed and wh ch represents the measurement n as near a ca brated form as poss b e Th s s gna can on y be obta ned from an nherent y ana og sensor
2) nherent y d g ta sensors shou d prov de an ana og output (4 to 20 mA or 1 5 V) of the s gna at the fastest sampl ng rate ava ab e (v a a D/A)
3) There shou d be a F e dbus connector ava ab e n the f e d so that any of the F e dbus s gna s can be assessed as we

Fieldbus Data Access

Data access on F e dbus shou d take nto account the mu t p c ty of users for each measurement at d fferent samp ng rates w th ant a as ng Th s can be done by prov d ng the unf ltered s gna at the fastest samp ng rate ava ab e say 10 m seconds together w th anti-a ased s gna s at a number of frequently used samp ng rates, say 10 m seconds 30 m seconds 100 m seconds 300 m sec onds, 1 second 10 seconds and one m nute There may be other preferred samp ng rates A ternat ve y, t may be poss b e to des gn the system "hand-shake" software to accommodate and recogn ze the spec f c samp ng and ant a as ng needs of each user At any rate once the specific samp ng rates have been estab shed, the sensor software must run the appropr ate samp ers and f lters to support and susta n the measurement at a of the samp ng rates that have been estab shed

8.0 Recommendations Summary

The fo ow ng recommendations apply to most of the d gita measurements n the pu p and paper ndustry

Sampling Rates

Process Trouble shooting: 10 to 100 m seconds/samp e

Process Control: 0 3 seconds/samp e or faster, 1 second/samp e can be to erated for s gna s with time constants of 3 seconds and s ower

Operator Console: Typ ca y 1 second update

Trending: Typ ca ly 1 second down to 10 second update

Archiving: Typ ca y 10 second down to one m nute update

Time delays (Measurement Processing or Data Transmission) Less than 5% of samp e update per od

Anti-Aliasing A m n mum of 12 dB of attentuat on at the Nyqu st frequency of each samp er dev ce

Filtering

Sensor nterna f ter ng shou d be ad ustab e beyond the needs of anti-a as ng The m n mum f lter time shou d be sma compared to the expected process time constant The f lter ng emp oyed shou d be near

Quantization

The m n mum acceptab e s gna quant zat on s 1 4096 or 12 bits equ va ent

Measurement Data Access.

Field Access

Each sensor shou d prov de f e d access v a an ana og output port to the ear est poss b e s gnif cant s gna Pure y d g ta sensors shou d prov de ana og output of the raw s gna There shou d a so be f e d access v a a d g ta port a ow ng fu access to a sensor outputs

Digital Output

D g ta sensors shou d support a of the samp ng rates potent a y needed n a typ ca m nsta ation as a set of mult p e channe s Each channe shou d provide the anti-a as ng f lter needed at the samp e rate n question

Documentation:

The documentat on for each sensor shou d c ear y exp a n the fo ow ng ssues

- measurement pr nc p e
- ca brat on procedure
- s gna path dynam cs from the raw measurement through to the sensor output shou d be spec f ed as a near transfer funct on
- samp ng rate or range of samp ng rates
- anti-a as ng process and attentuat on at the Nyqu st frequency
- s gna quant zation

Nomenclature and Symbols

A/D	ana og to-d g ta converter
dB	dec be Amp if cat on/attenuation expressed as 20 Log (Amp tude Ratio)
D/A	d g tal-to-ana og converter
DCS	D str buted Contro System
f_N	Nyqu st frequency the cyc c frequency at the samp ng rate
PID	Proportional-Integra -Derivative a gorithm user for contro
T_A	averag ng t me
T_S	contro er samp ng t me

References

1 B a kowsk W L *Keynote Address Control Systems 92 Conference* Wh stler British Co umb a Sept 1992 To be pub shed n Pu p and Paper Canada
2 B a kowsk W L and We don A D *The D g tal Future of Process Control its possib l t es limitations and ramific at ons* TAPP Process Contro Sympo s um, New Or eans Feb 1994 To be pub shed n TAPP Journa Vo 77 No 10 Oct 1994
3 Tapperson Gary *Evolution to Fieldbus the new reality of d gital connec tivity* Proceed ngs of the TAPP Process Contro Conference 1993 Nashv e TN
4 EnTech *Automatic Controller Dynamic Specification (Vers on 1 0, 11/93)*

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EnTech

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 Inc has specialized 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 outside specified limits, or when the variability characteristics of the products adversely affect the operation of the customer's secondary manufacturing, such as a printing press. Mill audits have shown that product variability is caused mainly by the combined behaviour of the many upstream process variables. In many cases it is possible to trace these causes right down to individual control valves. Of equal importance is the impact of process variability on manufacturing efficiency, in which small and subtle improvements in the process variability of key variables can have a large impact. Audits have identified that the cyclic *behaviour of control loops is a major cause for destabilizing pulp and paper production. To date about two-*thirds of all control loops surveyed were found to oscillate in automatic mode. Of these, well over half were found to oscillate directly due to the behaviour of pneumatic control valves. ***The undesirable behaviour 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 aimed at providing a guide for control valve dynamic behaviour that is suitable for the highly competitive process industry environment. It is intended as a guide for end users to gauge control valve performance in situ and also as a guide for both control valve purchase and sizing. While by no means 'the-last-word' in competitive control valve performance, the specification is intended to fill a void in an 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 originated to provide the end user with a 'how-to' guide for in-situ testing. Version 2.1 is a minor revision to Version 2.0 intended to stress three points which have caused some confusion. a) The Backlash/Stiction specification is by far the most important part of Version 2.1 and it is critical that the process variable moves in response to small input step changes (1% and less) indicating that an actual change in the valve Cv has occured. Valve stem movement alone is not an adequate indication. b) There is a need to clarify terminology as used in both Versions 2.0 and 2.1 regarding backlash/stiction. 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 Version 2.1 to imply "deadband". c) Finally, *ISA Specification S75.13 [3] is not equivalent to Version 2.1. The former applies to a test stand procedure* for valve positioners 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.

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2.0 Control Problem Definition - Linear Control

Most control valves are used as final control elements in simple feedback control loops with PID control algorithms. The feedback controller is essentially a linear device. This implies that if all elements of a control loop were also truly linear there would be few control problems. What does 'linear' mean in the control context? A linear dynamic element will respond to its input signal in a uniform fashion over its operating range with unchanging dynamics regardless of the size or amplitude of the change in the input signal. Hence, if the time constant of the step response is 2 seconds, then this will apply to steps of 0.01%, 1% and 100% equally well. If this is not true, then the system is nonlinear. In fact, all control valves are highly nonlinear. This is a major source of problems for control loop tuning and control loop performance because the apparent process dynamics to which the control loop tuning is matched keep on changing. As a result, the loop no longer works as intended.

Definitions:

Process dynamics: the way a process variable responds over time, to a change in controller output. This is best characterized by the step response, modeled by a transfer function and expressed in terms of Process Gain, Time constants and Deadtime.

Process Gain: the ratio of process variable change to controller output change after a step change in controller output.

Time constant: the time required for a first order process to reach 63% of the full change when a step change has been made.

Deadtime: the time after a step change prior to the start of the process reaction taking place.

Linear Dynamics: a single set of dynamic parameters (gain, time constants) applying equally well over the whole operating range.

Nonlinear Dynamics: dynamic parameters (gain, time constants) which change over the operating range.

Limit Cycle: a sustained control loop cycle or 'hunting' caused by a nonlinear element in the loop. Limit cycles cannot be corrected by loop tuning.

3.0 Control Valve Nonlinearities

Control valves have several forms of nonlinear behaviour which can be split into two categories:

- valve tracking nonlinearities,
- sizing and flow characteristic nonlinearities.

Valve tracking nonlinearities include behaviour such as 'stiction' and are primarily a function of valve design, shut-off requirements, actuator sizing, valve wear, maintenance, age, installation and other related factors. The common practice of demanding tight shut-off for a control valve often compromises control performance by greatly increasing the plug/seat friction and resulting stiction forces. In the future, high performance control may require control valves which do not have tight shut-off, hence requiring a separate additional shut off valve.

Valve sizing is generally a function carried out during the engineering phase of a project and results in the 'installed process gain characteristic' of the control loop. Ideally an installed loop process gain of 1.0 is desirable for self regulating processes, as it allows for full valve travel over the full span of the transmitter. However, in many cases valves have been severely oversized in the quest for more than adequate production capacity. This results in control valves with very high process gains which are nearly closed during normal operation. The normal operating region for these control valves is hence very narrow. The valve tracking nonlinearities can be magnified to the extent that good control is impossible.

Flow characteristic nonlinearities are a function of valve plug characteristics and the pressure drop distribution between the control valve and the rest of the process piping, pumps and other equipment. The impact of this is that the installed process gain will vary in a certain way over the operating range of the process. These process gain changes further complicate the control loop tuning and performance problem since a fixed gain controller cannot cope well with these changes.

3.1 Control Valve Tracking Nonlinearities

Valve tracking nonlinearities are caused by problems in the positioner/actuator/valve system with the result that the valve plug does not follow the input signal in a linear fashion. There are three regions of operation (referred to as A, B and C in this document). For a very small input signal step change (say 0.1%) the plug is not likely to move at all (Region A). Above an initial threshold (say 0.1% to 1%) there will be motion (Region B) but this will be subject to backlash or hysteresis or lost motion in the mechanical linkages together with stiction or transitions from static to dynamic friction causing sudden jumps in the plug position as successive step changes are called for. This behaviour (Regions A and B) is especially destabilizing because it causes the controller output to wind up while there is no response to the small actions requested. When motion does finally occur after stiction release, (stored potential energy release) the result is often an overshoot in the plug position. This overshoot can be even more damaging to control performance than the original stiction behaviour. When the input steps get much larger than the initial threshold, then a second threshold is reached (steps of say 10%) at which point the valve plug will respond to every change called for (Region C), but the motion will be velocity limited by the rate at which power air can flow to and from the actuator. In this region the valve is nonlinear in a different way because the dynamics get progressively longer with the size of step change. For process control regulation in steady state, the normal and most important region of operation is for step changes below this second threshold (Regions A and B). In this region, the problem with pneumatic control valves are nonlinearities of the backlash/hysteresis/stiction/overshoot type which may often cause the control loop to limit cycle.

3.1.1 Backlash/Stiction

Backlash and stiction behaviour is caused by mechanical tolerances and friction in the control valve/positioner/actuator system. The resulting limit cycles can only be corrected by reducing these. They cannot be addressed through loop tuning.

Specification

The combined Backlash and Stiction or Deadband is not to exceed 1.0%, as installed, during normal process operation. [reference 2].

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In-Situ Testing for Backlash/Stiction

In-Situ testing for backlash/stiction 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 signals at the terminal strip. Since the step changes for this test sequence will be small and the process signal is likely to be noisy, an analog input filter with a time constant of a few seconds is advisable. Set up to collect data at a rate of about 1 sample per second and start data collection (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 variability change from 'auto' to 'manual'.
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 as near to a clean step as possible. This is best done by entering a new value through a keyboard, rather then using a slew button. If a slew button or manual output adjustment must be made by hand, this should be a single move of approximately the right size, made as rapidly as possible. Start with a series of 0.5% bumps to begin with in one direction (either up or down). If no response is observed increase the size of the bump tests being used. When equal process reactions are noticed for two successive bumps, then the backlash has been taken-up in that direction. Reverse the direction of the bumps and repeat the test in the opposite direction. Estimate the backlash by counting the number of bumps after a direction reversal for which no response or only a partial response occurred. For improved accuracy, repeat the test procedure with smaller bump tests. It is important that these bumps be less than the initial motion threshold (Region A) in order to properly measure backlash and 'stiction'.
4) For the smallest bump test used in 3) above (say 0.2%), initiate a series of bumps in one direction and continue in that direction to observe the effects of combined backlash and 'stiction'. For small bumps the valve will not move initially; it will then move for the combined effect of all previous bumps since the last move. Once it has moved and stopped it will again be 'stuck' and several more bumps will be necessary to cause another move. Record the input signal and the process variable (PV) initial and final values in a table for each bump. Continue the test over at least two 'stiction' cycles as described above. Then reverse the direction by 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 initial value at the start of the test. Draw an XY plot of PV vs input signal similar to that of Figure 1. Estimate the combined backlash/stiction or deadband by estimating the maximum width of the resulting hysteresis plot as illustrated in Figure 1. Repeat the test, if necessary, to obtain a better estimate.

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 at least

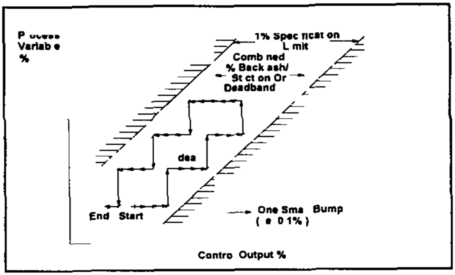

Figure 1

the speed of response listed in Table I and illustrated in Figure 2, over a range of step changes ranging from 10% of travel down to the combined backlash/stiction limit achieved above plus 1%. (Down to 1.5% if combined backlash/stiction is 0.5%)

Definitions:

Td	deadtime before any response is visible (seconds)
T63	time to reach 63% of a step change in input signal (seconds)
T98	time to reach 98% of a step change in input signal (seconds)
BW	bandwidth (Hz) the frequency at which the plug or trim position amplitude is attenuated by 6 dB. Bandwidth has been included in the specification for completeness even though it is difficult to test for in most cases

Speed of Response

Table I

Valve Size (inches)	(cm)	Td (sec)	T63 (sec)	T98 (sec)	BW (Hz)
0-2	0-5	0.1	0.3	0.7	1.6
>2-6	>5-15	0.2	0.6	1.4	0.8
>6-12	>15-30	0.4	1.2	2.8	0.4
>12-20	>30-50	0.6	1.8	4.2	0.27
>20+	>50+	0.8	2.4	5.6	0.2

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 2. Responses for larger valves are identical, except that time is rescaled appropriately.

Overshoot

Overshoots will be present if the step responses do not increase monotonically but rather overshoot and then 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 backlash/stiction limit plus 1%.

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In-Situ Testing for Speed of Response and Overshoot

In Situ testing for speed of response is very similar to that for backlash/stiction. The differences are as follows:

1) The data collection equipment should collect data at a rate which is at least 5 times faster than the expected time constant of the valve, typically 0.02 to 0.03 seconds. The input filtering 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 magnetic flow 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 position transmitter, to measure as close as possible to the valve plug position. Alternatively, it may be possible to measure line pressure some distance after the valve, or the differential pressure across the valve itself. In all cases the time constant of the measurement must be at least three times as fast as the expected time constant of the valve. (Table I, T63/3. For a 4 inch valve, time constant is 0.2 seconds, hence the measurement time constant should be 0.06 seconds)
3) Initially, steps of 10% of travel should be done. Plot the process variable response as in Figure 2. Calculate the % overshoot on the basis of the final process value. Tests can vary in direction to suit process conditions at the time although small steps should be done in the same direction so as not to incur backlash. Now repeat the tests for successively smaller steps, down to steps equal in size to the previously achieved combined backlash/stiction limit plus 1%. Measure Td, T63, T98 and % overshoot for each test. The results should comply to Table I and Figure 2.

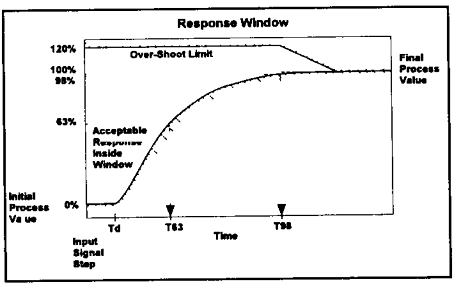

Figure 2

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 nonlinear flow characteristics. This results in the fact that the 'process gain', the relationship between input signal change and the resulting process change, itself changes with the operating point. The process gain combined with the controller gain determine how a control loop behaves. These changes in process gain are a major source of loop tuning problems and can often result in loop instability. Oversizing the valve makes this problem much more severe. To minimize this effect, the overall process gain for non-integrating process variables (eg. most flows, pressures, temperatures, but not tank levels) as determined by both valve sizing and transmitter span should have a nominal value of 1.0 and be kept in the range.

0.5 - 2.0 (% of span) / (% output)

eg. 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 signal span
2) **Speed of Response:**
   rate of change of valve position in response to step changes ranging in size from 10% in input signal (see Table I 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 - 2.0 %/%

References

1. EnTech™ Control Valve Dynamic Specification (Version 2.0, 3/94).
2. ANSI/ISA S51.1 1979 Process Instrumentation Terminology
3. ISA Standard S75.13 1989 Method of Evaluating Performance of Positioners with Analog Input Signals and Pneumatic Output

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