OPTIMIZING PAPER MACHINE DRYER CONTROL - A CASE STUDY

Doug Nelson, **EnTech Control Engineering, Toronto, Ontario**
Tom Gardner, **Gardner Systems Corp, Neenah, Wisconsin**

ABSTRACT:
The tuning of paper machine dryer section control loops represents one of the most complex tasks facing the paper machine control engineer. Improper tuning frequently causes substantial increases in machine downtime and product moisture variation, and is often responsible for chronic flooding of dryers on high speed machines. Good process design and control strategy design are shown to be fundamental requirements for successful tuning and control. The primary objectives are to minimize the non-linearities inherent in the process and to develop a tuning strategy that will optimize dryer section performance. Computer simulation can play a key role in achieving these objectives.

INTRODUCTION:
Approximately 200 paper machine audits have been conducted in the last decade. In the case of dryer section controls, tuning is difficult and the results are often inconclusive. This is because of the complex dynamics of the dryer pressure loop, the potential for wide variation in these dynamics, the built-in non-linearities which result from the split ranging of outputs and equipment characteristics which sometimes magnify the controller interactions. There is clearly a need for fundamental issues to be addressed.

In the first place there is little appreciation at the design stage for the dynamics of the dryer system, which commonly results in poor selection of control equipment. The control strategy typically defaults to traditional pressure and differential control without recognition of the dynamic consequences. Also, the amount of instrumentation provided is normally inadequate for the purposes of diagnostics and optimization. Furthering the dilemma, mill instrument crews rarely have the training required to tune loops with complex dynamics and the result is guesswork tuning which rarely produces successful results. Even experienced control engineers familiar with dynamics and control algorithms may not be able to overcome the dynamics of a poorly designed system.

Accordingly, the focus here is threefold: a survey of process design practice as it relates to controllability, the impact of control strategy on control performance, and controller tuning strategy and method. The work presented in this paper utilizes a case study approach, basing the process on the main section of a newsprint machine. Much of the data presented was generated with dynamic simulation, an indispensable tool for optimizing this complex, non-linear system.

PROCESS DESCRIPTION:
Figure 1 illustrates a standard process and control strategy for a main dryer section of a newsprint machine in which the blow-through steam is recirculated by a thermocompressor (TC). For the purposes of this case study, the section contains 14 dryer cans with equipment sized as indicated, machine speed set at 975 m/min, the dryers operating from 200 to 590 kPa (15 to 70 psig) with differential pressure in the range of 42 to 83 kPa (6 to 12 psid). Makeup steam, which is the primary source of steam, is supplied at 620 kPa (75psig). High energy motive steam to the TC is supplied at 1137 kPa (150 psig).

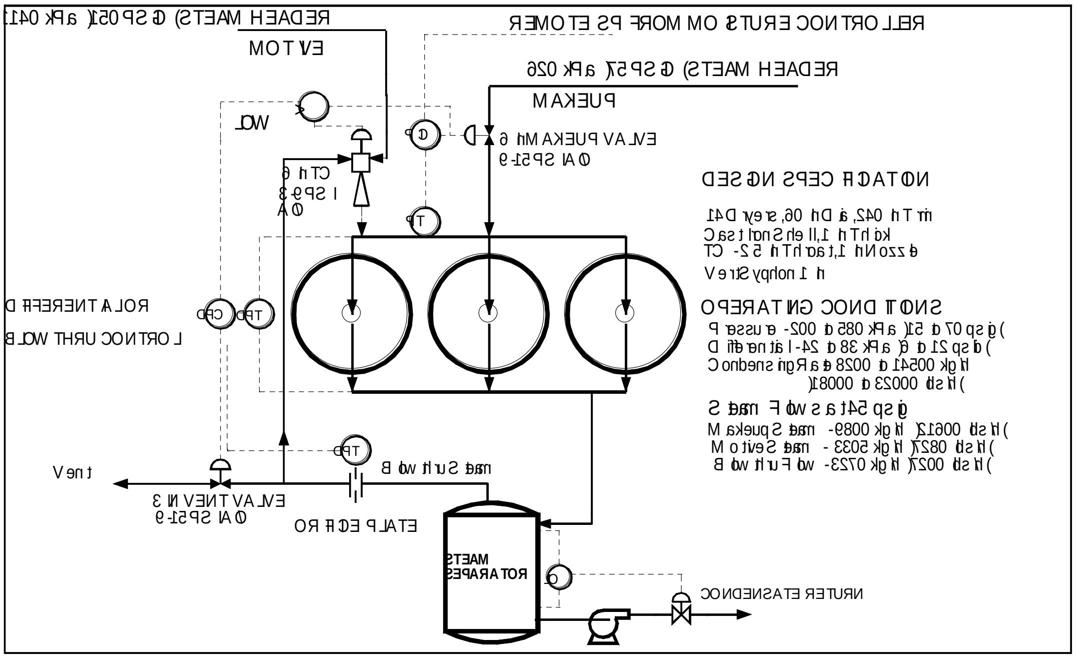

Figure 1 Main Dryer Section Of Newsprint Machine

There are three control loops employed in this system; Moisture control, Pressure Control, and either Differential or Blow-Through control.

Moisture Control Loop

The controller cascades a setpoint to the Pressure control loop. The dynamics of the Moisture control loop can be characterized adequately with a first order plus deadtime process model. The process time constant is governed by the thermal inertia of the dryer cans, but is also dependent on the tuning of the pressure controller.

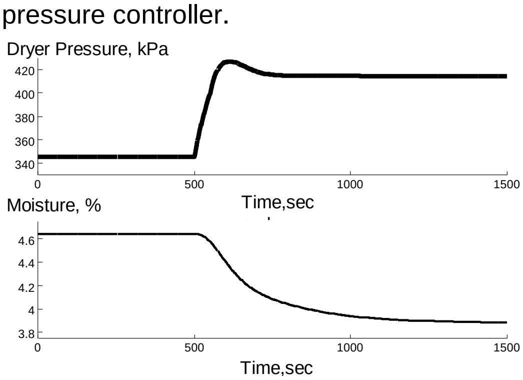

Figure 2 Moisture Response To A Pressure Setpoint Step

Pressure Control Loop

The pressure controller output is split ranged, stroking the TC from 0 to 50% output, and the Makeup valve from 50 to 100% range. Thus, the Makeup valve is closed at a pressure controller output of 50% and fully open at an output of 100%. The steam temperature increases with pressure, providing the driving force for heat transfer across the shell. The steam condenses inside the dryer at a rate approximately equivalent to the heat transfer rate divided by the latent heat of vaporization at the dryer pressure.

The pressure controller process dynamics response, best approximated by a second order model, can be thought of as the sum of two first order type processes in parallel. The fast dynamic is described by the heat transfer rate across the condensate layer. The second (and dominant) dynamic is governed by the large thermal inertia of the Dryer Shell. The increase in sheet temperature with Heat Flux slows the approach to steady state. It is important to note that the dominant time constant exceeds 300 seconds, typical of many dryer systems.

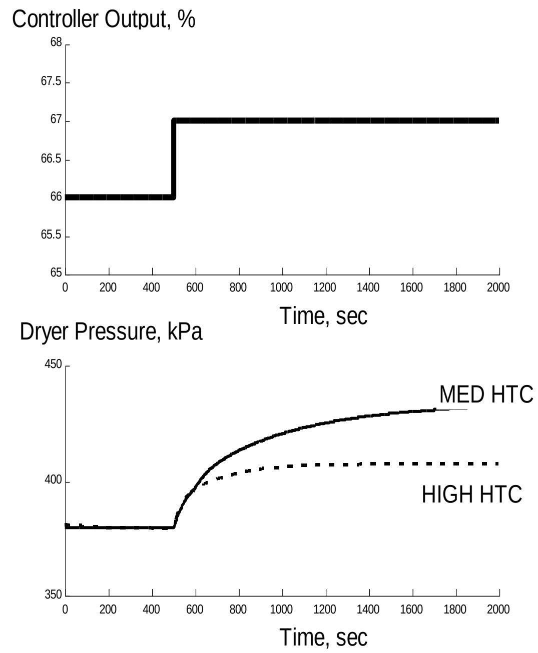

Figure 3 Pressure Response to a Controller Output Step

The process dynamics vary with operating factors which affect the Heat Transfer Coefficients (HTC), which include dryer felt tension, machine speed, condensing rate, and operating pressure [1], as indicated in Figure 3.

Differential Control Loop

The differential controller receives its input from a transmitter that measures the difference between the inlet and outlet manifolds of the Dryer Section. The control output is split-ranged, stroking the TC spindle from 0 to 50% and the vent valve from 50 to 100%. Increasing the Motive Steam Flow to the TC normally increases the blow-through flow, thereby increasing the differential pressure. If the differential is still below setpoint, the vent valve opens, providing a second path to further increase the blow through steam flow. 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 dryer differential is primarily a result of two phase flow friction loss and centrifugal pressure drop through the dryer syphon and drain piping [2]. Thus, the differential pressure is a function both of the blow-through flow rate and the condensate flow rate. A condensing rate change represents a disturbance to the differential loop, which must take control action to maintain setpoint. The process dynamics can be adequately represented with a first order model

Blow-Through Control Strategy

Blow-through control replaces Differential pressure control in almost all new dryer drainage systems. Pioneered by Gardner [3], it functions by controlling the ΔP on an orifice plate in the blow-through steam line instead of the differential pressure between the inlet and outlet manifolds of the dryer section. The gravimetric flow rate of blow-through steam increases approximately with the square root of the density of the steam, as does the steam condensing rate. Accordingly, the ratio of blow-through steam to condensing rate is nearly constant, which ensures enough blow-through steam to evacuate condensate at all times. The process dynamics of the Blow-through controller are adequately represented with a first order model.

IMPACT OF PROCESS DESIGN ON CONTROLLABILITY

The process and control design function involves designing the Dryer syphons and associated piping, selecting a Makeup Steam Valve package, and designing the TC. These design decisions contribute in a fundamental way to the controllability of the Dryer process system.

For the Dryer section described in Figure 1, the problem is that there are many built-in non-linearities. 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) is changed. In addition, the interaction between the Dryer System control loops is a key concern in the dryer system where the Moisture Loop cascades a setpoint to a Pressure loop which interacts with a Differential loop.

An important design objective should be to minimize the impact of these inherent non-linearities and interactions. However, as summarized in Table 1, the process design often does not recognize this requirement, and the result is a process that is very difficult to control. For example, oversizing the TC nozzle will result in the TC spindle operating in a near closed position where process dynamics are very non linear. This may ultimately result in cycling in the differential and pressure processes. An undersized TC Body can result in choking the blow-through steam at higher motive steam flows. This can result in a negative differential controller process gain, ultimately switching the TC to Pressure control, with the differential operating out of control range. Oversizing the Dryer syphons results in high blow-through flow rates to secure adequate differential for all Dryers in the section which in turn requires high motive steam flow demand and frequent dumping of steam through the vent valve. The design of the syphons, TC and control valves are thus critical and can only be done properly by computer modeling because of the large number of variables.

Even with the relatively good process design used in our case study, the process dynamics of the Dryer controllers vary significantly over the operating range, potentially resulting in poor control performance at certain operating conditions. The pressure controller process gain

TABLE 1 TYPICAL PROCESS DESIGN PROBLEMS

Design Problem	Process Result	Impact on Control	Impact on Tuning
Oversized TC Nozzle	TC operates near closed position, where dynamics are very non-linear.	Stiction and backlash problems magnified by high process gain.	High and Variable gain in DP loop.
Quick Opening TC Spindle	DP or Blow-Through controls oscillate.	DP loop cycles continuously.	Extremely high gain in DP loop.
Undersized TC Body	Blow-through flow choked down at higher motive steam flow, vent valve frequently opens, wasting steam.	Process gain on DP loop becomes negative, results in low select switching to PC .	DP loop operates out of control. PC controls both TC and Makeup valve. Dynamics of PC change.
Oversized Makeup Valve	Makeup Valve operates nearly closed.	Stiction and backlash magnified, non-linear regime of valve.	High Gain in PC loop. Controller Output falls below 50% on SP changes.
Oversized Syphons	High blow-through flow required to maintain minimum differential. High motive steam flow required, often wastes steam through vent valve.	High DP output signal, often over 50% , motive steam often the primary steam supply.	DP loop out of control. PC controls both Makeup Valve and TC.
Inefficient Separator Tank	Condensate carryover erodes TC and reduces performance, reduces drying, increases differential pressure.	Large errors in orifice flow measurement, DPIC output rises above 50%, non-linear, unstable process.	DP operates out of control range.

decreases significantly at high operating pressures, primarily as a result of the rapidly decreasing ΔP across the Makeup Valve. This is in spite of the fact that the makeup steam valve is sized conservatively (operates 10% and 50% open) to supply the maximum demand at close to line pressure, and has an equal percentage characteristic in this range .

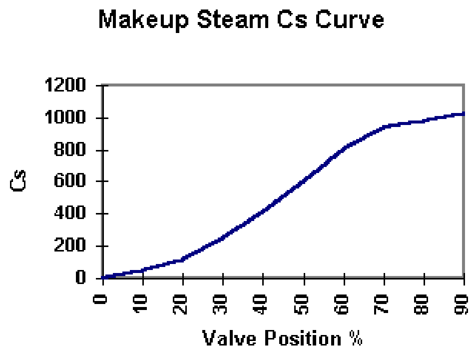

Figure 4 Makeup Steam Valve Cs Curve

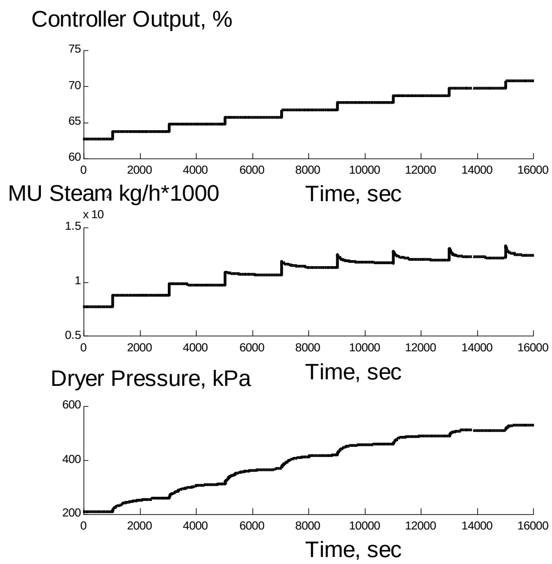

Figure 5 Pressure And Makeup Steam Flow Response To Controller Output Steps

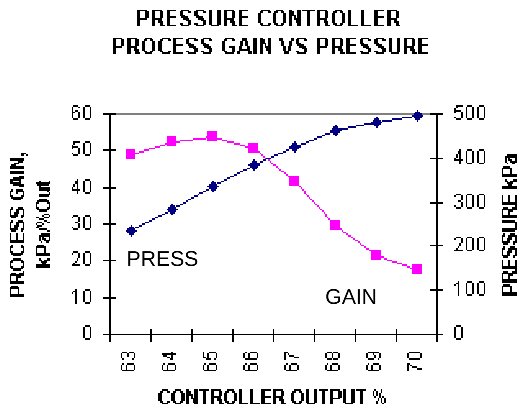

Figure 6 Process Gain Versus Dryer Pressure

The process dynamics of the Differential or Blow-through control loop also changes very substantially, and is a potential source of poor control performance. The recompression ratio characteristic, combined with the typical quick opening character of the TC spindle produces a high process gain at low differentials (or blow-through flows), since both the TC flow gain and the recompression ratio are relatively high. This range in dynamic response must be considered at the tuning stage. It is not unusual that the differential / blow-through control appears to be acceptable over the majority of operating conditions but cycles uncontrollably at low TC spindle positions.

Instrumentation Requirements

The ability to diagnose and eliminate Dryer control problems is often limited by a lack of on line process information. Troubleshooting control problems is often very difficult with only Dryer pressure and differential measurements available. Yet this is the norm for the majority of Dryer systems. Installing instrumentation to measure the Makeup, Motive, Blow-through steam flows, and steam header pressures would greatly facilitate the diagnostics of the types of process design problems listed in Table 1. An understanding of the factors affecting the dynamics of the Dryer loops would also be greatly enhanced.

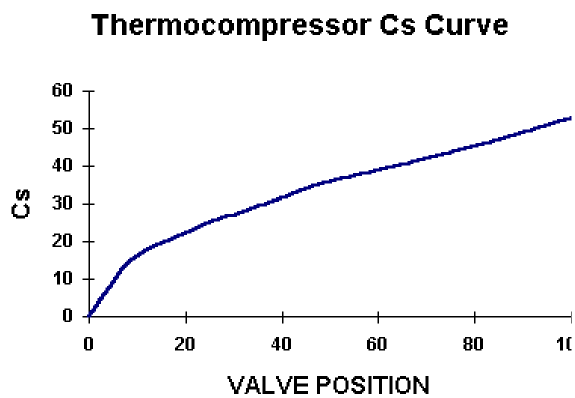

Figure 7 TC Flow Versus Valve Position

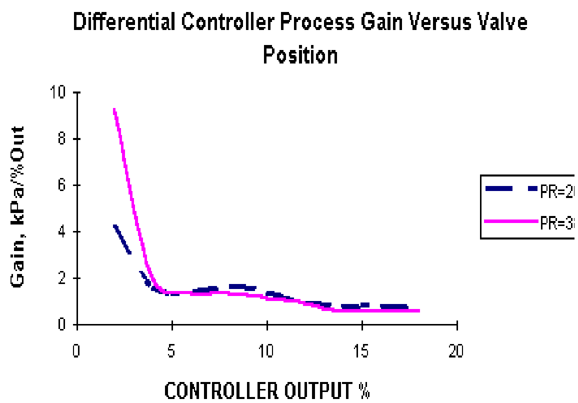

Figure 8 Differential Controller Gain Versus Operating Point

IMPACT OF CONTROL STRATEGY DESIGN AND CONTROLLABILITY

The control strategy design involves the selection of the controlled variables (i,e Pressure, Differential, Blow-through, Makeup Steam Flow) and manipulated variables (TC Valve, Makeup Steam Valve, Dump valve), the definition of the control algorithm and identification of measured variables. The design should be based primarily on process and operating requirements but must recognize the dynamics, the non-linearities and interactions inherent in the process system.

The primary process objectives of the control loops in Figure 1 are to minimize moisture variability and maximize machine efficiency. The objective of minimizing moisture variability requires a Moisture controller. The Differential or Blow-through controller is required to evacuate condensate adequately and permit adequate heat transfer. The Pressure controller linearizes (to some extent) the dynamics of the Moisture controller and attenuates disturbances in the Steam Header and condensing load more effectively than the Moisture controller.

A detailed evaluation of alternative control strategies to achieve these objectives is beyond the scope of this paper. The following discussion addresses the control strategy decision available in our case study ,that is, the choice between Differential and Blow-through Control.

Blow-through versus Differential Strategy

The Blow-through control strategy contains some fundamental advantages versus the Differential strategy. During normal operation, a constant Blow-through flow / Condensate flow ratio is achieved, ensuring adequate Blow-through flow to avoid flooding. The Blow-through strategy is more effective in limiting the impact of system non-linearities and controller interaction. On a sheet break, the Blow-through strategy decreases the Motive steam flow and maintains the Pressure Controller output above 50%. The pressure increase on the break is limited, and steam is not vented. On sheet recovery the response is relatively fast and predictable since the controller outputs are maintained close to their prebreak positions. Of particular significance is the fact that the low select relay will not be engaged on sheet break. The Blow-through strategy contains many of the benefits of a sheet break logic without the complexities.

The Differential control setpoints are often much higher than theoretical requirements, primarily to guard against flooding during transient surges of condensate. This requirement increases the motive steam flow rate and interaction with the Pressure control loop. On a sheet break or at low operating pressures, the decrease in condensing rate reduces the differential pressure. The Differential controller output increases, turning control of the TC over to the Pressure controller and opening the vent valve. This is done only in order to sustain high differential pressures which are in fact not needed. At the other extreme, if drainage stalls due to a large surge in condensate, for example on sheet recovery or a large pressure change, the differential pressure increases and the controller will reduce rather than increase the Blow-through flow. The Differential control action will increase the chance of flooding and machine downtime.

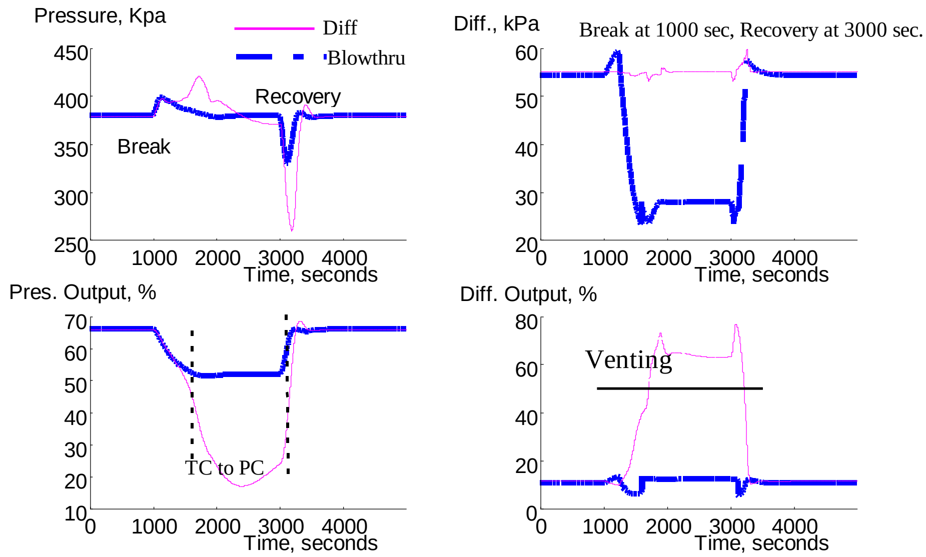

Figure 9 Blow-through Versus Differential Control Response to Sheet Break/Recovery

TABLE 2 CONTROL COMPARISON - DIFFERENTIAL VERSUS BLOW-THROUGH CONTROL

	Differential Control	**Blow-Through Control**
Advantages	1. Simple measurement	1. Effectively fixes Blow-through / Condensate ratio, which ensures adequate Blow-through flow at all Pressures 2. Excellent Response to sheet breaks. The PC will normally not assume control of TC on break. Better response time on sheet recovery. 3. Optimizes Energy efficiency.
Disadvantages	1. Increased Blow-through Flow, Motive Steam flow required, increasing interaction with Pressure controller. 2. Low Select relay engaged on sheet break, changing dynamics of pressure and differential loops. Also results in control deadband on sheet recovery, increasing time to reach setpoint. 3. Energy Inefficient since steam is vented. 4. Often requires sheet break logic.	1. High Process Gain makes controller more susceptible to tune by feel approach. 2. Measurement accuracy more difficult.

TUNING THE DRYER CONTROL LOOPS

Tuning the control loops should be viewed as the final step in optimizing Dryer control performance. The process and control strategy design defines the process dynamics and interactions in the system. The control engineer, working with this design, must develop a tuning strategy which will support the process objectives of minimizing moisture variability and optimizing machine efficiency. This is not a straightforward proposition. Many of the control objectives, summarized in Table 3, are competitive and compromises must be made in attempting to achieve the overall best control performance. For example, the pressure controller objective of responding quickly to sheet breaks (relatively fast tuning required) is fundamentally at odds with the objective of minimizing setpoint and load resonance.

Careful analysis is required to optimize the tuning because of the non-linearities, interactions and competing tuning criteria. Achieving adequate control performance for all operating conditions with a single set of tuning constants is often impossible, particularly if Differential control is employed. Sheet Break logic, adaptive gain and other strategies are sometimes required to achieve adequate performance over all operating conditions. For example, an adaptive gain strategy may be required if the Pressure controller process dynamics are highly variable. In this case, the Pressure loop may not be able to effectively track the Moisture controller remote setpoint for a specific grade, resulting in cycling in both the Moisture and Pressure loops. Process simulation is an effective tool in evaluating the potential benefits of a more advanced strategy.

On top of the issues of identifying a reasonable tuning strategy, the tuning of the pressure controller contains some special challenges. It is often difficult to identify the true process dynamics because of the long time constant. Convincing an operator to leave the pressure loop in manual mode while paper is being lost because the moisture is off target is often impossible. In addition the dynamics vary significantly making it necessary to gather process dynamics data over a wide range of operating conditions.

TABLE 3 CONTROL OBJECTIVES FOR SYSTEM LOOPS

Moisture	Pressure	Differential/Blow-through
Attenuate disturbances as effectively as possible	Track Moisture Controller Setpoint adequately to prevent interaction.	Respond quickly to prevent flooding.
Limit resonant peak to + 3 db.	Respond adequately to sheet breaks / recovery to minimize break time.	Minimize interaction with Pressure Controller
	Attenuate Steam Supply disturbances to maintain setpoint.	Limit Resonance to +3 db.
	Minimize risk of flooding resulting from fast, large changes to pressure.	Adequate Response to sheet breaks.
	Limit controller resonance to prevent increase in Moisture variability.	

Two tuning methodologies have been extensively used to deal with the overall challenges of pressure controller tuning. One approach is to represent the initial pressure dynamic as an integrator [4] (Figure 10). This approach leads to a tuning method designed for integrating processes. The often used near-integrating rule for setpoint tracking, while effective in tracking setpoints, is less successful in responding to sheet breaks/recovery since the tuning relies heavily on proportional action (Figure 11).

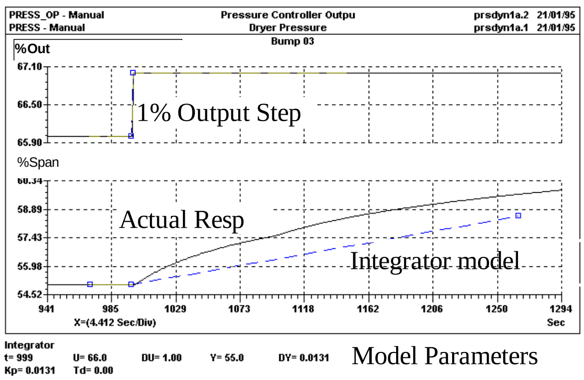

Figure 10 Representing the Pressure Response as an Integrator

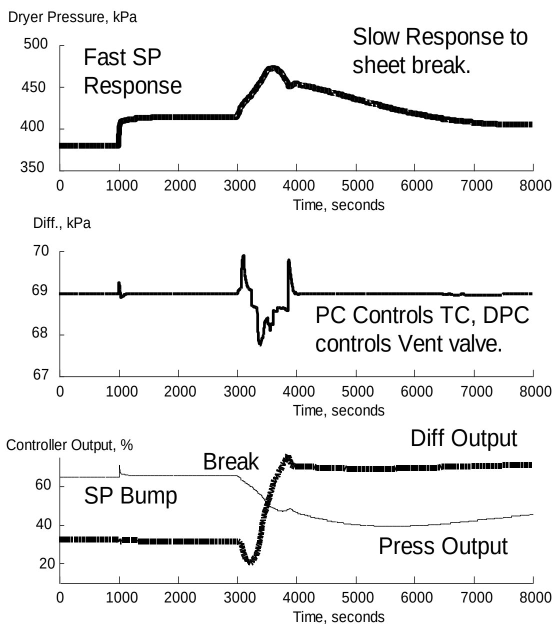

Figure 11 Near-Integrating Tuning - Response to Setpoint Change and Sheet Break

A second, more sophisticated, approach is to model the pressure response with a first order plus integrator dynamic (integrator lead model). This approach more accurately models the initial part of the pressure response. The tuning method involves canceling the process pole with the controller zero, and positioning the controller gain to achieve a non-oscillatory response with the desired closed loop time constant.[5]

This tuning method generally yields lower controller gains but much more aggressive integral action than the near-integrating tuning method. While this approach is superior for sheet break response (Figure 9), there is much more potential for setpoint and load resonance. This resonance is a potential source of increased Moisture variability.

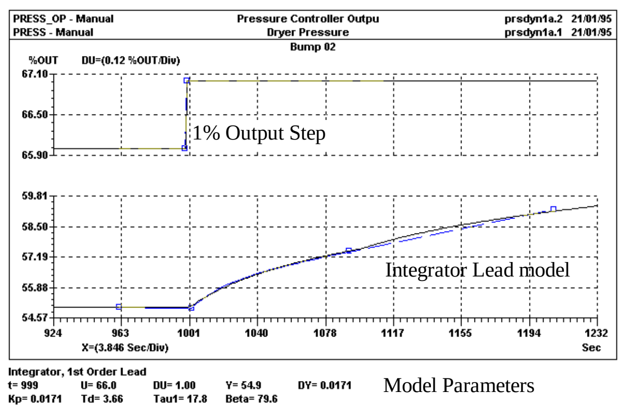

Figure 12 Modeling the Response as Integrator Lead

Identifying the process model parameters using either of the above approaches tends to be somewhat subjective. The basic problem is that the actual process response is second order and the best fit to an integrator lead model or an integrator model is highly dependent on the response time used in the identification.

Table 4 summarizes the tuning approaches described above. Based on the slow second order dynamics, a Lambda value of 1000 seconds would typically be considered a reasonable choice. This tuning, while robust, would result in almost no effective moisture

TABLE 4 COMPARISON OF PRESSURE CONTROLLER PROCESS MODELS AND TUNINGS

Process Model	Process Transfer Function	Lambda, seconds	PI Controller Tuning Kc, Gain %Out/%Span	T_R, Reset min	Tuning Equations
2nd Order Lead	$G_P(s)=\frac{7.79(108s+1)}{(351s+1)(71s+1)}$	1000	0.05	5.8	$\lambda = 3\ \tau$, $K_C=\frac{T_R}{K_P(\lambda+T_d)}$ $T_R=\tau$
Integrator	$G_P(s)=\frac{K_P}{s}=\frac{0.0131}{s}$	60	1.28	10	$K_C=\frac{1}{K_P\lambda}=1.28\%Out/\%Span$ $T_R=10\lambda=10\min/rep$ (Near integrating rule for setpoint tracking)
Integrator Lead	$G_P(s)=\frac{0.0171(79.6s+1)}{s(17.8s+1)}$	60	0.29	0.37	Pole-Zero cancellation , adjusting gain to achieve non oscillatory response.

control. Tuning to achieve a Lambda value of 60 seconds was calculated for both the integrator and integrator lead process dynamics. With this Pressure controller tuning, the Moisture controller Lambda needs to be 300 seconds or slower to prevent controller induced instability.

SUMMARY

This paper has focused on the key difficulties involved in tuning the Dryer control loops. We have stressed the fundamental impact of process design on control performance since audit experience has shown that poor design is very often the basic cause of poor Dryer control. It is an unfortunate fact that current process design practice often takes place without an appreciation of dynamic consequences which too often have serious consequences for the controllability of the process. In the Dryer section, the many non-linearities and controller interactions make an appreciation of process dynamics especially important. An integrated approach to process and control design is needed.

Control strategy design, instead of defaulting to the standard Pressure and Differential strategy, needs to recognize the dynamics of the controlled variables, the need to minimize inherent non-linearities, and controller interactions. The Blow-through strategy has been shown to be far superior to Differential control in this regard.

Successful tuning of the control loops requires an understanding of the process objectives of the control loops. The range of tuning criteria often makes this a difficult task. It is often true that adequate control performance cannot be achieved over the entire operating range without adding advanced control logic. The tuning of the pressure control loop is complicated by the complex nature of the pressure dynamics and the variation in these dynamics.

Computer simulation is an important tool in process design, control strategy design and tuning because of system complexity and non-linearity. Simulation tools are also an important component in developing optimum control strategies and tunings.

NOMENCLATURE AND SYMBOLS

Backlash	Lost motion in final actuator, normally resulting from 'slop' in valve linkage. Can cause limit cycling.
kPa	KiloPascals
kg/h	Kilograms/hour
Gp(s)	Process Transfer Function in the continuous Laplace domain
Gc(s)	Controller Transfer Function in the continuous Laplace domain.

Kc	Controller Gain (%Output/%Span)
Kp	Process Gain, usually (%Span/%Output)
Interaction	Controller Output affects adjacent processes in addition to it's own.
Lambda, λ	the desired closed loop time constant
MU Valve	Makeup Valve
Non-linear	Process response to a given input is not constant
PI	Proportional - Integral controller (Classical Form)
psi	pressure , pounds per square inch (1 psi = 6.89 kPa)
Resonance	amplification of input cycle in a given frequency range.
s	Laplace variable
Stiction	Static friction, the tendency of final actuators to stick. Often leads to limit cycles.
TC	Thermocompressor
Td	Process Deadtime, expressed in seconds
T_R	Controller Reset Time , usually expressed in minutes
τ	Process Time Constant, usually seconds.

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