Plant wide Control of Chemical Processes
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Lecture - 4
Common Industrial Control Loops and Advanced Loops

Good morning all of you, welcome back. Last time we looked at quite a few things covering essentials of process control fundamentals, and we covered the different modes of a controller with reverse action, direct action. And we also discussed a little bit about model fitting, process reaction curve method, auto tuned using the relay feedback test. I think I did not do a good job with regard to that process reaction curve method. So, I am I am going to try, and briefly cover it over here without the aid of power point.

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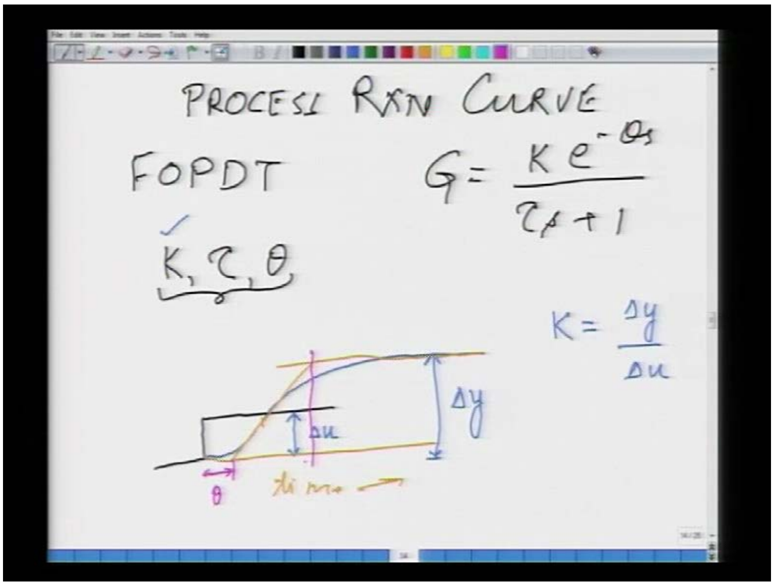

So, we going to redo or shall we do say, do again the process reaction curve method. So, what you have is, you want to fit, a first order plus dead time model to the step response the open loop step response of a system. First order plus dead time means the transfer function is G equals gain times e to the power minus dead time divided by tau s plus 1 and k, tau, and theta, the process gain first order time constant, and the dead time. These are the parameters we want to estimate the best values of this, and then one of the ways of doing it is the process reaction curve method.

And what we do in the process reaction curve method, is you give a step change of a certain magnitude, in response to this step change the process output responds like an s shaped curve, let us say something like this. It settles to a final steady state value, if this step change is for example, of magnitude delta, Delta for example, may be a 5 percent change in the position of a valve. And if this change as a percentage is what should I call this alpha, beta, gamma, delta well let us just call it delta y, the change in the output is delta y, change the input is in change in the input is delta u. Then the gain is delta y by delta u, ok.

Now, you want to so we got the gain, we want to estimate dead time and the first order time constant tau, in order to do that, what you do is you look at the inflection point in this s shaped curve, and the inflection point let us say somewhere over here. Draw a tangent at this point, that you locate the inflection point, at the inflection point you draw a tangent. So, this is the tangent at the inflection point and then of course, you also draw this horizontal line at the final steady state value, then the way to estimate tau and theta is if this is the time axis this is time, the way to estimate theta is the way to estimate dead time is this is theta, and what you do is you say that this is tau. So, this is the process reaction curve method and this is how you get the process gain, the dead time, and the first order time constant. There is also another popular method of doing this, and I will go on to that.

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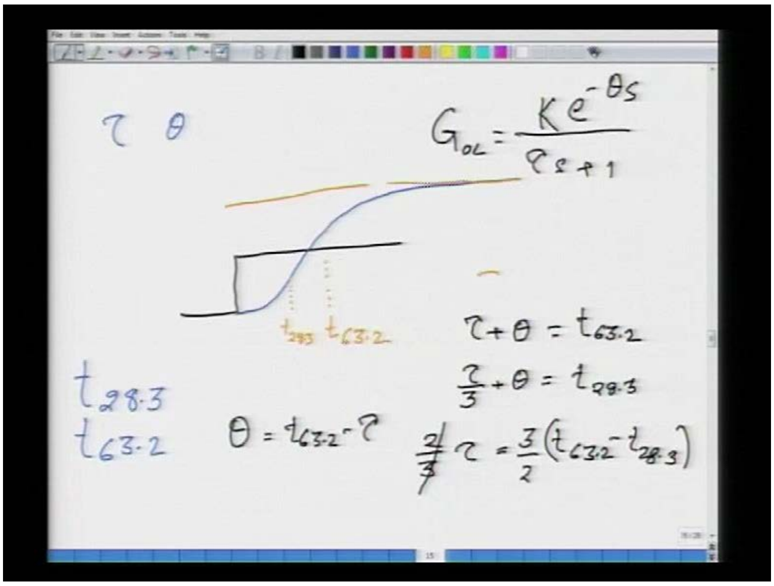

What we do there is you got the same thing, you gave the step and you recorded the step response, you get the gain the same way as before, what we are looking for is tau and theta. So, we are looking for what should tau be, and what should theta be, ok. What do what you do is you note the time it takes for the response to complete 28.3 percent, and you note the time it takes for the overall response to complete 63.2 percent.

Now, you will recall from your process control basic courses that, a first order response complete 63.2 percent in one time constant, and a first order response completes 28.3 percent in one-third of the time constant. So, if you know what is the time it takes for the response to complete 28.3 percent? and what is the time that the response takes to complete about two-thirds? What I am saying is, what is the time it takes, so the overall response total response is from 0 to… Here, if I see how much time it takes for it to comply go two-thirds of the way, 63.2 is about two-thirds of the way, this would be t 63.2, if I also look at how much time it takes for the response to complete about one-third, so little, little less than one-third. So, this is two-thirds one-third would be somewhere here, this would be t 28.3.

So, from the process reaction curve, you note the time it takes for the response to complete 28.3 percent, and the response it takes and the time it takes for the response to complete 63.2 percent. So, from these two what you have is tau plus theta should be equal to t 63.2 also, since the response completes 28.3 percent in one-third of the time constant, plus theta should be equal to t 28.3, and now if you subtract these two equations, what you will get is two-thirds of tau is equal to t 63.2 minus t 28.3, and what this implies is that tau is equal to one and half times the difference between these two times. So, once you have gotten this dead time can be esteemed estimated as t 63.2 minus whatever you calculated tau as alright.

So, this is the another way or popular way of estimating of force fitting a first order plus dead time model using a process reaction curve. Note that the controller is of in the process reaction curve method, you give a step change to the input; that means, for example, if cooling water is being used to used to cool a stream, then if the cooling water is 5 percent 50 percent open. Let us say you give a step and the cooling water valve is opened from 50 to 55 percent, so that is a 5 percent step to the input, then you look at the response of the temperature, this is the blue line is the response of the well let us say, it was t the blue line is the response of the temperature. And from there, you are trying to get you trying to get best estimates or reasonable estimates of K theta and tau, this is your this is your open loop process model.

Once, you have this model you can use classical techniques to figure out, what the ultimate gain is, what the ultimate period is, or you can use a simulator to figure out, what the ultimate gain and period are? And from that you can use standard tuning tables. For example, Ziegler Nichols or (( )) or khohin khone to figure out, what your controller parameters the controller gain the reset time and the derivative time should be? So, I think this should put everything in good prospective.

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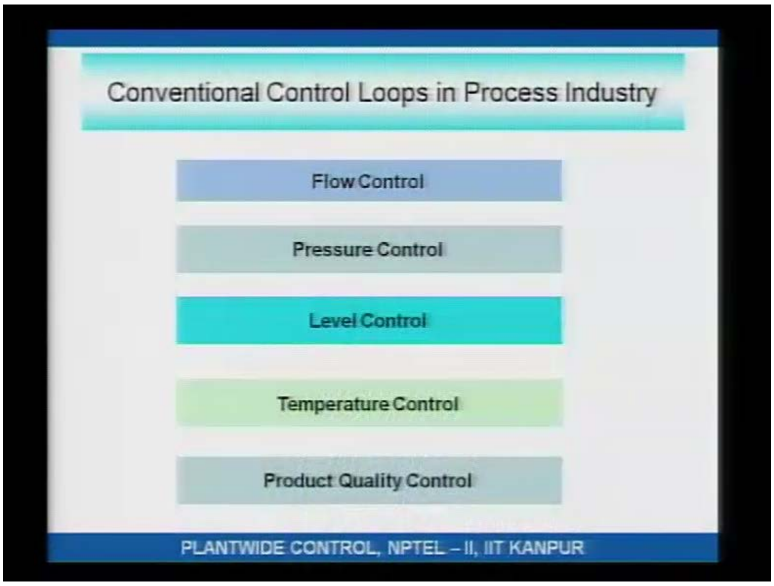

The common loops that are found in the process industry, so most of the control loops in the process industry will fall in any of these five categories: Flow control, Pressure control, Level control, Temperature control, and last but not least Quality control it could be product quality control, it could be raw material quality control, or it could be some kind of composition control somewhere inside the process. So, I would not say it is necessarily product quality control, but just quality control. So, these are the 5 most of the loops, that you find in industry in the process industry will file will fall in one of these four or five categories. So, now, we are just going to look at, what are the characteristics you know of flow loops, pressure loops, level loops, and, so on, so forth.

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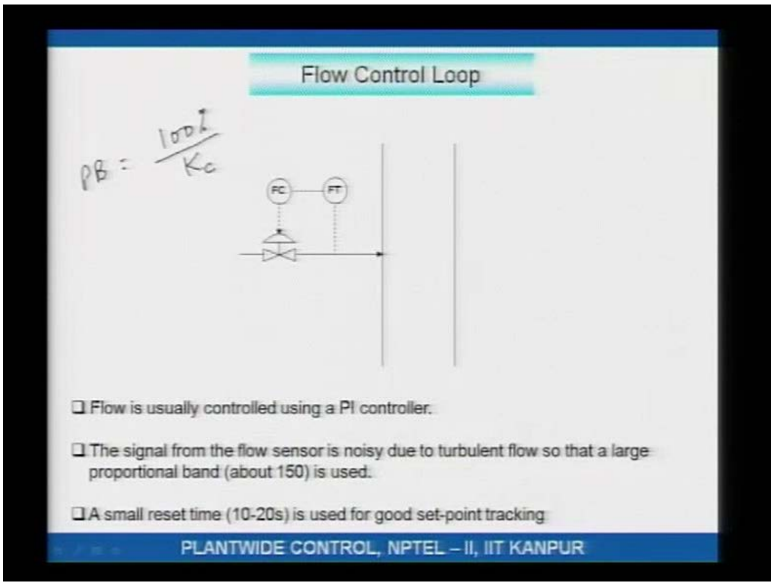

Flow is usually controlled using a PI controller, why do you not use the PID controller? Because, derivative action in amplifiers noise and flow sense or redeems are typically quiet noisy, why do you use integral action? That is because if I am saying that the desired flow rate is, so many kilograms per hour. I would like the actual flow rate to be exactly equal to so many kilograms per hour alright. So, an offset is not tolerable, an offset is not acceptable. Therefore, integral action is necessary, why do we have P action P action? So, that you get a fast and snappy response in the flow.

Since, the signal from the flow sensor is noisy de action is not used, and also proportional band is actually 100 by K c, proportional band is actually defined as 100 percent by the controller gain, that you implement. So, proportional band greater than 100 is actually is indicting that K c is less than 1. So, what we are saying is because you do not want the noise from the sensor to be amplified, you use a controller gain that is less than 1, typical proportional band will be of the order of 150.

Now, since flow responds immediately the moment, you change the valve position flow will respond almost immediately alright. So, therefore, the integral action the reset time is of the order of the dominant time constant of the open loop response. So, the reset time is of the order of 10, 20, 30 seconds and it is kept small. So, that you get a fast and snappy response, because you are interested in tight tracking of the flow set point. So, these are flow loops.

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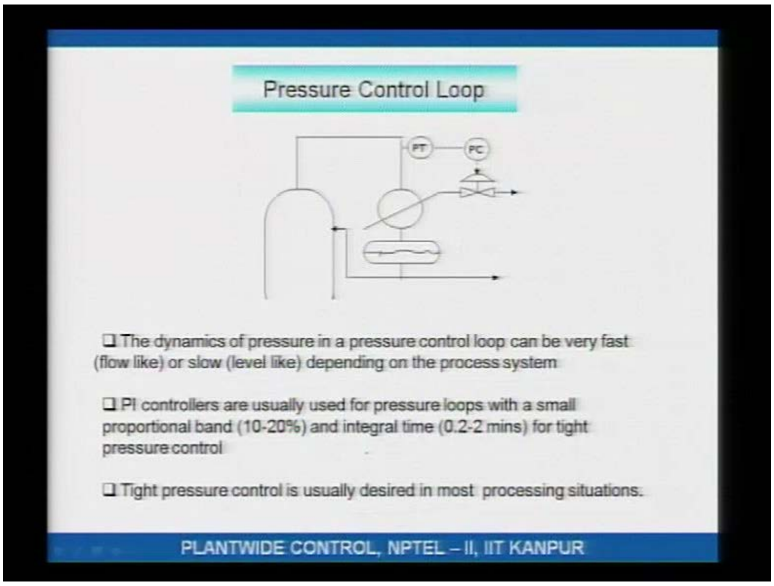

Pressure control loops, well dynamics of pressure loops can be as fast as slow loops for example, if your are for example, gas coming out of a cylinder or for example, gas coming out of a vessel you change the valve position, and the pressure responds immediately right. So, the dynamics of the pressure, open loop dynamics of pressure can be very fast; that means, the dynamics is flow like or it can be very slow; that means, the dynamic is level like depending on the process system.

So, also in most applications, if you are saying the pressure set point is x, you would like the advanced system settled down, settles down the pressure is actually x and not different. So, offset here is again not acceptable therefore, integral action is always there and proportional action of course, will be there to give you the desired speed of response, ok. Proportional band is small, and it is small a 10 to 20 percent, what that means is the controller gain is of the order of 5 to 10 integral time can vary from a few seconds 10, 20 seconds to a few minutes, depending on whether the open loop dynamics are flow like or level like. In most processing situations tight pressure control is desired, and that is why the pressure band is small, note that pressure senses are not as noisy as you senses, so, therefore, you can actually use higher K c.

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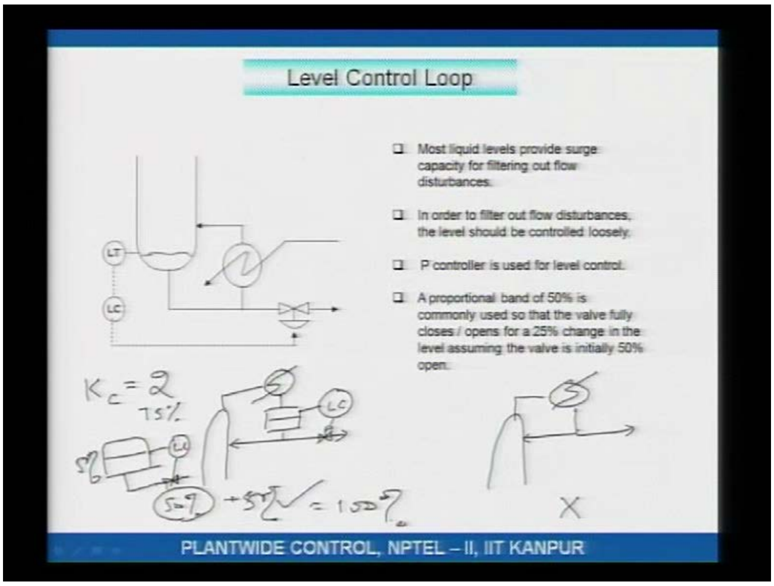

Most liquid levels provides surge capacity, what do you mean by surge capacity? surge capacity means a very common example for example, is the reflux drum in a distillation column, you see one of the main reasons well, as far as my understanding goes the main reason, why do you have a reflux drum in a distillation column? Very simple question, you will you would have seen this n number of times, but have you ever wondered, why do you have this reflux drum? Why do you not have a situation, where you know what ever gets condensed part of it is sent back to the column, and the other part is withdrawn as (( )) why is this more you know 99 percent of the time? Why is this the distillation column configuration? Why is this not the distillation column configuration? And the reason is you see the pressure of the column can fluctuate, and as the pressure in the column fluctuates the condensation, that is happening will fluctuate.

So, if you have a situation like this depending on pressure surges, and under surges how much is getting condensed will change, will vary, and as this varies what is being reflux and what is being drawn? Let us say you are having a 50 percent split, the reflux will fluctuate. Now, if the reflux is fluctuating, that means the flow the reflux into the column is fluctuating, if that reflux is fluctuating, you know that you are your chemical engineering, that if reflux varies the separation changes. So, what that essentially means is the purity of the distillate will actually vary, because the reflux is fluctuating alright to dampen out these fluctuations, this drum is provided which is called the reflux drum.

What does this reflux drum do? Even though the pressure is changing, and the condensation that is occurring is changing, you can hold the reflux constant and the distillate constant, and absorb the variability in the condensation rate as variability in the level of the reflux drum. So, the reflux drum is providing you surge capacity, you can handle pressure surges or under surges and, so on so forth. You can handle fluctuations inflow, you dampen out fluctuations in flow, so you are filtering out flow disturbances alright.

Now, since the purpose of such surge capacities is to filter out flow disturbances, the level should be controlled loosely. I do not think there is any brownie points for guessing that, why do you want to control it loosely? I mean for suppose, I have a level controller here, let us say I put in a level controller here, and it is tuned to be extremely tight, what that means is the level is not going to fluctuate at all, well that defeats the very purpose of putting in that reflux drum, because you are putting that reflux drum. So, that flow disturbances can be averaged out by allowing changes in the level.

Now, if you are controlling the level tightly, you defeated the very purpose of putting the surge capacity in there, so should everything be controlled tightly, here is an example no levels should be loosely controlled, why because they are benignant locations, they are like shock absorbers in a car, that take the bounce on the road. So, these levels are surge capacities, and these are provided to dampen out disturbances for smooth operation of the plant. So, tight control of surge capacities does not make sense.

Now, because you do not you know, because it is just a surge capacity, it really does not matter whether the level is 60 percent, 40 percent, or 50 percent. As long as the level is within a high to low range or a low to high range, it is just a long as the level does not go beyond 80 percent or does not go beyond below 20 percent operation is fine. So, since offset here is acceptable, many a times the level controllers are P only.

Proportional band of the order of 50 percent is commonly used, and what that essentially means is you see, what you have is proportional band of 40 percent 50 percent means, your proportional level controller gain is 2. So, what that means is let us say, at steady state your level is 50 percent. Now, if this level goes to 75 percent, so initially your level is 50 percent, and whatever is being used to control that level the valve, that is also 50 percent open, so that is my initial system at rest. So, valve is 50 percent opened initially, initially the level is 50 percent.

Now, with the gain of two let us say the level goes to 75 percent, if the level goes to 75 percent you have an error of 25 percent, and what that means is 25 percent, the output of the controller well the signal to the valve will be K c times e. So, K c is 2, error is 25 percent, so the signal will change output of the controller will change by 50 percent. So, initially my valve is at 50 percent, because the level has increased to 75 percent, my valve will be fully opened. So, what that means is, if your initial steady state is 50 percent, valve opening 50 percent level if the level changes by 25 percent in this direction or in that direction, the valve which is supposed to control that level moves from 50 percent to either fully open or fully closed. What that essentially means, is you are trying to control a level within a 25 percent band.

So, this is very typical of proportional controllers are all level controllers P only, you may guess that definitely not there are situations, where the level should be controlled tightly, and there PI controllers are used. There are situations where offset is not acceptable, and there you know, you use you do use PI controllers, there is another you reason for using P level controllers.

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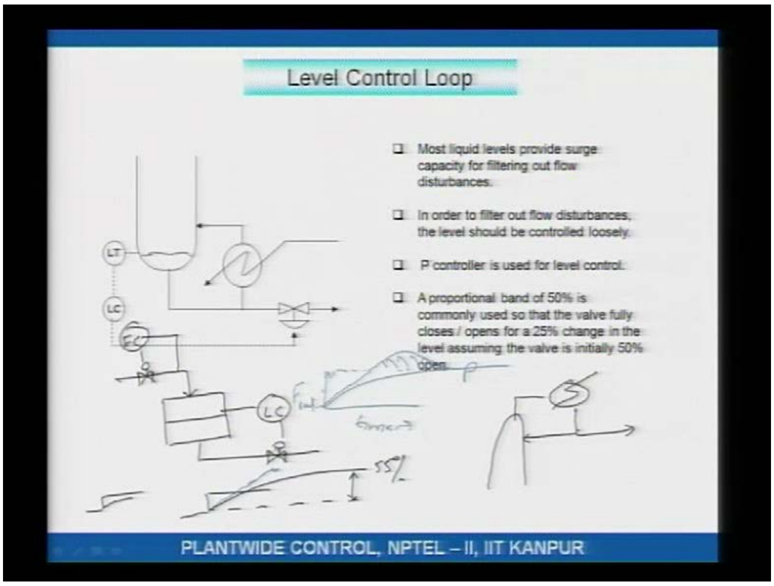

Let us say, I want to draw it again, and then I guess I will have to read it which is fine. Let us say this is a tank, you got something coming in, and you got something going out, and the way things stand. Let us say the level controller is in the direction of flow, it could also be other way around, let us say initially system is at steady stage, what that means is flow in is equal to flow out. So, the level is held constant, and let us just say it is held constant at some value 50 percent.

Now, let us say I give a small step change to the flow in, what that essentially means is the flow here goes up as a step, what would happen to the level? the level will start to raise, as the level starts to raise the level controller which let us say, the controller is proportional only, the level controller will start to open this valve, and what you what you would see is initially, the level was 50 percent. Because, the flow in increased the level increased, and this was the initial step the level increased, and it settled to some new value, and this new value is not 50 percent, there is some offset. Because, initially you are at 50 percent, but let us say the new level settles at 45 percent, so this is the response, that you will get using a P controller.

Let us see our controller is PI, now since you have integral action in there, the level must be returned back to 50 percent right, what does that mean in the initial period where the out flow was less than the inflow, level is building up. So, your response with the PI controller may look something like this, initially the level is this goes up keeps going up, and now whatever extra material has been accumulated inside the tank, this must be returned this must be drained out. So, if you are looking at the flow out with the P controller raises like this, and settles to some value.

If you looking at the flow outs using a PI controller, this is F out and what is out there is time, this is the P controller this is what it looks like, flow out increases and becomes the same as he inflow, with the inflow went up like a step, if you look at a PI controller flow out increases; however, the level has gone up beyond 50 percent. So, you have to drain more before it gets back. So, in a PI controller this area would be the same as this area. So, you see in order to bring the flow back this over shoot is necessary using a PI controller what that means, is if this tank is feeding a downstream unit, that downstream unit gets disturbed more, if you have a PI controller. So, that is an another reason why PI controllers are sometimes or are not used for surge capacities. So, we have essentially said that.

If you got surge capacities level should be controlled using a P only controller, with again that is about 2, PI controller should not be used, and we have discussed the various reasons for doing that.

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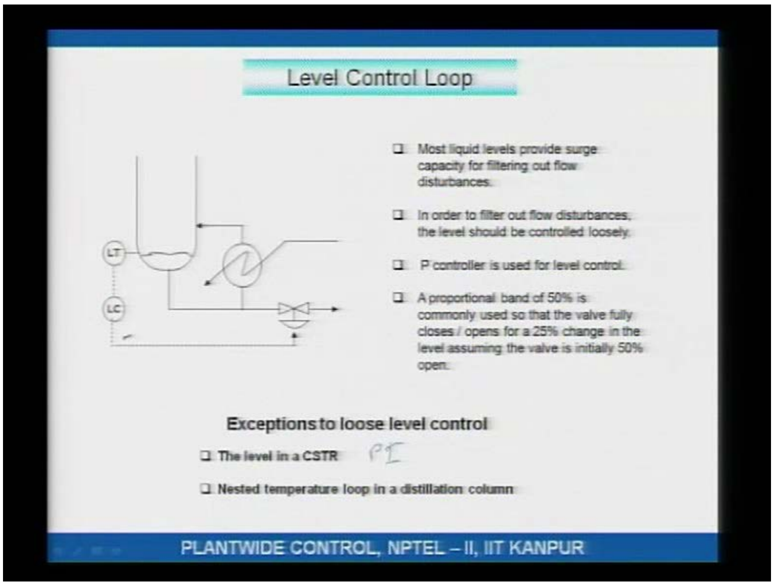

There are of course exceptions, an exception is to lose level controllers, level control in a CSTR, you see in a CSTR what is the tank level does effect the residence time? if you got more hold up, residence time is more, if you got more residence time, the conversion is more. So, residence time affects the conversion of the reactor, which effects the downstream units in terms of how much needs to be separated, which also effects the recycle and so on, so forth.

So, in order to hold the conversion constant level in a CSTR should be controlled tightly, ok, offsets here are not acceptable. So, therefore, level controllers in CSTRS may be of the PI type, there is also another example of a nested temperature loop in a distillation column, and I think we will get there when we get there just for the timing, just note that this is an exception when we cover distillation columns, we will discuss this at length.

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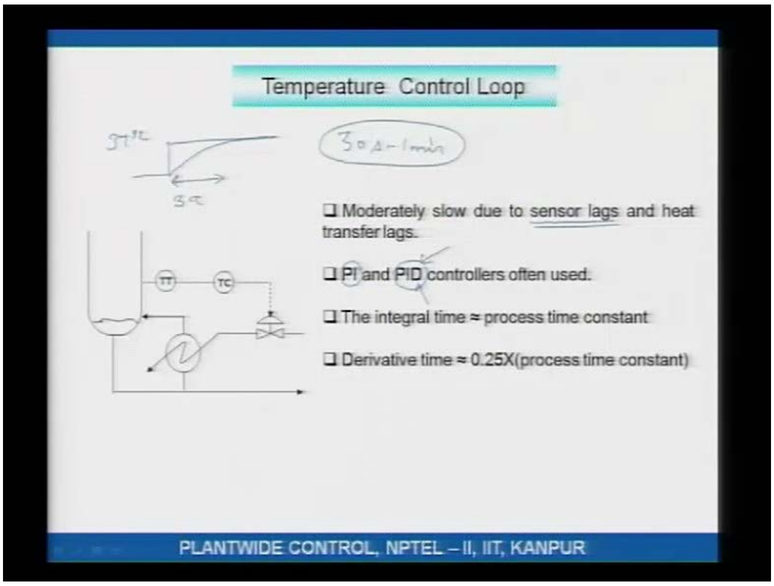

Temperature loops, well temperature loops are moderately slow, that is because the sensor shows lags. Also, when you are trying to control temperature, some sort of heating or cooling is required. So, for example, if you look at a heat exchanger, let us say the heat exchanger is being heated, you know the fluid is flowing in the tubes (()) exchanger, and steam is using is be is, use is being used at the heating fluid to heat temperature process stream. While, if you increase the steam flow temperature of the cell side would increase, then the tubes will get hotter and as the tubes get hotter, the fluid that is flowing through the tubes will see a hotter tubes, and it in turn will get heated, that see to the tubes are a large mass, and because they are a large mass of the order of tones, or a large heat exchanger it takes time for those tubes to heat up.

So, what that means, is if I increase the steam flow now, by the time I start seeing a change in the temperature, or the process fluid it is say for 5,4,5 minutes 2,3,4 minutes I see nothing. So, there are these lags because of thermal heat capacities involved in the process, also if you may you may, you may know, you know suppose you are measuring your own temperature, body temperature you put in a thermometer you wear it for 1 ½ to 2 minutes, before you take it out and read. So, what that means is, when I put the thermometer in, the temperature goes up from whatever is the ambient temperature, let us say the ambient is cold to whatever is my body temperature, which is let us say 37 degree Celsius, the mercury however, moves up from its initial ambient position to 37 degree Celsius in a slow way.

And this time, where it reaches you know, where it is almost equal the temperature indicated by the, by the mercury in the thermometer is almost the same as the body temperature, this time three time constants for 95, 95 percent response completion, it is of the order of 1 ½ minutes. So, the lag due to the sensor itself can be anywhere from 30 to 45, 30 seconds to a minute. So, the loops are slow, because of thermal lags thermal heat capacity lags as well as sensor lags.

Again in temperature control applications an offset is not acceptable. So, most of the controllers are PI, sometimes you also have PID controllers where tight control of the temperature is very necessary. So, you want the gain to be high, so you use de action to jack up the gain, which we which we saw last time, what is an example of PID controllers? for example, controlling the temperature of a highly exothermic reactor here, there is the possibility of a temperature runaway. So, you want to control the temperature tightly, that gives you a justification for de action, note that the temperature. In fact, in the reactors sometimes the measuring element is directly exposed to the fluid, in order to eliminate the lag associated with the sensor, to reduce the lag associated with the sensor, why is there the lag in the thermometer, that I put in my mouth, it is there because there is the metal, that gets in touch with my body that metal heats up, then the mercury heats up.

And therefore, there is a lag of you know that the time the first order time constant is of the order of 30 seconds to a minute, now in order to reduce that 30 seconds to only a few seconds. Sometimes, you expose the measuring element which could be an r t d filament directly to the fluid, this is particularly true for highly exit Thermic reactors.

What do you say the integral time to rule of thumb to the dominant process time constant, what is the dominant process time constant? Just for the sake of understanding, it is of the order of the time, it takes for the response to complete two-thirds of the way. Derivative time is typically set to about one-fourth the dominant time constant, the process time constant and this is just a rule of thumb, there could be exceptions to this. What is drawn here is steam being used to control the temperature of a tray in a distillation column. So, if the temperature of the tray is going down put in out steam, if the temperature of that tray is going up, put in less steam.

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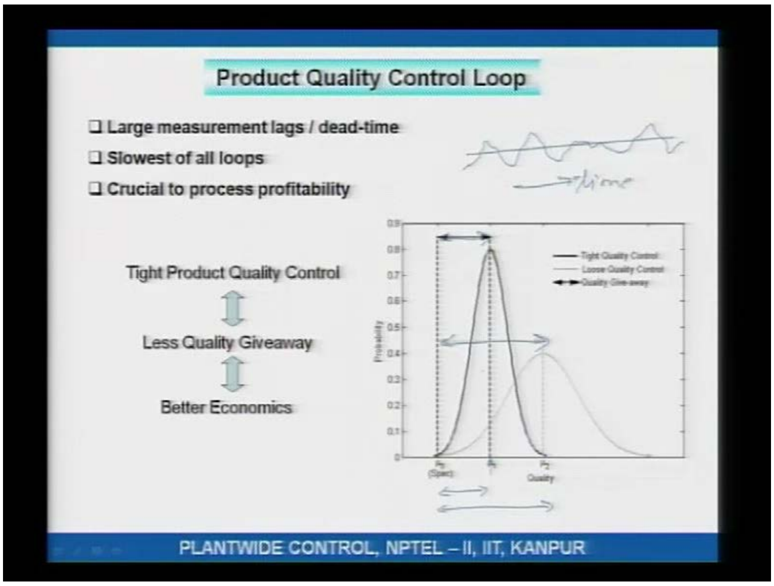

Quality control loops, well these are arguably the most important loops in the sense that, in terms of a production objectives you want to produce, so much of a certain amount of a certain quality. For example, you go to a gas station, gas station says guarantee 87 octane number fuel, this is a quality guarantee, that is that the vendor is giving you, and you get that super fuel whatever there you know, that costs 5 rupees extra or 3 rupees extra per liter, that super fuel guaranteed the octane number is more than 90 or 93, ok.

So, in terms of production objectives, quality loops are probably very important; however, measuring quality is not easy, I mean you would know from your chemistry labs, how difficult it is to measure composition of you know, even a simple mixture, you do a titration and you do it three times, and the reading fluctuates. You got these gas chromatographs or these liquid chromatographs HPLCS, and you put in the sample and by the time you know, what the composition is by the time, whatever component you are interested in by the time, that component eludes out from the column, if 15 minutes, 20 minutes sometimes hours. So, there are large lags or dead times associated with quality measurements in process plans, you take a sample from the pro you take a sample from the process, send it to a quality control lab laboratory, the fellow does the analysis, next shift fellow find out what was the quality of the sample that was taken it as ago alright. So, you know the quality of the sample, you know the state of the process, it as ago.

Imagine, measuring the molecular weight or a poly dispersity index of a, of a polymer you know these are quiet tricky or cumbersome measurement. So, therefore, because of these large measurement lags, or the dead time product quality control loops are arguably the slowest of all loops; however, these loops are crucial to process profitability, and why that is central to that, why it is show, you see it you arte controlling, your product quality tightly, less quality give away trans translates to better economics. And just to explain this point a little further, here are in a hypothetical two ways of operating a process , let us say I am saying that let us say this is octane number x axis, octane number I am guaranteeing that, my refinery or my process is producing gasoline of this octane number are more.

So, this specification guarantee let us say octane number 87, so any product that is taken from my refinery am guaranteeing that, it is going to be octane number 87 or more, now there is a smart operator, and there is a dumb operator. And the smart operator you see, because of variability, the product quality will not always be constant, it will fluctuate around a mean. So, you got time and you got whatever is coming out of the refinery, it is fluctuating around the mean, the product quality is fluctuating around the mean in time. So, this smart operator operates it in a way, such that these fluctuations are a very small range. So, the mean product quality is new one, and the fluctuations are over a small range, this is which we got a, you got a smaller standard deviation.

So, this spread in the dark line in the solid line, in the in the black line is less the other operator runs it, such that the quality control is not as tight and therefore, the spread is more. Now, in order to guarantee a product of minimum quality, this all the time the shift in the quality which is this for process one, and this for process two, the shift in quality where you are having lose control is more than the shift in quality, that is required for the tightly controlled process. Another way of looking at it is, if on average I operate my loosely controlled process at a mean product quality of this, then 99 or 99.9 percent of my product is guaranteed above my minimum spec. For the tighter control, if I operate my process on average at a quality of this, then 99.9 percent of the time my product quality that is withdrawn from the process is guaranteed to be above this minimum assured specification that I am saying.

Now, you would know for example, in a distillation column, if you want purer and purer bottoms, or distillate product your minimum reflux ratio goes up therefore, the operating reflux goes up therefore, the amount of steam that is required goes up. So, the tighter the quality spec the more the steam, you consume right this is a very simple example, but the point is the lesser the quality give away, the lesser the steam, you consume since you are consuming less steam per k g product, it translates to better economics. So, this quality give away concept is quiet fundamental to process operation.

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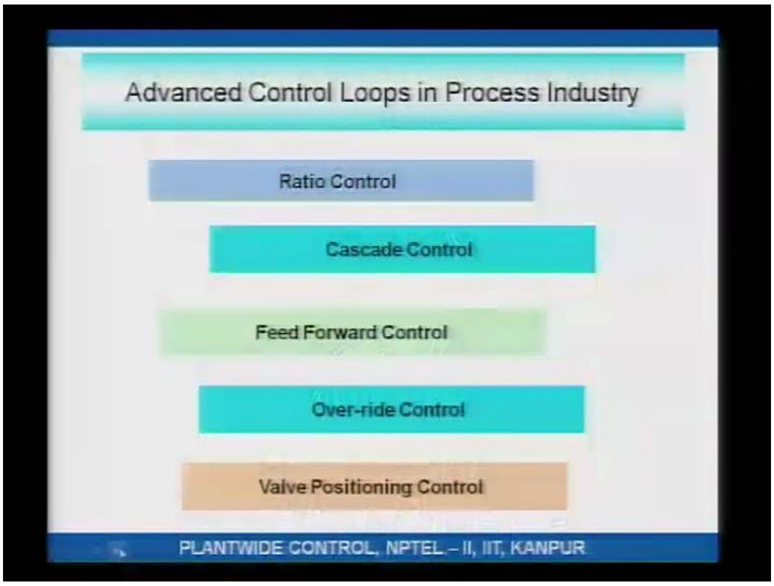

Now, there are these advanced loops in the process industry, that are used quiet often and these fall in the different, you know you got the standard feedback loop, that we have discussed. Then you got advance structures like ratio control, cascade control, feed forward control, override control, valve positioning or optimizing control, or sometimes it is also called constraint control. So, we have going to we are looking at common advanced control loops in the product process industry which are different form simple feedback loops. Ratio controllers, Feed forward controllers, Over ride controllers, and Valve positioning or Optimizing controllers.

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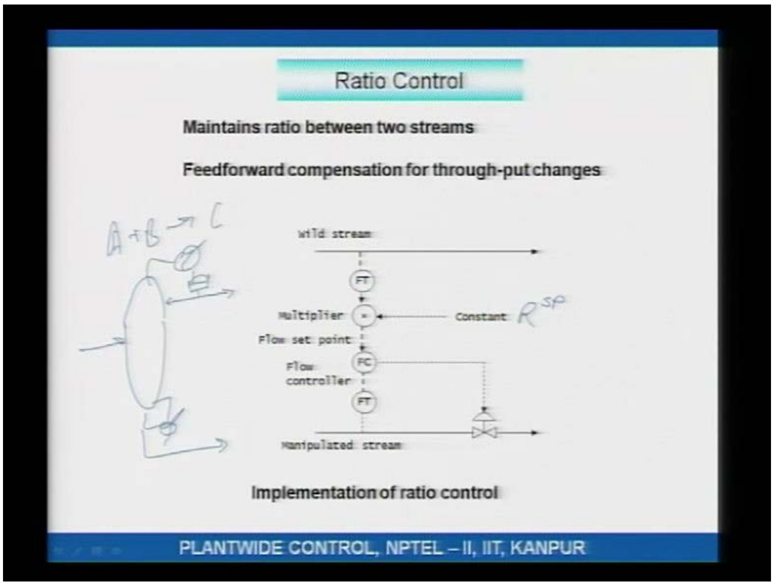

Ratio control, as the name indicates ratio control attempts to maintain the ratio between two streams, why do you want the ratio between two streams to be maintained? for example, let us say A plus B goes to C, this is the reaction that is occurring in a process system, reactions to A (()) dictates 1 mole of A require 1 mole of B. So, if you want to require more of C ultimately A has to go up. So, if A goes up 10 percent, B must also go up 10 percent, so that means is you would like to maintain B in ratio with a alternatively, you may want like to maintain A in ratio with B.

Similarly, if you look at for example a distillation column. So, you got some feed coming in I better show the reflux terms, that we talked about it, let us say the feed goes up 20 percent. Now, if in order to maintain the same distillate purity, reflux should also go up by 20 percent. So, what means is you would like to maintain the reflux in ratio with the feed. Alternatively, you may want to maintain, if more distillating is produced well, that means distillate, that means more stuff is coming in, more feed is coming in. So, what that means, is reflux should also go up perhaps, you may want to maintain the reflux in ratio with the distillate. But in industry what is more common is, if you are maintaining the reflux in the ratio with, if the feed has gone up 10 percent, you would be increasing the reflux 10 percent.

So, the purity of the distillate would not deviate by as much, because you are anyway putting as much reflux as it is necessary to maintain the purity. On the other hand, if you do not have a feed to reflux ratio, instead you have a distillate to reflux ratio, more feed comes in reflux is fixed. Because, more feed is coming in distillate will start to go up, so if the reflux is less than necessary impurity, here will go up and then because the distillate has gone up, and you are maintaining reflux in ratio with the distillate, then the reflux goes up; then the impurities that were coming out, the distillate start being start getting backing, and then the purity comes back. So, you see here, by maintaining reflux in ratio with the feed, you are actually in some sense compensating the disturbance in a feed forward way.

So, these ratios are quiet commonly used, and the most common implementation is as follows, you got the wild stream. So, for example, in the distillation example the feed is the wild stream, it can change to whatever it wants to change to… The reflux is the one, that is following the wild stream which is the feed, so if the feed goes up 10 percent the reflux must go up, go up 10 percent. If the feed goes down 10 percent, the reflux must go down 10 percent, how is it done? So, you got the wild stream, you measure its flow you multiply it by whatever is the ratio that you want of the manipulated stream, this multiply this is the ratio set point RSP. So, this flow times, this ratio tells you what the set point for the manipulated stream is, the flow set point for the manipulated stream is, and then a flow controller which is taking in this set point adjust this valve. So, that this flow rate is equal to this flow rate alright.

So, this is the implementation of ratio control we discussed, you know the cascade mode of a controller, and what the cascade mode is that the controller is actually getting its set point from a, from another higher level controller. So, the one that is getting the set point from above is called a slave, the one that is giving the set point is called the master. So, you got the master slave loop combination and just give you an example.

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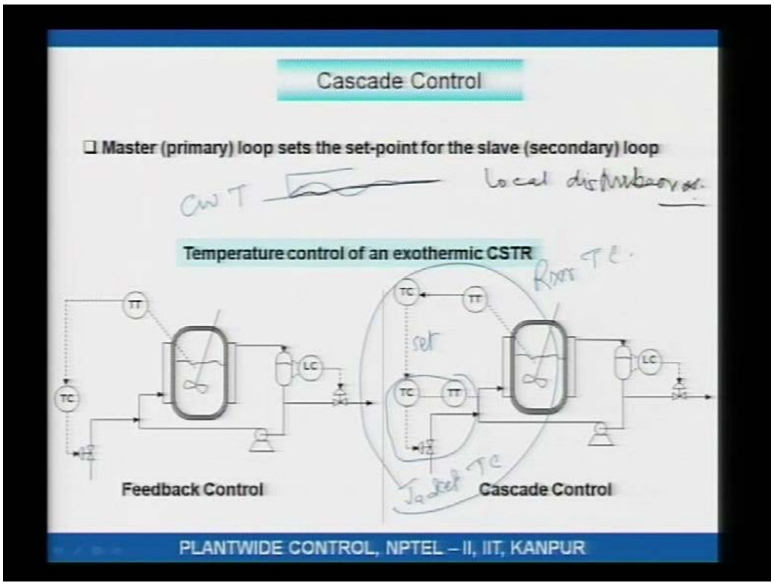

I have just taken a very common implementation in industry, here is a jacketed CSTR continuously start time reactor, an exothermic reaction is happening there, and in order to hold the reactor constant, you know you are you got fresh coolant coming in, the cooling circuit has got a high flow rate pump. So, the recirculation rate is very high of the liquid of the water, or whatever is the cooling liquid. So, this circulation rate is very high, so the jacket temperature raise is negligible. So, the CSTR the reactor essentially sees the jacket constant temperature, and that is why you got you put in this recirculation rate.

Now, let us say the temperature is going up, reactor temperature is going up, what you would like to do is reduce the temperature of the jacket, so that more heat gets removed, so that the reactor temperature that is going up gets back to set point. In order to do that, what you do is increase the flow of cold water into the circuit, so when you bring in more cold water, the temperature of the jacket will go down, more heat will get removed, and the jacket temperature that was going up should hopefully come back alright.

Here is this is a standard feedback control arrangement, never implemented in practice, for this kind of a system what is implemented in practice, for this type of a system is shown in the other figure, here what you have is you got a jacket temperature controller. This is the jacket temperature controller, this line, this guy is a jacket temperature controller and the other guy, this is the reactor temperature controller. Now, what is this jacket temperature controller doing, if the jacket temperature is deviating from set point, it will adjust the coolant flow rate, so that the jacket temperature stays at its value.

What does the master loop do? the master loop which is the reactor temperature control. So, what is this? it is setting, it is setting the set point of the slave jacket temperature controller. So, the master the reactor temperature controller is actually adjusting the set point of the jacket temperature controller alright. So, this is the arrangement and just to, just to just to tell you, why this is better than that? Imagine.

Let us say the temperature of this cooling water goes up or down, because of some upset in the cooling system which is supplying hot cooling water, so if this temperature goes up. So, let us say the jacket temperature the cooling water temperature goes up, cooling water temperature if the cooling water temperature goes up, what happens is this, in this scenario is cooling water temperature is gone up therefore, jacket temperature has gone up. Because, jacket temperature is gone up less heat is getting removed, because the driving force temperature driving force is less, because temperature driving force is less heat is getting removed.

Therefore, this reactor temperature if I look at the reactor temperature reactor temperature, actually goes up because you are removing less heat than necessary, now because the reactor term has gone up, this temperature controller says well, we need to put in more cooling water which is not as cool as before. So, as you putting more water the jacket temperature goes back to where it is supposed to go, and this causes the temperature to get back or let us say the temperature gets reactor temperature gets back. So, this is an conventional feedback control.

Now, let us say I have got in the second situation, I have got that cascade temperature control system, cooling water temperature goes up jacket temperature goes up, that is sensed by this slave loop. So, what this slave loop does is immediately increase the cooling water flow rate, so that the jacket temperature is brought back. So, what that essentially means is the reactor, even though the cooling water temperature has deviated essentially sees the jacket at constant temperature. So, because the reactor is seeing that jacket is at constant temperature, in this case reactor temperature may show some deviations, but not much, you will hardly notice any deviations in the, in the reactor temperature.

So, in this cascade arrangement what we are seeing is, local disturbances in the slave loop are removed, so the local disturbances are removed. Also, what we see is for this particular system, if you know a little bit of control theory, you know your open loop time constant may be whatever, if you put in a control loop, there the time constant of the closed loop system, can be jacket out can be reduced significantly. By making the loop appropriately tight significant, you know as tight as possible, so what does this loop can be really mean, made quiet tight, and overall for the master temperature controller, what it will essentially see is a faster system. Therefore, that will actually result in better control of the reactor temperature, even if there are nonlocal disturbances. So, this is summarized I think in the next line.

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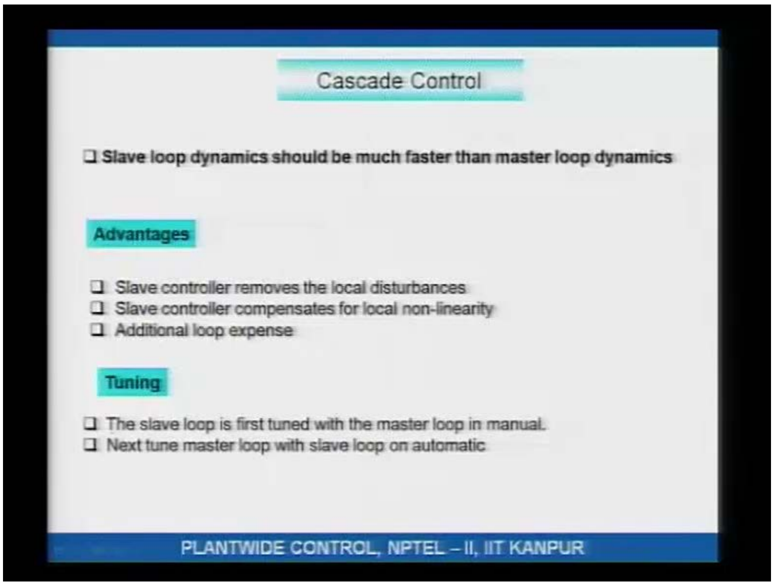

For the cascade control loop to for the cascade control system to be feasible, slave loop dynamic should be much faster than the master loop dynamics, and that makes sense master is saying give me the set point. But, if the master does not give the slave enough time to give it, that set point and if we keep changing the set point, the slave must be able to keep up with the master; that means the slave has to be fast enough to respond to the demands of the master. So, therefore, the slave loop should be much faster than the master loop, if that is not the case a cascade arrangement is advised, let us put it that way.

Advantages of the cascade control system, slave controller removes local disturbances, that I just explained, as slave controller also compensates with the local or non-linearity, you can actually think of non-linearity as a disturbance, ok. Well advantages a negative advantage is that mean the disadvantage is of course, there is an additional loop that needs to be maintained tuning, how do you tune a cascade control system? you first tune the slave loop with the master loop off, then with the slave loop tune for a fast and snappy response, you tune the master loop. So, that is the way to tune.

A many a times since, you are really not interested in the jacket temperature. For example, in the previous example is a secondary variable, you really not interested, whether the jacket temperature is at the set point being asked by the master controller. So, many a times the slave loop is actually a P only controller and not a PI controller, P only controller will give you tighter tuning, that a PI controller that we have seen before.

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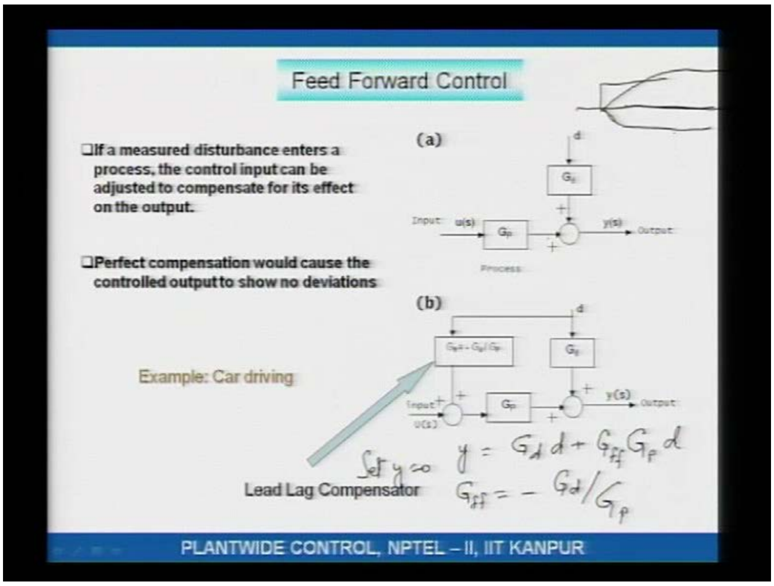

Feed forward controller is like driving a car, you use that a turn is coming, because that and then you take appropriate action. So, that your car always remains on the road. Feedback control, Feedback control is you have to have an error first, before some corrective action can take place right, so in feedback control the car will have to go off the road, before it is brought back. So, driving a car is more like feed forward control, and what it essentially translates to is that, if I measure disturbance enters a process, the control input can be adjusted to compensate for the effect of the disturbance. If you know that the disturbance is coming, you can make adjustment, so that whatever you want to control is not affected or affected very little.

Just to consider this, you got this disturbance d which effects the output, you got the control input which also affects the output alright. Now, let us say disturbance d comes, if disturbance d comes in as a step, let us say the output you know goes up. I have a measurement of this disturbance, and what I want to do is disturbance affect it like this, if there is a disturbance, this is what the output will do. I ask the inverse question, what should I do to this control input? So, that its effect is negative of this, so that when I add this and this output actually is 0, or stays where it is you know shows no deviation. So, what should I do to this? such that, this guy balances out this guy and this is done in this feed forward compensator.

And what we have is I mean, if you just look at this block diagram, y is equal to G d times d plus this guy G feed forward this transfer function, times G p times d, and since I want y to be 0, it just gives me a very simple equation, that says that G f f, if I set y equal to 0; that means I do not want any deviation in the output, set y equal to 0 what that gives me is G f f should be minus G d by G p. If I have this the output will not show any deviation, or let us just say very little deviation, if G d is the first order lag, if G p is the first order lag, G d by G p will be you know a numerator, tau s plus 1 and a denominator tau prime s plus 1 that is called a lead lag. So, this feed forward compensators many a times are simple lead lags, I think the tape is running out and therefore, we end here.

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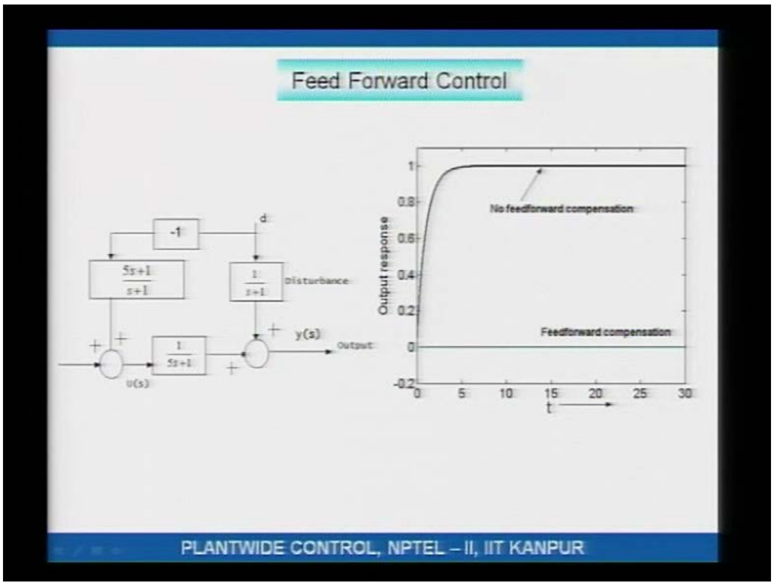

There are other types of… So, here is an example I give d, a step, this guy is not there, this fellow is you know the block is not there, if there is no feed forward compensation output goes like this. If I put a feed forward compensator, even though d goes up as a step, there is no change in the output, you do not notice any deviation in the output. So, were tight control is required, and you have measured disturbances, you can have feed forward compensation. So, that the output does not deviate or deviates very little.