Plantwide Control of Chemical Processes
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Lecture - 41
Cumene Process Plantwide Control

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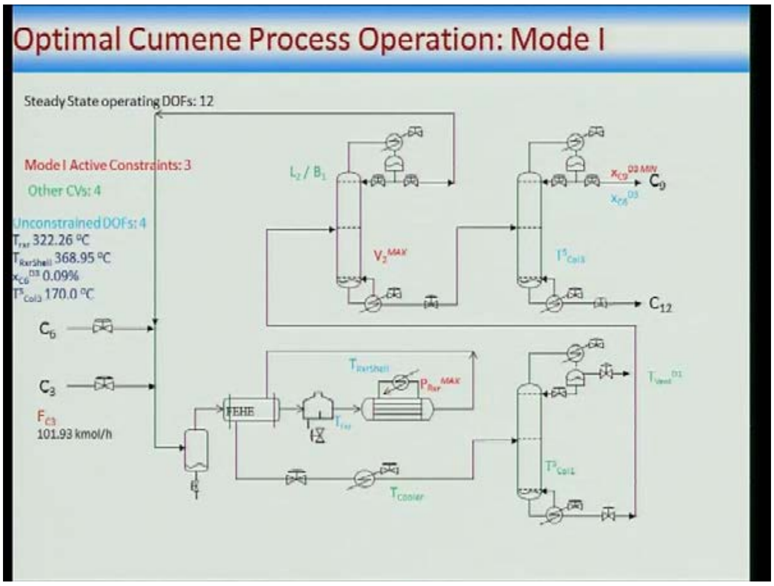

So, now that I have explained to you the mode 1 optimum, if I try and increase the throughput, to maximize the throughput which is mode 2.

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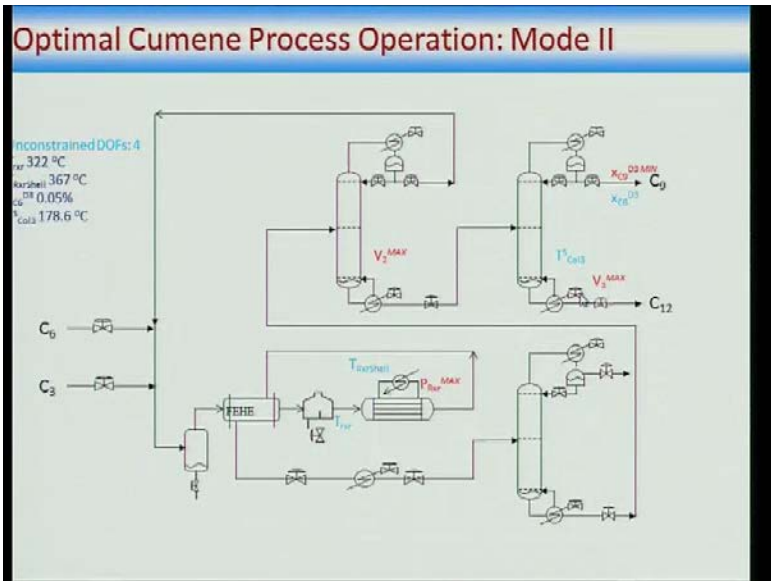

What happens is, in addition to those three constraints, V 3 max also goes active. The boil up on the product column goes active, and that limits my throughput. And of course, these four variables remain economically important, because again I want to make sure that my reactor operating conditions, are such that conversion is high, and selectivity is not too low, well that is that, the rest of the stuff is as in mode 1. So, it turns out that V 3 max is my bottle neck constraint, this is the constraint that limits my throughput, maximum boil up in the third column.

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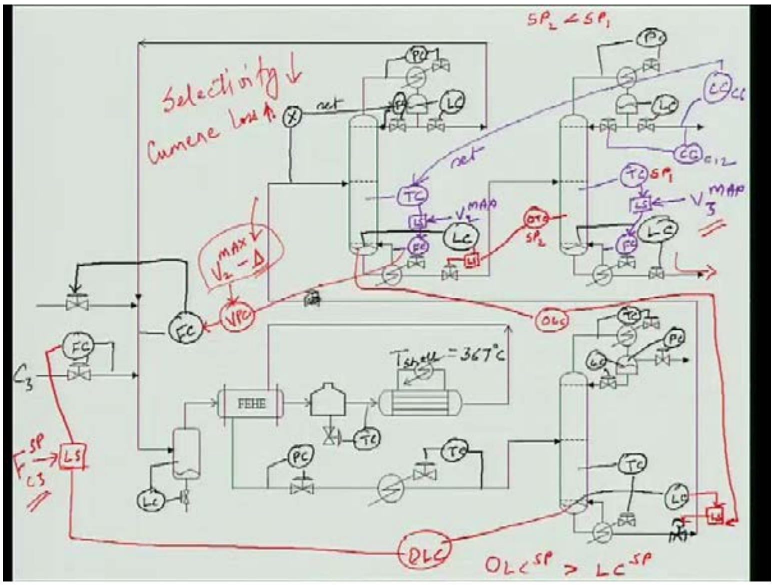

Now just to contrast, let us put up a control system. The conventional approach is, this is C 3. The conventional approach is you basically choose your throughput manipulator at one of the process feeds, and then put in your total material balance loops, and then the other loops. So, what do I do there, since the limiting reactant is C 3, I used that as my throughput manipulator. Then what I need to do is control the level, so if I want to control the level of the vaporiser, well the level has to go this way. Then I need to control, what. Let us put in all other total material balance control loops, so this one is needed to condense, so I will put a temperature controller this way.

Then I need to control the level of this chap, how do I control; man that valve is quite far away. So, let us just say this valve is not here, let us just say this valve is here, right here. So, what do I get, I basically get a level controller this way, I get a level controller for the top this way, I get a temperature controller this way, and I get a. This is the material balance control on the column, vapour, and liquid. Then on this column, I get level control this way. On this column, I get level control this way, pressure control this way, and then I get my pressure controllers and level controllers as before, and level control this way. Now what do I do with, this chap the conventional thing to do why did I put in the furnace there, it was put to control the temperature at of the reactor feed, so that is what I do using this.

Of course, what do I do with this valve, of course, I shall set that cooling rate such that; temperature of the shell is maintained constant at whatever was its near optimum value over the entire throughput range, which was 367 degree Celsius, or something like that. Then what do I use this valve for, well this valve will be used for keeping the pressure of the reaction section, how many valves are left now. I am left with this valve, that valve, you know 1 2 3 4 5, I am left with 5 valves. The total benzene flow reflects the inventory of the benzene, and therefore I do a total benzene flow control inside the process, and this make sure, that I do not put in too much benzene, or too little benzene I put exactly the amount of benzene; that is required in order to heat up the C 3; that is being fed. So, I have taken care of that.

Now let us start doing some stuff, I am going to set the reflux in ratio with the feed to the column. So, I take the reflux, the feed, I multiplied by the set point, and this is going to set the, reflux into the column, this is set. Then I need to control the, of course the inventory of, propylene dropping down is controlled by doing this, I also, but you see. On this column maybe I should change the colours, because. On this column what do I do, I say well boil up needs to be max, so if boil, but I do not really have not done the optimization. So, conventionally what I would do is, I will control a tray temperature, and there will be a V 1 max, or V 2, this is second column, so V 2 max, and of course between these two the low select is what will manipulate the steam. Well there is also, just a second, because it is just to be consistent, it is actually setting the boil up.

Similarly on the second column, on the product column, which is the third column, I have a similar arrangement, temperature controller goes to a low select, V 3 max, and then I have a boil up controller, and the low select output is setting the boil up. Then which valve is left, I have this valve that is, left why do I use it for; of course, I have to use it for product purity control, and therefore what I basically say that I want to do this composition control, which composition C 12, by adjusting the reflux. This is what I would do conventionally. Of course, I would also have another composition loop, because I have C 6 as another impurity here. So, this will be C 6, this composition loop will set, the set point of the temperature controller, so just a second. So, this guy is setting this chap. Now that I have done this, my control system is complete, now let us say I am increasing throughput.

Strictly speaking I want operate the process even in mode one, at as high as a boil up as possible, in the recycle column, so I want V 2 max, I want the temperature to be controlled, because if the temperature goes uncontrolled, what that will mean is that too much benzene can leak out and contaminate my product, I cannot afford to lose, benzene product impurity control. So, therefore, this temperature controller, on the recycle column, has to be always on; what that means is, I cannot have V 2 max ever going active, because if it goes active, then I am in trouble, I am losing benzene impurity control. What about this chap, the temperature controller on the on the third column. Well again I cannot afford this temperature control to be lost, because if it is lost then too much cumene will leak out the bottoms, and cumene is precious, that represents a significant economic loss, so now how do I manage it.

I say that well, in order to keep V 2 at max because that is what I want to do, now I am putting the optimal constraint controller, since V 2 has to maxed all the time, or close to max all the time, what I will do is, I will adjust I put a supervisory controller, that takes the boil up, do a valve positioning controller, which basically sets the total benzene, and each set point will be V 2 max minus delta. It will not be V 2 max, but slightly below V 2 max, so that, should there be transients that cause the temperature in the column to deviate. It should not happen that I lose the temperature control, because V 2 max got active. So, I will have to have some amount of back off to make sure that V 2 max never goes active, regardless of the worst case transient that this process has to deal with. So, the first disadvantage is, in this control system, you have to have back off in V 2 max.

Now as I am jacking up the through put; that means, the operator is increasing the C 3 set points, he does not know when to stop, what happens is V 3 max goes active. Once V 3 max goes active, either I put in a loop, that goes from here to here, that will be a very long loop, or I alter the material balance control structure, which is what I do, how do I a alter the material balance control structure, I am going to put that control system here. So, there will be a low select here, there will be an override temperature controller, because I cannot afford to lose, and of course set point 2, and this is set point 1, I will have like I discussed earlier set point 2 must be less than set point 1.

Now when level control is lost, I need an alternative handle, and that is what I do here. Again I have a low select, and I have an override level controller, how should I draw it, so there is an override level controller, whose output is going to this chap, and these are input's, this is the output. And once, this override level controller is lost. Well what I do is, what I do is over here, I also have a low select, what that does is, override level controller; that basically, and this is the operator input set point, flow set point, flow of the propylene set point. Now that I have drawn it, let me just explain it. See the set point of this override level cont. OLC set point is greater than LC set point, corresponding nominal level controller set point. So, the override controller set point is greater than the nominal level controller set point, the same level.

Now let us see how things go, at low throughput's V 3 is not active. So, because V 3 is not active, temperature control is using boil up, why is temperature control using boil up, because the override temperature controller set point is less than the set point of the nominal controller, and what that basically does, is that the override controller see's that look my temperature of the tray is actually higher. So, if the temperature is higher, I need to put in more feed to get the temperature back, and therefore the output of this controller is actually high. Because the output of this controller is high, the level control you know the level control of you know basically this signal gets passed. So, the feed to the product column is under level control of the recycle column, bottom to the re recycle column and so on so forth.

Now, let us say I am increasing this set point, as I am increasing this set point, what will happen is, V 3 max will become active. When V 3 max becomes active, temperature control will be lost of that sipping tray, this temperature will then, because temperature controller is saying increase the boil up, but V 3 max has gone active, boil cannot be increased, therefore temperature will start to go down. As temperature starts to go down, it will go below the set points of the override temperature controller. Now the override temperature controller will say oops, my tray temperature is lower; that means, I am putting too much feed, so this output of the override temperature controller will start to go down. Ultimately it will go below the level controller set point, and therefore my control will pass from the level controller, to the override temperature controller.

So, the feed to the column, to the product column, will now be under override temperature control. Now the level is uncontrolled on the recycle column bottoms. So, since this level is uncontrolled, and I am putting in too much feed, what will happen is. The level of this column, the level of this bottom, the bottom sump will start to go up. As this level is going up, what will happen is, this level will go beyond the set point of this override level controller. As it goes beyond the set point of this override controller, the override level controller will say oops, my level has increased too much, I need to reduce the feed, and because the override level controller output is now decreasing, it will ultimately go below, it will ultimately go small enough, so that low select selects it.

And therefore, what will happen is, the feed to the recycle column, will now be controlling the level of the bottoms of that recycle column. And now what will happen is, the level on the purge column bottom will be uncontrolled. Therefore, this level will start to rise, and ultimately this level will rise so much that the override level controller on the purge sump column, sump, on the purge sump, that override level controller will say oops, my level has risen too much, reduce the propylene feed, and ultimately this low select will pass the level controller set point to the propylene feed flow controller. And therefore, what happens is, once V 3 max goes active, I basically alter my material balance control structure from V 3 max all the way up to the propylene feed, this is what I do. What is the disadvantage; a, please note that you need three override controllers.

Also please note that because SP 2 is less than Sp 1, you will actually have more leakage of precious cumene, down the bottoms, that will happen, that is just to make it work. And the further, the difference the more the leakage of precious cumene downs the bottom, so you will have more loss of precious raw material. Also notice that in this scheme the other constraint which was V 2 max, you will have to operate at some amount of back off. This back off is delta, and therefore what this means is, in this control system you get greater cumene loss, in the d I p v bi product stream, as well as the process will operate on average at a slightly lower, V 2 at a boil up that is slightly lower than V 2 max, which basically means that the selectivity will be slightly lower. So, the selectivity will be slightly lower, and also cumene loss will be slightly greater.

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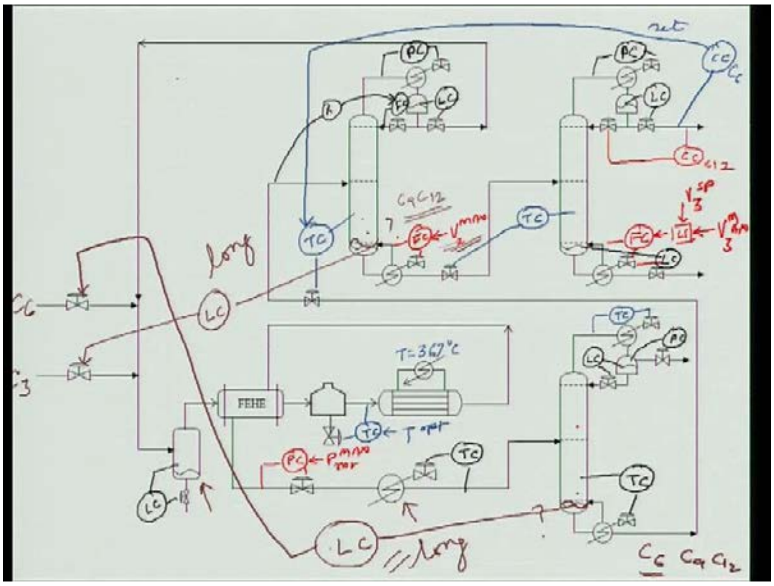

Now let us contrast it with our approach of doing things. How do we do things, well what we say is V 2. Let us like look at the maximum throughput constraints, all constraints that are active in maximum throughput need to be tightly controlled, if that is what I want to do, what I will do is I will basically put in a loop that does this, I will put in a loop that does this, and I will say this set point, because this control going to be very tight. This set point can be set at V 2 max without any back off, or negligible back off, this will be V 1 max. Then what was the other constraint that was active. The other constraint active was P reactor max, so this is going to be pressure this is the same as before. So, here I get, because I want to operate the reactor at maximum pressure, also I have this loop, because composition control of C 12, is by adjusting reflux, and this will a nice and tight loop, and this is being, let us see also.

So, these loops I have put in place, just based on constraints at maximum throughput, what are the other economically important things. The other economically important things are; I want temperature to be as small here as possible, so that I get minimum benzene loss; that is as P 4. I also want temperature tight control of the temperature, into the reactor, because the reactor is quite exothermic, so that will do that. I also set this temperature is equals to 367 degrees Celsius, so that. So, this is also set at its optimum value T optimum, whatever it was, I think it was 330 degree Celsius or something. So, the reactor inlet and the reactor shell temperature are set at appropriate near appropriate values, that ensures that the conversion and selectivity are maintained.

Now that I have done that, what else will I do, let us see. I need to tightly control the benzene leaking down the bottoms, so to do that, I put in. Economically important things, you see I want tightly control of this chap, however boil up is max, because boil up is maxed, control this temperature cannot be controlled using boil up. So, what are the options that I have, I can control it either using the feed or the reflux, since the feed is closer, it will have a faster impact, so what do I do, I control temperature this way. I also need the tight control of the benzene leaking down the bottoms of the, bottoms of the recycle column of this column.

However, its boil up is also at maximum, so therefore, I cannot manipulate the boil up. So, what do I do, similar logic I control temperature using feed to this column, and of course I have a benzene composition controller, composition of C 6, and that composition controller is essentially setting this set point, so this is setting that set point set. So, that takes care of economically important things, what else; of course, I want to make sure that, you know too much c three does not leak down the bottoms, that gets done using this. What else, I also need to make sure that whatever are the condensable, in the reactor effluent they are condensed, this ensures that, what else. I need to ensure that the reflux in the recycle column is high enough, so that not too much cumene leaks up the top of the recycle column.

So, what do I do, as I was doing before I do it the same way. Anything else, so this takes care of most of the all the economically important things, that have a steady state effect, I have put in the loops. Now let me put in the material balance control structure. So, for material balance control, I need to have pressure control, pressure control and all the columns will be basically conventional, so I got pressure control, and I have got same thing pressure controller. Now how about level controllers, well this is conventional level control, this is also. Well how do I put it, just put a level controller here, and it is conventional. How about level control of this chap well, I have got level control this way, what about level control of this chap, well I have got level control this way local pairing.

Now I have problem in the sense that, what about this level, you will notice that I have basically used up all valves, that are local to this column. What about the level of this chap, you will again notice that there are, well I forgot a loop, oh just a second. So, of course, there is level control here this way. So, once I put all these things in place, what I find is that the levels in brown, there are no close by valves, that are available to control those levels. So, how do I control this level, and how do I control this level. I can see that the only valves that I have available, are actually the too feeds; the C 6 feed, and the C3 feeds, C3 C 6. So, what do I do, let us see if it makes sense. Let us say I control the, what does this level represent, on the recycle column. This level actually represents C 9 and C 12 which are separated downstream, in the product column.

C 9 and C 12 are essentially reaction products, and if this level is increasing, what that means is, that I am producing too much C 9 C 12 in the reactor, which is accumulating in this column bottom. So, what that means is, I need to reduce the ethylene, to reduce the amount of C 9 and C 12 formed in the reactor, so that this level may go back; that makes sense, what about this level. Well this level is, because benzene is in large access in the column, so this is what accumulates here is a lot of C 6 C 9 and C 12, but C 6 will be the dominant component, so what I can actually do, is I can control this level, using a C 6, and I can control this level in this chap. You can see this level loops are long, contrast this with what we did previously, you see here I have a long economic loop, this is a long loop, for V 1 max, for ensuring V 2 max, column two is operated close to V 2 max.

And I also have a long override scheme, that alters, you know from here to here to here, well to here. I have a long override scheme that alters the control structure, when V 3 max goes active. Contrast this with this chap, did it go. Here I am having V 1 at V 1 max, no back off, V 2 at V 2 max, no back off, but what I have are, long level loops, two long level loops. So, there were two economic constraints, in the previous case what I had was, one of the economic constraints which is, sorry this is V 3 max, one of the economic constraints which was V 3 max; that was controlled using overrides. And the other economic constraints which was V 2 max, well you had to have back off their, and why did you have to have back off there, because this loop out here was a long one.

So, you had a long economic loop, and local level loops. In this case what do we have, we have local economic loops, and long level loops. So, I have got long level loops, and local economic loops. This is a long loop, this is a long loop. So, that is the difference in these two approaches, now the question is, will these level controller work in practice. Well it turns out, that, what do you have in these column you have a small amount of lag in the vaporiser, which is small, and then you essentially have flow through a pipe, because everything has flow through the pipe, because there is a small lag associated with this chap, there is a small lag associated with the condenser, but the rest of it just flow through a pipe.

So, really speaking, when I change the benzene, when this level controller changes the benzene, its effect will reach the bottom, actually let us say in 5 or 10 minutes. Since my sump is having a resistance time of 5 or 10 minutes, this level control should work reasonably well. Similarly the same logic also holds for this chap, and it turns out that we did the simulation, and it did turn out that for the disturbances that we considered which are pretty large. You are able to operate the process with these levels, fluctuating in the larger band, but not hitting high level or low level alarm limits. So, in this control system what I have done, is essentially transformed the variability to the levels. Levels do not affect my economic steady state.

Since levels do not affect my economic steady state, I would like to transform my variability as much as possible to the surge levels. This is a surge level, this is a surge level, and that is what this control system is doing. So, I hope this example actually brings out the difference in the two approaches. So, the trade off is basically this, you could have local level loops, and long constraint controlled loops, what should I say, active constraint control loops, this is the conventional approach. In our approach what we do is, what we are basically doing in our approach, is long level loops, instead of local level loops, and we get local active constraint control loops, instead of long active constraint control loops. So, because active constraints back off from an active constraints affect my economics significantly, this gives me significant economic advantage; alright.

Now let us see, just to complete this case study, what do I want to say, I want to say, if you are, let us say you want to reduce the throughput, see this control system will work at maximum throughput, if I want to reduce the throughput what do I do, I basically say well low select V 3 max, and of course the operator inputs a V 3 set point. So, when V 3 set point is less than the boiler the flooding limits of this column, if I reduce this set point below V 3 max, you see the boil up will reduce if the boil up reduces, temperature will reduce, if the temperature reduces the temperature controller here, will reduce the feed. If this feed reduces will, if this flow reduces the level here will actually start to go up, if the level, if the level here starts to go up, this level controller will basically reduce the propylene, so this is how throughput will reduce.

So, it is a very simple scheme, no alteration of the material balance control structure depending on which constraint is active, and which constraint is not active, regardless of where you are operating regardless of what throughput you are at, the basic inventory management scheme remains the same. So, that is that, I think this case study actually sort of contrast, brings out the contrast between the conventional pairing approach, and the actual pairing approach, and the economic plant wide control approach that we are proposing. So, I think that takes care of that, with that we have got still some time. So, let us see maybe we can do another case study, but I will leave, we have done enough numbers of case studies, may be what I want to do is, basically summarise what we have done in this course, and then just leave it at that. So, before I end, I just wanted to add one tick pick on override controls.

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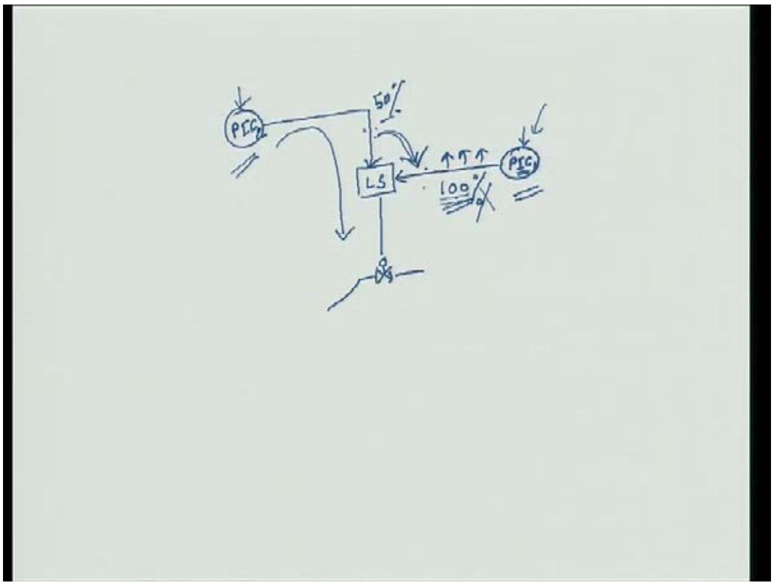

So, let us say I have got a low select, and let us say there are two controllers. I do not want to say what controllers, they are let us say, they are moving a valve, it is a general arrangement. And there is, what should I call it? Let us say there is a PI controller number one, and there is PI controller number two, which is controlling something it is not really relevant what you are controlling, but as long as it is clear to us that, this valve is being adjusted, to maintain either this or that, and that depends basically on which signal is greater of the two, now the a and both of these controllers are PI. If both of these controllers are PI, let us say what we have is, we have the operating condition is such that, this valve is under control of this chap. Then what happens is, this guys off.

So, whatever it is trying to control, it will move away, and by design what will happen is, in order to get its control back, what this guy will say is, its output will keep on increasing.

Because its output will keep on increasing, let us say this output goes and saturates at 100 percent. Let us say this output is at may be 50 percent. So, what happened was, when this output increased above 50 percent, control passed to this chap, and now because control has passed to this chap, and this chap, which is this chap is uncontrolled. To get control back what this guy is saying is, well move the valve, move the valve, and the movement of the valve, will actually be in the direction that takes this away from this. Now let us say that the operating condition changes, and now control has to pass from this guy to this guy. So, control has to pass from this guy to that guy. Now what happens is, this output is at 100 percent. The integral action in this controller, actually caused this output to reach its saturation limit. Now what has to happen is that this controller has to unwind from 100 percent to 15 percent to 50 percent, before control passes from here to here. So, there will be a very long period, where control did not pass from here to here. Shinskey has proposed what is called an external reset wind up scheme, and in the external reset wind up scheme, just to explain that, what I wanted to, just to explain that, let us see what external reset actually means.

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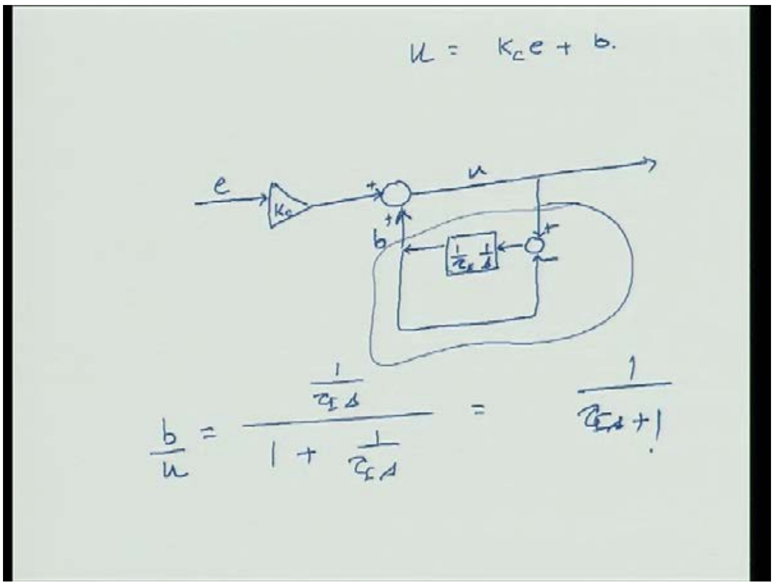

If you look at a PI controller, to have the error signal that comes, you multiply it by K c, and then there is a bias term that you add, a bias term. Let us just call it b, and this is the. So, u this is the, so u is equal to K c times e plus a bias term. Now let us say, I put in an integrator, with a certain gain associated with it, let us just say I put in 1 by tau i this, and I try and make sure that the bias matches the output of the controller. So, this is the bias, and this is the output of the controller. So, now if I look at this guy, this guy is actually, b is equal to by block diagrams 1 by tau s, whatever is in the forward path divided by 1 upon 1 by, whatever is in the forward path divided by 1 plus, whatever is in the forward path that is. So, b by u is equal to this chap, and if I solve this will come out to be actually 1 over tau i s plus 1. So, this means this chap over here, that is trying to make sure, that the bias is adjusted to match the u, this chap can actually be replaced by simply a lag.

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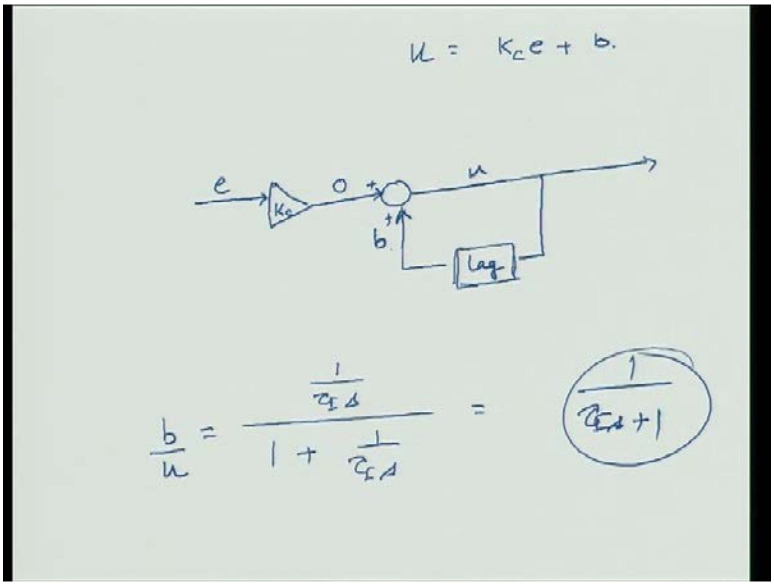

A simple lag, and this lag is actually this chap, lag of gain one, and time constant tau i. Now at final steady state, what will happen, at final steady state because this is just a lag, b will be actually be equal to u. If b is actual to u, what must the signal be the signal is actually K c e, K c is non zero. If K c is non zero, this guy at final steady state, this guy must be 0, unless this guy is 0, b cannot match u. So, this is another way of looking at PI controller, a PI controller may also be looked at in this way, where you are adjusting the bias so that it matches the controller output. So, this is again a PI controller, and just to convince you it is actually a PI controller, let me just show you.

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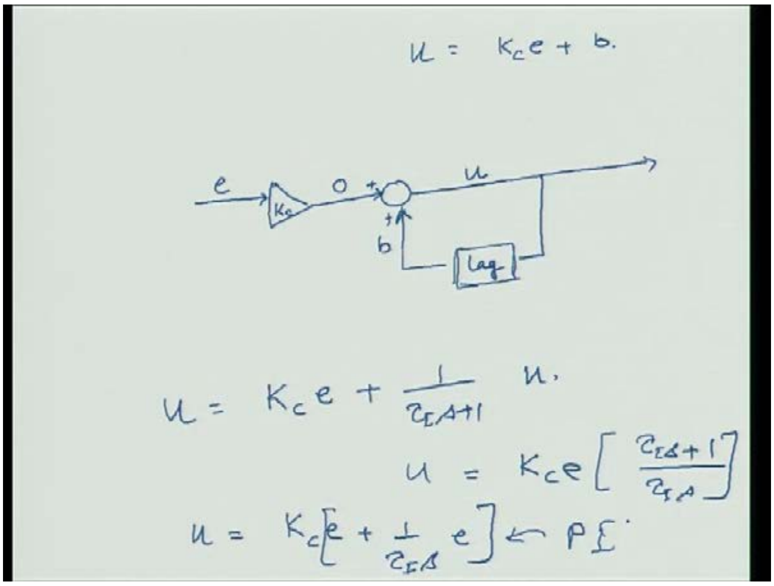

So, what is u, u is equal to K c times e plus b, and b is equal to u times the lag, and the lag is 1 by tau s plus 1 times u. If I collect the terms of u on one side, what I will get is, u times 1 minus 1 by tau i s plus 1, is equal to K c times e, and this will give, essentially tau s plus minus 1. So, one and one will cancel out, so what I will get here is, tau i s over tau i s plus 1, and then what I will get is, u is equal to K c e times tau i s plus 1 divided by tau i s, and then what I will get, is u is equal to K c e, K c times e plus 1 by tau i s of e. This is the PI controller transfer function, this is nothing but a PI controller. So, you can see that this block diagram that I have drawn is actually nothing but a PI controller. Now in terms of overrides, the things that I was talking about, when I am doing my overrides, and let us say I get back to the same thing as before, let us say what I have is two controllers competing for control of a valve; two PI controllers.

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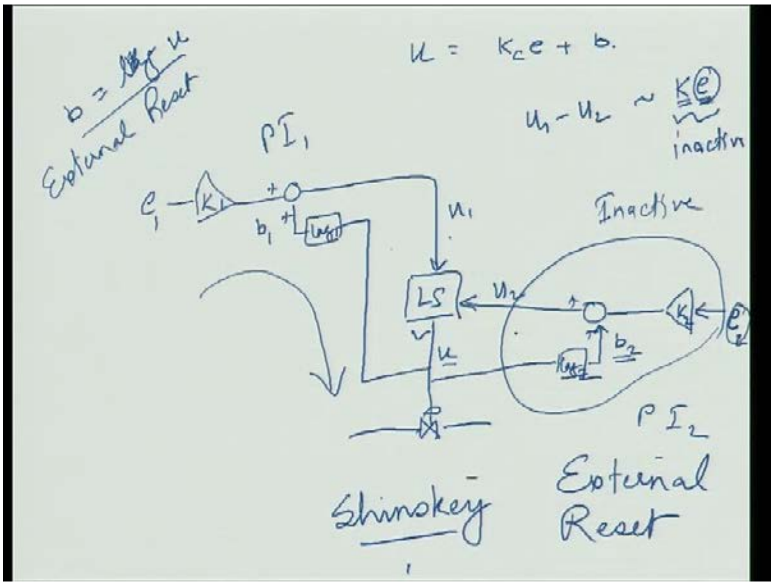

So, I have PI controller number 1, and I have PI controller number 2, and both of these let us say are competing through a low select or a high select. These are the two outputs, and this is basically going to a valve, the block diagram that I just drew, the reset action b, is taken from the implemented controller output u, case, well b is equal to a lag times u, so that is what I was doing. This is actually called shinskey calls it external reset. Now when two PI controllers are competing for the same valve, instead of taking your reset action from here, and from here. So, basically this PI controller will actually be, some K c, some K 2 times error in the second variable, and I have a lag, which gets added here. Similarly this PI controller can be seen as, you have got the error in the second variable, you multiply by, in the first variable sorry you multiply it by again, and then add to it lag one, and let us just call this lag two; these are the two controllers that I have. So, this is PI 1 sorry, this is PI controller two, this is same thing PI controller one.

Well if I take this external reset from whatever is implemented, you see this is the signal that is getting implemented. If I take external reset from whatever is implemented, then please note that this bias, and this bias, will keep getting updated to match, you know the bias will keep getting updated, so that the implemented signal the bi the b 2 will get updated, so that b 2 is close to you. Similarly b 1 will get updated, so that b 1 is close to you. So, when, let us say this guy is inactive, if this guy is inactive, if this chap is inactive; that means, u v you know the, this signal is getting passed, this signal is actually passing through. So, if this chap is inactive, the second controller is inactive. Then what will happen is, even though control this signal is getting passed, this bias b 2 will keep getting updated so that it matches u, what that means is, this signal will differ from this signal, only by whatever is the error here. And if this error remains small; that means, the deviation error does not change by much.

Then what that means is, this guy will not go too far away from that guy, and the problem that I was talking about that, this controller must unwind from 100 percent to the value here; that must happen, that need not happen, because now the passing over of the control is based purely on proportional action, not on the integral action. So, this is actually a very neat trick, this is called external reset by Shinskey, Shinskey calls it, by the way. Shinskey probably got everything right, Shinskey calls it external reset, and he came up with very simple solutions for problems that you would face, for practical problem that you would face, external reset. So, when you take the reset from the control signal that is implemented, then you will make sure that, if this is u 1, and this is u 2, and you basically make sure u 1 minus u 2 is essentially of the order of K times e of whichever, you know this K is of whichever controller is inactive, whichever is inactive of the inactive controller.

And if these e does not go too much, grow too much, the difference between what is implemented, and what is not implemented will not be too large, and then this reset wind up problem that basically requires controller to unwind substantially, the controller, because the controller outputs are substantially far away, that problem does not occur. This is one thing that I wanted to tell you about overrides. So, wherever a PI controller is competing for a valve, its reset action must be taken as an external reset of the signal that gets implemented, regardless of whether the controller is active or inactive. This was the one thing that I wanted to tell you, and I think with that it is time to summarise what we have done.

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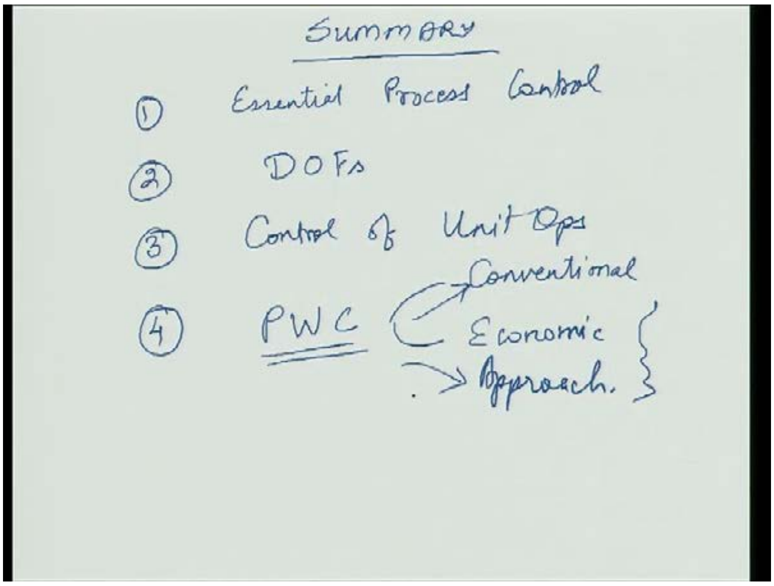

In summary all I want to say is, without getting too mathematical what we have covered are; essential process control fundamental, essential process control that is it. Things that are of use in practice, not too much of theory this way that way advance process control etcetera, but essentials, then we looked at degrees of freedom, and came up with a very intuitive way of figuring out, what the degree of freedom of a process is. Then we looked at, control of ordinary unit operations, control of unit ops. Finally, we looked at plant wide control, and what I have tried to show you, is that there is a conventional approach, in the conventional approach. And then there is the economic approach with constraints, and it is in this economic approach, that you pay a loop first for what is economically important, and then do the material balance control, and I hope that all of this, sort of convinces you that look designing a plant wide control system, is like piecing a puzzle together. How you piece the puzzle together, what approach you take to piece the puzzle together; one approach is the conventional approach, the other is the economic approach; that is where the what should I say, that is where the art is, that is where all the confusion is, and I hope after going through this course, at least some of the confusion will get clarified. I think that is a good place to end this course.

Thank you.