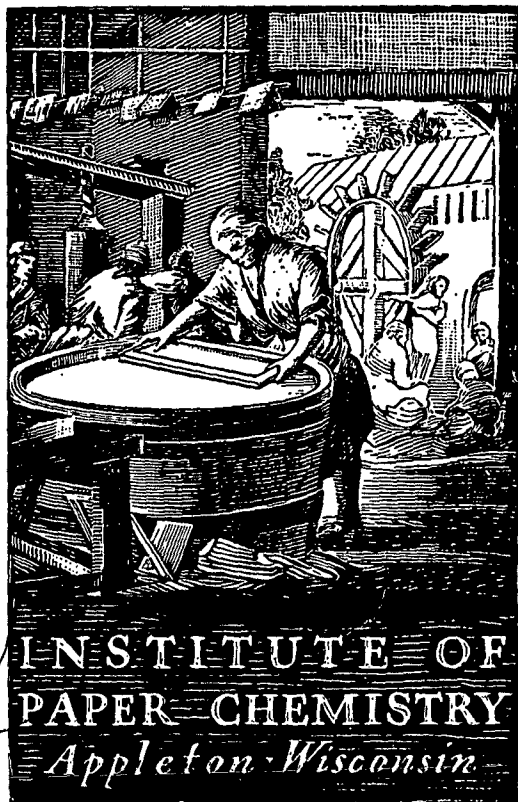


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**A STUDY OF EVAPORATOR SCALING**

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**Report Four**

**AN EVALUATION OF THERMAL DEACTIVATION**

**A Progress Report  
to**

**MEMBERS OF THE INSTITUTE OF PAPER CHEMISTRY**

**January 3, 1979**

THE INSTITUTE OF PAPER CHEMISTRY

Appleton, Wisconsin

A STUDY OF EVAPORATOR SCALING

Project 3234

Report Four

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MEMBERS OF THE INSTITUTE OF PAPER CHEMISTRY

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THE INSTITUTE OF PAPER CHEMISTRY

Appleton, Wisconsin

AN EVALUATION OF THERMAL DEACTIVATION

SUMMARY

Thermal deactivation is a process for controlling calcium carbonate scale in black liquor evaporators. The process consists of heating the black liquor to temperatures on the order of 300°F and holding it at this temperature for times up to 15 minutes. This converts the calcium in the liquor to a form which does not lead to scale formation when the liquor is subsequently evaporated.

Thermal deactivation is an effective method of minimizing calcium carbonate scales. Prior indications of effectiveness in laboratory tests were confirmed in pilot trials at a kraft mill. Scale accumulation was reduced about 90% during evaporation of deactivated liquor compared to the same liquor not subjected to deactivation.

Heating within the deactivation process can be accomplished either by direct steam injection or with an indirect heater. The latter is subject to scaling, which can be minimized by transferring the heat to recycled, deactivated liquor and mixing this with untreated feed liquor. Tests in the laboratory showed that the use of recycled liquor reduced scaling in the liquor heater by 90%. Direct steam injection has fewer operating problems, but is more energy intensive.

Commercial applicability of deactivation is determined by economic considerations. The key economic factors are the value of increased evaporator capacity, the severity of scaling, and the number of effects where significant calcium carbonate scaling is encountered. Deactivation economics also depend on the liquor heating method, the extent and conditions of by-product power production, fuel and electricity costs and the cost of boiler feedwater treatment.

Analysis indicates that, except for severe scaling, deactivation costs cannot be justified by savings in cleaning costs alone. It is necessary to be able to take credit for increased pulp production in order to generate a favorable return. In this case thermal deactivation must be compared with alternative methods for increasing capacity. Deactivation is most attractive for those mills where severe calcium scaling problems in the first one or two effects cause a bottleneck in pulp production. However, deactivation does not show a clear cut economic advantage over other methods of increasing evaporator capacity for the range of scaling rates normally found in black liquor evaporators.

## INTRODUCTION

Scaling of black liquor evaporators has been studied at The Institute of Paper Chemistry for a number of years. Results of work on calcium carbonate scales were reported in Progress Report Three (1), issued on November 15, 1977. It was found that the formation of calcium carbonate scales involved an interaction between calcium and organic components in the liquor. The scaling process is very sensitive thermally, with scaling rates doubling for as little as a 5 to 7°F temperature increase. We believe that calcium carbonate scales are due to the transfer of calcium from an organically bound state to calcium ion under thermal stress, and subsequent reaction with carbonate ion to produce  $\text{CaCO}_3$ .

During the course of the work it was found that calcium scales could be prevented from forming by subjecting the liquor to a process which we called thermal deactivation. The process involves heating the liquor to about 300°F, and holding it at this temperature for 10-20 minutes. This combination of time and temperature converts the scalable calcium in the liquor to a stable, nonscaling form. We believe that the stable form is suspended  $\text{CaCO}_3$  particles and that the deactivation process is directly analogous to the scaling process except that in deactivation the  $\text{CaCO}_3$  is precipitated in the bulk liquor rather than on the heated surface.

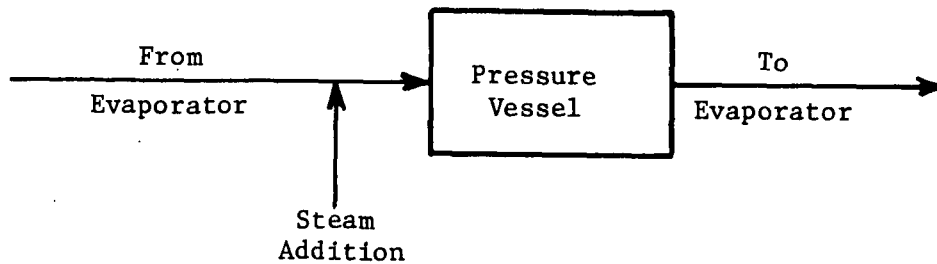
Extensive laboratory and bench scale tests on deactivation were carried out and are described in Progress Report Three (1). These showed that deactivation was workable and that the scaling tendency of treated liquors could be reduced by over 90%. The effects of key process parameters were determined. The results suggested that deactivation could have commercial applicability. Accordingly, the focus of work on deactivation shifted to resolving process design problems, testing the workability under mill conditions and evaluating process economics. The results of this work are reported herein.

## PROCESS DESIGN

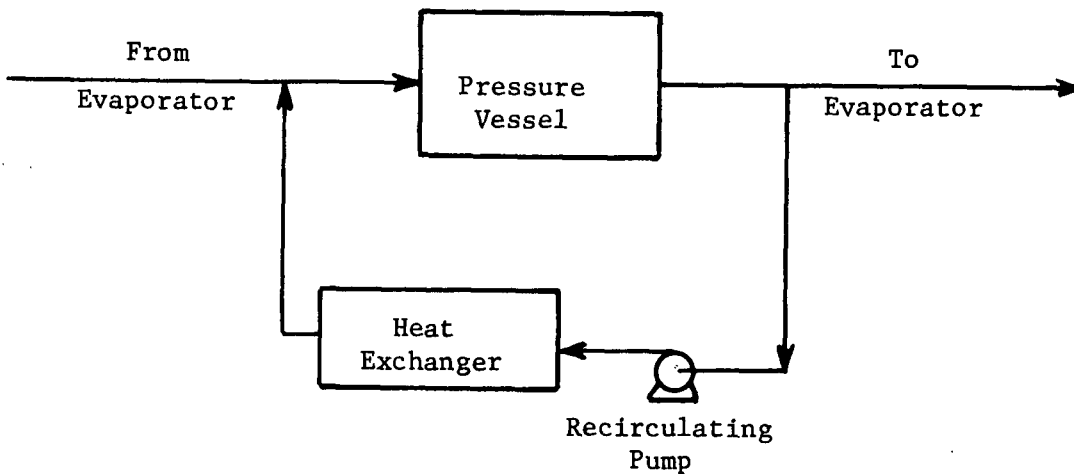
Deactivation is a simple process which involves raising the temperature of the liquor and providing a certain amount of residence time at the elevated temperature. Hence, a heater and a retention vessel are required. Since the necessary deactivation temperature is above the normal boiling point of the liquor, the system must operate under pressure. This requires a pressurizing feed pump and a pressure letdown device and flash tank for the treated liquor exiting the process. The equipment required for thermal deactivation is all standard, so that the process requires no new or untried equipment designs.

The only concern in deactivation process design was the liquor heater, since scale would be expected to form when undeactivated liquor is heated. One solution to this problem is direct steam injection since this would eliminate a hot surface prone to scaling. This approach is shown in Fig. 1a. There are disadvantages to the use of direct steaming, however. The condensed steam dilutes the liquor and increases the evaporator load, and the steam condensate is lost and not returned to the boiler. For these reasons, a method of indirect heating which would not be subject to scaling was desired.

The solution to this problem was the recycle heating concept. In this approach, heat is transferred to liquor that had already been deactivated. The hot, deactivated liquor is then recycled and mixed with feed liquor. This concept is shown schematically in Fig. 1b. It may be noted that the temperature of the recycled liquor is higher than that reached in the retention vessel where deactivation is completed. Since laboratory work had shown that the extent of deactivation depends on the temperature reached, there is a possibility for some scaling in the recycle heater where the liquor temperature is higher than the final deactivation temperature. This was the main question about the feasibility of using an indirect heater for deactivation.



a. Direct Steam Injection



b. Recycle Liquor Heating

Figure 1. Thermal Deactivation Process Configurations

## PILOT TRIALS

Previous work on deactivation can be divided into two categories (1):

1. Small scale (a few gallons) treatments using either a batch autoclave or direct steaming in a small flow reactor. The effectiveness of deactivation was determined by either measuring the "soluble" calcium content of the liquor or by scaling tests using electrically heated strips. Liquors from a number of different mills were used.
2. Larger scale (several hundred gallons) treatments in a flow system using a once-thru indirect heater. The effectiveness of deactivation was determined by runs on the single-tube evaporator at The Institute of Paper Chemistry. All of this work was done with liquor from a single mill.

At the completion of this work, a number of questions remained concerning the feasibility of the deactivation process. All of the liquor used in the program had been in transit or storage from a minimum of 1 week to several months. Since the deactivation process is closely analogous to an accelerated aging of the liquor, there were some questions about its workability on fresh black liquor during evaporation. There was also a need to prove that the recycle heating concept worked and to obtain experience with deactivation system performance over longer periods of time.

To provide answers to these questions and to demonstrate conclusively the technical feasibility of deactivation, a pilot deactivation unit was constructed. The pilot unit was capable of treating liquor at a rate of about 1 gpm. It was used in trials at a kraft mill. The objectives of these trials were to:

1. evaluate the effectiveness of deactivation using liquor taken directly from a mill evaporator train
2. obtain operating experience on continuous deactivation systems
3. evaluate the effectiveness of the recycle heating concept in minimizing scaling of the deactivation heater.

The pilot deactivation trials were conducted at Westvaco's Charleston, S.C. mill. This mill was chosen since its evaporators had a  $\text{CaCO}_3$  scaling problem and since a pilot evaporator system suitable for determining deactivation effectiveness was available there. Following the tests at Charleston, additional experiments on recycle heating were carried out at IPC.

#### PILOT SYSTEM

Figure 2 shows the deactivation system used for the Westvaco trials. It was designed for 1 gpm liquor flow at 10 min residence time. The unit had a liquor preheater (before the retention vessels), a recycle heater, and a steam injector for direct heating. A liquor cooler was used for simplicity of discharge liquor temperature control in the pilot trials. It would be replaced by a flash tank in a full-size deactivation system design.

The pilot evaporator system at Charleston consists of two parallel, single tube evaporators, constructed from 2-inch OD evaporator tubes with approximately 20 ft of heated length. Feed liquor from the second or third effect of the mill evaporator train was pumped to the pilot evaporators. Flow was split so that 1 gpm of liquor went to the deactivation system before one of the evaporators, and 1 gpm went to the second evaporator without treatment. This allowed a direct comparison of the scaling rates of untreated versus thermally deactivated liquors that were otherwise identical. Figure 3 shows the flow arrangement used.

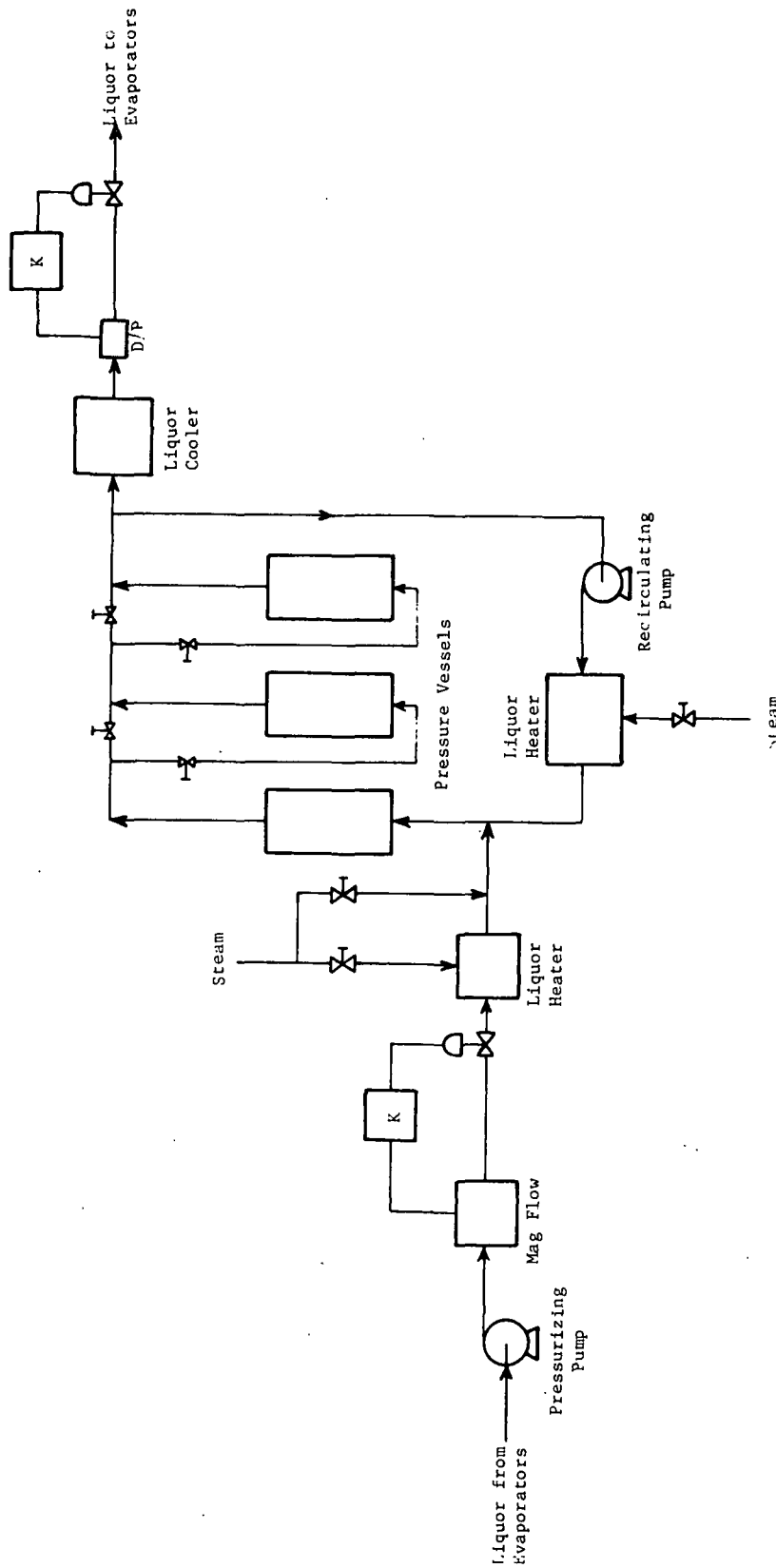


Figure 2. Pilot Thermal Deactivation System

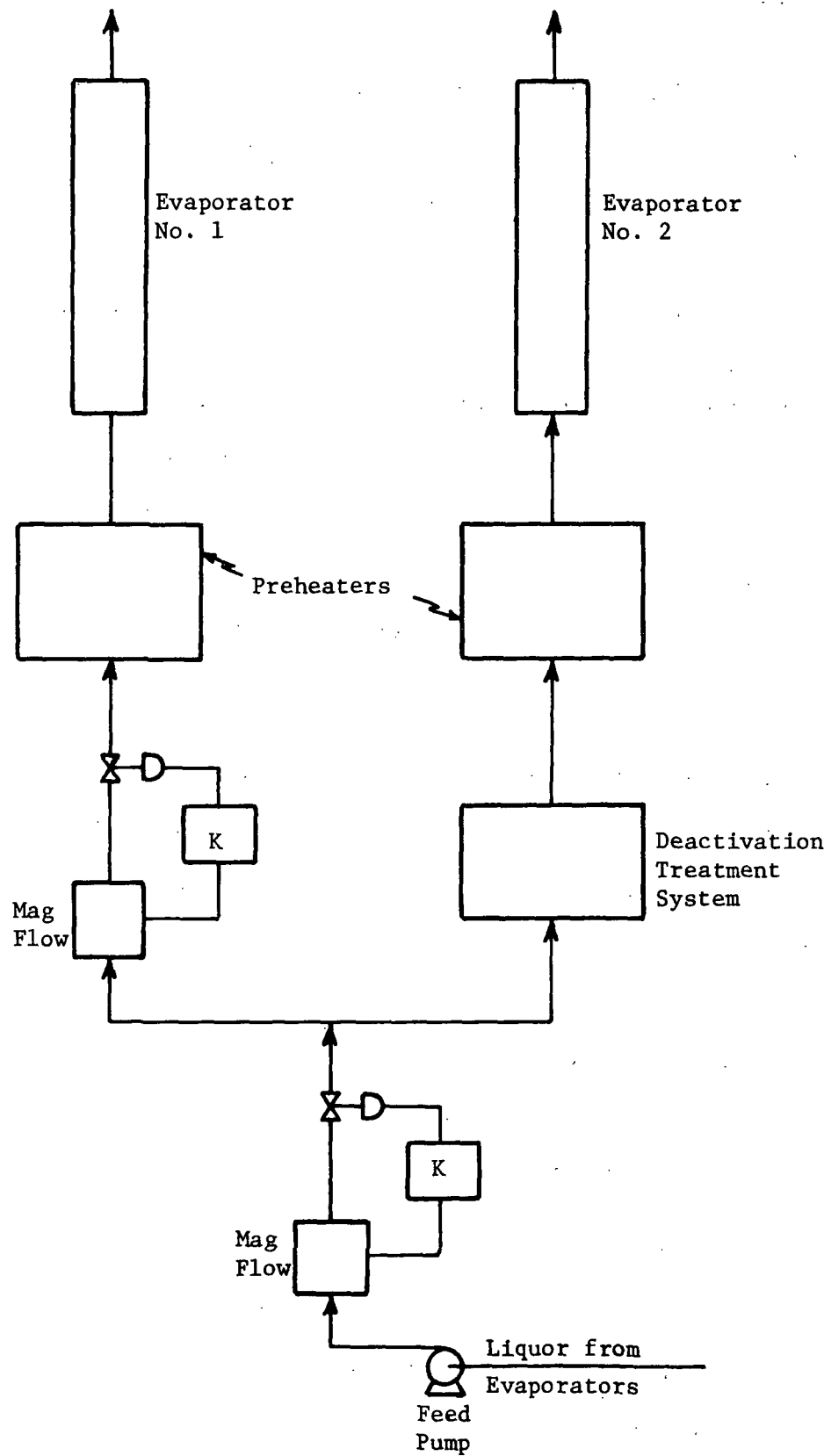


Figure 3. Westvaco Experimental Evaporators with Deactivation System

## TEST PROCEDURES

The pilot deactivator/evaporator was started up by circulating product liquor from either the second or third effect of a mill evaporator train through the system. The liquor flowing through the pilot deactivator was pressurized and heated until the exit liquor temperature reached 300°F. At that point, steam was admitted to the shells of both pilot evaporators. This began a deactivation run.

The evaporators were operated at nearly constant evaporation rate (1-1.5 lb/min) by increasing the shell side steam pressure as scale accumulated. Vapor head pressure was maintained at 15 psig during each run. Steam pressure to the evaporator preheaters was adjusted to give constant inlet liquor temperatures to both evaporators. Liquor flow was maintained at 1 gpm to each evaporator except during part of Run 3.

Three runs were carried out at Charleston. The first two runs employed direct steam injection in the deactivation unit. This diluted the liquor being deactivated from 33% solids to 30%. Run 1 was made without adjusting the solids of the untreated liquor, but in Run 2, water was added to the untreated liquor ahead of the pilot evaporator preheater to adjust the liquor solids to the same value as the deactivated liquor. Run 3 was made using indirect heating. The liquor solids to each of the test evaporators was the same since no dilution took place in the deactivation unit.

Operating data were collected from both evaporators and the deactivation system for calculation of heat transfer coefficients. The data collected included liquor flow rates, steam and vapor condensate rates, dilution water flow rate, shell and vapor head pressures, and liquor and steam temperatures. The ratio of softwood to hardwood liquor flow from the mill's weak liquor storage system was

monitored. At the end of Run 1, scale samples were scraped from the top of both evaporator tubes for analysis. The calcium contained in each of the tubes was determined after Run 2 by acid washing, collecting the wash liquor, and analyzing for calcium content.

#### RESULTS OF DIRECT STEAM INJECTION RUNS

Figures 4 and 5 show the results from the two direct steam injection runs. Evaporator heat transfer coefficient data are shown as reciprocal coefficient ( $1/U$ ) versus time. Since these runs were made at nearly constant evaporation rate, the slope of  $1/U$  versus time should be approximately proportional to scaling rate. The liquor solids and soluble calcium content of the liquor before and after deactivation, measured by the technique previously described (1), are given in Tables I and II.

The first direct steam injection run was made without adjusting the liquor solids of the untreated liquor. The heat transfer data (Fig. 4) show that the scaling rate in the untreated liquor evaporator was low for the first 11 hours, but increased by a factor of 4 during the last 14 hours. The increase in scaling rate corresponded to an increase in liquor solids during the same period. The heat transfer data indicate that the deactivated liquor evaporator did not scale during the run.

After the run, the inside surfaces of both columns were examined. The untreated liquor evaporator contained a scale typical in appearance to calcium scales observed previously. The deactivated liquor evaporator contained a very thin, light brown film. Samples of both deposits were analyzed for calcium. Each contained about 50%  $\text{CaCO}_3$ , based on the calcium analysis. The total amounts of deposit in each tube were not measured. However, the heat transfer data show that the total scale accumulation in the untreated liquor evaporator was enough to reduce evaporation

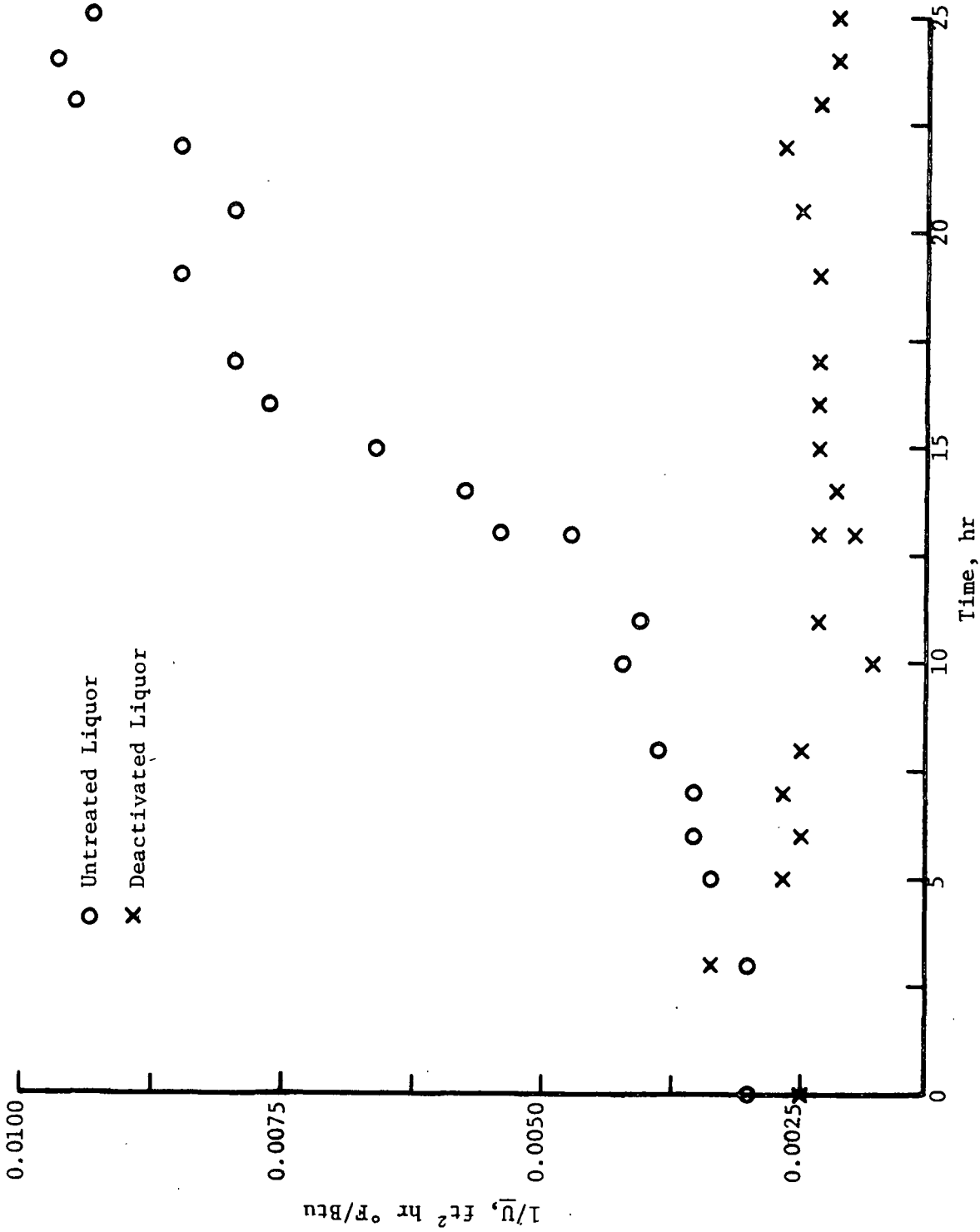


Figure 4. Reciprocal Heat Transfer Coefficient Versus Time for Direct Steam Injection Deactivation (Run 1)

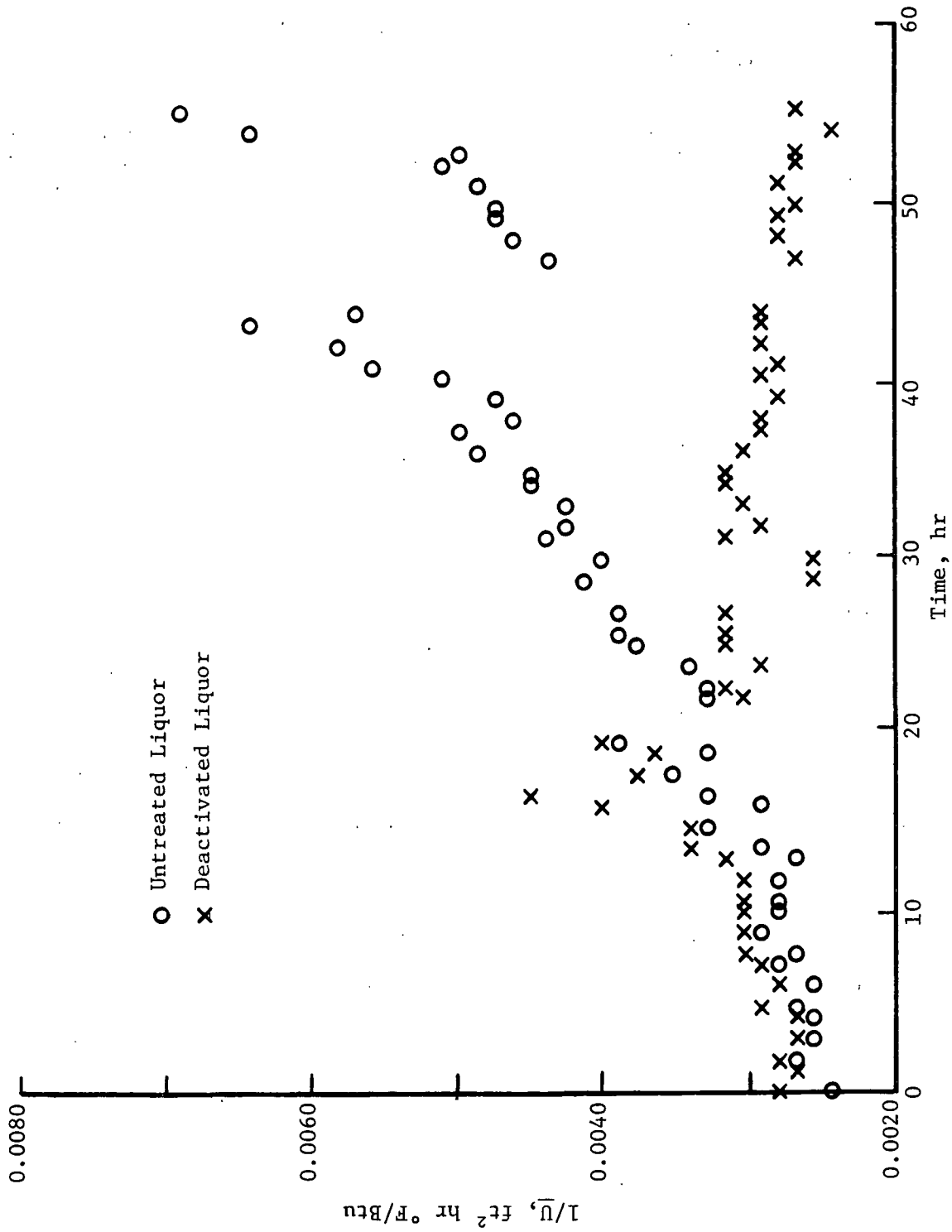


Figure 5. Reciprocal Heat Transfer Coefficient Versus Time for Direct Steam Injection Deactivation (Run 2)

TABLE I  
PERCENT LIQUOR SOLIDS FOR THE DEACTIVATION  
AND EVAPORATOR SYSTEMS (Runs 1 AND 2)

Time, hr	Inlet	Deactivation Outlet	Untreated Liquor Evaporator Outlet	Deactivated Liquor Evaporator Outlet
<u>Run 1</u>				
5.5	26.3	23.5	--	--
9.5	--	--	--	--
13.5	33.7	30.1	--	--
21.5	34.9	31.4	37.1	37.8
<u>Run 2</u>				
0	34.20	31.09	34.07	33.57
6	34.10	30.77	32.47	33.30
12	31.64	28.64	31.38	32.54
18	32.56	29.44	35.76	33.09
24	33.09	30.09	32.15	32.97
30	34.52	31.23	33.04	34.40
36	34.12	31.32	31.42	33.97
42	38.57	35.07	35.93	37.70
48	34.40	31.78	33.41	36.56
54	34.72	31.38	32.25	35.35

TABLE II  
LIQUOR CALCIUM ANALYSES DURING DEACTIVATION RUNS 1 AND 2

Time, hr	Total Calcium, % of BLS				Soluble Calcium, % of BLS			
	Untreated	Deactivated	Untreated Liquor Evaporator Outlet	Deactivated Liquor Evaporator Outlet	Untreated	Deactivated	Untreated Liquor Evaporator Outlet	Deactivated Liquor Evaporator Outlet
<u>Run 1</u>								
5.5	0.054	0.059	--	--	0.039	0.041	--	--
9.5	0.016	--	--	--	0.015	--	--	--
13.5	0.035	0.041	--	--	0.022	0.020	--	--
21.5	0.022	0.027	0.024	0.023	0.015	0.012	0.010	0.011
<u>Run 2</u>								
0	0.035	0.034	0.040	0.037	0.033	0.029	0.031	0.034
6	0.044	0.038	0.044	0.046	0.034	0.031	0.034	0.032
12	0.048	0.045	--	--	0.033	0.031	0.032	0.029
18	0.036	0.036	--	--	0.027	0.029	0.026	0.029
24	0.039	0.039	--	--	0.024	0.021	0.023	0.028
30	0.041	0.038	--	--	0.029	0.025	0.025	0.025
36	0.032	0.028	--	--	0.027	0.023	0.024	0.023
42	0.027	0.038	--	--	0.020	0.017	0.022	0.025
48	0.030	0.027	--	--	0.023	0.021	0.025	0.028
54	0.025	0.022	--	--	0.030	0.026	0.029	0.020

capacity by 67% over 25 hours (at constant overall  $\Delta T$ ), while that accumulated in the deactivated liquor evaporator did not reduce capacity.

During the second direct steam injection run, water was injected into the untreated liquor stream so that the liquor solids to both evaporators was the same. The heat transfer data in Fig. 5 show that scaling in the untreated liquor evaporator reduced evaporation capacity by 55% over 56 hours, but that no capacity was lost in the deactivated liquor evaporator. The sudden shift in heat transfer coefficient at 45 hours resulted from decreasing the liquor flow from 1.0 to 0.8 gpm.

The heat transfer data in Fig. 5 also suggest that the treated liquor scaled its evaporator from 10 to 20 hours into the run. However, the coefficient returned to nearly its unscaled value after 20 hours. The behavior may represent scaling followed by flaking off of the scale deposit, or it may indicate faulty heat transfer data during that period. The explanation is not apparent from the data available. However, the important result from these data is that the coefficient in the treated liquor evaporator is nearly constant over the entire run, indicating that no scaling occurred.

After the second steam injection deactivation run, the pilot evaporator vapor heads were removed and the evaporator tubes observed from the top. The untreated liquor evaporator had a brown deposit which covered the tube for at least the top two feet (limit of visual observation). The deposit was uniform but not heavy. The deactivated liquor evaporator contained patches of brown material, but not a uniform coating. Analysis of the HCl solution used to acid wash the tubes showed that the untreated liquor evaporator contained 8.7 g  $\text{CaCO}_3$ , while the deactivated liquor evaporator contained 1.5 g  $\text{CaCO}_3$ . Based on these analyses,  $\text{CaCO}_3$  scaling was reduced by about 85% with thermal deactivation.

The steam injector was inspected after Run 2. Scale had accumulated on the surfaces which contact hot liquor, including the liquor flow tubing downstream of the injector. The amount of scale formed was not determined. It was not enough to interfere with liquor or steam flow. However, this may be a problem with a larger scale system running over longer time periods. The steam injector, and piping immediately following, may have to be designed for ease in cleaning.

RESULTS OF INDIRECT HEATING RUN

The heat transfer data for Run 3, in which indirect heating was used in deactivation, are shown in Fig. 6. The heat transfer coefficients did not change substantially with time during this run in either evaporator. This result was surprising, considering the rapid scaling rates during Runs 1 and 2. The only apparent explanation for no scaling is that the liquor solids (25-30% during Run 3; 32-38% during Runs 1 and 2), were not high enough during the run to scale at the temperatures encountered in the evaporators. Table III shows the calcium analyses and solids data for Run 3.

TABLE III

LIQUOR SOLIDS AND CALCIUM ANALYSES FOR  
 INDIRECT HEATING (RUN 3)

Time, hr	Solids, %		Calcium Analysis, % of BLS			
	Inlet	Outlet	Inlet		Outlet	
			Total	Soluble	Total	Soluble
3	--	--	0.057	0.028	0.054	0.023
7	25.4	25.3	0.066	0.047	0.001	0.004
11	--	--	0.042	0.021	0.042	0.024
19	29.9	30.6	0.045	0.033	0.043	0.028
31	--	--	0.032	0.021	0.029	0.023

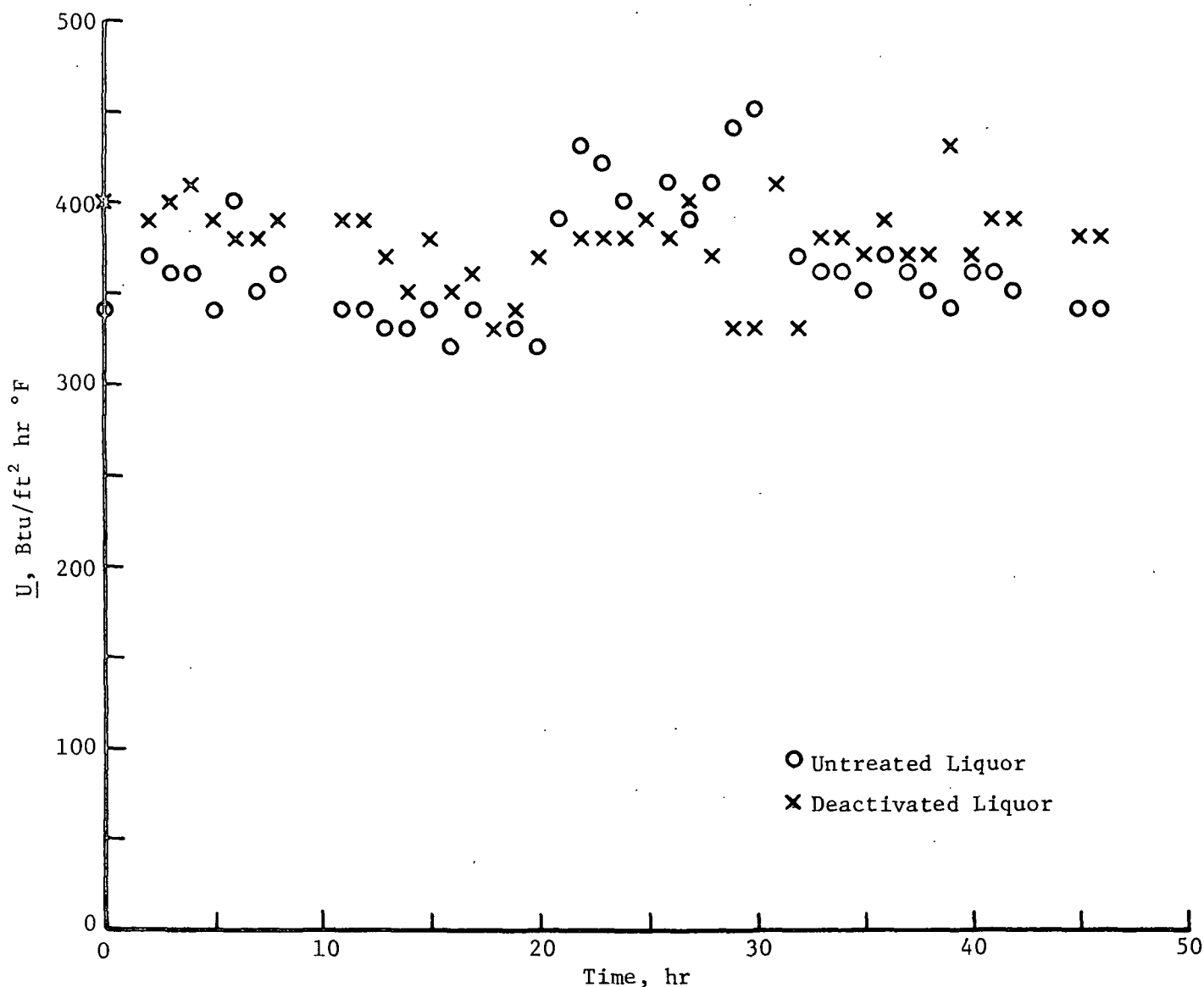


Figure 6. Heat Transfer Coefficient Versus Time for Indirect Heating Deactivation (Run 3)

The performance of the liquor heaters in the deactivation system did deteriorate, however. The recycle liquor heater was used initially to supply the entire deactivation heat load. Its performance deteriorated rapidly. When it could no longer transfer heat at the rate required for deactivation, the in-line heater was put into service to supply part of the load. The recycle liquor heater was taken out of service after 26 hours. From that point, the in-line heater supplied the entire heat load for the rest of the run.

Figure 7 shows the heat transfer coefficients for the recycle heater and the in-line heater during Run 3. According to these data, the recycle heater lost capacity so rapidly that it was totally ineffective after 25 hours of operation. When it was taken out of service, the heat exchanger tubes were inspected and were found to be plugged with fiber. The drop in heat transfer coefficient of the recycle heater resulted mainly from a drop in liquor flow rate through the heat exchanger (5.6 gpm at start of run; 0.4 gpm after 25 hr) rather than scaling resistance per se.

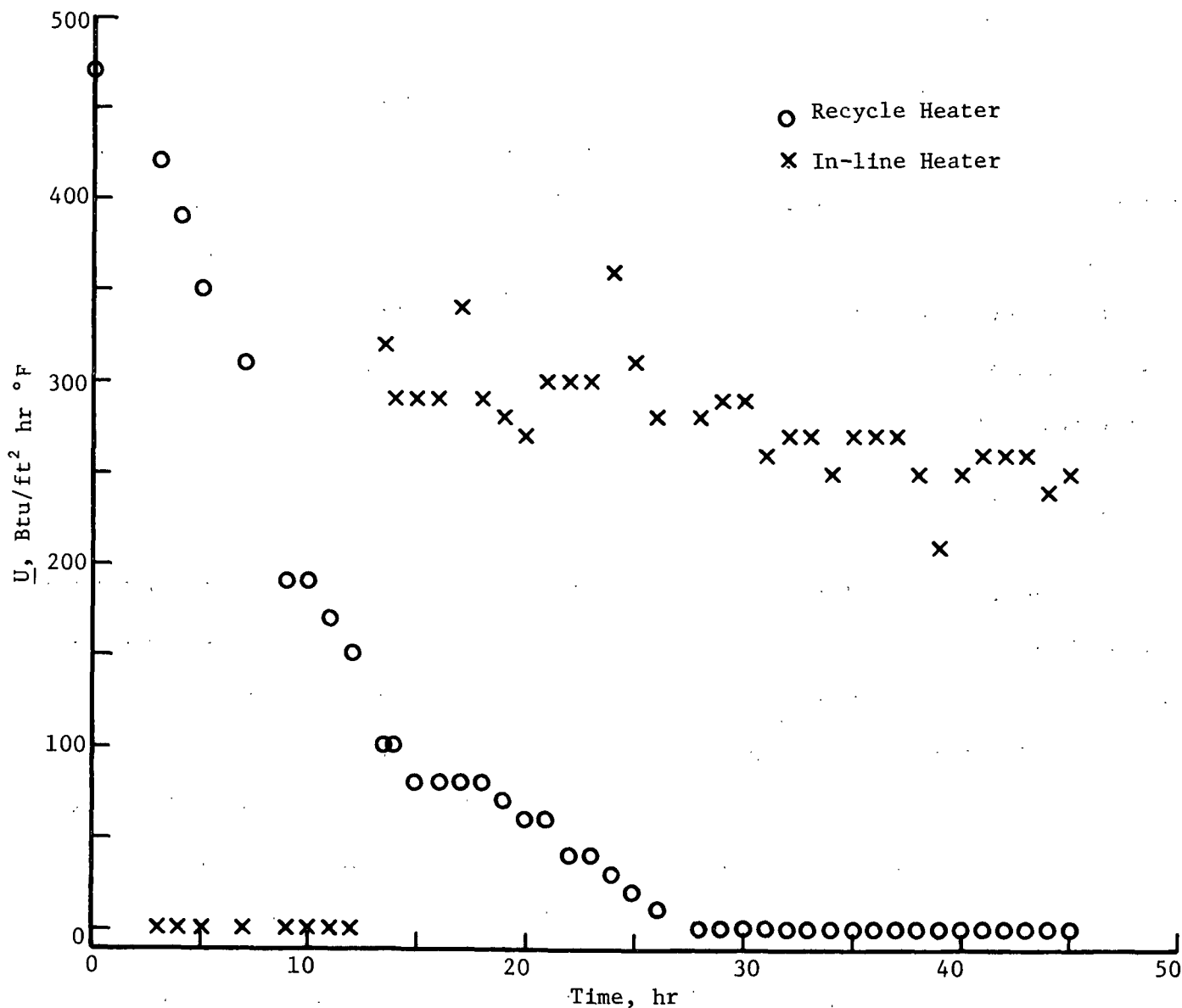


Figure 7. Heat Transfer Coefficients Versus Time for Deactivation System Indirect Liquor Heaters (Run 3)

The fiber plugging problem was related more to the heat exchanger design than to liquor handling and scaling problems. The heat exchanger for the pilot deactivation system was designed for high velocity flow to give high heat transfer coefficients with small volumes of liquor. This led to the use of 1/4-inch OD tubes. The small tubes and possibly an unusually high liquor fiber content during the run combined to create the fiber plugging problem. The fiber plugging problem is not expected to be a problem in production-scale systems with heater tubes on the order of one inch in diameter, or in which a fiber filter would be installed.

The in-line liquor heater was used to supplement the recycle liquor heater when the recycle liquor heater could no longer handle the deactivation heating requirement. It was used to supply the entire deactivation heat load once the recycle liquor heater was taken out of service. The heat transfer data show that the in-line heater coefficient dropped about 17% during its operation. However, the pressure drop across the heat exchanger increased rapidly near the end of the run, and eventually forced the shutdown.

#### ADDITIONAL RUNS AT IPC

The pilot trials at Charleston showed that heating did deactivate a calcium-scaling liquor, but they were inconclusive on the effectiveness of indirect heaters as a method of liquor heating. To resolve this, the pilot deactivation unit was tested at IPC using a hardwood kraft liquor which had been filtered to remove fiber. Water-soluble calcium salts were added to the liquor to promote scaling. The object of these runs was to compare scaling rates in the liquor preheater versus the recycle heat exchanger when each supplied the total heat load for deactivation.

Figure 8 shows the change in heat transfer coefficients for the two heaters as a function of time. The data are for one run made with the in-line heater and

three with the recycle heater. The liquor preheater scaled rapidly, with the heat transfer coefficient dropping to 45% of its initial value in 4 hours. The recycle heater coefficient dropped slowly in one run, and held constant in the other two. The average loss was less than 3% in 4 hours. Based on these results, the use of the recycle heating concept reduced scaling in the deactivation liquor heater by more than 90%.

#### SUMMARY

Thermal deactivation is an effective method of preventing  $\text{CaCO}_3$  scaling in black liquor evaporators. Scale accumulation was reduced by about 90% when a deactivated liquor was evaporated as compared with an undeactivated liquor.

Liquor heating for deactivation can be accomplished either by direct steam injection or with an indirect heater. The potential scaling problem with indirect liquor heaters can be handled by heating recycled deactivated liquor and mixing it with the untreated liquor. This method reduces the rate of scaling in the indirect heater by about 90%. The frequency of liquor heater cleaning when the recycled liquor heating concept is used should be determined with longer term mill trials, however.

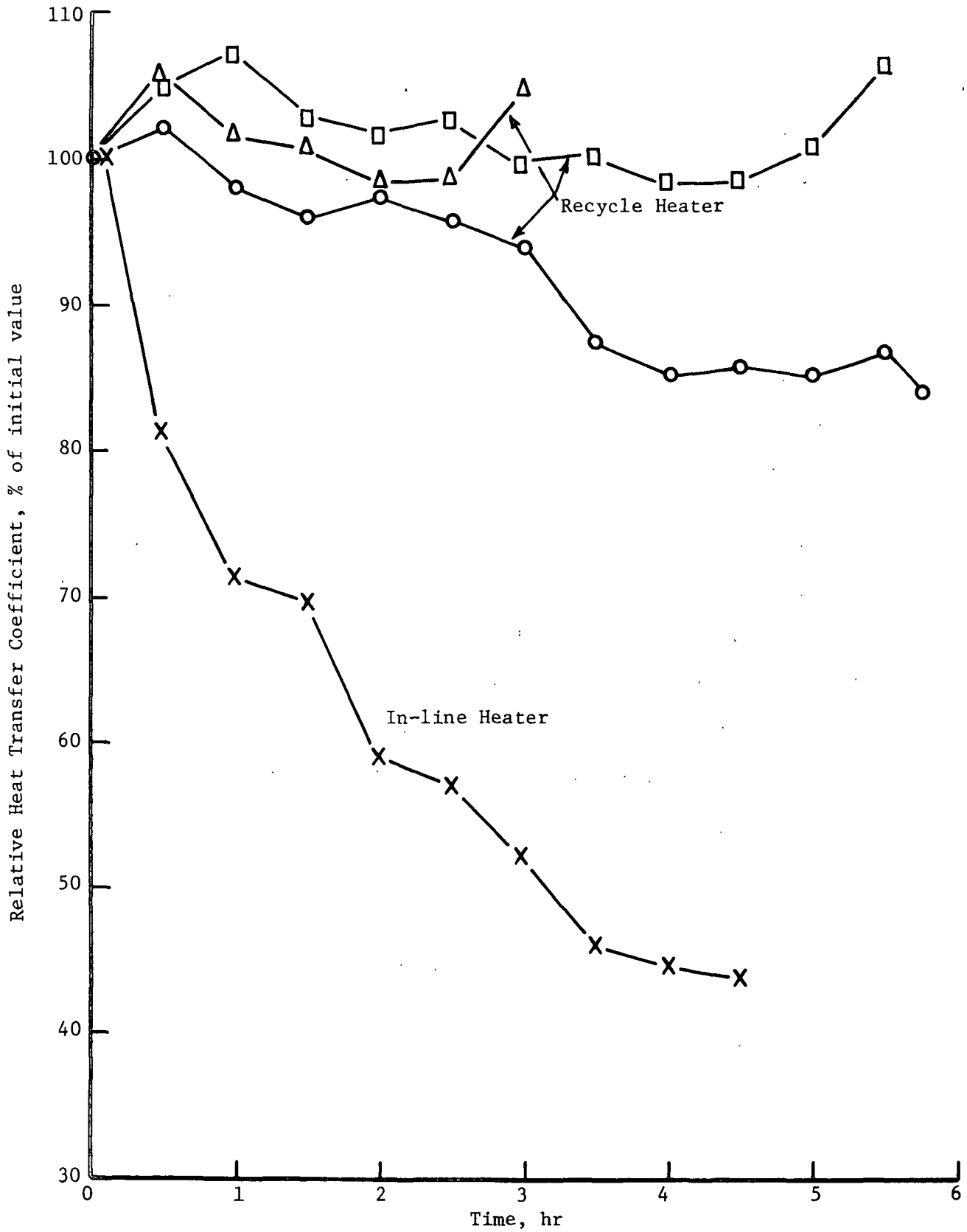


Figure 8. Relative Heat Transfer Coefficient Versus Time for Thermal Deactivation Liquor Heaters

## PROCESS ECONOMIC EVALUATION

### SYSTEM CONSIDERATIONS

Both the location of the deactivation system within the evaporator train and the liquor flow sequence in the evaporators affect costs. In general, the thermal deactivation system should treat the minimum amount of liquor required to prevent scaling. It should be located so as to treat liquor entering only those effects in which scaling is severe. Therefore, its location in a particular mill's evaporator train depends on the location and severity of evaporator scaling at that mill.

Liquor flow sequence should be chosen to maximize energy economy. Since heat is supplied to the deactivation system as process steam, the hot deactivated liquor should be returned to the first effect for the maximum energy economy. Flashed vapor can be introduced into the second effect. Figure 9 shows the recommended liquor flow sequence for a 6-effect evaporator with thermal deactivation before the first two effects.

### CAPITAL COSTS

Estimates of installed costs for the components of a thermal deactivation system are shown in Table IV. These data are for an evaporator train in a 500 BDT/day pulp mill, which concentrates liquor from 13.9 to 52.4% solids. The system design was based on a 300°F deactivation temperature, 15-min mean residence time, and 5/1 recycle ratio for indirect liquor heating. The heat exchanger was designed for a 10 psig pressure drop. Cost estimates for the process equipment were obtained from manufacturers in mid-1978, and modular factors (2) were applied to calculate the installed equipment costs in Table IV. Exponential factors for adjusting process equipment size are also given in Table IV.

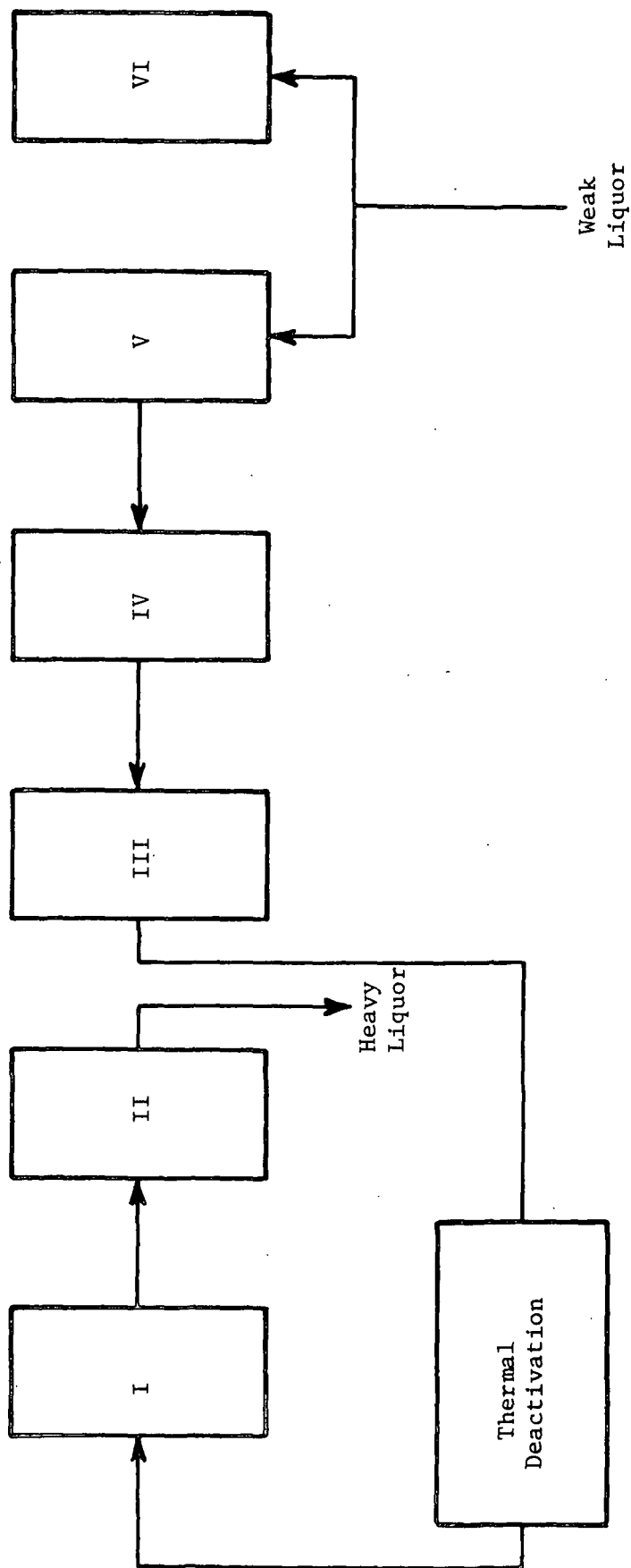


Figure 9. Liquor Flow Sequence for 6-Effect Evaporator with Thermal Deactivation Before the First Two Effects

TABLE IV  
 CAPITAL COST FOR THERMAL DEACTIVATION EQUIPMENT<sup>a</sup>

	Indirect Heating		Direct Steam Injection		Equipment Size Exponent <sup>b</sup>
	$\frac{6}{5} \setminus 4-3-2-D-1$	$\frac{6}{5} \setminus 4-3-D-1-2$	$\frac{6}{5} \setminus 4-3-2-D-1$	$\frac{6}{5} \setminus 4-3-D-1-2$	
Heat exchanger	\$165,000	\$257,000	\$ --	\$ --	0.67
Retention vessel	137,000	182,000	137,000	182,000	0.82
Flash tank	48,000	46,000	48,000	46,000	0.82
Primary pump	9,000	10,000	9,000	10,000	0.50
Recycle pump	31,000	35,000	--	--	0.50
Total installed plant cost	390,000	530,000	194,000	238,000	

<sup>a</sup>Equivalent to 500 BDT/day capacity. April, 1978 data for installed and instrumented equipment.

<sup>b</sup>
$$\frac{\text{Cost}(1)}{\text{Cost}(2)} = \left( \frac{\text{Size}(1)}{\text{Size}(2)} \right)^n$$
, where  $n$  = equipment size exponent.

The cost data in Table IV show that the capital cost for direct steam injection is less than half that for indirect heating. The ratio of capital costs for direct versus indirect heating becomes lower the greater the number of effects after deactivation. The liquor heater and retention vessel are the major cost items for thermal deactivation.

The temperature at which the thermal deactivation system is operated affects the capital equipment cost. Since thermal deactivation rate increases rapidly with temperature, deactivation at a temperature higher than 300°F will reduce the size and cost of the retention vessel. However, more heat transfer area will be required to allow for a higher heat transfer rate and lower thermal driving force at the same steam pressure for indirect heating.

Figures 10 and 11 show the total deactivation system capital costs as a function of deactivation temperature. For direct steam injection, capital cost drops rapidly as deactivation temperature increases, reflecting the large fraction of total cost represented by the retention vessel. For indirect heating, the liquor heater cost increases at about the same rate that the retention vessel cost drops. The result is a flat cost versus temperature curve with a shallow minimum at 315-318°F. The minimum does not correspond to the optimum deactivation temperature. Operating costs have not been considered, and they will push the optimum temperature lower.

#### OPERATING COSTS

When thermal deactivation is incorporated into an evaporator sequence, the energy economy of the system is decreased. The actual change in evaporation energy required depends on where the thermal deactivation process is included in the evaporator train, and on whether indirect heating or direct steam injection

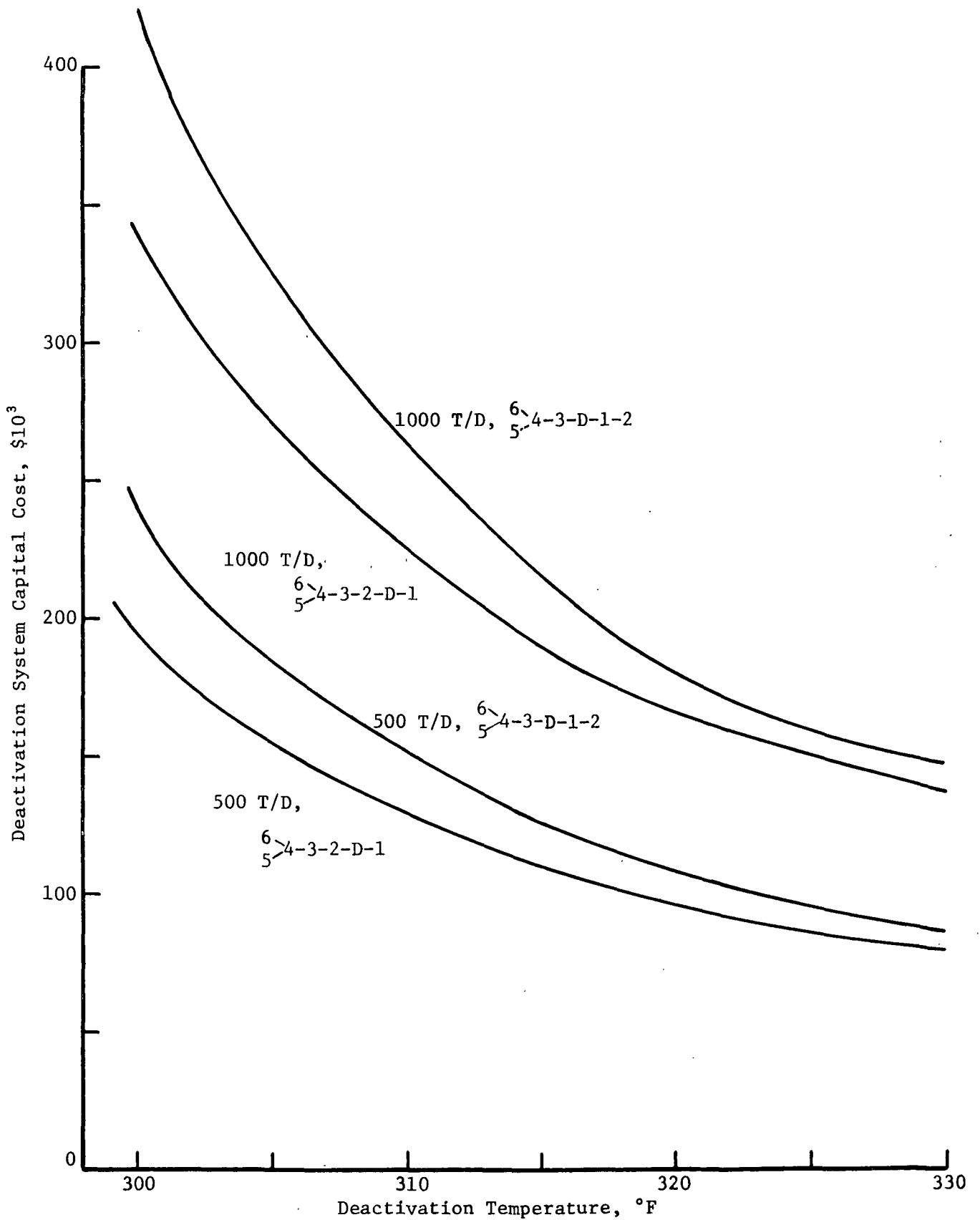


Figure 10. Deactivation System Capital Cost Versus Deactivation Temperature for Direct Steam Injection

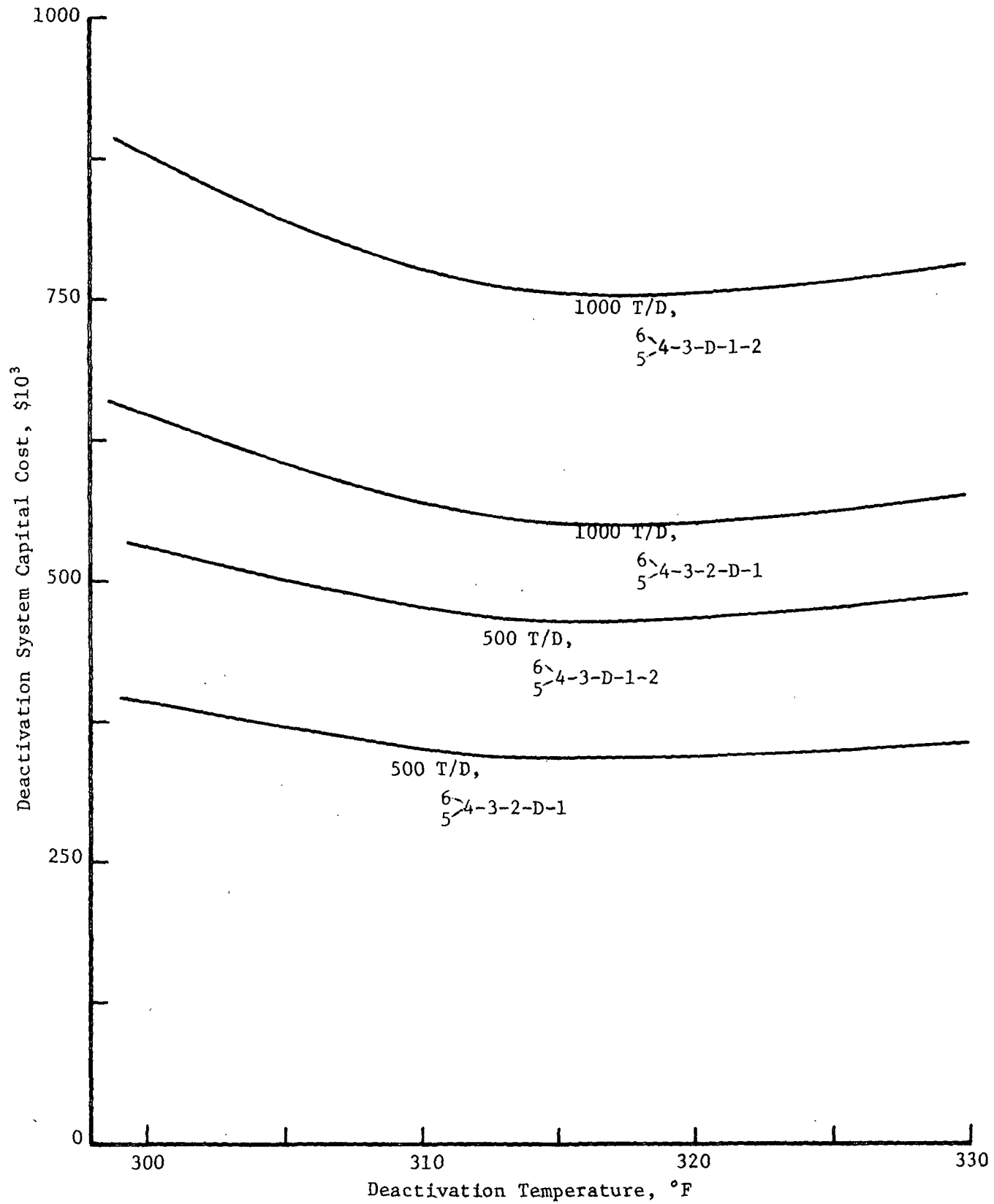


Figure 11. Deactivation System Capital Cost Versus Deactivation Temperature for Indirect Liquor Heating

is used. Thermal deactivation also requires higher pressure steam than does evaporation. The combined increase in total energy demand and shift to higher pressure steam changes the fossil fuel requirements and the electrical energy generation rate of the mill's steam and power system. It is the changes in by-product power and fossil fuel demand that are the energy costs of deactivation, not the increased process steam demand to the evaporators and deactivation process per se. The increase in boiler feedwater makeup requirement is a significant operating cost with direct steam injection. Finally, there is an energy cost to pump liquor through the deactivation system at approximately 100 psig, and to pump the recycled liquor in the case of indirect heating.

#### Process Steam Demand

The changes in process steam demand for evaporation and deactivation must be determined before the changes in electrical power generation and fossil fuel demand can be calculated. To determine the process steam requirements for deactivation, we simulated a 6-effect evaporator system with deactivation included, using a GEMCS computer model. The model is described in Appendix I. Liquor was treated before the higher solids effects only, since  $\text{CaCO}_3$  scaling generally is a problem only at higher temperatures. Liquor flow was countercurrent with steam and vapor flow before deactivation and cocurrent afterward. These procedures reduce the energy increment required when deactivation is used.

Figure 12 shows net evaporation energy required as a function of deactivation temperature for several alternative deactivation system locations, as calculated with the model. Indirect heating requires considerably less additional thermal energy for a given deactivation location than does direct steam injection. The reason is that, with indirect heating, the deactivation heater acts as additional first effect heat transfer area. Heat transferred in the liquor heater therefore receives the

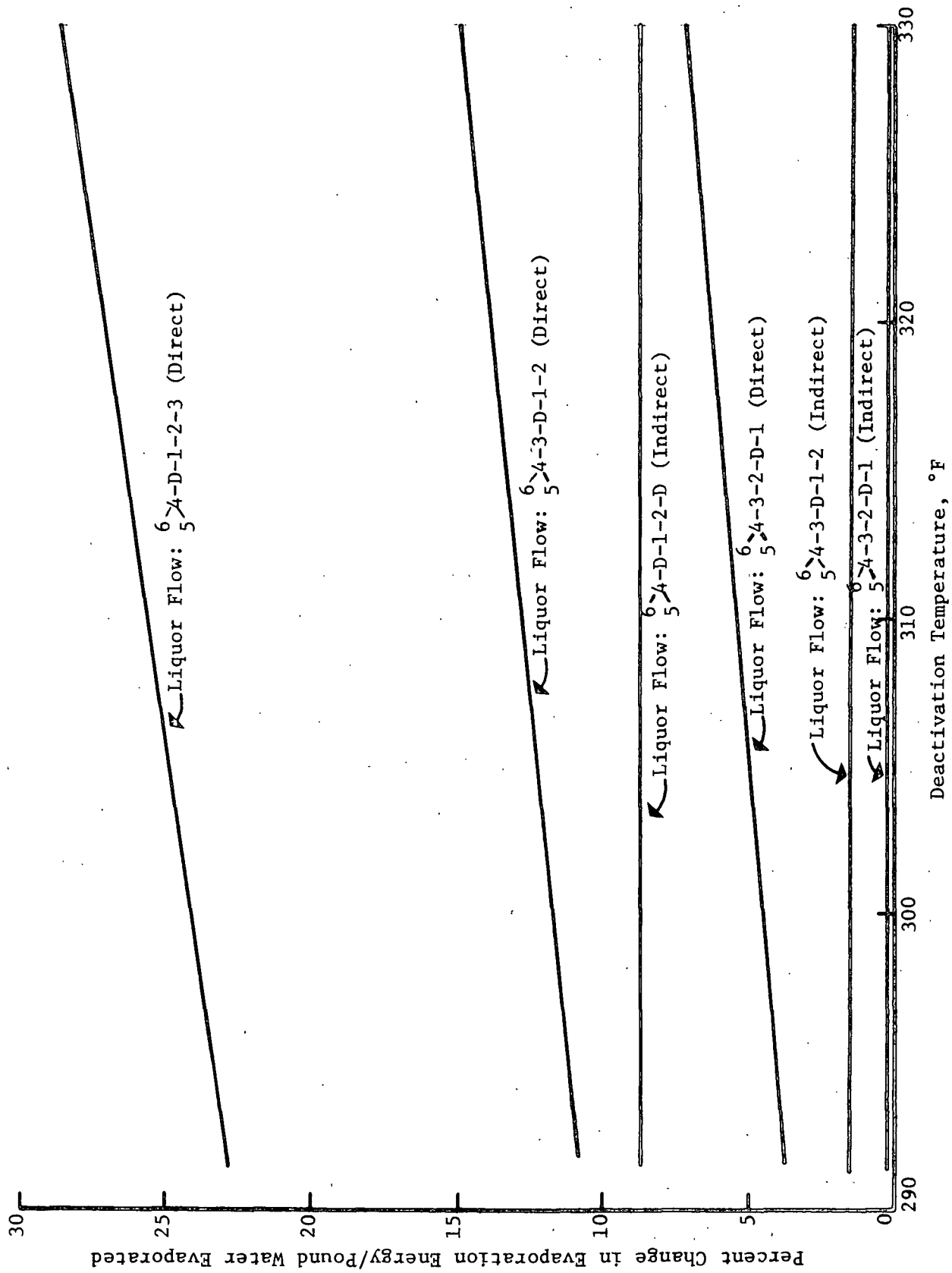


Figure 12. Percent Change in Evaporation Capacity Versus Thermal Deactivation Temperature

same steam economy as that transferred in the first effect. The increased total steam requirements with indirect heating, relative to that required for an evaporation train without deactivation, result from feeding colder liquor to the deactivation process as the deactivation system is moved nearer to the weak liquor effects. The evaporation energy is independent of deactivation temperature for indirect heating because the temperature of the liquor entering the deactivation system is independent of deactivation temperature, and because the overall system steam economy is not affected by the split of the total heat load between the first effect and the deactivation liquor heater.

With direct steam injection, the deactivation steam dilutes the liquor, but is flashed to the second effect shell before the liquor enters the first effect. Neglecting the sensible heat load, this is the energy balance equivalent of applying the deactivation steam directly to the second effect shell side. Therefore, steam used for direct injection receives the steam economy of five effects, rather than six effects. This accounts for the increase in evaporation energy with deactivation temperature for direct steam injection.

Figure 13 shows how the heating load is distributed between the deactivation liquor heater and the first effect.

#### By-product Power and Fossil Fuel Changes

Steam and power balances were made for mills with and without by-product power, for two deactivation system locations within the evaporation train. These were used to determine the changes in fossil fuel requirements, by-product power generation, and boiler feedwater makeup associated with thermal deactivation. The assumption on which the power system calculations were based is that the power system is operated for maximum electrical energy generation without condensing

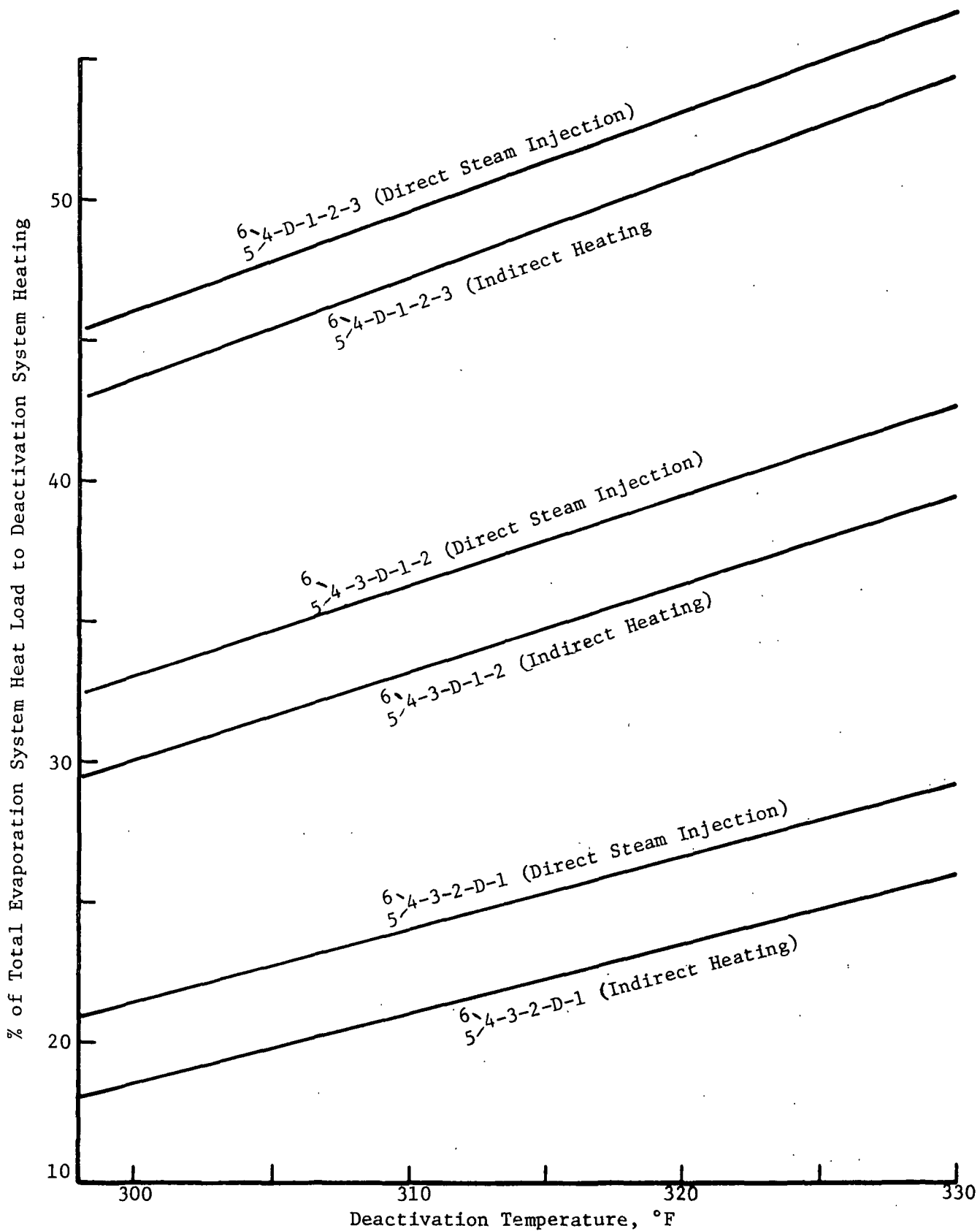


Figure 13. Split of Energy Consumption to Evaporators and Deactivation System Versus Deactivation Temperature

exhaust steam other than for process use. The basis for the calculations is included in Appendix II.

Table V shows the change in fossil fuel requirement and electrical energy generating capacity as a function of deactivation temperature. These results show that for mills with by-product power generation, indirect heating reduces the electrical generating capacity and usually reduces the fossil fuel requirement slightly. Direct steam injection greatly increases the fossil fuel required, but also increases the electrical energy generation capacity. Mills without by-product power generation are penalized slightly more than mills with by-product power generation when both incremental fossil fuel requirements and electrical power generation are considered. The relative positions depend strongly on fuel and power costs. Increasing the deactivation temperature strongly affects the energy cost with direct steam injection, but only affects the indirect heating energy cost slightly.

Table VI shows the change in boiler feedwater makeup requirement with thermal deactivation. The makeup requirements are a significant cost factor only for direct steam injection. Increasing the deactivation temperature by 30°F increases the makeup requirements by 60-75%, depending on where the thermal deactivation process is located in the evaporator system.

The data in Tables V and VI can be used to calculate the economic impact of thermal deactivation on the steam and power system, given the cost of electrical energy, fossil fuel, and boiler feedwater treatment. For reasonable costs, there is an economic penalty with either indirect heating or direct steam injection. However, the penalty is much greater with direct steam injection than with indirect heating unless the ratio of electrical power value to fossil fuel cost is extremely high. It also increases rapidly with increasing deactivation temperature for direct steam injection.

TABLE V

INCREMENTAL FOSSIL FUEL REQUIREMENTS AND ELECTRICAL POWER PRODUCTION AS A FUNCTION OF DEACTIVATION TEMPERATURE<sup>a</sup>

Deactivation Temp., °F	Indirect Heating			Direct Steam Injection <sup>b</sup>			
	6/5 4-3-2-D-1			6/5 4-3-2-D-1			
	Fossil Fuel	Elec. Power	Elec. Power	Fossil Fuel	Elec. Power	Fossil Fuel	
300	-25,000	-6.1	+14,000	+113,000	+4.4	+345,000	+12.9
310	-26,000	-6.8	-8,000	+382,000	+7.0	+618,000	+11.6
320	-30,000	-7.6	-16,000	+401,000	+7.0	+764,000	+15.8
330	-34,000	-8.4	-19,000	+454,000	+8.2	+916,000	+20.4
300	+2,500	--	+54,000	+81,000	--	+266,000	--

Mill with By-product Power Generation

Mill Without By-product Power Generation

<sup>a</sup> Incremental fossil fuel requirements as Btu/ADT pulp; incremental electrical power production as kw-hr/ADT pulp. Data are relative to a mill without thermal deactivation.  
<sup>b</sup> 65 psig steam used at 300°F; 150 psig steam used at 310°F and higher deactivation temperatures.

TABLE VI

INCREMENTAL BOILER FEEDWATER MAKE-UP REQUIREMENTS  
 WITH THERMAL DEACTIVATION

Deactivation Temp., °F	Heating Method	Incremental Boiler Feedwater Make-up, gal/ADT	
		<sup>6</sup> / <sub>5</sub> 4-3-2-D-1	<sup>6</sup> / <sub>5</sub> 4-3-D-1-2
300	Indirect	0.2	0.6
	Direct	44.0	75.8
310	Indirect	0.1	0.5
	Direct	62.4	101.3
320	Indirect	0.2	0.4
	Direct	69.8	111.4
330	Indirect	0.4	0.4
	Direct	77.2	121.5

Although the data in Tables V and VI show how steam and power system costs vary with deactivation system variables, they apply exactly only for the mill conditions defined in Appendix II. The effect of deactivation on steam and power system balances depends strongly on specific mill conditions. When actual cost data are needed, the results in Tables V and VI should be calculated for the conditions under which the mill in question operates. The calculation procedure that should be followed is shown in Appendix II.

Pumping Energy

The pumping energy requirements for thermal deactivation were calculated assuming an overall pump and motor efficiency of 0.72. Reference (3) was used to predict the liquor viscosity dependence on temperature and solids. Table VII gives the pumping energy requirements for different thermal deactivation system locations

within the evaporator train, for both heating methods. Pumping energy costs are small, representing less than 5% of the total operating cost for thermal deactivation.

TABLE VII  
PUMPING ENERGY REQUIREMENTS FOR  
THERMAL DEACTIVATION

Deactivation Temp., °F	Heating Method	Pumping Energy Requirements, kw-hr/ADT	
		$\frac{6}{5}$ 4-3-2-D-1	$\frac{6}{5}$ 4-3-D-1-2
300	Indirect	1.4	2.0
	Direct	0.9	1.3
310	Indirect	1.5	2.1
	Direct	1.0	1.4
320	Indirect	1.7	2.3
	Direct	1.1	1.6
330	Indirect	1.8	2.5
	Direct	1.3	1.8

OPERATING CREDITS

The capital and operating costs for thermal deactivation can be justified by a return on investment based on two factors: savings in scale removal costs and increase in evaporation capacity.

Scale Removal Savings

In calculating scale removal cost savings, credit should be given for reduced costs of evaporator cleaning by water boil out, acid washing, hydroblasting, or other scale removal treatments. For example, water boil out costs include the costs of water used to boil out, of steam consumed both during boil out and during evaporation of water used to boil out which is returned to the weak liquor system,

and of operating personnel during the boil out. Credit should also be given for longer evaporator tube life due to less frequent acid washing, when applicable.

For reasonable cleaning costs, we estimate that thermal deactivation cannot be justified by scale removal savings alone except in mills with extremely severe scaling problems.

#### Incremental Evaporation Capacity

Thermal deactivation affects evaporation capacity in two ways. It reduces or eliminates  $\text{CaCO}_3$  scaling, which increases the average heat transfer coefficient and therefore increases capacity. Also, adding additional heat transfer surface for indirect heating increases the net evaporation capacity. Direct steam injection has a slightly negative effect on net evaporation capacity, resulting from a shift in the heat transfer coefficient and thermal driving force product ( $\underline{U \cdot \Delta T}$ ) distribution.

The increase in evaporation capacity due to reduction of scaling depends upon the rate of scaling without deactivation, its reduction when deactivation is used, and the cleaning frequency in both cases. Harrison's procedure (4) was used to calculate the optimum cleaning frequency as a function of scaling rate, assuming:

1. linear decrease in capacity with time
2. 4-hour downtime for water boil out
3. water equivalent to 1 hour's evaporation capacity added to the weak liquor system during a water boil out.

From the scaling rate, optimum cleaning frequency, and down time for cleaning, we calculated the incremental evaporation capacity as a function of percent loss in evaporation capacity per month without deactivation. Percent loss in evaporation capacity per month is defined as the product of the percent loss in capacity

between cleanings and the cleaning frequency. The results are shown in Fig. 14 for cases where scaling is reduced by 100% and 85%. Figure 14 shows that the greater the rate of loss of evaporation capacity, the greater the gain in evaporation capacity by eliminating scaling. It also shows that the incremental capacity gained by eliminating scaling in a properly maintained evaporator train is not large, being less than 15% for most mills.

The incremental capacity changes associated with additional heat transfer surface (indirect heating) or changes in the UAT product distribution (direct steam injection) were calculated using the GEMCS simulation. The results are shown in Fig. 15 for several deactivation system locations and as a function of deactivation temperature.

The net increase in evaporator capacity is found by adding the incremental capacity changes from Fig. 14 and 15. For example, if liquor, which scales at a capacity loss rate of 125%/month, is deactivated using indirect heating before the first two effects, the incremental evaporator capacity with deactivation effective for 85% reduction in scaling is  $10.6 + 4.5 = 15.1\%$ . For direct steam injection under the same conditions, the capacity increase is  $10.6 - 1.2 = 9.4\%$ .

The value of incremental evaporator capacity depends on many factors. These include whether the evaporators are production bottlenecks, and the value and marketability of pulp. Since these factors vary tremendously from mill to mill, the value of incremental evaporator capacity must be determined for each specific mill.

#### PROCESS ECONOMICS SUMMARY

Factors which affect the cost of thermal deactivation can be divided into two categories: process design cost parameters and external cost factors. The process design cost factors are those over which the design engineer has some control

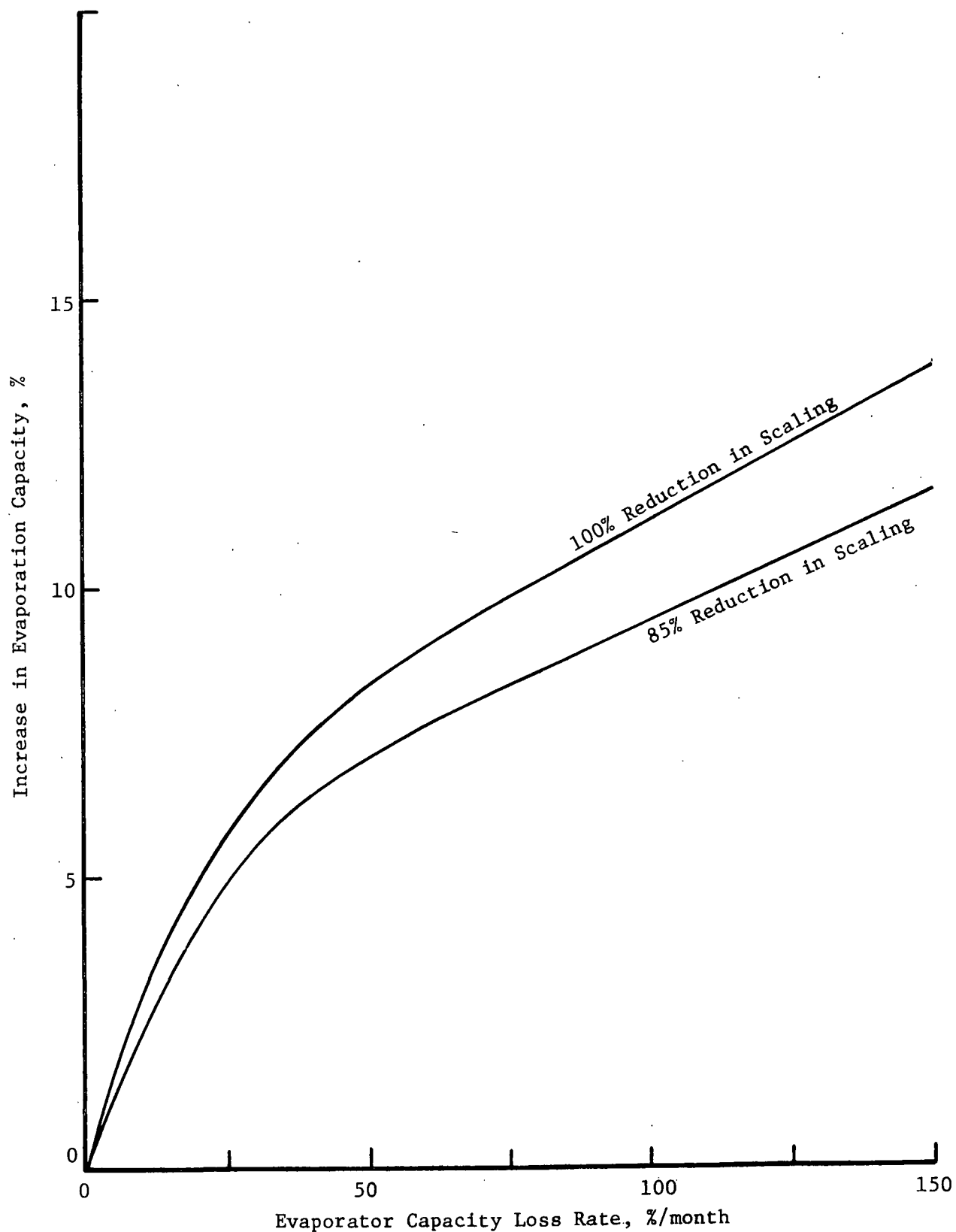


Figure 14. Incremental Evaporator Capacity Due to Scaling Rate Reduction Versus Scaling Rate Before Deactivation

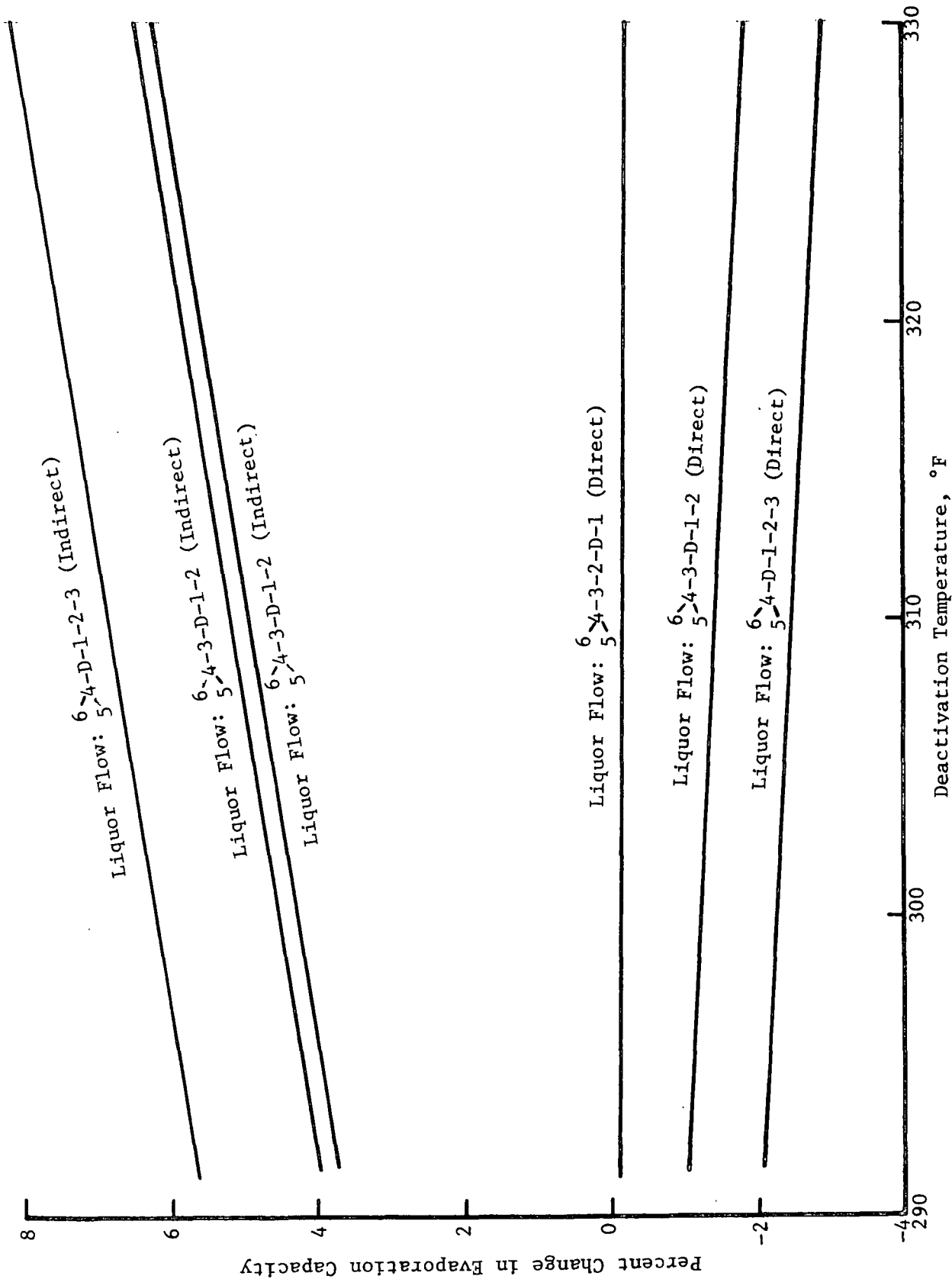


Figure 15. Percent Change in Evaporation Capacity Versus Thermal Deactivation Temperature

in optimizing the process design. For thermal deactivation, they include choice of heating method, location in the deactivation train, and deactivation temperature. External cost factors are those over which the design engineer has no control, but which do affect the cost of deactivation. They include the steam and power system operating conditions, whether the mill generates by-product power, cost of fossil fuel, electric power, and boiler feedwater treatment, savings in scale removal costs, and the value of incremental evaporation capacity. The effects of variables from the two categories cannot always be separated. How each of these factors affects the cost of thermal deactivation is considered in the following sections.

#### Process Design Factors

The choice of liquor heating method affects both capital and operating costs. Although capital costs for deactivation with indirect liquor heating are about double those with direct steam injection, the higher operating costs for direct steam injection offset the capital advantage. At current fossil fuel, electric power, and boiler feedwater treatment costs, thermal deactivation at 300°F before one effect costs \$0.50-0.80/BDT pulp, and before two effects \$1.00-1.50/BDT pulp (Appendix III), regardless of heating method. These figures are approximate ranges of cost, including amortization of capital, and without credit for scale removal savings or value of incremental evaporation capacity.

The choice of heating method depends strongly on the costs of fossil fuel, electric power, and boiler feedwater treatment. The indirect liquor heating method is capital intensive and less sensitive to operating costs. Doubling energy and water treatment costs increase the total deactivation cost by 20-30% for indirect heating. Direct steam injection is more energy intensive, so that doubling the energy and water treatment costs increases the total deactivation cost by 50-80%.

The location of the deactivation system in the evaporator liquor flow sequence strongly affects deactivation cost. The total cost of deactivation increases nearly in proportion to the volume of liquor treated, so that liquor should be deactivated before only those effects with severe scaling. Since total cost increases rapidly with liquor volume treated, a practical limit to the number of effects after deactivation is probably two.

Increasing the deactivation temperature has little effect on the economics for indirect heating, but does affect direct steam injection costs. There is no net capital cost advantage to higher deactivation temperature for indirect heating, since the decreased retention vessel cost is completely offset by the increased heat exchanger cost. Operating costs for indirect heating are independent of deactivation temperature also, if the same pressure turbine extraction steam is used for liquor heating at either temperature.

With direct steam injection, capital costs drop rapidly as deactivation temperature increases, but fuel, power and water treatment costs all rise more rapidly. The result is that deactivation at 300°F is usually more cost effective than deactivation at a higher temperature. The economics are particularly sensitive to turbine extraction steam pressure. If the deactivation temperature can be increased without using higher pressure extraction steam, then the total cost increases 10-25% per 10°F increase in deactivation temperature. If a higher pressure extraction steam must be used, the rate of cost increase is substantially higher, and depends on the available extraction steam pressures.

#### External Cost Factors

The most important factor in justifying thermal deactivation is the value of incremental evaporation capacity. The value of incremental evaporation

capacity is also the most difficult factor to evaluate. When pulp production is severely limited by evaporation capacity, deactivation or almost any other means of increasing capacity can be justified economically. In this situation, the cost of deactivation must be compared with that of an additional evaporator body, or with other alternatives. The comparison must be made for the specific mill situation. Appendix III shows the results of return on investment calculations for incremental pulp production valued at \$10/ton and \$100/ton.

Scale removal cost savings will also affect the comparison between thermal deactivation and other scale control means. In general, the thermal deactivation process becomes more competitive with adding an additional evaporator body as scaling becomes more severe.

Whether a mill generates by-product power or not affects the cost of thermal deactivation and the choice of liquor heating method. Indirect heating generally costs less for a mill without by-product power than for a mill with by-product power. The opposite is true for direct steam injection. Whether deactivation increases or decreases by-product power generation, and the relative costs of fossil fuel and electrical power are the important variables in this case.

#### ALTERNATIVES TO THERMAL DEACTIVATION

In terms of evaporator operation, thermal deactivation accomplishes two results. It provides additional evaporation capacity and it reduces downtime and expense for scale removal. When considering a thermal deactivation process to accomplish either or both of these, one must also consider alternative methods and make an economic comparison of the choices.

##### Additional Evaporator Body

One method of obtaining extra evaporator capacity is to add another evaporator body. For maximum capacity increase, an additional body should be

put in parallel with the lowest coefficient effect in the evaporator train. Doing so shifts the  $\Delta T$  distribution so that the higher coefficient effects have more  $\Delta T$ . Figure 16 shows the capacity increase that can be achieved when this is done. The data in Fig. 16 were obtained using the GEMCS simulation discussed in Appendix II, and apply strictly only to that set of heat transfer coefficients and operating conditions. Figure 16 does show that the capacity increase with an extra body is greater than proportional to the increase in heat transfer area for up to 12% increase in total evaporator area. This results from the shift in  $\Delta T$  to higher coefficient effects.

Adding an evaporator body does provide additional capacity, but does not reduce or eliminate cleaning costs. A mid-1978 estimate of the purchased cost of an evaporator body is \$130,000 for 10,000 ft<sup>2</sup> of heat transfer area. The installed cost is \$429,000. Evaporator body cost varies with the 0.7 power of heat transfer area. For economic comparisons, Fig. 16 can be used to estimate the size of an additional evaporator body required to give the same capacity increase as obtained with thermal deactivation. The ROI calculations in Appendix III show that thermal deactivation can have a slight advantage over an additional evaporator body for high scaling rates.

#### Lower First Effect Temperature

It should be possible to reduce or eliminate scaling by operating the first effect at a lower shell side temperature. For example, our previous results (1) showed that scaling is very temperature sensitive, and that below a steam temperature of 265°F, scaling may not be significant. The capacity gained by eliminating scaling offsets at least part of the capacity lost by lowering the thermal driving force. As a rough estimate of how lowering the first effect steam temperature affects capacity, consider an evaporator train which is limited by scaling to an average

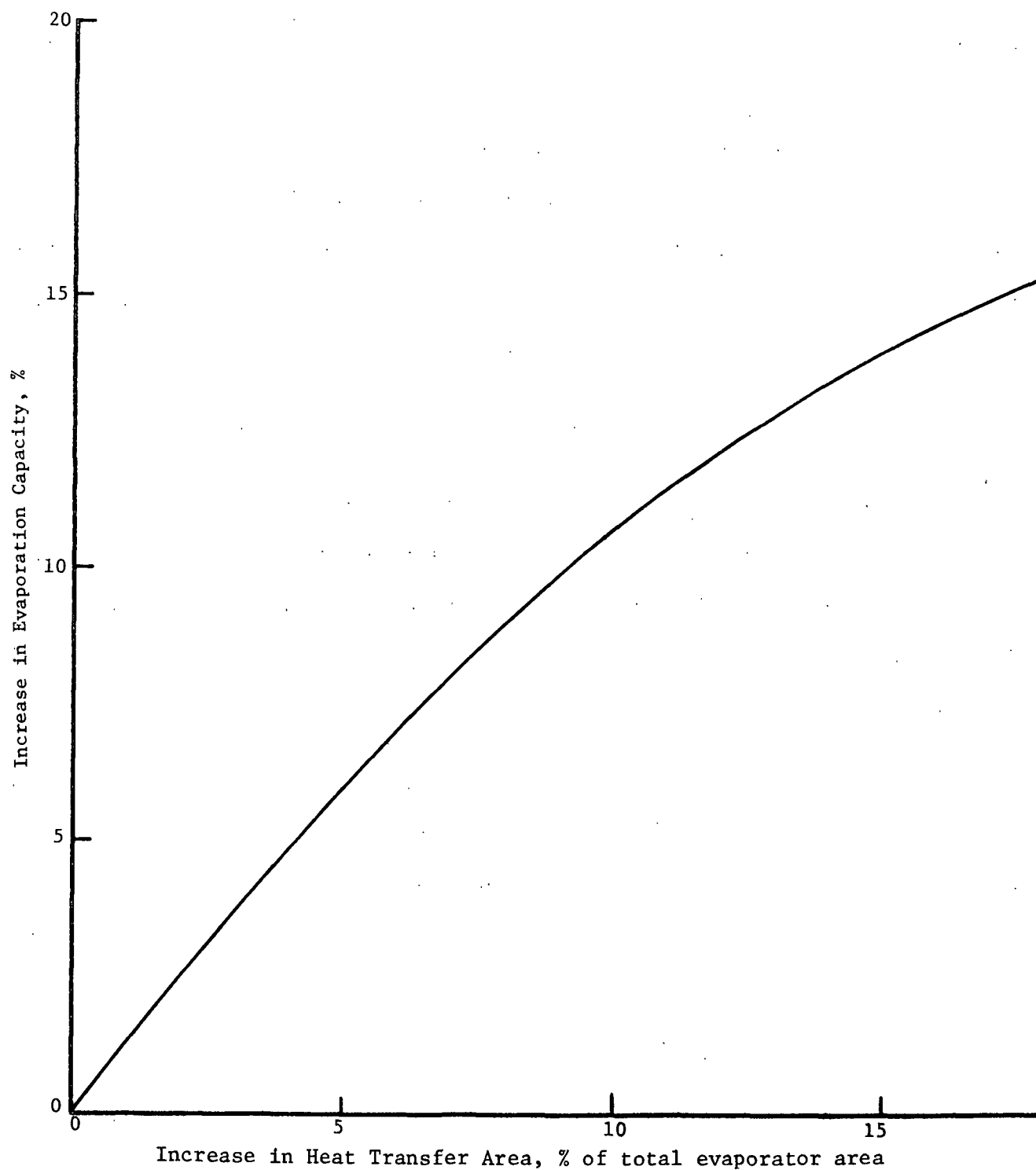


Figure 16. Incremental Evaporation Capacity Versus Incremental Heat Transfer Area (All Areas Added to First Effect) for 6-Effect Evaporator

capacity of 90% of its unscaled capacity. If the system scales rapidly at 285°F steam temperature but not at 265°F, and if the effective thermal driving force for the system is 130°F, then the system would have  $110^{\circ}\text{F}/130^{\circ}\text{F} \times 100 = 85\%$  of its unscaled capacity if the first effect steam temperature were lowered by 20°F. The net results are 5% less capacity than when operating at 285°F under scaling conditions. Figure 13 can be used to estimate the additional heat transfer area required to increase capacity to the required level when operating at lower first effect steam temperature.

#### SUMMARY

The cost of thermal deactivation depends primarily on the value of increased evaporation capacity, severity of scaling, and the number of effects where scaling is significant. The liquor heating method, extent and conditions of by-product power production, and the costs of fuel, electricity, and boiler feedwater treatment also affect the cost of deactivation.

Except for severe scaling, deactivation costs cannot be justified by savings in cleaning costs alone. It is necessary to be able to take credit for increased pulp production in order to generate a favorable return. In this case, thermal deactivation must be compared with alternative methods for increasing capacity. Deactivation is most attractive for those mills where severe calcium scaling problems in the first one or two effects cause a production bottleneck. However, deactivation does not show a clear cut economic advantage over other methods of increasing evaporator capacity for the range of scaling rates normally found in black liquor evaporators.

#### ACKNOWLEDGMENTS

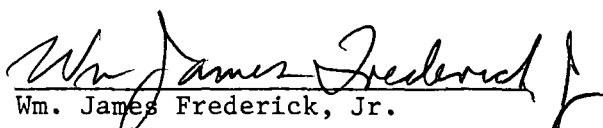
The authors would like to thank the Westvaco Corporation for their assistance and cooperation in the pilot study of thermal deactivation. We owe a special thanks to Dr. Ray E. Harrison and his group at Westvaco Corporation's Charleston Research Center for their assistance and many suggestions during the pilot work.

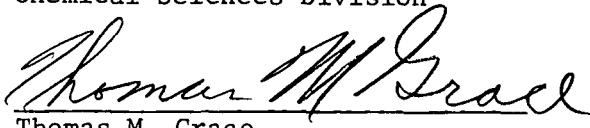
We also acknowledge the assistance of Orlin C. Kuehl in designing, constructing, and operating the pilot deactivation system, and the assistance of Bruce D. Andrews, Harold J. Grady, and Donald G. Sachs for their contributions to the project.

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## APPENDIX I

### GEMCS 6-EFFECT EVAPORATOR/DEACTIVATION MODEL

A GEMCS 6-effect evaporator model was used to calculate how thermal deactivation affects steam economy and evaporation capacity when added to an evaporator train. Figure 17 shows the flow diagram for the model with thermal deactivation before the first two effects. The model uses 6 effects with equal heat transfer area in each effect. Weak liquor feed is split equally between effects 5 and 6. The heavy liquor effect is a two-pass effect with respect to liquor flow, but with a common steam shell. The heat transfer area is equally divided between each pass.

The heat transfer coefficients for the 6 effects were assumed to be constant with respect to liquor solids. In practice this meant that the same set of heat transfer coefficients were always used, and that their order of assignment followed the liquor flow sequence. Since the flow sequence for effects 4-6 was never changed, their coefficients were always the same. The order of coefficients for effects 1-3 was changed to correspond to the liquor flow sequence. Table VIII shows the set of heat transfer coefficients used in the order of liquor flow.

The solution scheme for the 6-effect evaporator/deactivation model was based on achieving a fixed heavy liquor solids and operating at a fixed surface condenser saturation pressure. Figure 18 shows the logic used to reach a converged solution. The scheme involved two nested control loops. An initial estimate of steam and weak liquor flows, and of all internal streams was input to the model. When the model reached a converged solution, the saturation pressure of the 6th effect vapor (equivalent to surface condenser pressure) was checked. If it



differed by more than a specified amount from the required value, the first effect steam flow was adjusted in proportion to the error. This was repeated until the required 6th effect vapor pressure was reached.

TABLE VIII

HEAT TRANSFER COEFFICIENTS FOR 6-EFFECT  
EVAPORATOR/DEACTIVATION SIMULATION

Effect Number	Order of Liquor Flow	$\underline{U}$ , Btu/ft <sup>2</sup> hr°F
6	1 (weak liquor)	190
5	2 (weak liquor)	240
4	3	316
--	4	386
--	5	392
--	6A (split effect, heavy liquor side)	174
--	6B (split effect, weak liquor side)	242

The second control loop was used to reach the required heavy liquor solids. It operated by adjusting the weak liquor flow in proportion to the heavy liquor solids error. Weak liquor flow adjustments were made only after the required 6th effect vapor pressure was reached. However, adjusting the weak liquor flow meant that the steam flow had to be reset also, which required another set of iterative calculations to converge for the steam flow-surface condenser pressure loop.

The convergence limits for heavy liquor solids and surface condenser pressure affect both solution accuracy and convergence time. To determine what the tradeoffs were, runs were made at different values of each convergence limit. The results in Table IX show that when the convergence limits were varied by a factor of 2 for liquor solids and a factor of 6 for surface condenser pressure,

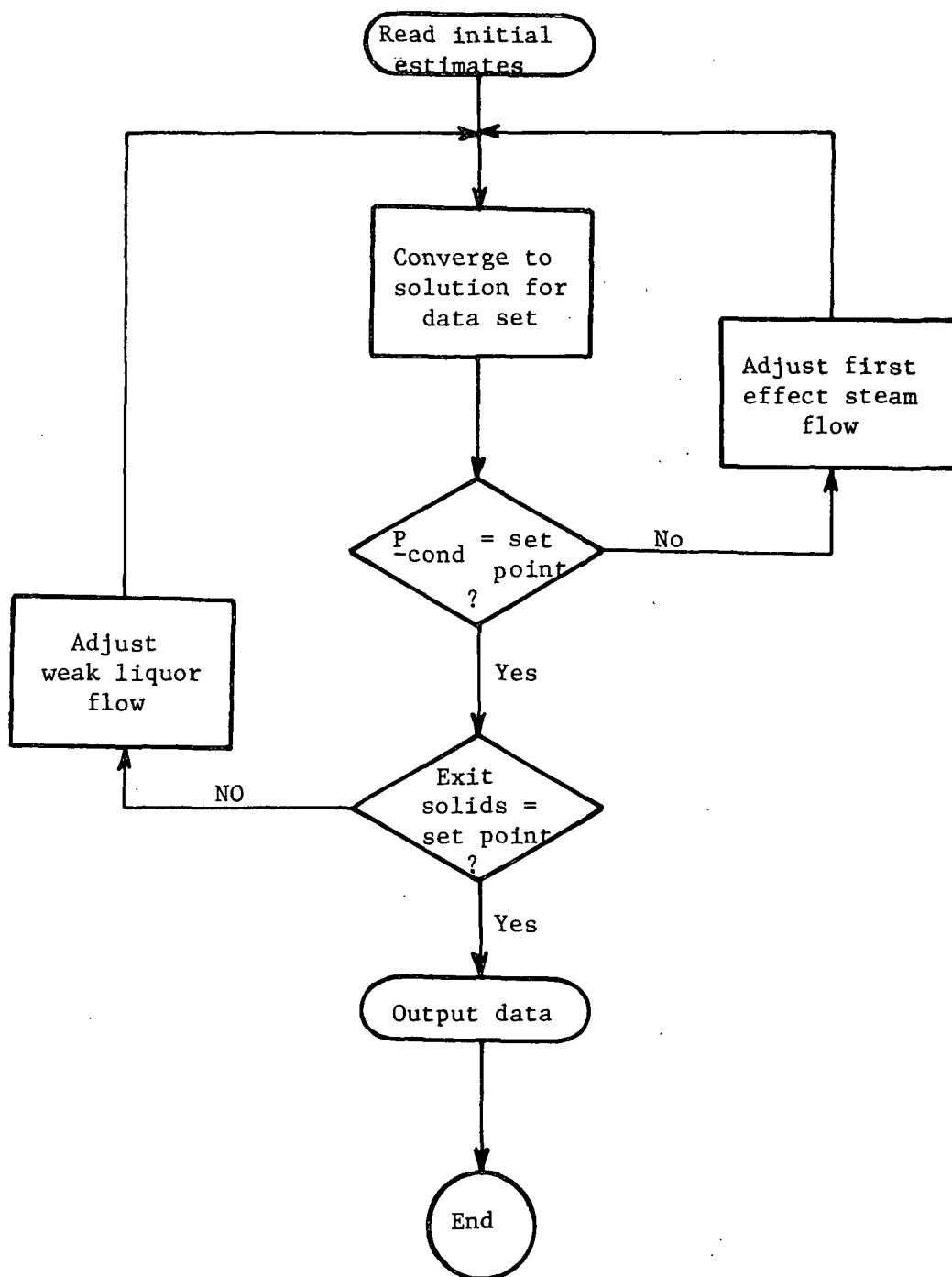


Figure 18. Computation Logic Diagram for GEMCS 6-Effect Evaporator/Deactivation Model

TABLE IX  
 COMPARISON OF CALCULATED FLOW RATES, SOLIDS CONTENTS,  
 AND TEMPERATURE PROFILES VS. CONVERGENCE LIMITS  
 FOR GEMCS 6-EFFECT EVAPORATOR SIMULATION

<u>Limits:</u>	I	II	III	IV
Heavy liquor convergence criterion, % liquor solids	0.1	0.05	0.1	0.1
Heavy liquor concentration, % liquor solids	±0.1	±0.05	±0.1	±0.1
Condenser temp. (pressure) °F (psia)	±0.2(0.01)	±0.1(0.005)	±0.36(0.02)	±0.5(0.03)
<u>Operation:</u>				
Weak liquor, lb/hr	330,598	330,594	330,168	331,933
Weak liquor solids, %	13.90	13.90	13.90	13.90
Heavy liquor flow, lb/hr	87,610	87,753	87,681	88,132
Heavy liquor solids, lb/hr	52.45	52.37	52.34	52.35
Evaporation rate, lb/hr	242,987	242,841	242,487	243,801
Steam economy	4.766	4.765	4.765	4.766
Boiling temperature, °F				
1A	252.5	252.5	252.5	252.4
1B	248.4	248.4	248.4	248.3
2	224.3	224.4	224.4	224.3
3	202.6	202.6	202.7	202.5
4	181.2	181.2	181.3	181.0
5	159.3	159.3	159.5	159.0
6	129.9	129.9	130.1	129.7
<u>Number of Interactions to Reach Convergence:</u>	187	279	150	62

the evaporation rate, steam economy, and liquor flow rates varied less than 0.6%, while the convergence time increased by a factor of 4.5. The first set of convergence limits shown in Table IX were chosen for the calculations with the deactivation system included.

## APPENDIX II

### STEAM AND POWER SYSTEM ENERGY COST CALCULATIONS FOR EVALUATING THE IPC THERMAL DEACTIVATION PROCESS ECONOMICS

The IPC deactivation process for calcium-scaling black liquors requires an increase in energy input when incorporated in an evaporator train. The energy costs for deactivation are three: process heating, liquor pumping, and loss in by-product power. Pumping energy depends on liquor flow rate, pressure drop, and liquor properties. Its calculation is straightforward. However, since the deactivation process becomes an integral part of the evaporation system, its energy flows affect the entire by-product power system. To calculate the process heating energy and by-product power penalty for deactivation, we must consider the entire power system, rather than the evaporators and deactivation process alone. This appendix shows how to calculate liquor heating energy and by-product power costs when a deactivation process is included in an evaporator train.

The net energy penalty for deactivation depends upon the mill power system operating conditions. For that reason, the energy cost for incorporating a deactivation system must be calculated separately for each mill. There are, however, two general categories of mills to consider: those with by-product power generation, and those without. The following two examples illustrate the procedures for calculating the deactivation energy penalty in each case. Remember that these examples are for specific mill operating conditions, and the particular numbers calculated do not apply in general to all mills.

## CALCULATION PROCEDURES

### Evaporator/Deactivation Energy Balances

The energy required as process heat for deactivation must be determined before the power system balances can be calculated. Figures 12 and 13 show the energy required as process heat when deactivation is incorporated into an evaporator train. The data are relative to the energy required for evaporation in a system without deactivation. They were obtained by simulating the evaporator/deactivation system using GEMCS. Although the GEMCS simulation results were obtained for a specific set of evaporator operating conditions (liquor properties, liquor solids, etc.), they will be applied directly to any 6-effect evaporator sequence regardless of liquor properties or weak or heavy liquor solids.

### Mills with By-product Power Generation

With by-product power generation, the energy costs for deactivation include a loss in by-product energy as well as an increase in liquor heating energy. The incremental energy cost for liquor heating can be calculated directly from the data in Fig. 12 and 13. However, extracting intermediate pressure steam from the turbines for process heating changes the entire power and steam balance. Therefore, the energy balance around the entire power and steam system must be recalculated to determine the change in boiler fuel requirements.

The basis for the calculations is the conventional recovery system described in the Kraft Recovery Alternatives Forum Supplement (5). The mill data represent state of the art recovery and power generation technology. The fuel, steam, and power balance for the conventional process is given in Table X. Figure 19 shows the key elements and energy flows for the system.

TABLE X

BASE CASE

Fuel, Steam and Power Balance

Basis:	Main Steam Conditions:	
	Pressure, psig	950
	Temp., °F	925
	Enthalpy, Btu/lb	1464
	Turbine efficiency, %	67
	Turbine exit enthalpy, Btu/lb	
	400 psig	1389
	150 psig	1322
	65 psig	1277
	Hog fuel available, BDT	0.06
	Heat to steam from hog fuel, 10 <sup>6</sup> Btu	0.500
	Power boiler efficiency, %	80
	Process steam demand, lb	
	<u>150 psig Sat.</u>	
	Paper drying	5100
	<u>65 psig Sat.</u>	
	Bleach plant	2500
	Miscellaneous	1000
	Condensate Return:	
	From Process:	
	% Return	60
	Enthalpy, Btu/lb	250

TABLE X (Continued)

BASE CASE

From within Recovery Envelope:

% Return	95
Enthalpy, Btu/lb	
400 psig	425
150 psig	333
65 psig	273

Feedwater make-up is calculated from the balances and condensate losses specified above.

Boiler blow down and other losses have not been included. Process condensate return is an estimated average, and is not specific for the individual departments.

Feedwater make-up temp., °F	80
Reference temp., °F	32

TABLE X (Continued)

BASE CASE

Conventional Recovery

•Steam and Condensate Flows (Except BFW Heating)•

Point of Use	Sat'd. Steam, lb	Steam Press., psig	Turbine Flow, lb	De-sup. Flow, lb	Concd. Return, lb	Concd. Enthalpy, 10 <sup>6</sup> Btu
Digesters	3593	150	3158	435	2156	0.539
Paper mach.	5100	150	4482	618	3060	0.765
Air htr.	1049	150	922	127	997	0.332
Liq. htr.	72	150	63	9	0	0
Shat. jet	12	150	11	1	0	0
Evap'n.	3158	65	2861	297	3000	0.819
Misc.	1000	65	906	94	600	0.150
Bl. plant	2500	65	2265	235	1500	0.375
Soot bl'g.	<u>600</u>	400	<u>600</u>	<u>0</u>	<u>0</u>	<u>0</u>
Total	17084		15268	1816	11313	2.980
		400	600			
		150	8636			
		65	6032			

•Feedwater Make-up•

$$17084 - 11313 = 5771 \text{ lb}$$

•Deaerator Balance•

Flow into Deaerator

Returned Condensate:

$$11313 \text{ lb} \qquad 2.980 \times 10^6 \text{ Btu}$$

Make-up:

$$5771 \text{ lb} \qquad 0.277 \times 10^6 \text{ Btu}$$

65 psig Turbine Exhaust

$$X \text{ lb} \qquad X (1277 \text{ Btu/lb}) \text{ Btu}$$

TABLE X (Continued)

BASE CASE

Flow Out of Deaerator

$$(17084 + X) \text{ lb} \quad (17084 + X) \cdot (282 \text{ Btu/lb}) \quad \text{Btu}$$

Deaerator Balance

$$(2.980 + 0.277) \times 10^6 + X(1277) = (17084 + X) (282)$$

65 psig Turbine Exhaust to Deaerator

$$X = 1569 \text{ lb}$$

•Total 65 psig Turbine Exhaust•

$$6032 + 1569 = 7601 \text{ lb}$$

•Heat to Power Generation,  $10^6$  Btu•

400 psig Extraction:

$$600 \text{ lb} (1464 - 1389) \text{ Btu/lb} = 0.045$$

150 psig Extraction:

$$8636 \text{ lb} (1464 - 1322) \text{ Btu/lb} = 1.226$$

65 psig Exhaust

$$7601 \text{ lb} (1464 - 1277) \text{ Btu/lb} = \underline{1.421}$$

$$\text{Total} = 2.692$$

•kw-hr Generated•

$$2.692 \times 10^6 \text{ Btu} \times \frac{1 \text{ kw-hr}}{3414 \text{ Btu}} = 788.5 \text{ kw-hr}$$

•Total Heat Added to Feedwater at Boiler•

Total Turbine Throttle Flow:

$$15268 + 1569 = 16,837 \text{ lb}$$

Heat Added at Boilers:

$$16661 \text{ lb} (1464 - 282) \text{ Btu/lb} = 19.901 \times 10^6 \text{ Btu}$$

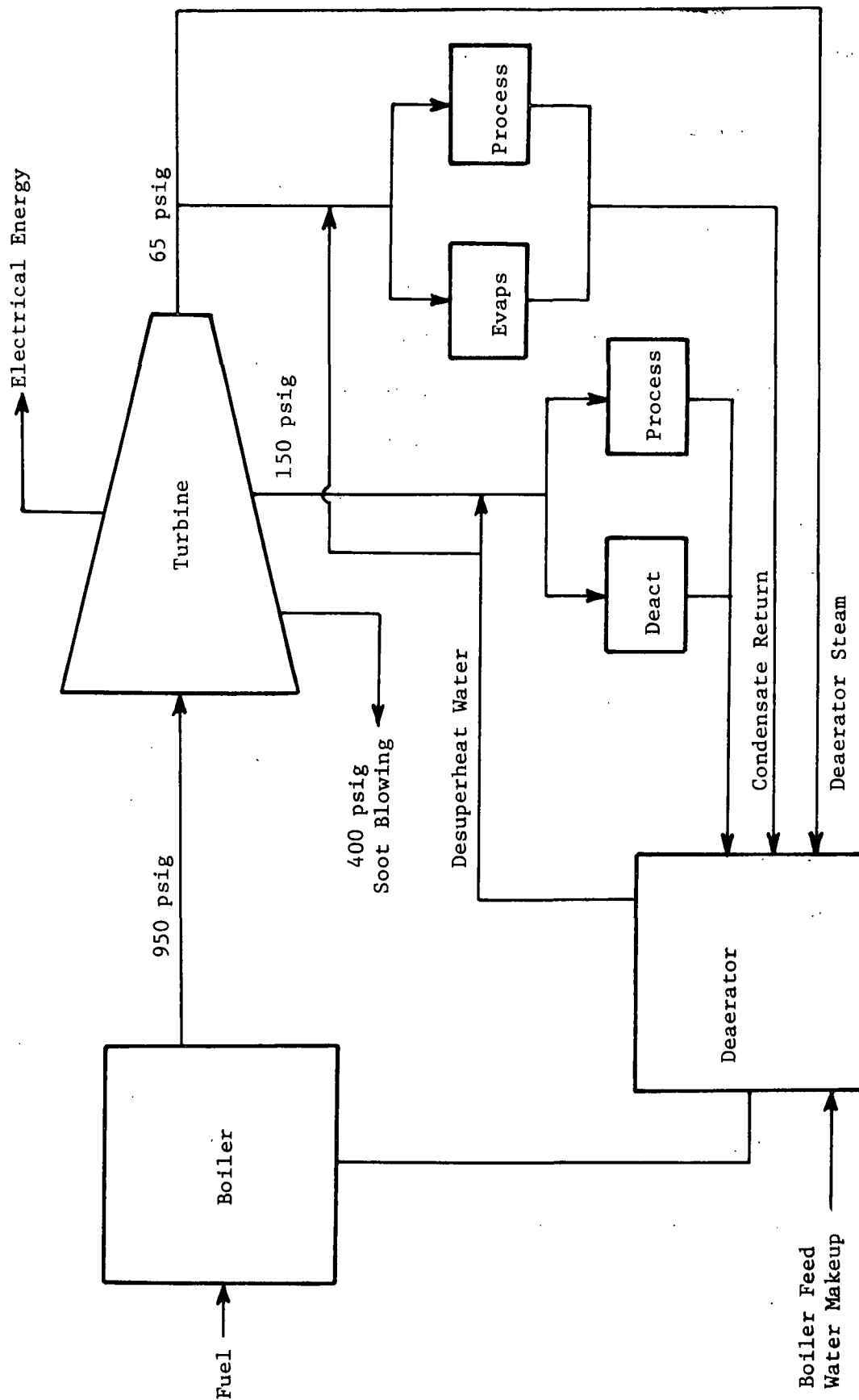


Figure 19. Power and Steam System: Steam, Condensate, and Electrical Energy Flows

Modifying the fuel, steam and power balance for the conventional system to include deactivation requires the following assumptions:

1. The power system is operated for maximum electrical energy generation as a noncondensing system. This means that extraction steam is condensed for process use only. The 65 psig turbine exhaust steam flow is limited by the heat required for process demands and at the deaerator. Steam to the deaerator is limited because the specific enthalpy in the deaerator cannot exceed that of 65 psig condensate.
2. The deactivation process and evaporator energy demands are defined in Fig. 12 and 13.
3. The evaporators and concentrator can be separated from each other, so that the data in Fig. 12 and 13 can be applied to the evaporators only. The evaporators were assumed to have a steam economy of 4.8 when operated separately. The liquor is assumed to reach 50% solids in the evaporators. The 65 psig steam required for the evaporators is 2118 lb/ADT, while that required for the concentrator is 743 lb/ADT.

Energy calculations for the deactivation process between the first and second effects for indirect heating are shown as follows. The energy flows are calculated on an airdry ton (ADT) basis for consistency with the base case data.

A. Steam and condensate flows to deactivation system and evaporators:

1. 65 psig extraction steam demand without deactivation = 2118 lb/ADT

2. with deactivation,

65 psig extraction steam to evaporators

$$= 2118 \times 0.816 = 1728 \text{ lb/ADT}$$

150 psig extraction steam to deactivation

$$= 2118 \times 0.186 \times \left( \frac{1277-282}{1322-338.5} \right) = 399 \text{ lb/ADT}$$

3. Total extraction steam flows for process heating

400 psig: 600 lb

150 psig: 399 + 8636 = 9035 lb

65 psig: 1728 + 3914 = 5642 lb

4. Desuperheat water flows (desuperheat water enthalpy = 282 Btu/lb):

to 150 psig steam: 9035 × 0.1378 = 1245 lb

to 65 psig steam: 5642 × 0.1038 = 586 lb

B. Deaerator balances

Condensate returned from evaporators:

$$0.95 \times 1728 \times 1.1038 = 1812 \text{ lb} \times 273 \text{ Btu/lb} = 0.495 \times 10^6 \text{ Btu}$$

Condensate returned from deactivation heat exchanger:

$$0.95 \times 399 \times 1.1378 = 431 \text{ lb} \times 333 \text{ Btu/lb} = 1.44 \times 10^6 \text{ Btu}$$

Other condensate returned: 6213 + 2879 = 9092 lb

$$0.738 + 1.636 = 2.374 \times 10^6 \text{ Btu}$$

Total condensate returned: 11,335 lb

$$3.013 \times 10^6 \text{ Btu}$$

Flow from deaerator to turbines and desuperheaters = 17,108 lb

Makeup feed water = 17,108 - 11,335 = 5773 lb

$$5773 \text{ lb} \times 48 \text{ Btu/lb} = 0.277 \times 10^6 \text{ Btu}$$

65 psig Exhaust steam to deaerator:

$$3.013 \times 10^6 + 0.277 + 1277X = (17,108 + X) (282 \text{ Btu/lb})$$

$$X = 1542 \text{ lb}$$

C. Power generation

400 psig extraction: 600 lb (1464 - 1389) Btu/lb =  $0.045 \times 10^6$  Btu

150 psig extraction: 9035 lb (1464 - 1322) Btu/lb =  $1.283 \times 10^6$  Btu

65 psig extraction: 7184 lb (1464 - 1277) Btu/lb =  $1.343 \times 10^6$  Btu

Total electrical generator =  $\frac{2.671 \times 10^6 \text{ Btu}}{3414 \text{ Btu/kw-hr}} = 782.4 \text{ kw-hr}$

Incremental electrical energy =  $782.4 - 788.5 = -6.1 \text{ kw-hr/ADT}$

D. Boiler fuel requirements:

Boiler steam flow =  $16,820 \text{ lb} \times (1464 - 282) \text{ Btu/lb} = 19.881 \times 10^6 \text{ Btu}$

Incremental electrical energy to steam

=  $19.881 \times 10^6 - 19.901 \times 10^6 = -20,000 \text{ Btu}$

Incremental purchased fuel =  $\frac{-20,000 \text{ Btu}}{0.8} = -25,000 \text{ Btu/ADT}$

E. Boiler balance

	lb/ADT	$10^6$ Btu/ADT
Steam flow and heat added at power boiler	16,820	19.881
Hog fuel heat to steam	--	0.500
Liquor heat to steam	--	14.454
Fossil fuel heat to steam	--	4.927
Purchased fossil fuel required (80% boiler efficiency)		6.159

Table V contains the results of these calculations for the two deactivation system locations within the evaporator train, for both direct and indirect steam heating.

Mills Without By-product Power Generation

For mills without by-product power, there is no by-product power penalty with deactivation. The increased energy demand with deactivation satisfies the additional liquor heating requirement and increased condensate losses. The mill power system shown in Fig. 20 is the basis for the following calculations. The deaerator enthalpy was held constant (282 Btu/lb) in these calculations. This required that the boiler steam flow vary to supply the required deaeration steam, as well as to satisfy the evaporator/deactivation system demands. The calculations are based on air-dry ton of pulp produced.

A. Steam and condensate conditions:

Main steam	Pressure	300 psig
	Temperature	460°F
	Enthalpy	1230 Btu/lb
First reduction	Pressure	150 psig
	Temperature	423°F
	Enthalpy	1230 Btu/lb
Second reduction	Pressure	65 psig
	Temperature	399°F
	Enthalpy	1230 Btu/lb

Condensate return and makeup feed water conditions the same as by-product power generation.

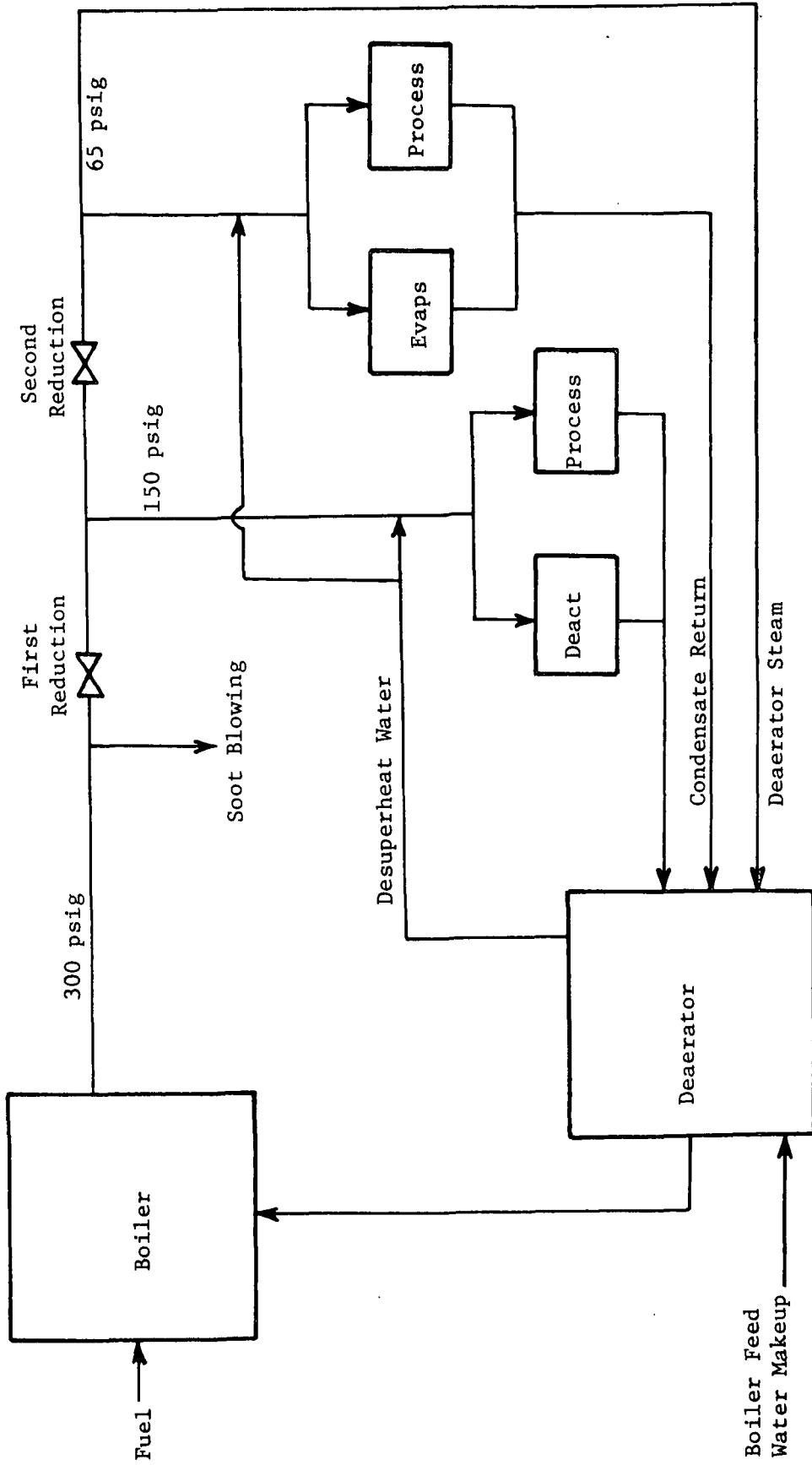


Figure 20. Steam System: Steam and Condensate Flows

B. Desuperheat steam flows

$$\begin{aligned} 150 \text{ psig steam: } & 1 \text{ lb (1230 Btu/lb) + X(282 Btu/lb)} \\ & = (1 + X) (1196 \text{ Btu/lb}) \end{aligned}$$

$$X = 0.0372 \text{ lb/lb superheated steam}$$

$$65 \text{ psig steam: } 1 (1230) + X(282) = (1 + X) (1183)$$

$$X = 0.0522 \text{ lb/lb superheated steam}$$

C. Process steam flows

1. soot flowing steam:  $600 \text{ lb} \times 1230 \text{ Btu/lb} = 0.738 \times 10^6 \text{ Btu}$

2. without deactivation

$$\begin{aligned} 150 \text{ psig process steam (sat'd): } & 9826 \text{ lb} \times 1196 \text{ Btu/lb} = \\ & 11.752 \times 10^6 \text{ Btu} \end{aligned}$$

$$\begin{aligned} 65 \text{ psig process steam (sat'd): } & 4320 \text{ lb} \times 1183 \text{ Btu/lb} = \\ & 5.111 \times 10^6 \text{ Btu} \end{aligned}$$

$$\begin{aligned} 65 \text{ psig evap. steam (sat'd): } & 2338 \text{ lb} \times 1183 \text{ Btu/lb} = \\ & 2.766 \times 10^6 \text{ Btu} \end{aligned}$$

Desuperheat water flow:

$$\text{to 150 psig steam: } 9826 \text{ lb} \left(1 - \frac{1}{1.0372}\right) = 352 \text{ lb}$$

$$\text{to 65 psig steam: } 6658 \text{ lb} \left(1 - \frac{1}{1.0522}\right) = 330 \text{ lb}$$

Total desuperheat water flow = 682 lb

300 psig steam flow:

$$\text{to 150 psig steam: } 9826 - 352 = 9474 \text{ lb}$$

$$\text{to 65 psig steam: } 6658 - 330 = 6328 \text{ lb}$$

3. With 2-D-1, indirect:

$$\begin{aligned} 65 \text{ psig evap. steam (sat'd)} &= 0.816 \times 2338 = 1908 \text{ lb} \times 1183 \text{ Btu/lb} \\ &= 2.257 \times 10^6 \text{ Btu} \end{aligned}$$

$$\begin{aligned} 150 \text{ psig deact. steam (sat'd)} &= 0.186 \times 2338 \text{ lb} \times \frac{901 \text{ Btu/lb}}{857 \text{ Btu/lb}} \\ &= 457 \text{ lb} \times 1196 \text{ Btu/lb} = 0.547 \times 10^6 \text{ Btu} \end{aligned}$$

Desuperheat water flow

$$\text{to 150 psig steam: } 10,283 \text{ lb} \left(1 - \frac{1}{1.0372}\right) = 369 \text{ lb}$$

$$\text{to 65 psig steam: } 6228 \text{ lb} \left(1 - \frac{1}{1.0522}\right) = 309 \text{ lb}$$

Total desuperheat waterflow = 678 lb

300 psig steam flow:

$$\text{to 150 psig steam: } 10,283 - 369 = 9914 \text{ lb}$$

$$\text{to 65 psig steam: } 6228 - 309 = 5919 \text{ lb}$$

D. Deaerator balance

1. Without deactivation

Input:  $X = \text{lb 300 psig stm (1230X Btu)}$

Condensate return:

$$150 \text{ psig processes: } 6213 \text{ lb (1.636} \times 10^6 \text{ Btu)}$$

$$65 \text{ psig processes: } 2879 \text{ lb (0.738} \times 10^6 \text{ Btu)}$$

$$65 \text{ psig evap: } 2221 \text{ lb (0.606} \times 10^6 \text{ Btu)}$$

$$\text{Total condensate return: } 11,313 \text{ lb (2.980} \times 10^6 \text{ Btu)}$$

Output:

$$\text{Boiler feed water} = (16,402 + X) \text{ lb (282 Btu/lb)}$$

$$\text{Desuperheat water } 682 \text{ lb (0.192} \times 10^6 \text{ Btu)}$$

Mass balance for makeup feedwater = Y

$$Y + 11,313 + X = 16,402 + X + 682$$

$$Y = 5771 \text{ lb} \times 48 \text{ Btu/lb} = 0.277 \times 10^6 \text{ Btu}$$

Energy balance for 300 psig steam flow:

$$0.277 \times 10^6 + 2.980 \times 10^6 + 1230X = (16,402 + X) (282) + 0.192 \times 10^6$$

$$X = 1646 \text{ lb} \quad (2.025 \times 10^6 \text{ Btu})$$

2. With 2-D-1, indirect

Condensate return:

$$150 \text{ psig from deact.} = 434 \text{ lb} \quad (0.145 \times 10^6 \text{ Btu})$$

$$65 \text{ psig from evaps.} = 1813 \text{ lb} \quad (0.495 \times 10^6 \text{ Btu})$$

$$\text{Total condensate return} = 11,339 \text{ lb} \quad (3.014 \times 10^6 \text{ Btu})$$

Boiler feedwater: 16,433 + X

Desuperheat water: 678 lb (0.191 × 10<sup>6</sup> Btu)

Mass balance for makeup feedwater (Y)

$$Y + 11,339 + X = X + 15,833 + 678$$

$$Y = 5772 \text{ lb} \times 48 \text{ Btu/lb} = 0.277 \times 10^6 \text{ Btu}$$

Energy balance for 300 psig steam flow

$$0.277 \times 10^6 + 3.014 \times 10^6 + 1230X = (16,433 + X) (282) + 0.191 \times 10^6$$

$$X = 1618 \text{ lb} \quad (1.990 \times 10^6 \text{ Btu})$$

E. Boiler balance

1. Without deactivation:

$$\begin{aligned} \text{Total boiler heat load} &= (16,402 + 1646) \text{ lb} (1230 - 282) \text{ Btu/lb} \\ &= 17.110 \times 10^6 \text{ Btu} \end{aligned}$$

2. With deactivation:

$$\text{Total boiler heat load} = (16,433 + 1618) (1230 - 282)$$

$$= 17.112 \times 10^6 \text{ Btu}$$

$$\text{Incremental heat load} = 0.002 \times 10^6 \text{ Btu}$$

$$\text{Incremental fossil fuel required} = \frac{2000 \text{ Btu}}{0.8} = 2500 \text{ Btu}$$

Table V also summarizes the results of the energy balances for two evaporator/deactivation configurations, for both direct and indirect steam heating.

Extension to Other Mills

The results in Table V show how deactivation system configuration and heating method affect purchased fuel requirements for a particular set of boiler, extraction, and exhaust steam conditions. These results do not apply to power systems operated at other conditions. A new set of calculations must be made to evaluate the energy costs associated with deactivation for other systems. However, we can predict qualitatively how changing the boiler, extraction, or exhaust steam pressures will affect incremental energy cost for deactivation.

1. Increasing or decreasing the steam pressure from the boiler should not affect substantially the incremental energy cost for deactivation. The deactivation energy costs result from (a) higher heat input to the deactivation/evaporation process and (b) less electrical energy produced when steam is extracted at a higher pressure. Neither of these is affected substantially by the high pressure steam conditions.
2. Decreasing the turbine exhaust steam pressure increases the energy cost for deactivation, all other factors held constant.

The electrical energy generated by expanding steam from 150 psig to 35 psig, for example, is about 50% greater than when expanding from 150 psig to 65 psig.

3. Increasing the extraction steam pressure used for deactivation has the same effect on deactivation energy penalty as does decreasing the exhaust steam pressure.

APPENDIX III

EXAMPLES OF ECONOMIC ANALYSIS FOR THERMAL DEACTIVATION

1. Cost of thermal deactivation on a \$/BDT pulp basis.

Operating cost parameters: Fossil fuel           \$3.00/10<sup>6</sup> Btu  
                                  Electrical power   \$0.030/kw-hr  
                                  Boiler feedwater makeup treatment \$1.84/1000 gal

Amortization of capital: 20% internal rate of return, straight line depreciation  
                                  over 10 years, 48% tax rate

$$\begin{aligned} \text{Annual cost} &= \left(0.2 + \frac{1-0.48}{10}\right) \times \text{capital cost} \\ &= 0.252 \times \text{capital cost} \end{aligned}$$

Using the data in Tables IV, V, VI, and VII for a 500 BDT/day pulp mill with indirect heating deactivation at 300°F for liquor entering the first effect only:

$$\text{Annual cost of capital} = \frac{\$390,000 \times 0.252}{500 \text{ BDT/day} \times 350 \text{ days/yr}} = \$0.56/\text{BDT}$$

Steam and power system costs:

$$\text{Incremental fossil fuel cost} = -25,000 \frac{\text{Btu}}{\text{ADT}} \times \frac{\$3.00}{10^6 \text{ Btu}} \times \frac{1 \text{ ADT}}{0.9 \text{ BDT}} = -\$0.08/\text{BDT}$$

$$\text{Incremental electrical power cost} = \frac{6.1 \text{ kw-hr}}{\text{ADT}} \times \frac{\$0.030}{\text{kw-hr}} \times \frac{1 \text{ ADT}}{0.9 \text{ BDT}} = \$0.20/\text{BDT}$$

$$\text{Net steam + power system cost} = \$0.20/\text{BDT} - 0.08/\text{BDT} = \$0.12/\text{BDT}$$

$$\begin{aligned} \text{Boiler feedwater makeup cost} &= 0.2 \text{ gal/ADT} \times \frac{1 \text{ ADT}}{0.9 \text{ BDT}} \times \$1.84/10^3 \text{ gal} \\ &= \$0.0004/\text{BDT} \end{aligned}$$

$$\text{Pumping energy} = 1.4 \text{ kw-hr/ADT} \times \frac{1 \text{ ADT}}{0.9 \text{ BDT}} \times \$0.030/\text{kw-hr} = \$0.05/\text{BDT}$$

$$\text{Total cost} = \$0.73/\text{BDT}$$

2. Discounted cash flow (DCF) analysis.

The following calculations are made to compare thermal deactivation with an additional evaporator body as a means to increase evaporation capacity. The comparison is made for a 500 BDT/day mill, using indirect heating thermal deactivation before one effect. A value of \$10/BDT incremental pulp capacity is assumed in one case, and of \$100/BDT in the other. These values bracket the \$27/ton of paper reported by Rogers (6). The rate of loss of evaporation capacity was taken as 150%/month.

For the comparison, incremental capacity was calculated using Fig. 14 and 15, assuming an 85% reduction in scaling rate. Scale removal savings were calculated as follows, assuming the evaporator cleaning frequencies shown:

Water boil out frequency [calculated according to Harrison (5)]:  
15 times/month

Cost: \$424/boil out (includes steam, water, labor)

Acid wash frequency: once/month

Cost: \$640/acid wash (includes inhibited acid, water, labor)

Hydroblast frequency: once/yr

Cost: \$10,000/effect

Total annual cleaning cost = \$94,000

Figure 16 was used to calculate the heat transfer area of the additional evaporator body. The cost of an evaporator body was calculated as

$$\text{Installed cost} = \$429,000 \left( \frac{\text{Heat transfer area}}{10,000 \text{ ft}^2} \right)^{0.7}$$

Equipment costs were depreciated by the straight line method.

Table XI summarizes the cost data used and the procedure followed in the analysis.

The results show that either method of increasing capacity shows a favorable

return on investment, and that thermal deactivation has an advantage for this particular case.

TABLE XI

DISCOUNTED CASH FLOW ANALYSIS OF THERMAL DEACTIVATION VERSUS  
AN ADDITIONAL EVAPORATOR BODY FOR INCREASING EVAPORATION CAPACITY  
(500 BDT/DAY PULP MILL, INDIRECT HEATING DEACTIVATION BEFORE  
1 EFFECT, 150%/MONTH RATE OF EVAPORATION CAPACITY LOSS)

Extra Capacity by:	Thermal	Deactivation	Additional Evaporator Body	
Value of incremental pulp production, \$/BDT	10	100	10	100
Additional capacity, %	15.4	15.4	15.4	15.4
Capital investment	390,000	390,000	470,000	470,000
Scale removal savings	94,000	94,000	0	0
Incremental pulp capacity value	269,500	2,695,000	269,500	2,695,000
Operating costs	(24,500)	(24,500)	0	0
Depreciation	(39,000)	(39,000)	(47,000)	(47,000)
Taxable savings	300,000	2,725,500	222,500	2,648,000
After tax savings (50% tax)	150,000	1,362,750	111,250	1,324,000
Cash flow	189,000	1,401,750	158,250	1,371,000
Discount rate for zero net present value	47.5	365	31.5	292

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