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Designing and Operation of Reactors for Suspension Polymerization of Vinyl Chloride

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INTRODUCTION

For suspension polymerization of vinyl chloride there are considerable economic incentives to develop improved operational techniques. For future plants a method to optimize readily both the design and operating techniques is needed. In such cases, consideration must be given to several design variables, operating parameters, and economic considerations. For existing plants the preferred procedures for polymerization may change at frequent intervals because of variations in costs, availability of materials, and demands for specific PVC products.

The objective of the present investigation is to develop a generalized computer model that can be used in such optimizations for both new and existing PVC units. In addition, such a model can be used to evaluate readily the potential importance of certain changes in variables in order to determine the magnitude of benefits that might be realized relative to both capital and operating costs. Such an evaluation should be helpful in identifying promising areas for future research.

The approach used here was developed to a considerable extent based on earlier Purdue investigations [1, 2]. Two or three variables

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that have been largely overlooked in the past have been found to be of key importance.

TESTING OF MODEL DEVELOPED

The calculation details used in the model are described in the Appendix. Over one thousand computer runs have been made in order to demonstrate the relative importance of over 35 key variables (design, operating, or economic variables) and in an effort to indicate improved reactor design and preferred operating conditions. The general procedure employed can be summarized as follows:

(a) Certain operating conditions, reactor details (including volume of reactor, height-to-diameter ratio, type of reactor wall, and number or size of baffles), and type of heat transfer surface used to control the polymerization temperature need to be specified. Table 1 lists the values of these variables used in the so-called base case of this investigation.
(b) Costs to calculate both investment (or capital) costs and polymerization costs must be specified. Table 2 lists the costs used in the base case.
(c) Kinetic data must be provided. In the base case it was assumed that the shape of the curve of rate of polymerization versus time was identical regardless of the total time of polymerization; the maximum rate was assumed to be 1.8 times the average rate. Normally, kinetic data can be obtained from either laboratory or plant data.

Half-pipe jackets were used in the design of reactors in the present investigation. They result in high overall heat transfer coefficients [6]. Values in the range of 85 to 110 Btu/h, sq. ft. °F were estimated for the vessels designed here; such values are significantly higher than those in some commercial units now in operation. Half-pipe jackets provide high rates of heat transfer and also result in only relatively low pressure drops of coolants. If sufficient information were available so that overall heat transfer coefficients could be calculated for other types of jackets, the latter jackets could also be considered.

In the present investigation the ability of the computer model developed was demonstrated for an example in which a new plant was to be built having an annual capacity of 300 million pounds PVC. The relative importance of the numerous design variables and operating parameters were identified by comparison to the so-called standard runs whose conditions are specified in Tables 1 and 2.

In calculating comparative polymerization and capital cost information, it was first necessary to specify the internal volume and the

TABLE 1. Standard Conditions Used in Sample Optimization[a]

Operating Conditions for Polymerization and Reactor Operation:

1) Temperature for polymerization, 140° F (60° C)
2) Volume fraction of vinyl chloride in suspension, 0.47, or water-to-vinyl chloride weight ratio of 1.25:1.
3) Sufficient agitation and suitable reactor design to give an inside heat transfer coefficient of 200 Btu/h, sq. ft, °F. It was assumed 1 horsepower nominal energy was required for every 100 gal of reactor volume.
4) Vinyl chloride conversion, 90% per batch.
5) Four hours provided as down time (i.e., time between end of one polymerization and start of next).
6) Rate of heat release as function of polymerization time is such that the ratio of maximum rate is 1.8 times the average rate. (This variable is, of course, strongly affected by the initiator (or catalyst) employed and the polymerization recipe).

Reactor Design:

1) Walls constructed of low-carbon steel but clad on inside with stainless steel.
2) Four water-cooled baffles per reactor.
3) Highly polished inner surfaces in order to obtain high heat transfer coefficients as indicated above.

Jacket Design and Operation:

1) Half-pipe jacket with 4 in. half pipes and 0.75 in. spacing.
2) Cooling water flow is once through.
3) Maximum pressure drop of water in jacket, 25 psi.
4) Maximum water velocity in jacket, 15 ft/s.
5) Temperature of cooling water to jacket, 86° F (30° C)

[a]Any of the conditions can easily be modified as desired.

height-to-diameter ratio of the reactor. Then the following were calculated in the order specified below:

(a) Reactor diameter (picked to nearest 0.5 ft).
(b) Reactor height.
(c) Wall thickness.
(d) Number of parallel passes of half-pipe coil needed around circumference of reactor so that pressure drops of cooling water in jacket and the maximum flow velocity of water in jacket were close to but less than the following constraints: 25 psi

TABLE 2. Cost Information Used in Sample Optimization

Investment Cost

1) Purchase price of reactors and water-cooled baffles as supplied by Brighton Co., Cincinnati, Ohio.
2) Piping, foundations, structures, instruments, electrical wiring, insulation, painting, installation, and general overhead for building and installation were estimated as fraction of purchase price of reactor. The fractions used were 0.30, 0.15, 0.05, 0.12, 0.10, 0.05, 0.08, 0.10, and 1.4, respectively.
3) The sum of 1) and 2) indicates the total investment cost for the polymerization section only of the plant.

Operating Costs

1) Labor and supervision: \$20/h. It was assumed that one operator could handle four reactors because of good instrumentation and controls. A cost of \$5/h per reactor was considered reasonable.
2) Cooling water at 86° F, 30 ¢/1000 gal.
3) Electricity, 3 ¢/kWh.
4) The annual costs of depreciation, maintenance, property taxes, insurance, and general services were calculated as a fraction of the total investment cost. The fractions used were 0.10, 0.08, 0.02, 0.01, and 0.05, respectively.

pressure drop and 15 f/s velocity. Up to 10 passes were needed for some of the larger reactors. As a further constraint, the parallel jackets were designed so that the inlet and outlet openings of the several parallel passes were one above the other; such arrangement simplifies the piping arrangement for the cooling water.

(e) Time required for polymerization of a batch of vinyl chloride.

(f) Total cooling water requirements for each batch run and also the maximum flow rate of cooling water during the batch run. Such a maximum flow rate occurs when the rate of polymerization is highest.

(g) Number of batch reactors required for production of 300 million pounds/year of PVC. In addition the actual predicted capacity per year of the plant; in all cases this capacity was slightly greater than 300 million pounds/year since an integral number of reactors must be chosen.

(h) Investment costs for reactor portion of plant with annual capacity of 300 million pounds.

(i) Cost of polymerization reported on cents/pound PVC basis.

Polymerization costs ranged over rather wide ranges depending on the specific conditions chosen. Costs ranged from as low as 1.0 cents per pound PVC produced up to about 2.0 cents for conditions that are thought to be similar to those used in at least some industrial plants. Capital (or investment) for a 300 million pound per year unit ranged from as low as about $9,000,000 to as high as $20,000,000. Duplicates of current commercial units were often much higher than the lower value. In all cases there was a close correlation between low polymerization costs and low capital costs.

In making these economic evaluations, several assumptions were made as follows:

(a) The costs of vinyl chloride, initiator (or, as it frequently is referred, catalyst), stabilizing agent (or surfactant), and process water (added as suspending fluid for vinyl chloride and PVC) are identical in comparative runs. These costs were not included in the polymerization costs reported here.
(b) The investment and operating costs applicable to other portions of the PVC plant besides the polymerization section are identical. The other portions of the plant include storage facilities for vinyl chloride, initiator, stabilizing agents, processing water, and PVC; and the separation, recovery, and processing sections involved with PVC, unreacted vinyl chloride, stabilizer, and process water.
(c) The quality and properties of the PVC product are identical in all cases investigated regardless of the reactor design and difference in operating conditions. This assumption is, of course, not always true [5].

For the base case with a 20,000-gal reactor having a height-to-diameter ratio of 2.0 to 1, the factors contributing to the polymerization costs are shown in Table 3.

Total investment for the polymerization portion of the plant to produce 300 million pounds PVC per year was $11.3 million for the above base case. A total of 12 reactors was needed (actual estimated production was 311.8 million pounds PVC per year).

As a general rule for PVC reactors, if improved heat transfer can be realized, the following two options are available:

(a) Use of less cooling water.
(b) Use of faster rates of polymerization and hence shorter times for batch polymerization; such faster rates can be realized by changes in the type and amount of initiator used and possibly other changes in operation.

Obviously an optimum must be sought between these two options. As a general rule, emphasizing the search for faster rates is the preferred route toward the optimum. Faster rates reduce all polymerization costs listed above, except for the cost of cooling water.

TABLE 3. Factors Contributing to Polymerization Costs

	Cents/lb	% of total
Depreciation	0.327	25.4
Maintenance	0.261	20.3
Property taxes	0.065	5.1
Insurance	0.033	2.6
General services	0.163	12.7
Labor and supervision	0.152	11.8
Electricity	0.107	8.3
Cooling water	0.179	13.9
Total	1.287	100.0

Relative to faster rates of polymerization, many commercial PVC plants operate with polymerization times (from the start of reaction to 90% conversion) for a batch run as low as 7 to 10 h [4]. Times as low as 4 h are also known to be used for certain industrial polymerizations. Two techniques are or have been employed to obtain adequate temperature control for such fast polymerizations:

(a) Small reactors, such as 2000 gal, with high surface-to-volume ratios. Such an approach is no longer economically feasible.
(b) Large reactors up to at least 20,000 gal and perhaps even 50,000 gal that employ either reflux condensers or a judicious combination of initiator, level of agitation, and the use of chilled water.

IMPORTANT REACTOR DESIGN PARAMETERS

The method of controlling temperature, or of removing the heat of polymerization, is of major concern for designing a reactor. As a general rule, the reactor should be operated so that heat transfer is almost rate controlling. Initially reactors will be considered in which all heat is transferred from the reactor through the jacket and water-cooled baffles. Later the use of reflux condensers will be considered. Table 4 illustrates several typical comparisons indicating the importance of several design or operating variables. For reactors without reflux condensers, the following were found to be of major importance.

TABLE 4. Select Comparative Costs (base case values used unless specified)

Variable investigated	No. of reactors	Annual capacity in millions lb/yr	Annual investment in millions dollars	Operating cost, ¢/lb PVC
	For 10,000 Gallon Reactors			
Height/dia. ratio:				
1:1	21	302	17.0	1.51
2:1	18	317	10.0	1.26
3:1	16	311	8.9	1.16
For 20,000 Gallon Reactors (3:1 ratio height to diameter)				
Wall construction:				
Solid SS	14	302	14.3	1.66
Clad SS	10	309	9.4	1.14
h_i = 160 Btu/h, sq. ft, °F	11	307	10.3	1.25
= 220	10	322	9.4	1.11
Cooling water temperature:				
41°F	7	344	6.6	0.57[a]
86°F	10	309	9.4	0.93[a]
	h_i = 140 Btu/h, sq. ft, °F			
No reflux condenser denser	11	329	10.4	1.22
Reflux condenser to remove 20% of heat	9	321	8.4	0.99
Ratio of maximum to average rate of heat release:				
2:1	13	307	12.3	1.45
1.5:1	11	329	10.4	1.22

[a]Operating costs do not include those of cooling water.

Reactor Wall Design

Reactor walls constructed with stainless steel clad on low carbon steel are greatly preferred as compared to either walls constructed of solid stainless steel or glass-clad on low carbon steel. Both investment and operating costs can be reduced by about 25-40% when stainless steel clad walls are used as compared to either of the other two walls; Table 4 indicates such a comparison for a 20,000-gal reactor.

Height-to-Diameter Ratio

In the past, reactors have been designed with ratios of approximately 1:1 to perhaps as high as 1.5:1. There is no indication that higher ratios have ever been considered. It has now been found that 20 to 40% reductions in both operating costs and investment costs are generally possible as the ratio is increased from 1:1 up to 3:1. Such a comparision of 10,000 gal reactors is shown in Table 4. Such improvements occur because of the increased area available for heat transfer through the wall. In making these calculations it was assumed that investment costs of a given volume reactor do not change as the height-to-diameter ratio changes. Based on information supplied by the Brighton Corp., such costs may actually decrease slightly with larger height-to-diameter ratios. Counterbalancing this, however, may be slightly greater costs due to multiple impellers with longer agitator shafts in the taller reactors. It would seem that height-to-diameter ratios of at least 3:1 could be used without serious limitations because of agitator design.

Reactor Volume

The optimum size as determined from operating and investment cost considerations depends to a considerable extent on the height-to-diameter ratio. Limiting the discussion for the moment to reactors that are not provided with reflux condensers and that have height-to-diameter ratios of 3:1, the preferred size for a reactor is generally either 20,000 or 30,000 gal for the values of parameters used in the base case. The lower operating cost of a 20,000-gal reactor as compared to a 10,000-gal reactor is shown in Table 4.

Water-Cooled Baffles

Use of up to four water-cooled baffles in each reactor increases the annual production capacity of a given reactor on an annual basis. Small but nevertheless significant reduction of operating costs are also realized. A reduction of 6-8% in operating cost occurs for reactors containing four water-cooled baffles as compared to ones with essentially identical but noncooled baffles.

ECONOMIC IMPORTANCE OF SEVERAL OPERATING VARIABLES

For reactors in which heat transfer is exclusively through the jacket and water-cooled baffles, several operating variables are found to have a large effect on the economics of PVC production. Although such effects were realized previously in a qualitative manner, quantitative information has apparently never been previously reported.

Initiator(s) Used

Changing the amount and especially the type of initiator (or catalyst) has an important effect on the flatness of the heat release rate or the ratio of the maximum rate of heat release to the average rate for a run. A ratio of 1.0:1 would signify a completely flat rate, i.e., a constant rate. Based on data supplied by Pennwalt in Technical Bulletin 30.90 [7] for various initiators, ratios of the maximum to the average rates are possible for as low as about 1.1:1 whereas other initiators give ratios as high as 2:1. Ratios from 2:1 down to 1.0:1 were tested. In making these calculations, account was always taken of the energy introduced by the agitator; in many cases 5-10% of the energy that has to be removed from the reactor is energy added by the agitator.

Economic saving of at least 10-30% can be realized by changing to initiators causing flatter rates of heat release. Such a savings is demonstrated in Table 4 as the ratio of maximum to average rate of heat release is reduced from 2:1 down to 1.5:1. Even greater savings can be obtained at lower ratios. In making these calculations it was assumed that initiator costs in all cases were identical which is obviously not quite correct. Initiator costs could easily be added to the computer model if desired.

Level of Agitation

Agitation has a significant effect on the inside coefficient of heat transfer (h_i). In general more agitation will increase h_i and, hence, the overall heat transfer coefficient (U). Although several industrial companies do not seem to know the exact values of h_i in their reactors, values of 160 to at least 200 Btu/h, sq. ft, °F seem possible. (Based on information published by Shinetsu Chemical Co. and on information obtained from other PVC manufacturers, these values seem reasonable. They are, however, considerably higher than values reported by Cameron et al. [4] for what is probably a Conoco PVC unit using a reflux condenser.) It is, of course, realized the h_i values vary as the batch run progresses and as PVC is produced. If sufficient agitation can be provided to obtain h_i values of 220 as compared to 160, both capital and operating costs are reduced by about 10% as indicated by a comparison shown in Table 4. Even greater reduction would be possible with h_i values of 240. In making these calculations it was assumed that electrical costs for operating the agitator did not change. Actually these costs would presumably increase with higher h_i values so that the apparent economic savings are perhaps only in the range of 6-8%. Of course, the character and quality of the granule PVC product would also change with the level of agitation [5].

DESIGN AND OPERATION OF JACKET

Several factors were found to have important effects on the economics of PVC production. For example, there are two methods of flowing the coolant through the jacket [3]:

(a) A recycle pump is provided and from 0 to 100% of the water is recycled depending on the heat transfer load. At high rates of heat transfer, the amount of water recycled is low or even zero.
(b) No water is ever recycled and the cooling water is once through. Variations in the amount of heat to be transferred are controlled by varying the flow rate of coolant to the reactor.

The latter method, i.e., once through coolant or water, was found in all cases investigated to be the preferred method.

(a) In almost all cases, less cooling water was required per batch run with once through water [2]. If refrigerated or quite cold water is used, this conclusion may not be true.
(b) There is a substantial expense for recycling water because of the investment costs to provide a large pump and the cost of operating it.

The jacket in each case was designed for the maximum rate of heat release during the batch run. In most cases as part of this design, it was assumed that the temperature rise of the cooling water in the jacket was 9°F (or 5°C). The same jacket was provided for both once through and recycle water because it was assumed that no recycle occurred during the period of maximum rate of heat release;

in other words, the coolant flows were identical for this time period for both recycle and once through water. A temperature rise of 9°F (actually the minimum temperature rise during the batch run) was found in general to be approximately optimum. A large temperature rise results in lower cooling water costs but investment costs are substantially higher. There is still a need, however, to investigate in more detail the minimum temperature rise that is most desirable for initiators which provide relatively flat rates of heat releases and for those plants in which cold water is available as a coolant.

Although not tested here, the effect of spiral baffles in a jacket could easily be predicted using the current model. Such a baffled jacket would undoubtedly be very good from a heat transfer standpoint. It is, however, unclear as to the capital and maintenance costs for such a jacket.

The temperature of the cooling water has a large effect on the amount of cooling water required per batch run and on the theoretical capacity of a given reactor. Ignoring water costs, there is a 30-40% reduction in operating costs possible as the cooling water temperature decreases from 86°F (30°C) to 41°F (5°C); one such comparison is shown in Table 4. These results suggest that a PVC manufacturer might find it economically desirable to change his operation (such as operating procedure and/or initiators used) during the winter when water temperatures are lower.

Although not attempted, the model developed could easily be adapted to test and optimize the use of chilled (or refrigerated) water as a coolant; refrigeration of the cooling water would sometimes be economically feasible. Selecting the level of refrigeration could easily be accomplished with the model once the cost of chilled water is known.

Volume Fraction of Vinyl Chloride in Suspension

The optimum volume fraction of vinyl chloride can easily be determined using the model. Preliminary tests indicate that a fraction of 0.47 (or a water-to-vinyl chloride weight ratio of 1.25:1) is close to the optimum. A smaller fraction of vinyl chloride reduces the amount of PVC produced per batch but permits shorter times for polymerization. A larger fraction that increases the amount of PVC produced per batch does so at the expense of longer times for polymerization.

Temperature of Polymerization

Both operating and investment costs are significantly reduced as the temperature of polymerization is increased. The properties and quality of the PVC product, however, are known to change significantly with temperature.

ECONOMIC ADVANTAGES RELATIVE TO REFLUX CONDENSERS

Large economic savings are possible if a reflux condenser can be used to remove a portion of heat of reaction and of agitation from the reactor. These condensers can be built sufficiently large so that normal cooling water can be and normally is used. The following three options are possible for reactors equipped with a reflux condenser:

(a) Use the reflux condenser only when the rate of heat transfer is at or near the maximum.
(b) Use the jacket for heat transfer only when the rate of heat transfer is at or near the maximum.
(c) Use both the reflux condenser and jacket during the entire run.

Comparisons were made of the latter option when 0, 10, and 20% of the total heat was removed by the reflux condenser. Both capital and operating costs were reduced by 10 to 15% as the percent of the heat removed by the reflux condenser increased from 0 to 20%. Table 4 indicates for one comparison that 20% heat transfer by the reflux condenser reduces the operating cost from 1.22 to 0.99 ¢/lb PVC. The time for polymerization was for this specific comparison in a 20,000-gal reactor reduced from 12.2 to 9.6 h. Even greater reductions would be possible if more heat was removed by means of reflux condensers.

Use of a reflux condenser requires that fouling in the reflux condenser be minimal; and several companies have developed satisfactory procedures. Chemische Werke Hüls AG apparently transfers over 50% of the heat through the reflux condenser [8].

Use of a reflux condenser results in some additional design factors such as the location of vaporization zone in the suspension or emulsion, condensation of vinyl chloride and water in the reflux condenser, method of returning liquid condensate to the reactor, and the remixing of vinyl chloride reflux with initiator. Some companies claim that very high quality PVC product is obtainable in reactors using a reflux condenser; there may, however, be differences in quality in other reactors.

DISCUSSION OF RESULTS

The comparative results shown here have identified variables that have important effects on the economics of vinyl chloride polymerization. As general rules the following should all be employed to obtain the preferred polymerization:

(a) Use batch reactors that are large, probably in at least the 20,000 to 30,000 gal size, have fairly high height-to-diameter ratios, and employ walls constructed of low carbon steel clad with stainless steel.
(b) Use reflux condensers to remove at least 20-50% of the heat released during periods of maximum heat release.
(c) Provide highly effective jackets such as half-pipe jackets or jackets provided with spiral baffles. Water-cooled baffles are also helpful.
(d) Use initiators that provide relatively flat kinetics and also provide relatively high rates of polymerization.

Based on the general rules specified above, polymerization costs at least as low as 1.0 ¢/lb PVC and investment costs as low as $9 million are predicted which is a significant saving as compared to the base case.

In designing a new PVC plant or in developing improved operation in an existing plant, the model will obviously need to be modified in order to incorporate the best available information relative to plant design and operation, kinetics of polymerization, and cost information. Although the assumptions made in the present investigation are thought to be reasonable, they certainly are not applicable in all cases. The model developed is flexible enough so that many modifications can easily be made. Once the model is customized to a given plant or company, optimization investigations can quickly be made. The model can also be used as a guide for future research or developmental work relative to process improvements; conditions at which significant economic advantages possibly can be realized and can easily be identified. For example, the potential advantages that could be realized with better automation and hence reduced labor could easily be evaluated.

The optimization technique reported here can easily be adapted to other batch polymerizations by either suspension or emulsion techniques. Preliminary tests have already been made for the suspension polymerization of styrene. It seems apparent that the model can also be used for other batch reactors including batch hydrogenerators such as those used to produce oleomargarine and shortening stocks and for fermentators.

APPENDIX: OUTLINE OF COMPUTER PROGRAM

(1) Values for various operating, design, and economic variables were fed to the computer. Generally the values shown in Tables 1 and 2 were used.
(2) Calculation of reactor dimensions. After the reactor volume and height-to-diameter ratio were specified, the program calculated height, diameter, wall thickness, and total heat transfer area for both jacket and baffles.

(3) Design of jacket. Kinetic information for polymerization was specified relative to the heat release rate (caused by polymerization) as a function of time. The ratio of the maximum rate of heat release to the average rate of heat release was also specified.

(a) U value was assumed; usually a value of 100 Btu/h, sq. ft, °F was set as a first approximation.
(b) Log mean temperature difference for heat transfer was calculated (as a rule the inlet and outlet temperatures of cooling water were set at 86 and 95°F, respectively, when the rate of heat release was maximum).
(c) $Q_{max} = UA\Delta T_{ln\ mean}$ = maximum rate of heat release.
(d) Calculate maximum rate of flow of the cooling water, $w_{max} = Q_{max}/(Cp_{water})(\Delta T_j)$, where ΔT_j was usually 9°F (95-86°F).
(e) A subroutine designed the number of passes for w_{max} (setting the number of passes to maintain a pressure drop for water in jacket of less than 25 psi and to maintain a flow velocity in jacket of less than 15 fps).
(f) Calculate heat transfer coefficient in jacket, h_0 (using Dittius-Boelter equation).
(g) Calculate U (using h_0, h_i, and wall resistance).
(h) Repeat until U values converge to desired limits.

(4) Time required for batch run (from the start of polymerization to a predefined conversion):

(a) Determine maximum rate of heat release because of polymerization (the energy transferred to reaction mixture because of agitation or pumps must be subtracted from Q_{max}).
(b) Knowing ratio of maximum heat of polymerization to average heat of polymerization and the total heat of polymerization for batch run, the time required for a batch polymerization can be calculated.
(c) Calculate time for complete batch run.

(5) Reactors required for plant. With plant capacity set, the following calculations were made:

(a) PVC capacity for a batch run.
(b) PVC capacity for year in single reactor.
(c) Number of reactors required.

(6) Cooling water requirements:

(a) The heat that needs to be transferred as a function of time during batch run was calculated using kinetic data for polymerization and energy added by agitator and recycle pump (if used).

(b) For a given time period and using once through water, assume temperature of cooling water leaving jacket.
(c) Calculate $\Delta T_{\ln\ mean}$.
(d) Calculate U.
(e) Calculate flow rate of water needed to remove heat.
(f) Calculate velocity of water in jacket.
(g) Calculate h_0.
(h) Calculate U.
(i) Repeat calculations (a) through (h) until the desired convergence is realized.
(j) Determine flows of cooling water at other time periods (and at other rates of heat release).
(k) Calculate total water flows per batch run using the trapezoid (or Simpson) method, i.e., integrate water rates over entire batch run.

(7) Economic calculations. Based on information in Table 2.

(a) Calculate total capital cost.
(b) Calculate water costs per batch run.
(c) Calculate electrical cost to operate agitator motor for batch run (the time of polymerization and horsepower of motor were both known).
(d) Calculate labor costs.
(e) Calculate other operating costs.
(f) Calculate total cost of polymerization.

(8) Evaluation of a given operating or design parameter. Any specific parameter can be evaluated at several levels or a combination of several parameters can also be evaluated.
(9) Optimization procedure. The optimum operating or design parameters can be determined for any combination of three parameters selected.

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