Designing Ion Exchange Systems

Part III: Rating The Resin

By C. F. Michaud

Designing an ion exchange system requires both the proper chemical design as well as the proper physical design considerations.

Parts I and II of this series offered assistance in establishing a starting point (the water analysis) and the end point (desired quality, resin selection) of our designing journey. Part III will help describe the vehicle (How am I going to get there?) by which we will travel from Point A to Point B.

The heart of an ion exchange system design lies in our ability to properly rate (determine capacity, leakage and regeneration scheme) the resin system. The economic consequences of having a "few" extra cubic feet of resin may cost tens of thousands of extra dollars in regenerant costs and waste neutralization over the life of the system. On the other hand, a "few" feet short might cause short runs or compromised quality that could lead to premature equipment failure down the road.

Proper design dictates conservative decision making, yet accurate performance projections. Being very conservative is not nearly as disastrous as being overly optimistic regarding a resin's capabilities. There is little risk of overdesign on a small D.I. system - say up to 10 gpm other than the risk of not getting the job because your design is too costly.

Equipment manufacturers differ in their approach to rating a resin. Consideration for complexity is proportional to the size of the system. What makes good sense at 300 gpm may be too costly a concept at 30 gpm. A practical solution for 100 gallons per day may be impractical at 100,000 gpd. The reason behind this thinking lies in the fact that the relative cost of instrumentation, controls and engineering becomes an increasingly large proportion of the expense of a smaller system. At some point in the scale up or scale down, one has to go more complex or simplify the design, rather than simply increase or decrease the diameter of the column.

The proper design of an ion exchange system will offer the customer the most economical design while conservatively assuring the quality of the effluent for three years.

The economics of a system design must look not only at the capital costs for the new equipment, but must factor in the operating costs as well. In addition, the op erators' skills at the customer's plant will influence ou choices in design. To set out to describe in detail all o the considerations for every type of ion exchange sys tem would be beyond the scope of these articles. There fore, we are going to choose one water composition an one desired effluent quality and design two system that differ only in respect to the complexity of th system.

Complexity vs. Simplicity

For this example we will use the same water analy sis as in Part II. (See Table 1)

Table 1

Ion	ppm as $CaCO_3$	Ion	ppm as $CaCO_3$
Ca^{++}	150	HCO_3^-	150
Mg^{++}	50	$SO_4^=$	50
Na^+	200	Cl^-	200
Total Cations	400		400
		Silica	35
		Total	435

Our influent water is low in turbidity and 75° F and onl 0.2 ppm of free chlorine. Our problem is to design a deionization system to satisfy boiler water make-up feed having the quality found in Table 2.

Each system will produce 200 gpm and operate 24 hrs/day. One system will economize on capital expendi ture, the other will economize on operating costs.

The Rating

We have a starting point of 23.4 grain/gallon cati ons, 25.4 gr/gal anions. We have approximately 38% Ca^{++}, 12% Mg^{++}, 40% Na^+, 35% Alkalinity, 57% FMA (free mineral acidity) and 8% silica. Our Cl^- ratio (Cl^-/FMA) is 0.8. These ratios must be determined in orde to determine the ion exchange capacities and correction factors.

Table 2

Ion	Max ppm as $CaCO_3$
Ca^{++}	< 0.05 ppm
Mg^{++}	< 0.05 ppm
Na^{+}	< 2.0 ppm
HCO_3^-	< 0.1 ppm
SO_4^{--}	< 0.05 ppm
Cl	< 0.1 ppm
Silva	< 0.05 ppm
Hardness	< 0.05 ppm
Specific Conductivity	15 micromhos/cm
pH	7.0 to 8.5

Our effluent quality suggests that no mixed bed polisher is required in that only 15 micromho is required. 15 micromho is approximately 3 ppm for SAC/SBA two bed or 6 ppm for SAC/WBA two bed. Since our effluent quality dictates a max of 0.1 alkalinity and low silica with a neutral or alkaline pH, our choice is SAC/SBA.

To convert specific conductivity (micromhos) to ppm as $CaCO_3$, divide your value by 4.99 for a two bed with SBA and by 2.53 for a two bed with WBA. To convert specific resistance (ohms) to conductance, divide 1,000,000 by your ohms. One micromho = 1 million ohms or 1 megohm. 0.1 micromhm is 10 meg.

There are times when we may have soft water available to use in regeneration. *Soft water or decationized water must be used for dilution and rinsing of caustic.* If soft water is not available, we can include a softener for regenerating the anion. OR, we can include the extra cation capacity into our cation design. If we include the softener, we can regenerate the cation (using raw water) and anion (using soft water) at the same time. This saves time, but adds to the expense. Soft water is, however, cheaper to produce than decationized water. If large quantities of water are required, and/or turn around time is short, use softened water. Otherwise, build the extra capacity into your cation unit. Regenerate the cation first, then your anion using decationized water from the SAC unit for backwashing, dilution and rinsing the anion unit.

Without having done much work, we have chosen to use a SAC and SBA. We do not have a highly oxidative condition (only 0.2 ppm Cl^-) nor hot water (75°F). Therefore, we will use a conventional SAC, such as Purolite C-100. For the SBA, we can use either a Type I (Purolite A-400) or Type II (Purolite A-300). Since our silica requirement can be met with either a Type I or II SBA, we will rate the system with a Type II because we know we will get higher capacity. Let's look at the simpler system first.

Simplicity

Our design demands 200 gpm for 24 hrs/day, a total of 2,880,000 gallons. Since a 24-hour operation is required, we will need a twin system with alternate operation. A single system could be considered if sufficient storage, 4 hours, can be provided. This would require nearly 50,000 gallons in this case. For our example, we will opt for the twin system. Also, we have opted to use decationized water for regeneration of the anion. First, we size the anion:

$$\text{Anion requirement: } (200\text{ gpm})\left(\frac{60\text{ min}}{\text{hr}}\right)\left(\frac{6\text{ hr}}{\text{cycle}}\right)\left(\frac{25.4\text{ grains}}{\text{gal}}\right) = 1{,}828{,}800\text{ grains}$$

As a starting point, we have selected a 6 hour cycle for each train. A Type II SBA (from the literature) would require 6 lbs of NaOH/ft to meet our 0.05 ppm effluent silica requirement with 75°F regeneration temperatures. Warming the caustic to 95°F would reduce our regenerant requirement to approximately 5 lbs/ft and a further reduction to 4 lbs/ft could be possible at 120°F. We would have to look at this potential cost savings in our design.

At 75°F and 6 lbs, our capacity (from the literature) is determined to be 19.3 kilograins per ft (K gr/ft^3). That capacity, however, is for new resin. We can expect a loss of at least 8-10%/year or 30% over three years. We therefore must build an "engineering" factor or allowance into our design to handle this decline in capacity. For a Type II, we'll use a 20% downgrade and rate the resin at 15.44 Kgr/ft.

To achieve a 1,828,800 grain/day removal, we need (1,828,800 / 15,400) = 118.45 ft^3 of SBA II resin. This fits nicely into an 84" diameter vessel with a cross section of 38.5 ft^2. (Use 119 ft^3).

This vessel will backwash at 2.7 gpm/ft^2 (104 gpm) for 10 minutes (1040 gallons). We will regenerate at 6 lbs NaOH at 0.5 gpm/ft^3 (60 gpm) using 4% NaOH (0.3481 lbs/gal) using 50% NaOH as our stock caustic (6.364 lbs/gal). Since we need a total of 6 lbs x 119 ft^3 = 714 lbs 100% NaOH per regeneration and our 4% contains 0.3481 lbs/gal, we therefore need (714 / 0.3481) = 2050 gallons of 4% NaOH. Our feed solution will be (714 / 6.364) = 112.2 gallons of 50% NaOH. Therefore, our dilution water is (2050 - 112) = 1938 gallons of decationized water to be supplied from the cation unit. In addition, another 1938 gallons will be supplied for the slow rinse cycle. We will use another 6000 gallons for the fast rinse (200 gpm for 30 minutes). Therefore, the total additional cation throughput must produce an extra (1040 + 1938 + 1938 + 6000) = 10,900 gallons.

(200 gpm)(360 min) =
72,000 gal + 10,900 =
82,900 gal/cycle
(82,900)(23.4 gr/gallon) =
1,939,860 grains/cycle

To meet our sodium leakage requirements of 2.0 ppm max requires that we regenerate our SAC with a minimum amount of acid to produce only 0.5% cation leakage. For H_2SO_4, this will require in excess of 10 lbs/ft^3 (10lbs/ft^3 will produce an effluent with an average Na^+ leakage of .95% or 3.8 ppm at 50% Na^+ and 35% Alkalinity). Best we look at using HCl.

Our literature indicates that 0.5% leakage can be obtained with about 10 lbs of HCl. This also gives us a capacity of 31.8 Kgr/ft^3. We will apply only a 10% down grade for this resin and rate it at 28.62 Kgr/ft^3.

To achieve our 1,939,860 grains/cycle cation removal, we therefore need (1,939,860 / 28,620) = or 69 ft^3 of SAC. (A viable option for some OEMs would be to go to counter-current regenerations with HCl at a level of 5 lbs using 80 ft^3 of SAC. With CC regen, our leakage would be less than 0.5 ppm Na^+ even at regen levels of 3 lbs/ft^3 or less.)

Our cation vessel will be a 66" diameter with 69 ft^3 of SAC. It will backwash at 142 gpm for 10 minutes. Regeneration with HCl will be with 6% HCl (1340 gallons). This assumes 10 lbs and 69 ft^3. Using a 30% HCl feed solution will require 240 gallons of acid and (1340 - 240) = 1100 gallons of dilution (raw) water. Total regen flow rate will be 45 gpm (assumes 30 min. contact time). Use 30 min. for slow rinse and 20 min. fast rinse. Total cation water required is (1420 + 1100 + 1100 + 4000) = 7620 gallons.

Table 3
Hydraulic Design Summary

	Cation	Anion
Flow	200 gpm	200 gpm
Cycle	6.9 hrs.	6.0 hrs.
Resin Vol.	69 ft^3	119 ft^3
Resin Type	SAC-8%	SBA-Type II
Regen. Level	10lbs HCl/ft^3	6lbs NaOH/ft^3
Regen. Flow	.65gpm/ft^3	.5gpm/ft^3
Contact Time	30 min	35 min
Service Flow/ft^3	2.9 gpm	1.7 gpm
Service Flow/ft^2	8.4 gpm	5.2 gpm
Tank Diameter	66 in.	84 in.
Bed Depth	35 in.	37 in.
Resin Rating/ft^3	28.62Kgr	15.44 Kgr

Neutral Regeneration Waste

An optimum design would allow the excess acid and caustic to neutralize each other. This is not always possible on the first try.

Our acid requirement for cation regeneration is 28.62 Kgr/ft^3 or 1.31 equivalents of (HCl) / liter. The equivalent weight of HCl is 36.5 gms. Therefore, we need:

eg. gms

$$\frac{(1.3\ eg)}{liter}\frac{(28.3\ l)}{ft^3}\frac{(60\ ft^3)}{cycle}\frac{(36.5\ gns\ HCl)}{eg}\frac{(1\ lb\ HCl)}{454\ gns} =$$

gms

206 lbs HCl/cycle eg.

In our example, we used 690 lbs of HCl. Therefore, the regenerant waste will contain (690 - 206) = 484 lbs of HCl that will have to be neutralized.

Our caustic requirements for anion regeneration is 15.44 Kgr/ft^3 or .71 equivalents of NaOH / liter of resin. (21.8 Kgr /ft^3 as $CaCO_3$ capacity is equal to 1.0 meg/ml or 1.0 eq / l). The equivalent weight of NaOH is 40 gms. Therefore, we need:

$$\frac{(.71\ eg)}{liter}\frac{(28.3\ l)}{ft^3}\frac{(119\ ft^3)}{cycle}\frac{(40\ gns\ NaOH)}{eg}\frac{(1\ lb\ NaOH)}{454\ gns} =$$

211 lbs NaOH/cycle

Since we used 714 lbs of NaOH to regen our anion, we have an excess of (714 - 211) = 503 lbs of NaOH. It takes 40 lbs of NaOH to neutralize 36.5 lbs of HCl. To neutralize 484 lbs of HCl will require:

$$484 \times \frac{40}{36.5} = 530 \text{ lbs NaOH}$$

In our example, we will need to add (530 - 503) =27 lbs of NaOH to our mixed regen effluent. Our decationized regen water, being acidic, will actually neutralize an additional 13 lbs of NaOH. Our total NaOH requirement is therefore 40 lbs. It would suffice to increase our anion regeneration by .33 lbs/ft (.33 x 119 ft = 39.6 lbs) and take advantage of the regenerant value. Also note that most plants would use lime or soda ash as lower cost neutralizers.

Our simple system is illustrated in the diagram on page 68.

Earlier, we mentioned a potential savings on caustic by warming the regenerant. A savings of 1 lb of NaOH / ft^3 times 4 cycles/day is equal to 173,740 lbs of NaOH/yr. At $.30/lb, this is equal to $52,000/yr. At 2 lbs/ft^3, the savings is over $100,000/yr.

Our regenerant costs are approximately:

Anion:

(6)(119)(4)(365)($0.30/lb) = $312,700/yr for NaOH

Cation:

(10)(69)(4)(365)($0.30/lb) = 302,200/yr for HCl
TOTAL = $614,900/yr
OR: $5.85/1000 gallons

Complexity

Let's now compare these figures to that of a more complex 200 gpm continuous system on the same water supply.

Since we need continuous production, we obviously need two complete systems. One will carry the full load while the other regenerates. Does each have to operate 12 hours? Absolutely not. It only takes about two hours to regenerate, four if they are done sequentially. A six-hour service cycle allows for some error.

Anion Design

Our anion would produce 200 gpm for six hours or 72,000 gallons (1,838,800 grains).

We have moderate alkalinity (35%), moderate FMA (57%) and not a lot of silica (8%). Let's look at using a degassifier followed by a weak base anion, then a strong base anion. We'll use cation effluent to regen the anion and assume 100 gallons / ft^3 for the anion regenerants. Since we'll be increasing the percent silica that the SBA sees, we'll use a Type I resin such as Purolite A-400.

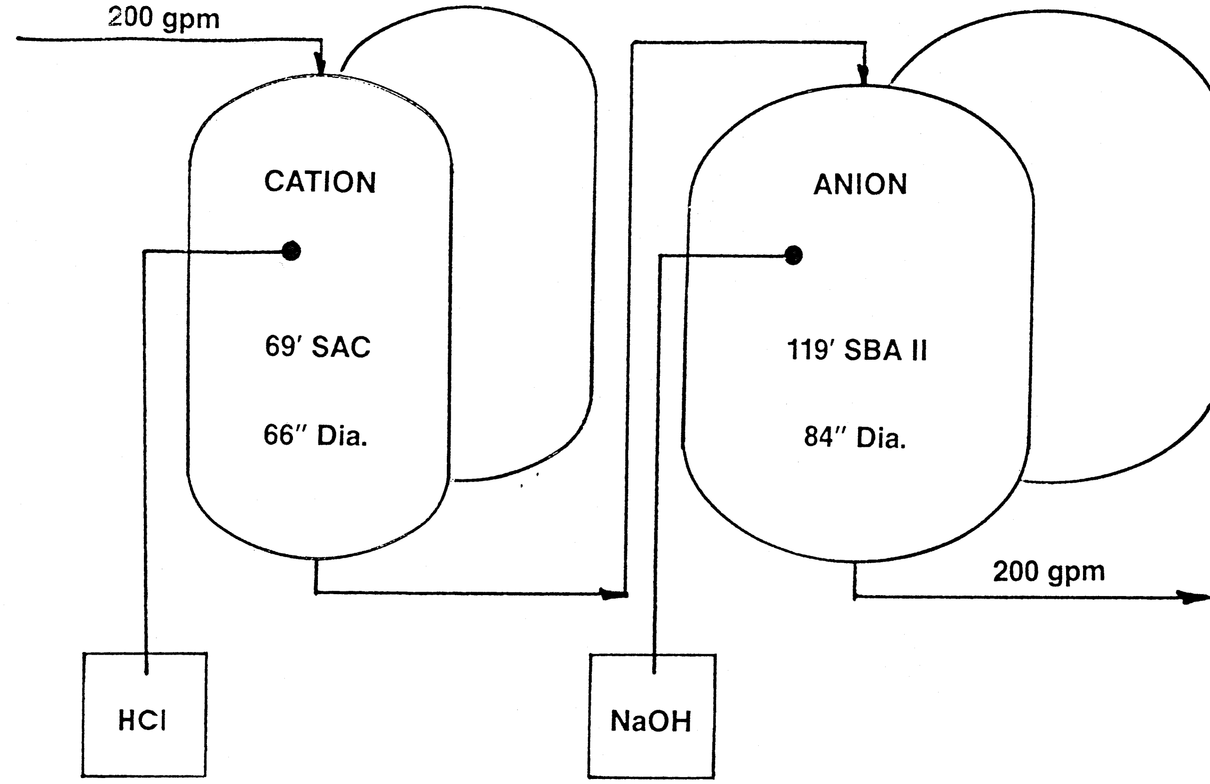


Figure 1

Exiting the cation, all of our anion constituents exist as their acid forms. That means that HCO_3^- is really CO_2 plus water. An atmospheric degassifier will reduce that CO_2 to 15-20 parts. Let's use 20.

Our weak base now sees some CO_2 and silica along with 250 parts of FMA. Although the WBA can remove FMA very effectively, let's assume 95%. Our SBA therefore is going to see the following:

	ppm as $CaCO_3$	%
HCO_3^-	20	30%
Cl^-	12.5	18%
Silica	35	52%
TOTAL	67.5 = 3.95 grains	

To treat over 200 gpm for 6 hours requires the removal of 284,000 grains. Since the silica is over 50% of the anion loading, we'll use 120°F water for regeneration. The literature tells us we can expect to see an average silica leakage of 0.08 ppm at 6 lbs and 0.04 ppm at 8 lbs. We'll use 8 lbs. At this level we can expect 18.0 Kgr/ft^3. Because our chloride ratio is high (1.0), we must use a downgrading factor of .9 and an engineering factor of .85. This gives a corrected capacity of 18.0 x .85 x .9 = 13.77 Kgr. To remove 284,400 grains requires only 20.7 ft of anion. We have a dilemma. Prudent hydraulic design dictates that we would have at least a 20 ft^2 vessel and 30" bed depth. That's 50 ft^3 of resin or enough to run for 15 hours. To operate for 16 hours would require 55 ft^3 of SBA.

Let's move on. For our WBA selection, we will select a durable macro such as Purolite A-104. The WBA can be regenerated with the spent caustic from the SBA and, according to the literature, we can expect 22.0 Kgr/ft^3 capacity for this WBA.

The WBA sees the following ionic loading:

250 ppm of FMA = 14.62 grains/gal.

To operate for 6 hours would require : 14.62 x 200 x 60 x 6 = 1,052,640 grains removal which takes 47.8 ft of resin. To operate for 16 hours would take 127.6 ft WBA. A 72" diameter vessel with 4.5' bed depth will do.

To regenerate our WBA requires 3 lbs/ft of NaOH or 384 lbs. Our SBA with a rated capacity of 13.77 uses only 1.6 lbs/ft^3. At an 8 lb regen level, we have an excess caustic level of 352 lbs (8 - 1.6) x 55. We must therefore increase our NaOH level by .6 lbs/ft^3 in order to completely regenerate the WBA with spent NaOH. (We need not be afraid to increase our NaOH level. Recall that in the prior example, we were nearly neutral. Since we are using "spent" caustic to regen our WBA we will have to add alkali to neutralize our final waste. So let's go ahead and increase the caustic level to 10 lbs.)

By increasing our caustic to 10 lbs, we increase our capacity on the SBA slightly, about 5%, to 14.54 Kgr/ft^3. This reduces our resin requirements to 53 ft^3. Excess NaOH from SBA regen becomes 442 lbs/cycle of which 320 is used by the WBA, leaving 122 lbs to neutralize any excess HCl.

Cation Design

We could apply the same design logic to the use of the cation by using WAC followed by SAC. The WAC would remove 150 ppm of hardness and use spent acid to regen. A WAC gel resin such as Purolite C-105 would exhibit a capacity of 60 Kgr/ft^3 with a hardness to alkalinity ratio of > 1.0 at 75°F.

Since our anion system contains 181 ft^3, we will assume 100 gal/ft decationized water for regeneration (181 x 100) = 18100 gallons. Our cations must therefore treat 210,100 gallons. This means that our WAC must remove 150 ppm of hardness (8.77 gr/gal) or 1,842,600 grains. At 60 Kgr/ft^3, this will require 31 ft^3 of WAC.

Our SAC must therefore remove 3,072,000 grains. However, we've changed our water composition. Now it's 80% sodium. However, we can now tolerate 0.8% cation leakage as sodium and 10 lbs HCl/ft will still make it. Our cation capacity at 10 lbs HCl (with 80% Na^+) is 34 Kgr/ft^3. Applying a .9 engineering factor gives us 30.6 Kgr/ft^3. Therefore, we need 100 ft^3 of SAC.

Stratified Beds

It is possible to combine both the WBA and SBA as well as the WAC and SAC in the same bed. Since prudent design would dictate a max flow/ft^2 of 8 gpm, we need a 72" diameter vessel. This has a surface area of 28.26 ft^2. The WAC portion of a stratified bed should be at least 18" deep. Therefore, we need about 45 ft.

At a flow rate of (200 / 45) = 4.45 gpm/ft^3, our WAC capacity is down graded to about 41 Kgr/ft^3. Therefore, we can still remove the (45 x 41,000 = 1,845,000) grains per cycle. Our SAC will have a 3.5' bed depth.

For maximum efficiency, the stratified bed will regenerate upflow. 4.7 lbs of HCl is required to regenerate 1 ft^3 of WAC, (plus a 10% excess.) Therefore, our SAC must supply 233 lbs of excess acid. Since we are using 10 lbs HCl/ft^3 = (1000 lbs for the SAC) and are utilizing 3.19 lbs/ft (319 lbs total), we have total excess HCl of (1000 - 319 - 233) = 448 lbs. Our excess caustic will neutralize 111 lbs leaving 337 lbs excess HCl. To neutralize this will require 258 lbs of lime (Ca).

Our stratified anion bed is also in a 72" vessel with 128 ft^3 of WBA sitting atop 55 ft^3 of SBA. Regen is upflow to take advantage of the regen efficiency of the WBA using spent caustic from the SBA. Our system now looks something like one illustrated in Figure 2.

Operating Cost Comparisons

Our more complex system uses 550 lbs of caustic per cycle and regenerates 1 1/2 times per day. Our cation uses 1000 lbs of HCl per cycle and also regens 1 1/2 times/day. Our approximate regenerant costs are:

Anion:
(550)(1.5)(365)($.30.lb) = $90,300/yr for NaOH

Cation:
(1000)(1.5)(365)($.30.lb) = $164,400/hr for HCl

Neutralize:
(258 lbs lime)(1.5)(365)($.05/lb) = $7100/yr for CaO

TOTAL = $261,700/gr
OR: $2.49/1000 gallons

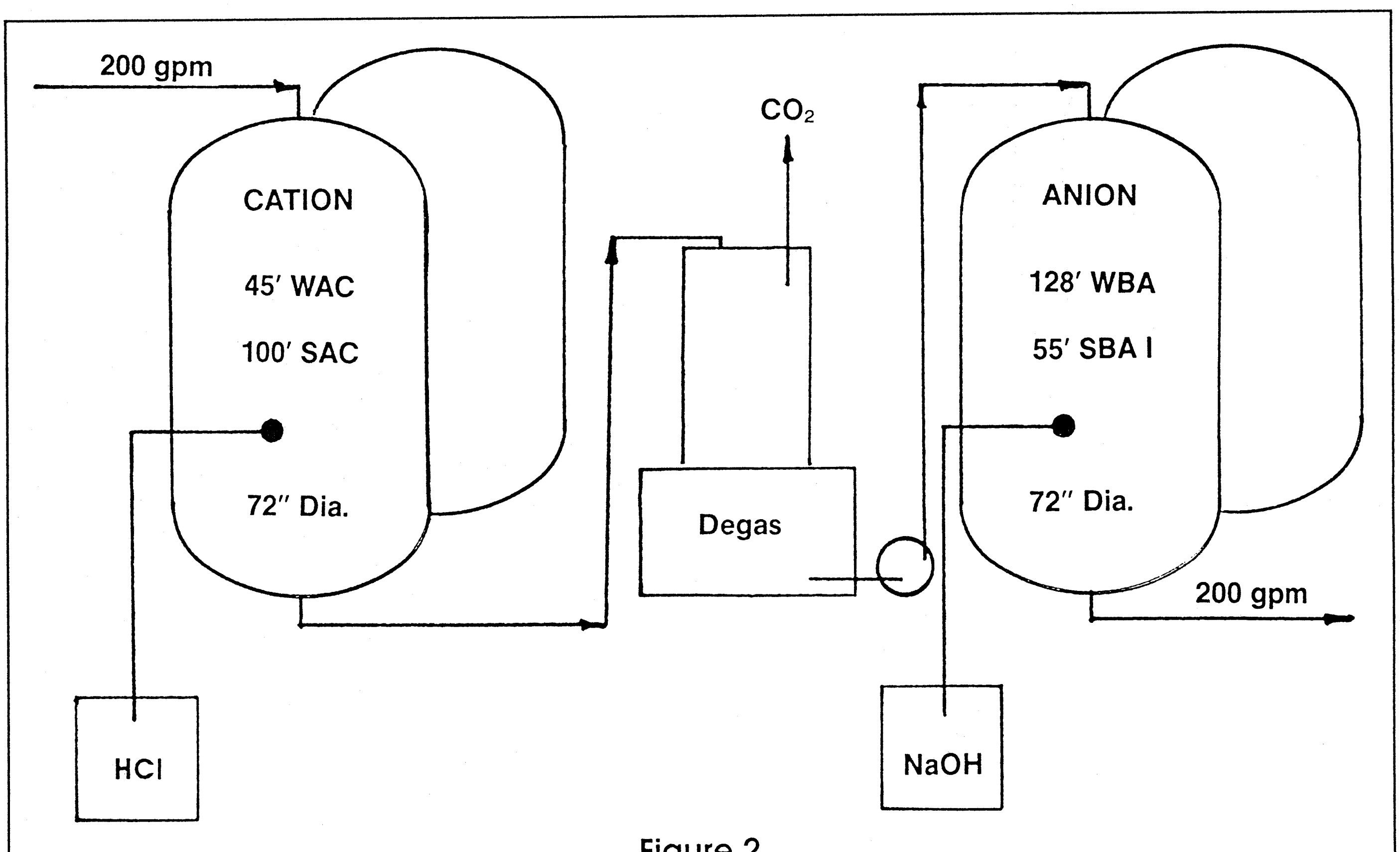


Figure 2

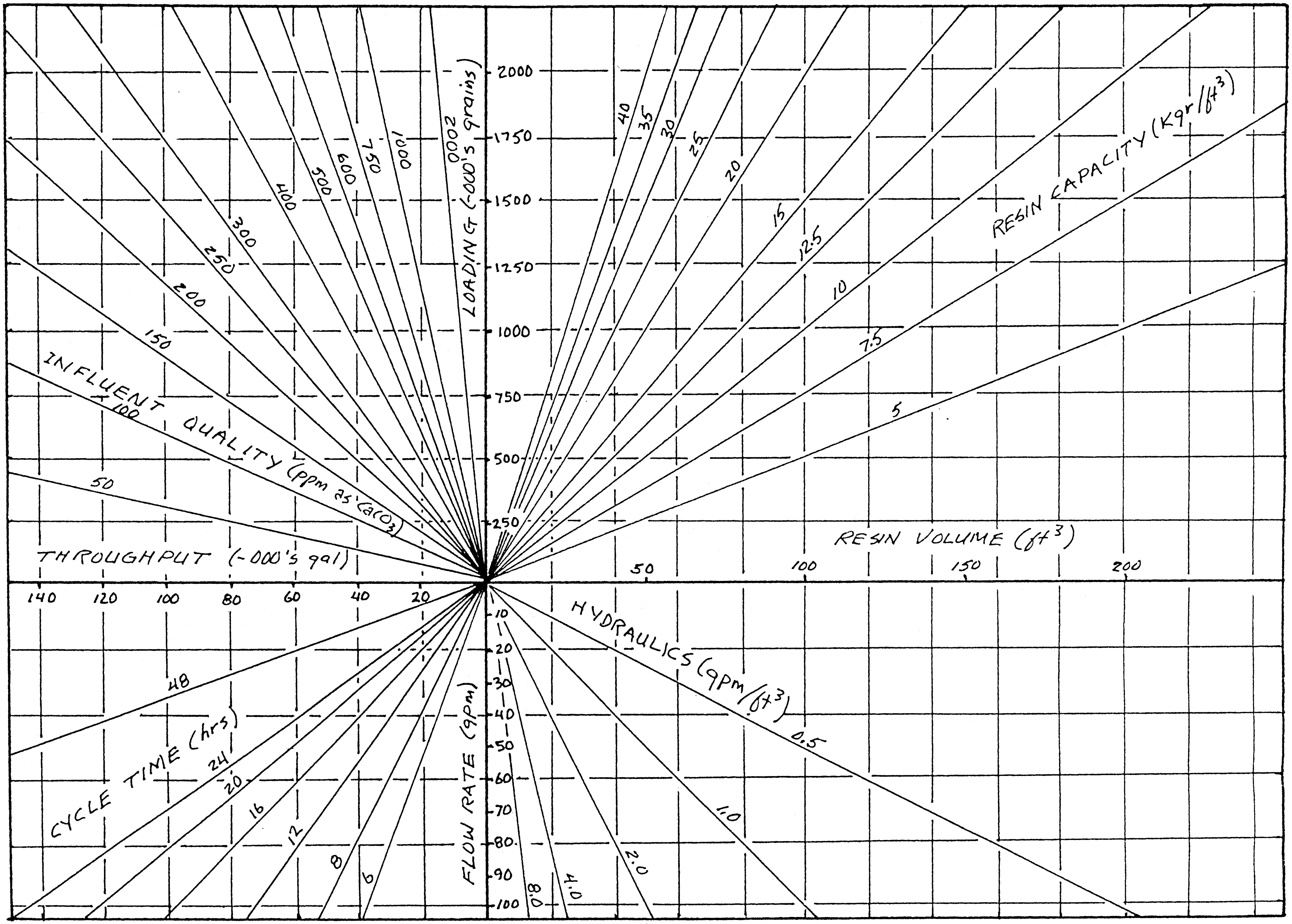
Table 4
Hydraulic Design Summary

	WACation	SACation	WBAnion	SBAnion
Flow	200 gpm	200 gpm	200 gpm	200 gpm
Cycle	17.5 hrs.	17.5 hrs.	16 hrs.`	16 hrs.
Resin Volume	45 ft^3	100 ft^3	128 ft^3	55 ft^3
Resin Type	WAC-gel	SAC-8%	WBA-MR	SBA - I
Regen Level	*15 lbs HCl	10 lbs HCl	*3.5 lbs NaOH	10 lbs NaOH
Regen Flow/f^{t3}	1.44 gpm	.65 gpm	.27 gpm	.65 gpm
Contact Time	30 min	30 min	45 min	45 min
Service Flow/ft^3	4.4 gpm	2.0 gpm	1.6 gpm	3.6 gpm
Service Flow/ft^2	7.1 gpm	7.1 gpm	7.1 gpm	7.1 gpm
Tank Diameter	72"	72"	72"	72"
Bed Diameter	19"	42"	54"	24"
Resin Rating/ft^3	41.0 Kgr	30.6 Kgr	32.0 Kgr	14.5 Kgr

* equivalent amount of regenerant passing through

Table 5
Capacity

Resin	Regenerant	4 lbs	6 lbs	8 lbs	10 lbs
SAC	HCl	20.0	25.2	28.4	30.0
SAC	H_2SO_4	11.3	12.9	15.8	17.0
SAC	NaCl	14.0	20.0	24.0	27.0
WAC	HCl	50.0	NA	NA	NA
WAC	H_2SO_4	50.0	NA	NA	NA
SBA I	NaOH	10.5	12.0	13.3	14.0
SBA II	NaOH	15.3	16.4	17.3	18.2
SBA II	NaCl	6.0	10.0	14.0	16.0
WBA	NaOH	20.0	NA	NA	NA
WBA	Na_2CO_3	20.0	NA	NA	NA



LOADING (-000's grains)
2000
1750
1500
1250
1000
750
500
250
RESIN CAPACITY (Kgr/ft3)
40
35
30
25
20
15
12.5
10
7.5
5
RESIN VOLUME (ft3)
50
100
150
200
INFLUENT QUALITY (ppm as CaCO3)
2000
1000
750
600
500
400
300
250
200
150
100
50
THROUGHPUT (-000's gal)
140
120
100
80
60
40
20
HYDRAULICS (gpm/ft3)
0.5
1.0
2.0
4.0
8.0
FLOW RATE (gpm)
10
20
30
40
50
60
70
80
90
100
CYCLE TIME (hrs)
48
24
20
16
12
8
6

Figure 3

Summary

The actual design of an ion exchange demineralizer can be very complex. There are many factors that must be considered. As the cost comparison above points out, the simpler systems chemical demand costs over $350,000/year more to supply. This is partially offset by more expensive and larger equipment, controls and engineering. Nonetheless, the complex system will end up costing the customer less over its life which is how prudent buyers evaluate the true system cost.

Estimating Your Needs

Although there are many factors that make up the "rating" of an ion exchange resin, we can "ball park" our designs using Table 5 and the nomograph found in Figure 3. The capacities listed are appropriately downgraded in keeping with conservative practice and should be used only with the nomograph for making estimates.

Instructions for Use of Nomograph

1. Follow flow rate down to desired value.

2. Follow line left to desired cycle time.

3. Follow line straight up to intersect influent quality.

4. Follow line right to rated capacity (Use Table 5).

5. Bring line straight down to find resin volume.

6. Continue line down to cross original flow rate line. This will give you hydraulic flow rate. Target is 1 - 2.5 gpm ft^3.

NOTE: Multiples may be used at any point, i.e. for 200 gpm, use 100 gpm and multiply all other values by 2.

Example: 50 gpm softener, 20 hr cycle, 200 ppm hardness, regen with 8 lbs NaCl.
Answer: From Table 5, capacity is 24 Kgr/ft^3.

1. Follow 50 gpm left to 20 hrs.
2. Follow intersection up to 200 ppm. Note cycle is 60,000 gallons.
3. Follow right to 24 Kgr. Note total loading is 70,200 grains.
4. Follow line down to cross resin volume at 30 ft^3.
5. Hydraulics are 1.7 gpm/ft^3.
6. Like we said, it's only an estimate.

The November issue will include a Trouble Shooting Guide and a Glossary.

About the Author

Chubb Michaud is a chemical engineering graduate of the University of Maine and has been involved in ion exchange water treatment for over 10 years. He is founder of Systematix Co. of Brea, CA and represents Purolite Co. in the Western states.