TABLE I: ANALYSES OF WELL WATERS IN PARTS PER MILLION

	1935, ppm								
Well No	747	748	750	1333	1334	1335	1336	1468	1469
Capacity, gpm	100(1)	120	300	170	173	250	95	100	170
pH value	5.9	5.4	5.9	4.1	4.3	4.9	4.0	5.1	4.6
Bicarbonate (HCO_3)	4	6	4	5	5	5	0	4	3
Sulphate (SO_4)	4	14	7	43	14	7	65	1	4
Chloride (Cl)	5	16	7	87	60	32	173	3	4
Silica (SiO_2)	9	7	8	9	6	9	10	7	7
Iron (Fe)	2	0.3	1	0	3	0.8	8	0.2	2
Aluminum (Al)	0	Tr	0	Tr	Tr	Tr	Tr	Tr	0
Manganese (Mn)	0	Tr	0	0.1	Tr	0	0.3	0	0
Calcium (Ca)	2	6	3	8	6	4	15	0.7	0.2
Magnesium (Mg)	1	4	2	6	5	2	12	0.7	0.1
Sodium (Na) Calc.	2	4	3	57	36	15	103	2	3
Total Dissolved Solids(2)	26	55	33	216	140	71	385	15	22
Suspended Material	2	0	2	Tr	0	Tr	3	Tr	Tr

	October 30, 1944, ppm				
Well No	1333	1335	3700	3929(4)	4241(4)
Cap. gpm	50	75	300	300	300
pH value	5.8	5.5	5.4	4.0	5.3
CO_2	0	0	0	0	0
HCO_3	0	4	4	2	4
SO_4	28	22	3	135	37
Cl	22	12	13	35	Tr
SiO_2(3)	13	10	39	10	6
SiO_2(4)	8	8	6	9	3
Fe(4)	1.1	0.3	0.5	0.9	0.1
Al(4)	2	1	2	2	1
Al(3)	0	0	0	0.2	0
Ca(4)	4	3	4	27	3
Mg(4)	1	1	1	0.5	0.3
$CaCO_3$(4)	14	13	14	70	9
Tot D S(5)	91	63	38	264	46
Sus Mat	10-12(7)	2-4(7)	74(7)	6-8(7)	2-4(7)

(1) Estimated
(2) Sum of the components as determined by chemical analysis
(3) Drilled 25 ft away from old well No 1468
(4) Not used in boilers
(5) Total in water under test
(6) Filtered, total after filtration
(7) Estimated for water under test
(8) X-ray examination of suspended material: Calcite, quartz and aluminum silicate; probably sillimanite ($Al_2O_3.SiO_2$)
(9) Sum of components determined by analysis

water in the hot process would dissolve aluminum silicate and carry aluminum to the boilers.

Thus the most important step in preparation of makeup water was removal of suspended material by coagulation, preferably with a ferric-iron salt, followed by filtration at normal water temperature and at a pH of no more than 7 or 8 as an absolute maximum.

Following clarification, we considered three procedures, (1) base exchange (2) hot-process treatment with phosphate (3) 2-stage hot-process treatment for silica removal and phosphate softening.

Objections to the base-exchange system were as follows:

(1) It did not fit well into the water system. Before reaching the boiler house, make-up water passes through heat-exchange equipment in plant process. Here it attains a temperature 160 to 170 F. This meant base-exchange softening would have to occur before the makeup water was used for cooling.

(2) Further changes in well-water composition might mean use of more city water or poorer quality water such as that from Well 3929 (Table I). This development would pose two problems: need for increase in base-exchange capacity; possible need for hydrogen-cycle exchangers in addition to sodium cycle for adjustment of alkalinity if city-water consumption increased.

(3) There would be no provision for silica removal. While additional removal beyond clarification might not be necessary, the specter of changing well-water composition dictated otherwise.

(4) Good wells were being pumped as heavily as possible. Any increase in water use at the power plant meant greater consumption of city water. Thus the considerable quantity of water needed for rinsing base-exchange units was objectionable.

Single-stage hot-process phosphate softening would reduce treated-water hardness to a low point and minimize amount of sludge developed in boiler water. But should silica removal prove necessary, this system of softening would provide therefor. Further, if makeup water changes made preliminary hot, lime-soda softening desirable, the system would not be suitable.

The most logical system considered was 2-stage hot-process treatment. The first-stage hot-process unit would give some silica removal; the second stage, softening with phosphate. Should well water go bad and more city water or harder well water be needed, preliminary lime-and-soda-ash softening could be accomplished in the first stage and additional phosphate softening in the second.

The equipment selected is complete and provides for any eventuality except intolerable increases in concentrations of ions—sulphate and chloride.

Equipment Installed. Clarifying equipment, Fig. 2, includes a 40,000 gal tank where well water collects and mixes with enough city water to meet plant demands. This mixture is pumped through a meter and control valve to the 63,000-gph clarifier, where it is flocculated for removal of suspended material. Positive-displacement pumps feed ferric chloride and caustic soda, the flocculating agents, from separated dissolving tanks to the clarifier. Ferric chloride is automatically proportioned to the flow of water to the clarifier, and caustic-soda feed is controlled by pH value of clarifier effluent.

From the launders of the clarifier flocculated and settled water flows to a clearwell of 4000-gal effective capacity. From here it is pumped through eight 8-ft dia, vertical, pressure filters operating in parallel. Filtering material is Anthrafilt.

Filters are backwashed with filtered water and the backwash returns to the inlet section of the clarifier. No pumps are used; the pressure in the filter-effluent line serves to force water at controlled rate of flow through the filter beds.

Filtered water goes to heat-exchange equipment in plant process, and there it returns to a 40,000-gal tank in the boiler house at 160 to 170 F. Clarified, filtered water under control of a float-operated valve maintains a constant level in the tank.

From this tank filtered water is pumped through a meter, an air-actuated pilot-operated inlet-water controller and a vent condenser to the spray-type heater on the 63,000-gph first-stage hot-process unit, Fig. 3. The plant employs thoroughfare deaerator. Steam vented from the deaerator heater heats makeup water to 220 F in the first-stage hot-process spray heater. Noncondensable gases go from the vent condenser.

Silica removal, for present makeup water, constitutes the prime function of this first-stage hot-process unit. Caustic-soda solution and dolomitic-lime slurry are used in the treatment are mixed in separate chemical tanks. They are then pumped into the sedimentation tank in proportion to the flow of water. Recirculation of sludge helps sedimentation and silica removal.

TABLE II; TREATED-WATER CONDITIONS CLARIFIER-FILTER EFFLUENT, 1947

Date	pH value	Hardness $CaCO_3$, ppm	Silica SiO_2, ppm	Iron Fe, ppm	Aluminum Al, ppm	Turbidity ppm
12/5	7.4	29	8	Neg	Neg	Neg

	First-Stage Hot-Process					Second-Stage Hot-Process					
Date	A(1) reading, ml	MO(2) reading, ml	Hardness $CaCO_3$, ppm	Silica SiO_2, ppm	pH value	A(1) reading, ml	MO(2) reading, ml	Hardness $CaCO_3$, ppm	Phosphate PO_4, ppm	Silica SiO_2, ppm	Chloride Cl, ppm	pH value
11/1	—	—	—	—	—	1.6	3.6	9	16	3	17	10.0
11/2	1.6	3.2	20	3	10.2	1.5	3.6	7	10	3	20	10.2
11/3	1.0	3.4	20	3	9.6	1.0	3.6	7	3	4	23	9.6
11/4	1.1	3.1	20	4	9.7	1.1	3.4	7	6	3	17	9.6
11/5	0.8	3.6	26	4	9.3	0.9	3.0	5	6	4	14	9.3

(1) Ml N/20 acid per 100 ml sample, phenolphthalein indicator. (2) Ml N/20 acid per 100 ml sample, methyl-orange indicator.

Settled water from the uptake cone of the silica-removal unit passes to the 63,000-gph second-stage hot-process unit for further treatment with anhydrous disodium phosphate. This unit operates flooded. Disodium phosphate solution is pumped from a dissolving tank to the sedimentation tank, in proportion to water flow. Phosphate-treated water flows from the uptake cone of the phosphate softener filters mentioned above.

Two 600-gpm pumps and rate-of-flow controllers handle filter backwashing with softened, filtered water. Backwash returns to the top of the phosphate softener.

Condensate returns to a collection tank at a rate of about 8000 lb per hr. From there it is pumped to the condensate compartment of the deaerating heater. Plans are progressing for recovery of about 35,000 lb per hr of plant condensate.

The thoroughfare deaerator is designed to heat 720,000 lb per hr of feedwater to a minimum of 220 F with 10-psig steam. Oxygen concentration of effluent is less than 0.03 ml per liter by the Winkler test.

Operating Results. Only one major operating difficulty has developed. Water quality has not been affected seriously but the operators have had to be extra vigilant.

The clarifier-filter installation where aluminum- and silica-rich suspended matter is removed from the well-city water mixture is the heart of the treating system. Floc formation has been good with as little ferric chloride as 6 to 8.5 ppm, but normally 15 to 20 ppm; enough caustic soda to hold pH close to 7 is needed.

At flow rates approaching the capacity of the equipment the slurry level has been too high and then additional load has resulted in carryover of an excessive amount of floc to clearwell and filters. This has happened particularly in the morning during filter backwashing and chemical-tank filling. The alert action of the operators has staved off trouble and kept the filters from being seriously fouled. They have been bypassing raw water into the clarifier launders, which reduces flow through the equipment and gives the slurry level a chance to drop. In spite of this bypassing procedure filtered-quality has not suffered because filter effluent contains negligible amounts of iron, aluminum and suspended material.

Laboratory tests indicate that weighting ferric-hydroxide floc with a few ppm of activated silica or bentonic clay materially increases settling rate. It is believed this procedure may solve the difficulty described above.

Silica removal and some softening with hydrated dolomitic lime and caustic soda at a pH close to 10.0 takes place in the first-stage hot-process softener. Chemical consumption varies because of irregularities in composition of well-city water mixture. Recent figures over a 15-day period are approximately 40-ppm dolomitic lime and 40-ppm caustic soda. Since silica concentration of clarified, filtered water is not high, removal is limited to a few ppm. It is arbitrarily based on holding concentration in boiler water to about 30 to 40 ppm with blowdown between 5 and 10% of makeup.

Anhydrous disodium phosphate at a pH in the neighborhood of 10 gives final hardness removal in the second-stage hot-process softener. Phosphate consumption changes with variations in hardness and suspended material present in the first-stage effluent. For a time it ran about 40 ppm, but the recent 15-day period averaged high, almost 55 ppm.

Hardness of the phosphate-softener effluent runs higher than expected, normally between 5 and 10 ppm as calcium carbonate. While efforts will be continued to reduce hardness as well as phosphate consumption the condition is not alarming. Sludge content of the boiler water is not high or troublesome.

Table II shows typical water conditions for a week in Dec 1947. Turbidity, and iron and aluminum concentrations of the clarifier-filter effluent were negligible. Hardness as calcium carbonate was 29 ppm; silica concentration, 8 ppm.

Hardness of the first-stage effluent ran about one-third lower than that of the clarifier discharge. Silica concentration was reduced 50% or more even though pH value dropped as low as 9.3 toward the end of the week.

Silica concentration did not change in the second-stage effluent. Hardness was normal, between 5 and 10 ppm but, strangely, it was higher during the first part of the week when phosphate concentration and pH were higher. Explanation is not readily apparent since filtered, second-stage effluent is normally free of suspended sludge.

A word about chemical costs. Some variation can be expected. During the first 15 days of 1948 chemical costs per million lb of treated water amounted to $5.70. This is 4.8¢ per 1000 gal.

Boiler Results. The new 240,000-lb-per-hr steam generator has been on the line steadily since Feb 1947 and, by Feb 1948, has generated 1,512,729,000 lb of steam at an average rating of 230,000 lb per hr, peak rating 300,000 lb per hr.

There have been no tube losses or development of analcite scale in the older 125,000-lb-per-hr boilers. Occasional inspections have shown nothing but a little watery sludge.

Using the internal condition of the older boilers as an index of conditions in the new unit is a unique feature—particularly so in view of past experience with the older boilers and the fact that heat release per cu ft of furnace volume is considerably greater therein than in the new boiler.