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FINAL REPORT

MARTIN VORUM
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STEVEN W. GOERING
EUGENE A. FRITZLER

February 1978

Coury and Associates, Inc.
7400 West 14th Avenue, Suite 2, Lakewood, Colorado 80214

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Idaho Operations Office, Division of Geothermal Energy
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ABSTRACT

Information on the geothermal resources of the San Luis Valley, Colorado, has been gathered and reviewed and a preliminary, quantitative assessment of the magnitude and quality of resources present was carried out. Complete process designs were developed for the processes of producing crystal sugar from beets and for malting barley for use in the brewing industry, in each case adapting the processes to use a 302°F geothermal water supply as the main process energy source. A parametric design analysis was performed for a major pipeline to be used to ship geothermal water, and thus deliver its heat, out of the San Luis Valley to three major Colorado cities along the eastern threshhold of the Rocky Mountains. Cost estimates for capital equipment and energy utilization are presented.

The analyses of the two process applications indicate favorable economics for conversion and operation as geothermally-heated plants. A major geothermal water pipeline for this region is seriously limited on achievement of the economy of scale by the physical absence of significant demand for heat energy. Finally, the development and utilization of Colorado's San Luis Valley geothermal groundwaters hold the potential to contribute to the prudent and beneficial management of that area's natural water resources systems.

LIST OF FIGURES

LIST OF TABLES

I. BACKGROUND

A. General Objectives

This study of the geothermal resources of the San Luis Valley, Colorado, and representative industrial and commercial alternatives for the utilization of these resources, was conducted for three major purposes. First, the disparate data on the geology and hydrology of the San Luis Valley was identified and reviewed, and it served as the basis for an estimate--the most comprehensive to date--of the quantity, quality, and accessibility of the Valley's hydrothermal resources. Second, two energy-intensive industrial processes, presently found operating in various parts of Colorado on indigenous food crops, were each considered in a detailed process design evaluation in order to determine the comparative economics of basing plant designs on the use of a geothermal energy source, as opposed to the use of standard, fossil-fuel energy sources. Third, an estimate was developed of the potential present and future markets for energy, specifically for geothermal energy, in the regional vicinity of the San Luis Valley. This estimated market potential was then used in evaluating the feasibility for a major pipeline delivery system.

In the most general terms, however, the goal of this project is to develop detailed, pertinent information that will serve in generating the simple momentum of awareness and interest in a significant alternative energy resource. Specifically, this work is meant to, first, provide commercial developers with the means to rapidly assess their own applications concepts, and second, to quantitatively determine the potential impact of system design parameters on the economic and technical viabilities of such concepts. Many studies conducted to date have gone only as far as to perform rather superficial evaluations of the applicability of geothermally-derived energy to domestic and commercial space heating, industrial process heating, and varied uses in agriculture, aquaculture, and animal husbandry. In order to bring the concepts for use of geothermal heat to fruition at significant levels in our total consumption of energy, it is necessary to make detailed application analyses as soon as possible. The sort of evaluation undertaken in this project still must be carried further in terms of design detail and resource evaluation to satisfy the basic conservative posture of business and industry towards innovation.

B. Subjects for the Present Study

The hydrothermal resources of the San Luis Valley are a focal point in our analysis. Although the geology and hydrology of the Valley have been repeatedly dissected and monitored for most of this century, much of the work has remained uncollected and is available only in disjointed bits and pieces. Especially little is known about the Valley sediments below the upper 2,000 feet of strata. In addition, these various geological studies have presented somewhat contradictory interpretations of common data. Therefore, information from these sources has been reviewed and evaluated, and the findings are summarized in this report.

The designs of two agri-industrial processes were developed in detail, specifically modified to derive all possible process heat from a 302°F geothermal water source. The processes are the production of crystalline sugar from sugar beets and the production of brewer's malt from barley.

Both beet and barley farming, as well as malting and sugar production, are major Colorado industries. As of 1974, there were ten beet sugar plants in Colorado out of some 52 plants in the continental U.S. However, in 1977, four Colorado plants were moth-balled because of economic problems in the sugar industry as a whole. Barley, meanwhile, is the largest-acreage crop in the San Luis Valley, being planted on somewhat more than 30 percent of the Valley's arable land. The Coors Company of Golden, Colorado, has plant facilities near Denver for conversion of barley to malt sufficient to supply all their beer brewing needs, and annually contracts for about half of the total San Luis Valley barley crop.

With few exceptions, geothermal resources in the U.S. are found many miles from major centers of population, and hence, removed from their best potential domestic, commercial, and industrial markets. Additionally, much of the known resources are not of high enough temperatures to be used economically solely for electrical generation. A number of prior studies to date have considered pipelining geothermal water to nearby, small communities, but this severely limits the consumer base for such a highly capital intensive effort. Near-future use of significant amounts of geothermal energy, then, depends on transporting geothermal fluids to major population centers; developing significant user facilities at or near remote resources is a very long-term proposition. A parametric approach to pipeline design and cost estimation was followed in this project, considering a line with variable diameters from 2 feet through 5 feet. The system envisioned would be run from the San Luis Valley, over the Sangre de Cristo Range in the eastern Rocky Mountains, and thence to a modest metropolitan triangle comprising three cities: Canon City, Colorado Springs, and Pueblo. This area includes in excess of 300,000 inhabitants, is industrialized, has experienced strong population and economic growth for more than a decade, and plans to judiciously accommodate further growth predicted generally for Colorado and other Western States. Demographic data for the three-city complex has been used to estimate potential heat energy demands, and thereby assign some level of confidence to the practicability of the various pipeline capacities.

C. Investigative Procedures

The work performed on this project was conducted essentially according to the division of subjects outlined in the previous section. That is not to say that the subjects were not considered with respect to possible interrelationships. On a continuing basis, our findings and current questions were reviewed comparatively. This served to maintain a coordinated, directed effort, as well as highlighting common problem areas needing further attention.

The analysis of data on the geology and hydrology of the San Luis Valley, and estimation of the geothermal resources were performed by Dr. Richard W. Davis, a consulting geologist specializing in hydrology. A broad range of data sources were accessed, including well and spring analyses, drilling records from both hydrocarbon and water wells, and collected results and interpretations of geological and geophysical studies that either included or focused on the Valley. Individuals with current, working knowledge of the Valley and its geologic characteristics were consulted, as well.

The evaluation of industrial process applications was given the most comprehensive treatment. At the outset, a consultant on processes, Robert Bailie, Robert Bailie and Associates, developed basic process design and performance data for the beet sugar and malting processes. This provided us a general picture of industrial production of sugar and malt as it is practiced. Thereafter, we made an intense effort to ascertain the critical process variables that either allow or constrain variations in process design, and which in turn have strong bearing on achieving the design of an economical, geothermally-heated plant. Through these efforts, a design optimization, sufficient to initially demonstrate the real degree of practicable industrialization of moderate-temperature geothermal resources, has been completed.

The acquisition and preliminary assessment of demographic data for the Valley area and for the Front Range cities were performed by consultant Steven Weiner of Weiner and Associates. Historical population growth patterns and recent projections of growth to the year 2000 A.D. were acquired. Total energy use and distributions of energy loads among various consumers were also assessed with the cooperation of entities in the municipal governments of Canon City, Colorado Springs, and Pueblo, and with assistance from appropriate utilities. Cumulatively, this data serves as a basis for estimation of the market potential of hot water in quantities of many thousands of acre-feet. That estimate, in turn, allows a rough evaluation of the relative economic viabilities of the pipeline capacities for which design calculations were performed.

D. Supplement

Appendix E is a supplementary section related to Chapters V and VII. It presents much more detailed preliminary engineering design and cost evaluation for both the barley malting process and the sugar beet refining process. The cost elements include production wells, brine delivery lines, brine pretreatment system for acidification, degasification and pH adjustment, surge tanks, brine discharge lines, and injection wells. In addition, the sugar refining process includes flash tanks for production of steam for the evaporators. The costs of pumps and miscellaneous equipment such as instruments and controls are included.

II. RESULTS AND CONCLUSIONS

The project work as contracted with the Energy Research and Development Administration,* and as outlined in the preceding chapter, was to conduct an investigation into the potential for the economic utilization of the apparent geothermal energy resource located in the San Luis Valley of Colorado. The results of that work are presented in this report and are summarized below.

1. The analysis of the hydrothermal energy potential of the San Luis Valley of Colorado indicates that a substantial resource exists in that region capable of supporting large-scale industrial applications. A reservoir is estimated to contain 30 million acre-feet of water ranging from a minimum of 250°F for well depths of 8,500 feet to 10,000 feet; temperatures of 400 to 500°F are tentatively projected to exist at 12,000- to 13,000-foot depths. The dissolved solids content indicated between 8,500 and 10,000 feet is of the order of 2,000 parts per million. The reservoir mechanics would probably facilitate drilling at depth with little difficulty, as well as the maintenance of good well flows and circulation within the hot, confined aquifers without disturbing the unconfined aquifer. It may be possible to draw excess water from marsh areas at low points in the Valley to inject, along with recycled geothermal waters, assisting in maintaining artesian pressures in the confined aquifers, at the same time controlling salt content and undesirably high water tables at points near the surface. These considerations are probably the most significant sources of concern to area farmers. Substantial irrigation is indispensable to the agriculture industry. In fact, beneficiation of the entire San Luis Valley water system is a plausible complement to the development of commercial and industrial-scale utilization of the geothermal resources.

2. Based on a limit on the geothermal brine temprature of 302°F, a comprehensive engineering design was developed for a geothermally heated 5,000-ton-per-day beet sugar processing facility. A facility of that size using conventional fossil fuels most commonly exhibits a heat load ranging from 177 million Btu/hr to 312 million Btu/hr, depending on the overall plant configuration and its associated steam-to-sugar beet energy efficiency. These factors are largely a function of the age of a given plant, the more recent designs being, usually, the more efficient, too. The present design for a process heated via geothermal fluids is less efficient, requiring 365 million Btu/hr for a balance between in-plant capital costs and the costs of reservoir development, including large pipeline delivery systems. Fuel economy in a beet sugar plant is achieved at the cost of increased capital investment per pound of sugar-producing capacity, which results from requirements such as increased heat transfer surface area, additional pumping capacity, large vacuum systems, and larger or more sophisticated purification and sugar extraction systems. Further optimization of this design may lead to 10 to 30 percent reductions in process heat consumption, with corresponding reductions in geothermal fluid consumption and well costs.

3. Costs of a geothermal energy system are clearly competitive with fossil fuel systems for sugar beet industry applications. Capital costs were developed based on replacement costs of the steam heat as supplied by contemporary fossil fueled plants. The efficiency of a sugar plant is defined as a percentage in terms of the pounds of steam consumed, at a 212°F reference temperature, per pound of beets processed. Thus, a "highly efficient" plant actually would have a relatively low efficiency value. The geothermal energy-supplied plant resulted in costs ranging, on an equivalent production basis, from $1.77 per million Btu's for an average, existing plant in the United States (75% efficiency) to $2.85 per million Btu's for an extremely efficient (44% efficiency) plant. For a new plant such as would probably be built in the U.S. today, approximating a 55% efficiency, the equivalent replacement fuel

*Now the Department of Energy

cost would come to \$2.34 per million Btu's. Significantly, these costs are based on the characteristically short 120-day operating period for most beet sugar plants in this country. These figures imply that an economical geothermal beet sugar plant favors a relatively "inefficient" plant. As mentioned in the preceding paragraph, improved efficiency results in higher capital costs. The use of a moderate temperature heat source, such as 302°F geothermal water, adds to the same factors that increase capital costs in attempting to improve process efficiency.

These cost figures are highly encouraging of themselves, and the more so because they are based on amortization of plant costs during only 120 days of each year, corresponding to the approximate annual operating campaign for most beet sugar plants. Other process applications with much higher annual on-stream factors will yield even more attractive cost factors. Utilization of geothermal resources by complementary processes also would work strongly to reduce costs below those derived for a single sugar plant.

4. Two additional, important conclusions pertain to the plant cost figures presented herein. First, the geothermal energy cost is strongly a function of the capital well costs, based on a conservative 10,000-foot well depth. Capital costs for conversion of a beet sugar plant to geothermal energy were about 25 percent for additional, in-plant process equipment, and 75 percent for wells and distribution piping to the plant. If the proposed resource development produces adequate water temperatures at shallower depths, the costs of the energy supply system could be substantially reduced, as a result just of savings on well drilling costs. Second, for a new plant a geothermal energy system would eliminate the need for a main plant steam boiler, thus greatly reducing both capital and labor costs. Credit for this last factor was not included in the cost analysis presented here.

5. Certainly it is true that currently there are available fossil fuel resources which are competitive with the energy cost figures presented. These include natural gas, fuel oil, and coal. However, such resources continue to be managed from the perspective of a national energy shortage; the costs of these resources are projected to rise in ranges of 7 to 10 percent per year, which presently would account mainly for inflationary price increases and allows little room for supply-and-demand forces raising fossil fuel prices. The geothermal system, on the other hand, should not experience this increasing cost trend to a corresponding degree, depending more nearly exclusively on inflation in labor and capital goods costs. Additionally, the continued use of natural gas for industrial purposes is contrary to national energy development and conservation programs currently being outlined in Congress. Finally, the combustion of either oil or coal increasingly incurs significant, additional capital and operating costs related to more sophisticated environmental control systems. Particulate collection, sulfur dioxide removal, and water quality control requirements are the most prominent of these factors. Considering that the use of geothermal energy often times does not exhibit these environmental control problems with the same severity, a properly designed geothermal system is seen as a more attractive and reliable alternative.

6. Based on the hydrothermal resource identified in the San Luis Valley, a comprehensive and detailed design for a 12,000-bushel-per-day barley malting facility has also been developed. An average fossil fuel heat load for such a facility is approximately 48 million Btu per hour. Barley malting is a process more readily identifiable as a potential user of geothermal energy in the Valley for several reasons. Barley already is the primary crop grown in the San Luis Valley; also, barley malting is a year-around industry, andprocess complexity and temperature requirements (190°F) are relatively low compared to a sugar plant.

7. The cost evaluation for a geothermally heated barley malting plant was found to be very favorable and meriting further industrial consideration. For a delivered brine temperature of only 260°F, the replacement energy costs would come to only $2.63 per million Btu's. A vapor recompression system, to more efficiently utilize the available low-moderate temperature brines, was investigated and was found to be uneconomical for the slight additional efficiency realized as reduced brine consumption.

8. A considerable potential exists for utilizing geothermal resources in this region of the country. The market for agricultural goods, by establishment of a larger regional foods processing industry, would benefit agriculture in general and, particularly, such areas as the San Luis Valley that are below national averages in terms of economic criteria for standard-of-living assessments. Other manufacturing and refining processes were considered briefly as candidates for in-depth analyses of energy-source conversion costs. A list of several promising industries was drawn together and is presented below. Note the ranges of operating temperatures listed, and the heating media. These are also industries presently well established in or about Colorado, if not specifically in the San Luis Valley.

Process	Range of Maximum Operating Temperatures	Heating Medium
Concrete Block and Brick	165-350	Steam
Borax	140-210	Steam/air
Bromine	225	Steam
Chlorine	150-300	Steam
Phosphoric Acid	250-320	Steam/air
KCl	200-250	Steam/air
Sodium	240-275	Steam/air
Glass	70-350	Steam/air
Lumber	150-350	Steam/air
Synthetic Rubber	250	Steam/air
Steel	150-220	Steam/air
Textiles	70-300	Steam/air
Soup Canning	140-210	Steam/air
Dairy	140-212	Steam/air
Gasohol	120-280	Steam/air

In the context of the San Luis Valley, an agricultural area with economic difficulties, the gasohol process is a very attractive item to consider. This process is currently receiving considerable attention as a possible source of hydrocarbon fuel. Agricultural products, such as grain, corn, or beet sugar, are biologically converted to alcohol and distilled, producing a liquid fuel that burns well with gasoline. However, this is a net energy consuming process in terms of the fuel energy produced versus fuel energy used. It is difficult to justify the consumption of hydrocarbons such as coal or fuel oil to produce fewer equivalent Btu's as the convenient liquid fuel. A geothermally heated gasohol plant, however, would not give a net hydrocarbon fuel debit.

9. From present information the San Luis Valley geothermal reservoir promises to be capable of delivering very large amounts of industrially-useful heat, as well as water requiring little treatment. This could actually be realized with significant attendant possibilities for achieving environmental improvement to the San Luis Valley natural water systems and agricultural lands as an integral aspect of geothermal resource development. This was mentioned briefly in Item 1 of this chapter. Specifically, vast amounts of surface water are lost to the atmosphere by evaporation without profitable use. In addition, the Rio Grande River Compact, an agreement among the States sharing access to the River, constrains the States to allow certain minimum amounts of water to leave their boundaries via the River. Weather patterns periodically reduce River flows such that withdrawal of water is extremely limited within the River Compact stipulations. If the energy content of geothermal water makes acquisition of the water attractive, then proper management of the overall Valley water system with that supplemental source would make the water supply go much further toward satisfying the varied demands.

10. The inland, western portion of the U.S., in general, is in a growth trend exceeding those in most other parts of North America. This is expected to continue particularly in Colorado, due to the proximity of vast coal and other mineral resources. This factor has been recognized in the public and private sectors, and in many municipalities planning has been considered or actually instituted to accommodate cultural and economic growth with due attention to management of industrial, commercial, and domestic development. The distribution of water and energy resources to regional cities is a major concern.

Population growth and the water and natural gas-supplied energy consumption for the San Luis Valley and the Front Range cities of Canon City, Colorado Springs, and Pueblo were analyzed. Cautious estimates of population growth that can be expected in the near term, and tentatively projected to the year 2000, predict that the population for the above study sites will approximately double with respect to 1975-1976 census data. Assuming a fairly homogeneous distribution of growth among various sectors of the economy, and assuming some stability of per capita resource consumption, energy and water consumption can also be projected to approximately double in the next quarter century.

11. A major, overland pipeline to transport geothermal water to a remote site was preliminarily designed. A range of flow capacities was considered, and user costs were estimated on the usual basis of dollars per unit of heat delivered. The demographic data on present and predicted heat and water consumption assisted in evaluation of the prospective feasibility of various pipeline sizes. A 60-inch line, for example, would deliver about 100 percent of the heat energy presently consumed in the tri-city Front Range area investigated, in correspondence to the estimate of nearly 100 percent population growth by the year 2000. Costs for such a trunkline system would be of the order of \$3.65 per million Btu's, in present-day dollars. For a 24-inch diameter line, delivering about 84 percent less water and thermal energy than a 60-inch pipeline, associated energy costs are estimated to be about \$4.50 per million Btu's.

For comparison of energy costs, electricity in the Valley region is available at about 2 1/2¢ to 3 1/2¢ per kilowatt-hour, which is in the range of \$7 to \$10 per million Btu's. Propane, a widely-used fuel locally, costs more than \$4.00 per million Btu's. Natural gas is scarce in the Valley, but a good cost figure, subject to availability, would be in the range of \$1.75 to \$2.00 per million Btu's. Thus a 24-inch pipeline would lead to energy costs of the order of costs prevailing in the San Luis Valley now. A very large pipeline would deliver energy at lower costs not very much greater than more competitive energy costs for areas outside of the Valley.

12. These geothermal energy cost estimates are based on data which pertains only to a portion of the geothermal resources in the San Luis Valley. It is very likely that higher temperatures, possibly 400 to 500°F, are available at depths to 13,000 feet, as compared to the estimated 300°F temperatures predicted at about 10,000 feet. Only modest increases of water temperature above 300°F are necessary to noticeably reduce either process equipment costs or costs for wells to produce a given amount of heat, or both. Additionally, the scope of feasible applications would increase dramatically with a higher resource temperature of only 350°F, compared to the constraints imposed by the assumed maximum resource temperature of 302°F. Temperatures of such magnitude begin to make electrical generation an attractive option that would further enhance the economic feasibility of process and space heating applications.

13. The costs derived are based entirely on the use of geothermal fluids for energy. In the case of a major geothermal pipeline, the energy-based pipeline costs might be substantially reduced by providing for consumption of the cooled water instead of discarding the water irretrievably. Considering the modest 2,000 ppm dissolved solids content predicted, treatment costs for this possible source of industrial or municipal water would not add much more heavily to the pipeline costs than if the same amount of water were supplied, instead, by the usual surface water gathering systems, presumably delivering a better quality water.

III. HYDROLOGY

The following sections summarize what has been learned about the hydrology and geology of the Valley in the course of this study. Much of the information in this summary is taken from the final report on the hydrology of the San Luis Valley by Davis. His complete report is included as Appendix A of this volume.

A. A General Description of the Valley

The San Luis Valley is at the northern head of the Rio Grande Rift System, which runs south and southeast to the Gulf of Mexico. Figure 1 depicts a view of the Valley. Two mountain ranges form the eastern and western boundaries. These are the Sangre de Cristo and San Juan Ranges, respectively. Within Colorado, the Valley extends about 90 miles north of the New Mexico border; at the widest point, the Valley is about 35 miles across, and the approximate elevation is 7,700 feet. The Rio Grande River enters from the San Juan Mountains at Monte Vista. The watershed area servicing the River, including that part in the mountains, amounts to about 8,000 square miles. The Valley floor, itself, covers about 3,200 square miles.

The maximum depth to crystalline basement rock is variously estimated to be 20,000 to 30,000 feet. Valley fill consists of varied strata of alluvium and volcanic debris, frequently forming interbedded layers with considerable variability in geologic characteristics. An approximate cross-sectional diagram in Figure 2 depicts the profile of the Valley floor from a point near Monte Vista to Crestone Peak. In the upper several thousand feet of strata there are significant overlapping strata of fine silt, clay, and perhaps shale, forming an effective, low-permeability barrier on top of deeper strata, which are filled with water. These covered strata are termed a "confined" aquifer because of limited upward mobility of the trapped water. The confined aquifer is a prime target as a geothermal resource. The upper strata are "unconfined" in that water is essentially free to diffuse upward and evaporate to the atmosphere. This zone of strata is called an unconfined aquifer.

Unconfined and confined aquifers contain at least two billion acre-feet of water. The shallow unconfined aquifer occupies almost the entire Valley surface and extends to 50 to 100 feet below the surface. The water table is at less than 12 feet almost everywhere, and is at the surface in large areas that are presently marsh wastelands. The confined aquifer underlies most of the Valley and has sufficient head to cause artesian flow of water to the surface. The waters in the confined aquifer are usually of better quality; the analysis of a large number of shallow well samples has shown concentrations of 70 to 437 ppm (parts per million) of dissolved solids in parts of the confined aquifers as compared to 52 to 13,800 ppm in the unconfined aquifer. More recent data from several deep wells has shown the confined aquifer also contains water about an order of magnitude more concentrated than 437 ppm.

The total annual water supply to the San Luis Valley in Colorado is about 2.5 million acre-feet, of which about 60% is by stream flow from snow melt on the adjacent mountains. The remainder is from precipitation on the Valley floor. The Valley discharges about 2 million acre-feet by evapotranspiration, about half of which is nonbeneficial from uneconomic plants such as phreatophytes. Another 0.5 million acre-feet is lost by stream flow across the State line.

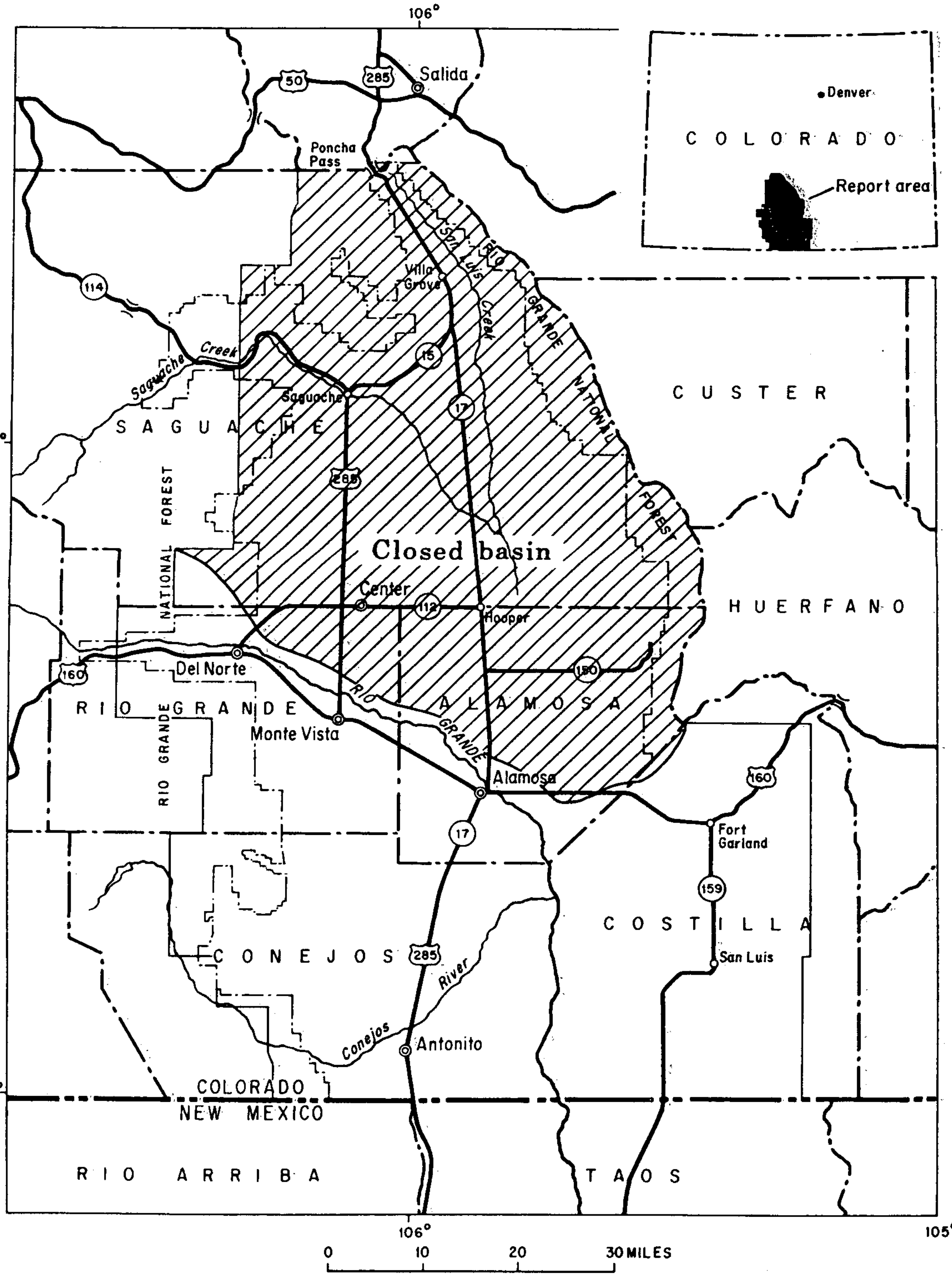


Figure 1.--San Luis Valley Area Map

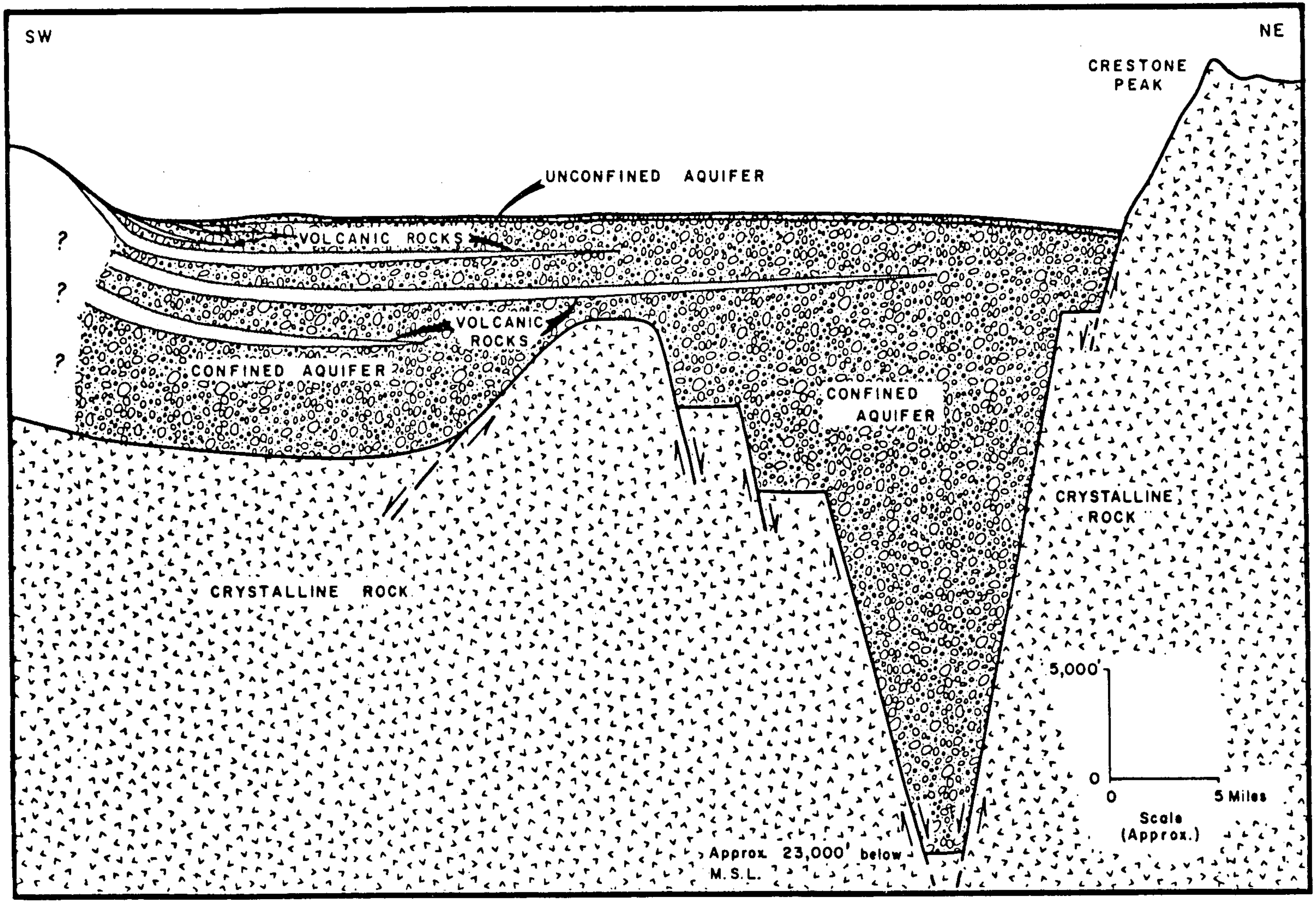


Figure 2

SW-NE cross section through San Luis Valley showing relationship between geologic conditions and aquifers (Modified from: Emery and others, 1969; and Gaca and Karig, 1966)

Withdrawals for irrigation of about 750,000 acre-feet per year were recorded in the 1960's. Despite withdrawals at this rate, there is much more water in aquifer storage than there was before large-scale irrigation began between the years 1880 to 1890; in 1900, the water table was 50 to 100 feet below the surface. This is approximately at the depth of the first substantial clay strata bounding the confined aquifer. The Valley could accommodate much greater rates of water withdrawal from the unconfined aquifer. Parts of the Valley are waterlogged, so that a reduction in the water table would be desirable.

Vertical movement of groundwater is restricted by the interbedded volcanic flows and interspersed clay beds. Accordingly, a net production of water from the deeper confined aquifer will have little or no effect on the viable withdrawal rates from the unconfined aquifer. Thus, an increase in water production to support a geothermal industry would probably not be detrimental to the present irrigation supplies. This consideration is discussed further in the next chapter.

B. Evidence of a Geothermal Reservoir

Various studies have been made of the geology and the groundwater systems in the San Luis Valley. These include the drilling of an exploratory geothermal well near Alamosa to a 9,500-foot depth. Data from this well and from numerous oil wells, as well as data from geophysical studies, have been analyzed and are summarized here.

These data have been interpreted to indicate the existence of a major aquifer in the northern and eastern portion of the Valley, approximately underlying the Sand Dunes National Park. This is shown as the cross-hatched area in Figure 3. The aquifer changes considerably in physical characteristics with depth. The most promising zone begins at a depth of 8,600 feet and extends to at least 9,500 feet. This information was acquired from logs of well #1-32, Mapco-State, located near the eastern periphery of the Valley and slightly north of Alamosa (see Figure 1).

In this depth zone the porosity is about 25%, and Bouguer gravity data yielded an estimated 225 square-mile surface area for the aquifer. On these bases, it is calculated that the aquifer contains 30 million acre-feet of water.

Extrapolation of geothermal gradient data from other sources (5) indicates a temperature of at least 250°F at about the 8,000-foot depth. Geothermal and geochemical data also predict elevated temperatures. Geothermometry methods used by the Colorado Geological Survey yielded source temperature estimates ranging from about 195° to 395°F for thermal wells and springs in the San Luis Valley. Other well data include a temperature of 110°F at 4,200 feet, and 250°F below 8,500 feet in the Mapco well. This 250°F value was obtained during drilling when thermal equilibrium probably had not been attained. Therefore, higher temperatures than 250°F are anticipated at this depth range.

The permeability from 8,600 to 9,500 feet appears to be sufficient to enable production and reinjection rates exceeding 500 gallons per minute (250,000 lb. per hour). Potentiometric data further indicate the total dissolved solids content is about 2,000 parts per million in that depth range. This salinity value suggests that substantial circulation of the contained water prevents chemical degradation by dissolution of minerals, and it also agrees with the observation of high porosity and permeability values for the aquifer. The available data points to the area just discussed as the most attractive for further exploration and possible development.

The same area also contains shallower zones of less favorable, water-bearing strata having lower temperatures, higher salinities, and lower permeabilities. For example, between 5,700 feet and 6,600 feet, permeability is notably less than that below 8,600 feet. Two water samples were analyzed in the upper zone; one analysis was performed on-site, and the other was done in a laboratory. The results gave 9,400 and 5,800 ppm, respectively, of dissolved solids. These values do not agree with data presented by Barret and Pearl (6) for other wells and springs in the Valley. It is felt that these water samples were taken from an aquifer high in silt or shale content, possibly explaining the high solids contents.

In addition, the overall body of data points to another aquifer in the extreme northern portion of the Valley with temperatures near 220°F. This aquifer lies at depths between 5,000 and 10,000 feet, and contains 2 million acre-feet of water. The hydrological analysis also noted that the northern aquifer could be a part of the much larger aquifer that underlies the Sand Dunes to the southeast.

In conclusion, this data reflects a high probability of attaining feasible production rates of moderate temperature geothermal brines from the San Luis Valley. The potentially major economic problem associated with exploitation of this resource is related to the rather great depths involved, of about 8,000 to 10,000 feet. With respect to this point, however, it should be remembered that the temperature logs that were run were made shortly after drilling was completed, and it is possible that full equilibrium was not attained. Thus, higher temperatures at shallow depths may yet be found. Furthermore, a range of possible temperature profiles predicts temperatures as high as 400° to 500°F, between depths of 12,000 to 15,000 feet.

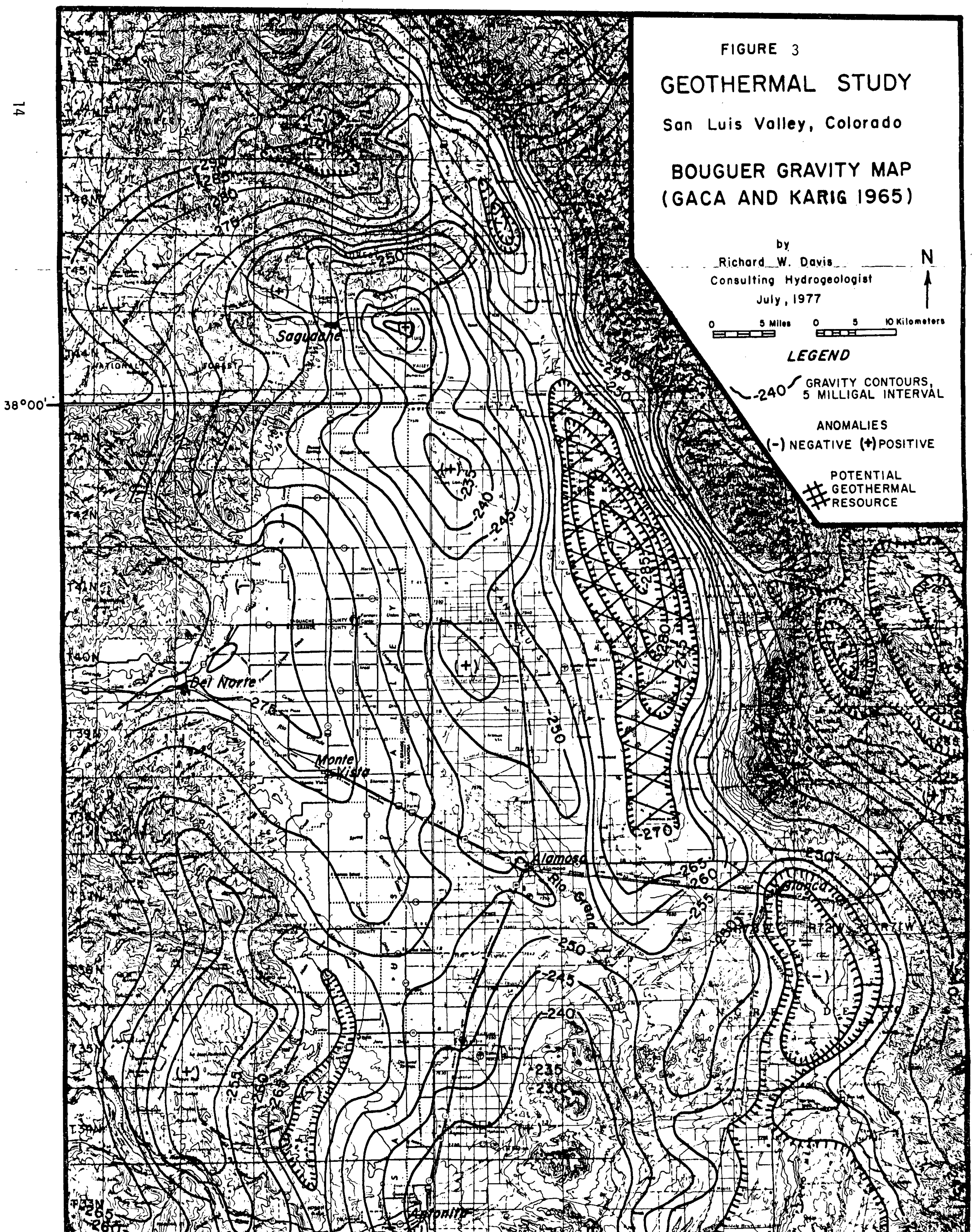


FIGURE 3
GEOTHERMAL STUDY
San Luis Valley, Colorado
BOUGUER GRAVITY MAP
(GACA AND KARIG 1965)
by
Richard W. Davis
Consulting Hydrogeologist
July, 1977
N
0 5 Miles 0 5 10 Kilometers
LEGEND
240 GRAVITY CONTOURS, 5 MILLIGAL INTERVAL
ANOMALIES
(-) NEGATIVE (+) POSITIVE
POTENTIAL GEOTHERMAL RESOURCE
Saguache
Del Norte
Monte Vista
Alamosa
Rio Grande
Blanca
Antonito
COLORADO
NEW MEXICO
38°00'
37°00'
106°00'

IV. THE SAN LUIS VALLEY AS A REGIONAL WATER SYSTEM

A. The Context for Consideration of Geothermal Resources

The resources of the San Luis Valley should be considered in relation to their interactions with or effects on related resources in other regions. This chapter discusses the role of the surface and groundwater resources within the Valley, alone and also in terms of the hydrological role of the Valley as a component of the Rio Grande River system. The discussion considers the uses of these water systems, problems involved therein, and possible ameliorative measures that are based on the inclusion of the development of geothermal resources at significant levels. Discussion in Chapter IX, on the design and costs of a major geothermal pipeline, will be related to quantification of potentially "significant" levels of use of San Luis Valley geothermal resources. Although this discussion is qualitative for the most part, a number of qualitative factors that could be considered as well are outside the scope of a study of this order. An example which was considered briefly concerns the interaction of hydro-resources with environmental systems such as atmospheric systems. There are insufficient baseline data and theory to understand and predict atmospheric interactions with the various water bodies of a region, except in extremely general terms (9). Accordingly, this discussion includes only the hydrological resources.

First, an elemental review of the San Luis Valley natural hydrological systems is suggested: Figure 4 is a cross-sectional flow schematic for water. The Valley is a collecting point for about 8,000 square miles of watershed, including parts of the surrounding mountains. Water enters the Valley directly as precipitation to become surface water; additional surface water enters by stream flow from the mountains. Some surface water moves underground by way of faults, or it percolates from the surface downward through predominantly course, highly permeable strata at the margins of the Valley, entering the lower strata that make up the confined aquifer. Little is known of the subsequent flow paths of water in the confined aquifer, except that minor amounts of water do migrate to the surface through or around the lenticular, overlapping clay and shale beds above; other small amounts may migrate horizontally to the south, but it is thought the path for such flow is probably highly restricted in cross-sectional area. Well water from both the confined and unconfined aquifers is added to surface waters via irrigation. That surface water which is not carried across the Colorado-New Mexico border either soaks into the upper strata, entering the unconfined aquifer, or it evaporates. There is a dynamic balance between the sources of in-flow to the unconfined aquifer, stream flow out of the Valley, and evaporation from both ground and surface waters. The evaporation, also called evapotranspiration, is identified according to its source--that which is caused by crops, and by the action of non-beneficial, non-agricultural plant life. Evaporation of groundwater from barren, desert land is comparatively negligible.

B. Problems in the San Luis Valley Water Systems (7, 8)

The groundwater reservoirs and surface waters in the Valley, including the Rio Grande River, are resources for which the desired uses tax the systems' capabilities. The use of the Valley for farming, requiring substantial amounts of irrigation, along with the growth of communities and agriculture along the Rio Grande River, have resulted in serious constraints on the balance between the functioning of the collective human systems and the environmental water systems. In this sense the Valley

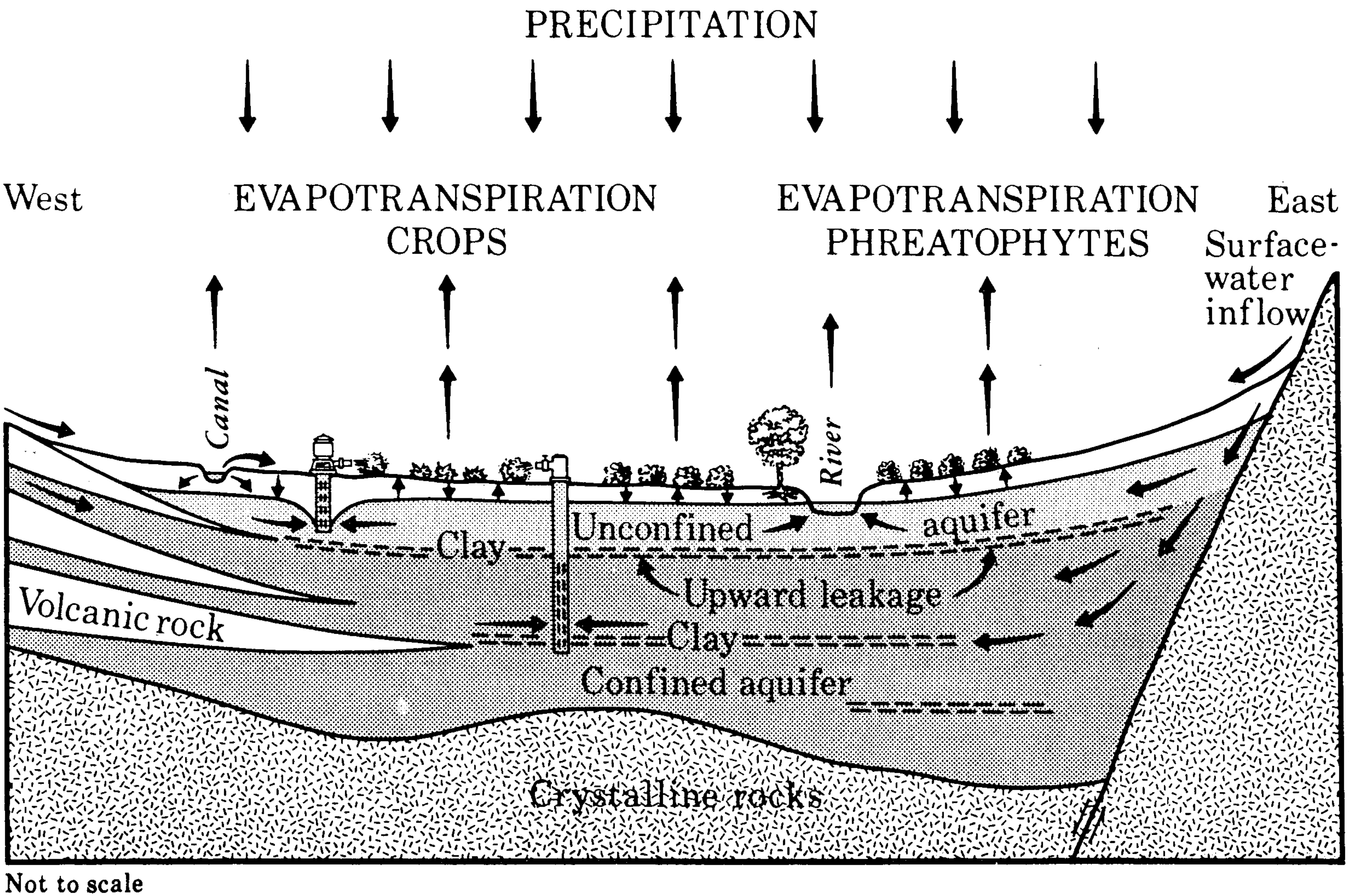


Figure 4.--Cross-Section of San Luis Valley Water Systems

has two outstanding problems so related.

First, because of the continual increases in the amounts of water withdrawn from the Rio Grande River for irrigation and other uses, downstream inhabitants were suffering, or in danger of suffering, severe water shortages in the early decades of this century. As a result, in 1929 the States of Texas, New Mexico, and Colorado signed the Rio Grande Compact on the management of Rio Grande waters, delineating annual River flows which the several States would guarantee to cross their common borders. In actual practice this limits the amount of irrigation that can go on at the expense of the residual volume of the Rio Grande flow. Under the terms of the Compact, Colorado is presently indebted for approximately one million acre-feet of water because of excessive withdrawals in past years.

The second problem also results, to a large extent, from the heavy regional dependence on irrigation. A portion of the Valley northeast of Alamosa, called the Closed Basin, is the lowest point in the Valley and is hydraulically bounded to the south by alluvium of the Rio Grande River. Consequently, it tends to act as a sump and both surface waters and groundwaters from much of the northern portion of the Valley migrate to the Closed Basin. The area presently is a salt marsh. About one million acre-feet of water evaporate non-beneficially from this area annually. The land is presently largely unused, though it is an important water fowl habitat. As noted briefly in Chapter III, due to increased irrigation the water table in much of the Valley has risen markedly in this century and significant amounts of land have been lost to the marshes area. That fact and the concomitant accumulation of salts in upper soil levels, even where marshes have not formed, require further constraints on irrigation in areas draining to the Closed Basin and, hence, restrictions on cultivation of usable farm land. Irrigation, drainage, and water-table depths are of a somewhat lesser concern to the south of the Closed Basin.

C. The Role of Geothermal Resources in Valley Water System Management

A potential solution to the above problems has been studied, approved, and recommended more than once since the signing of the River Compact in 1929. It is called the San Luis Valley Closed Basin Project, and it would result in an annual drainage of about 100,000 acre-feet of water from the Basin to the Rio Grande River. This would be of great benefit to the availability of land in the Basin watershed, and it gives consideration to a variety of integral factors, calling for careful protection and management of the wildlife habitats, among other concerns. Farm land management would be aided and Colorado's liability to the River Compact would be significantly supported and eased. Otherwise, massive, non-beneficial evapotranspiration will continue; deterioration of arable land may progress; and less of Colorado's Rio Grande water debt will be serviced. The Closed Basin Project has been reconsidered several times by State and Federal representatives since the signing of the Rio Grande Compact almost 40 years ago, when the concept was endorsed by Colorado and New Mexico officials. We did not learn the reasons this proposal hasn't been implemented in some form. As presented in the past, the Closed Basin Project would not involve geothermal resources. However, there are attractive options afforded by incorporation of geothermal resource utilization to promote the desired effects of water systems management in the San Luis Valley.

Utilization of geothermal resources in the Valley could benefit management of the water systems, either alone or in conjunction with the measures included in the past-proposed Closed Basin Project, in several conceivable ways. First, geothermal water used industrially for heating will probably be flashed to produce steam. The resultant condensate might then pick up some slight impurities, but for all intents and purposes the condensate will be pure water that could be added to the Rio Grande River after cooling. Second, recalling the stipulation set in this study for a maximum 302°F temperature source, residual geothermal water after flashing steam for process heating must be expected to amount to about 85 to 90 percent of the total water flow before flashing. Given the low salinities anticipated for deep hydrothermal reservoirs in the Valley, desalting of the residual water to acceptable levels for addition of even greater amounts of water to the River is also a possibility; less rigorous desalting might also be practicable for use of geothermal water in place of Rio Grande River water for irrigation. With respect to statutes governing the temperature effects of effluent additions to surface waters, though, it would still be necessary to remove on average, about 100 Btu of heat per pound of brine. Third, withdrawal of confined aquifer water is not expected to have much, if any, effect on the water table because of the degree of aquifer integrity created by the complex, shale-silt-clay confining strata in probable drilling locations. Finally, if geothermal water were transshipped out of the Valley by pipeline, marshland water could be withdrawn for reinjection, assisting in drainage of the Closed Basin and other areas where water table depths may be judged as undesirably shallow.

D. Public Awareness of Water Problems

The Valley populace, particularly the farmers, are acutely aware of the perennial water supply and use problems. Since most water wells access the unconfined aquifer, a serious drop in the water table depth would be very harmful to farming operations. These problems are endemic to agricultural operations, and the people there must live with that fact. Nevertheless, any proposal to change the environmental status quo generates readily discernible tension because of the delicate balance already involved. The Valley economy, in general, lags behind that of Colorado and the nation as a whole. Substantive improvement of the natural water systems would be welcomed as a means of facilitating a stronger regional economy. However, the likelihood that the use of geothermal resources from the confined aquifer will not result in dried-up irrigation wells in the unconfined aquifer, nor obviate compliance with the terms of the Rio Grande Compact, will have to be substantiated much more thoroughly to gain the local, public acceptance necessary for implementation.

V. THE APPLICATION OF GEOTHERMAL ENERGY TO THE BEET SUGAR REFINING INDUSTRY

A. Introduction and Industry Background

The growing of sugar beets and the associated beet sugar refining process without question constitute important segments of the Colorado agricultural and industrial community. Sugar beets were at one time grown in the San Luis Valley in significant quantities, but barley has recently become the largest single crop grown in that region. Most of the Colorado beet sugar refining facilities are now located along or east of the Front Range of the Rocky Mountains. The sugar refining process is a complex, agri-industrial, energy-intensive process with significant potential for future geothermal application. Federal Energy Administration data shows that beet sugar processing consumes more energy than any other of the food products industries (11).

In this chapter, the technical details of the refining process for recovering sucrose from beets are reviewed; a special emphasis is given to the evaporation section of the process. This process flow description is important in that it frames the basis for our selection of operating conditions throughout the geothermal process design. Consideration is then given to the specific application of geothermal energy to this process, keeping in mind the 302°F resource temperature constraint established in the study guidelines. The design methodology and the economic evaluation of the resultant conceptual process design are then discussed.

As noted earlier, the beet sugar refining operation is comparatively complex. In recognition of that fact, a comprehensive and quite detailed description of the entire flow process has been developed separately and is presented in Chapter VI. That section is fundamental to this study in that it explicates the many operational constraints implied in the interrelated physio-chemical and biological processes inherent in the industry. Such constraints ultimately determine the viability of conversion to a geothermal energy source. The reader may choose to bypass that detailed treatise, depending on how thoroughly one wishes to familiarize himself with the numerous considerations which are encountered in developing the design bases for the geothermal facility. The evaporation section, however, is discussed alone in the present chapter because it is the point of introduction of process energy, whether the heat source is a boiler or the Earth. The parameters controlling modification of the evaporator train to accommodate the use of geothermal heat could not be presented in the format of "optional reading!"

B. The Sugar Beet Refining Process (1, 2, 3, 4)

The intended methodology for the study was to first develop a process design representative of current industrial practices, thereby establishing the operational basis for making adjustments enabling substitution of a geothermal heat source for boiler steam. In so doing, we found that in the beet sugar industry there is general disinclination to speak in terms of a "typical" beet sugar plant because of the considerable variability in plant capacities and ages; in addition, geographical differences in plant locations require that individual sugar plants must be operated especially for beets having physio-chemical qualities peculiar to cultivation in given regions. Consequently, an intensive analysis was made of all phases of the entire beet sugar recovery process to ensure that our "representative" process design did not include any gross errors of generalization. The following general description is complemented by details contained later in this Chapter and in Chapter VI.

Viewed in its totality, the beet sugar refining process is a straightforward operation. The general process flow is depicted in Figure 5. Given the raw materials and the end products of this process, there are few changes that can be made that significantly affect the basic flowsheet of a beet sugar plant. Indeed, in the past 100 years the major process modifications have been due mainly to advances in the technology of heat transfer equipment and purification chemistry.

Sugar beets are cleaned and sliced into thin strips, which are then steeped in warm, low-salinity water in the diffuser battery. The sugar and various impurities enter into solution through the beet cell walls by a process very similar to osmosis. These solutes and suspended impurities also pass directly into the water from cells that have been broken or crushed in handling.

Two product streams then leave the diffuser. The insoluble beet pulp is washed, pressed, and then dried in a fossil-fuel fired kiln. This dryer consumes a significant fraction of the energy needs of the sugar refinery, but temperature requirements for pulp drying are normally in the range of 1200°F, which contributed to exclusion of this process step from the present geothermal design. The second stream leaving the diffuser is the raw juice, a dilute and impure sugar solution. Its dissolved solids concentration is in the range of 10 to 16 Brix; i.e., approximately 10 to 16 percent solids by weight, of which about 90% is sucrose.

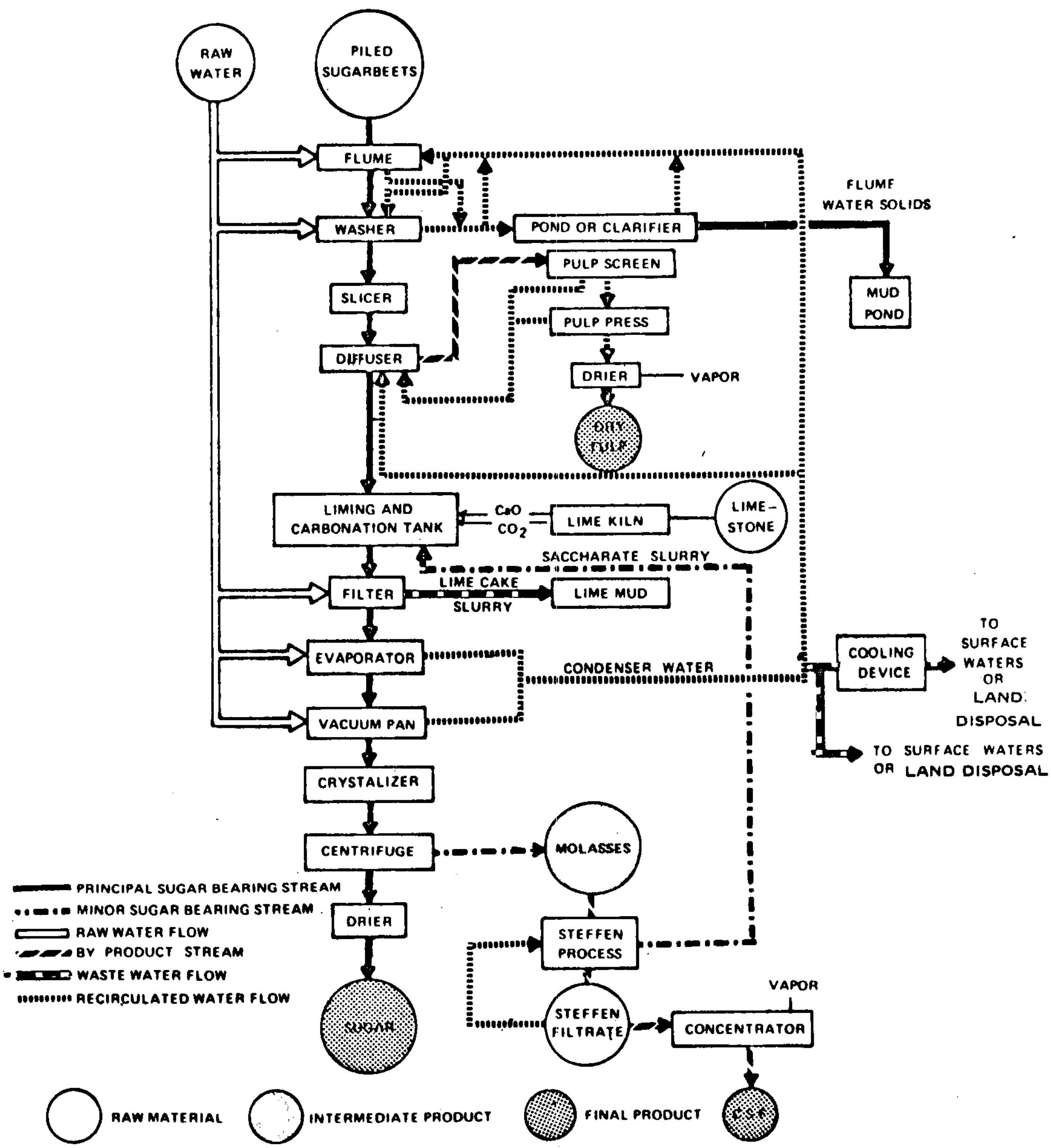
The raw juice passes through three basic purification stages separated by intermediate filtration and settling operations, to produce a pure "thin juice" with about the same solids concentration. Water is then evaporated from the thin juice until a solution of between 60 to 70 Brix (thick juice) is obtained. The evaporation steps represent the most economically viable points for geothermal energy application. The sugar solution has been processed on a continuous basis to this point. The thick juice then enters the "pan section" of the plant, consisting of two to four vacuum crystallizers operated in series-batch mode. The sugar solution is first concentrated to a labile, but not self-seeding, condition. Seed crystals are added and allowed to grow with gradual addition of more juice. The pans are operated such that when finally filled to capacity, the magma of crystals and liquor is at 90 to 95 Brix. Crystals and liquor are then separated by centrifugation and the crystals are washed.

Sugar is crystallized from the liquors fed to the pans in as many as three or four successive stages or "strikes," each stage producing a lower-purity batch of crystals and residual liquor. Only the first-strike sugar, sometimes blended with a portion of the second, leaves the process as product sugar at 99-100 percent purity. The low-purity sugars are melted and blended back with the thick juice from the evaporators. The various impurities taken into solution with the sugar cannot economically be totally eliminated in the purification section. Consequently, as in many other processes, the practice of recycling streams raises the recovery rate of the desired product; final impurities are removed in the byproduct molasses stream at the end of the process.

The molasses from the last "strike" will range between 70 to 90 Brix, with a sugar purity of about 60 percent. The stream may be further refined to extract sugar and may be used in making MSG (monosodium glutamate), or it may be applied directly to stock feed for its significant nutrient content.

Figure 5

MATERIALS FLOW IN BEET SUGAR PROCESSING PLANT WITH TYPICAL WATER UTILIZATION AND WASTE DISPOSAL PATTERN



As taken and modified from Beetsugar Technology, Second Edition, Edited by R.A. McGinnis, Beetsugar Development Foundation, Fort Collins. Colorado (p 645), 1971, (65)

C. Evaporation in the Beet Sugar Refining Industry

The concentration of the purified sugar solution via the evaporation step is clearly the most energy-intensive in the sugar refining process; this process section is comprehensively discussed here.

1. Process Details

The purified sugar solution entering the evaporator section is called the thin juice and will have a total dissolved solids content in the 10 to 15 weight percent range that was present in the raw juice leaving the diffuser. This concentration is then raised to 60 to 70 percent in the evaporators. The level of 60 to 70 Brix concentration of the thick juice leaving the evaporators is the optimum concentration level for subsequent blending with low-purity, remelted sugar from the crystallization section to make feedstock for the white sugar crystallizing pans. The dissolved solids in the blended feedstock are then approximately 94 percent sucrose, unless the beets have been seriously degraded during storage.

As the sugar solution increases in concentration through the evaporators, the increased dissolved solids concentration raises the juice boiling point temperature relative to the condensing temperature of steam evolved. In the last evaporator effect, this can amount to a difference of 8 to 10°F. As the vapors from the effects are subsequently used for heating, the boiling point elevation causes a decrease in the available temperature driving force between those vapors and the streams they will heat. The juice viscosity also increases by a factor of between roughly 25 and 100, depending on the precise thick juice Brix out of the last evaporator body. This physical change significantly affects the heat transfer coefficients in the evaporators. Based on data acquired for operating sugar plants, the first effect can be expected to have an overall heat transfer coefficient of 500 to 600 Btu/hr-ft^2-°F, while in the last effect the coefficient will be of the order of 100 Btu/hr-ft^2-°F.

Most of the dissolved gases in the juice stream, principally ammonia and some carbon dioxide, come out of solution in the first evaporator effects. These gases can either be purged from the steam chests of subsequent evaporator bodies, if those are under positive pressure, or they must be drawn out by the vacuum condenser systems. It is desirable to prevent re-absorption of the non-condensable gases into the condensate in those cases where the condensate is reused for makeup water.

2. General Design Considerations

The use of evaporators in a few sugar beet processing plants has fairly well paralleled the development of the best off-the-shelf evaporation technology. Economic conditions in recent decades have prevented most plants from retrofitting any of their various process sections on a continuous basis, and only a very few plants are fully modernized because the capital costs for doing so have been unjustifiable. However, while the most current equipment has not always been used, most plants are in step with efficient process design. For example, multiple-effect evaporator trains have long been in use throughout the industry.

In a simple n-effect, forward-feed evaporator system, the vapor from each effect, 1 through n-1, is used to evaporate approximately an equal amount of steam in the next following effect. Only the first effect is supplied heat via boiler process steam. As "n" increases, the amount of process steam required for the first effect decreases. In opposition to the virtually limitless, theoretical potential or energy savings by increasing the number of effects, the costs of capital equipment and associated labor again force a design compromise. In the beet sugar industry, the compromise has been for three to five evaporator effects; four- and five-effect systems are used in many of the most up-to-date plants. Some plants also use three-effect evaporators, all operating at positive pressure, with good steam efficiencies.

The evaporator section removes water from the thin juice to a concentration slightly less than that which enables spontaneous crystallization of the sugar. This simple function, however, obscures the importance of the evaporation section with respect to the entire plant's energy efficiency. In any process involving evaporation of part or all of a liquid stream, efficient use of energy also depends very strongly on recovering the heating potential of the evaporated vapors and the remaining liquid stream to heat the incoming liquid stream as much as possible. This practice, called regenerative heating, has been thoroughly incorporated into the process schemes of today's beet sugar technology. Again, a balance is drawn between energy savings and costs for capital and labor necessary to accomplish the savings. It will be shown that the fact that the evaporators operate at the highest temperatures in the entire plant will enable the regenerative heating scheme to be very simply and effectively applied using a minimum of geothermal water in the new design.

The first limitation on ideal application of regenerative heating is inescapable in any industry. A finite temperature driving force must exist between the incoming (warming) stream and the outgoing (cooling) stream; otherwise infinite heat transfer surface area would be required. Thus, some nonregenerative sensible heating of the inlet stream is unavoidable in order to establish the necessary temperature differential.

The beet sugar process suffers further inefficiency of recovery of heat because of exiting streams of particulate solids and high-viscosity liquids exhibiting low heat transfer characteristics. Sugar leaves the process, to be cooled, at about 180°F. This requires that considerable sensible heat must be added to replace the heat rejected in this product stream. In addition, the molasses product exits the process at about 135°F. It is also not suited as a heating medium because of its viscosity.

Evaporation is generally performed down to about 140°F, using steam that condenses 20 to 30°F hotter. The capital costs to be able to utilize steam at lower temperatures are prohibitive. The capital costs for using outgoing liquid streams in liquid/liquid heat exchangers are also prohibitive, and thus, no regenerative heating occurs below 160 to 170°F in the hot fluids.

Finally, the energy consumption schemes for beet processing are further complicated by the branching and by the cooling and reheating of process streams. Generally speaking, most of the vapor produced in all but the last evaporator bodies is used for the regenerative heating of the various process streams. The steam efficiency of preheating, however, is partially reduced by heat losses resulting from poor insulation of the various process equipment and piping. The crystallizing pans producing white sugar (white pans) are typically serviced by vapors from the first evaporator body, and the pans producing brown sugars (raw pans) are serviced by vapors from the second and sometimes third evaporator bodies. These crystallizing pans produce low-temperature vapors that condense at about 140°F, but these vapors, too, are sent directly to vacuum condensers rather than being used as a heating medium elsewhere. This again contributes to the unused portion of the potential for regenerative heating. In conclusion, the process flow scheme for multiple-effect evaporators, white pans, and raw pans described above constitutes a set of intertwined and parallel multiple-effect heating trains of varying length, all of which are integral to the entire process. To successfully apply geothermal heating to these systems, one must match the capability of the hot brine resource to the various, interdependent energy demands of these systems.

D. A Geothermally Heated Sugar Beet Process

1. Scope of Applications to the General Energy Consumption Pattern

As noted in preceding discussions, two principal energy consumptive steps in the refining process, pulp drying and evaporation, consume as much as 85% of the total heat energy demand in a typical plant. Approximately 35% of this is associated with the pulp dryer and the remainder with the evaporation process. A schematic drawing of typical energy inputs for the sugar process is given on Figure 6. Pulp driers are typically gas or oil-fired rotary kilns directly utilizing the burner exhaust gases at about 1200°F as the heating medium. If geothermal brine, on the other hand, is available at a maximum of 302°F, the volume of heated air used for the drying medium would have to increase almost by a factor of six in order to provide the same amount of drying capacity. The reason for this, of course, is the greatly reduced heat source temperature, which limits the maximum attainable drying medium temperature. Since the specific application to pulp drying entails a major redesign of one piece of equipment, the pulp drier was not considered in the present study. In addition the high temperature lime kiln is excluded from consideration.

2. Special Geothermal Design Considerations

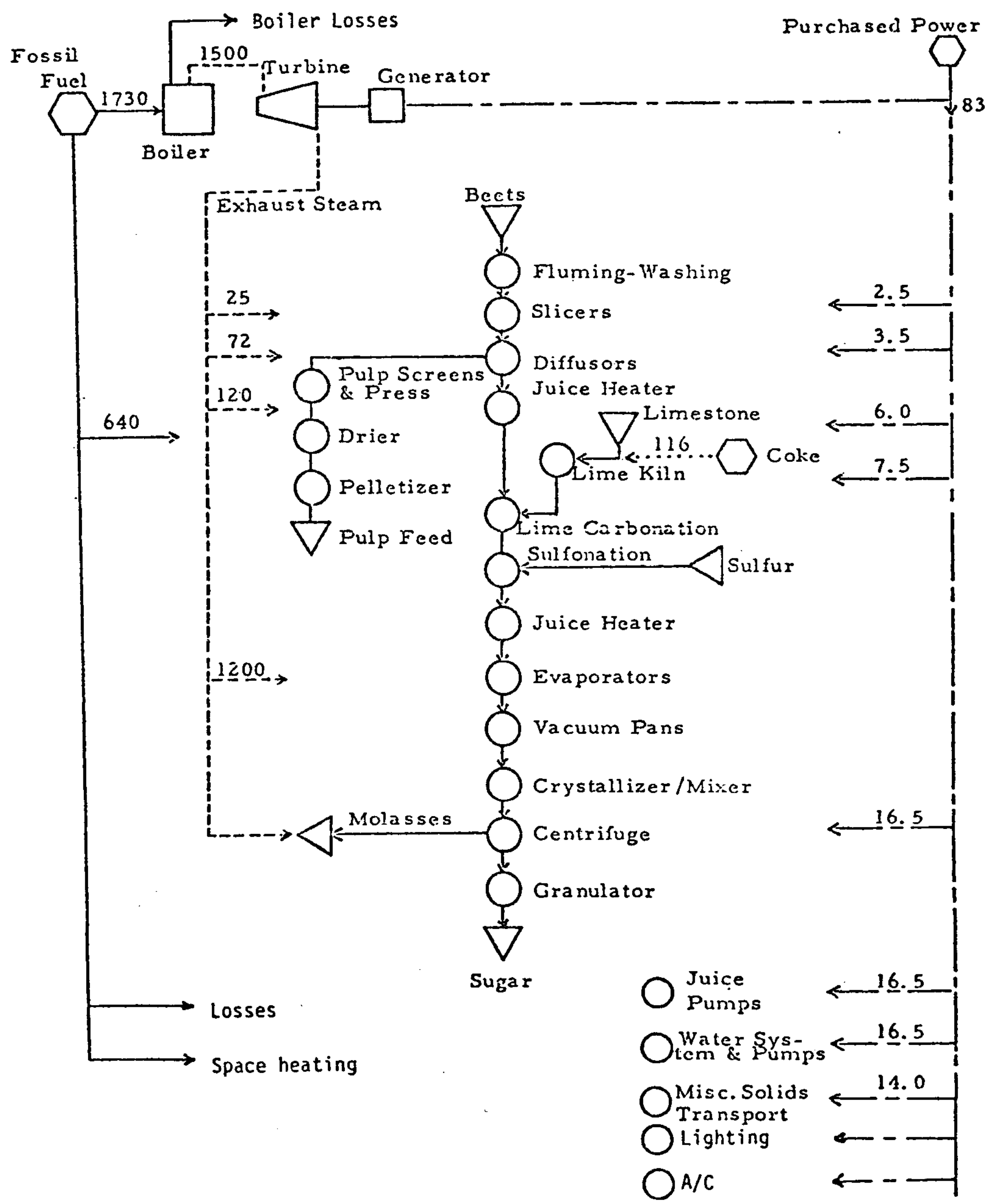
a. Boiler Steam Versus Geothermal Water

The efficiency of energy usage in a contemporary beet sugar plant using steam produced in a boiler is affected by four factors. First, the upper steam temperature is limited by the tendency for degradation of the sucrose at high temperatures. Second, economics have thus far institutionalized energy waste by making thorough insulation relatively too costly. Third, the relative costs of capital goods versus fuel have also contributed to the loss of energy as heat rejected at elevated temperatures, obviating more effective regenerative heat conservation. And finally, the complexity of process design effectively raises the optimal levels of energy consumption. The first and fourth of these factors are dealt with in this section.

In designing a beet sugar process to use geothermal energy from a low to moderate temperature temperature resource, the above limitations are still in force and further economic considerations must be dealt with which will also affect steam efficiency. Now, however, the steam efficiency, per se, will have a different significance because the cost of geothermal steam in dollars-per-million-Btu's has a different basis than do the energy costs for a fossil-fueled process. The geothermal resource temperature is a major factor in the cost of geothermal steam. In

Figure 6

Material and energy flow for a typical beet sugar processing plant.
(Energy inputs in units of thousands of Btu /ton beets sliced.)



conjunction with the temperature factor, the process complexity will fix the cost of energy for a geothermally heated process. The role of these factors in constructing the process flow scheme are discussed next.

The present design study calls for the use of a moderate temperature, liquid-dominated, geothermal heat source, which has been stipulated to have a maximum temperature of 302°F (150°C). In contrast to this criterion, the standard beet sugar process is supplied process heat in the form of steam at about 280 to 300°F. As a first approximation, let us assume we can cool the geothermal water from 302°F down to 150°F, giving about 152 Btu per pound of water. On the other hand, steam at 300°F has a latent heat value of 910 Btu per pound. Accounting also for the sensible heat of the condensed steam if it is cooled to a final temperature of 150°F, as well, the steam has a total of about seven times the usable heat content of the geothermal water. In addition,the first six parts of that seven-fold heat advantage for steam will be recovered in condensing the steam, allowing for heat transfer coefficients several times larger than are possible using water as the heating medium. The dissolved mineral content of geothermal water also tends to pose a serious, potential scaling problem for heat exchange surfaces. In consideration of the above factors, the most desirable way to use the geothermal brine from the wellhead is to flash the brine to a lower temperature, segregate the steam so formed, and then direct the steam to process units as needed. The geothermal vapor provides higher heat transfer coefficients and a less aggressive chemical environment than if the geothermal water itself were used directly in the process equipment. However, it should be recognized that the flash to a lower temperature does impact on efficient overall process design.

b. Geothermal Economics

The cost of a geothermal heat source is directly related to the number of production and reinjection wells that must be drilled; that is, the cost is almost a direct function of the quantity of geothermal brine required. For a single flash of the brine, the required brine flow rate is approximately determined by the following equation:

$$B = S \div \frac{h_{BW} - h_{BF}}{H_{SF} - h_{BF}} \qquad (1)$$

Where: B = required flow rate of wellhead brine, pounds per hour
S = required geothermal steam flow rate to the process, pounds per hour
h,H = brine or steam enthalpy, Btu per pound

and subscripts refer to:
BW = wellhead brine
BF = brine leaving the flash vessel
SF = steam leaving the flash vessel.

The term $(H_{SF}-h_{SF})$ in the above fraction is the latent heat of vaporization of water, which is approximately constant over the temperature range of interest. Thus, the brine flow rate is minimized as the steam flow rate (S) is reduced, or as the liquid enthalpy difference term $(h_{BW}-h_{BF})$ is increased. Unfortunately, as shown below, for fixed reservoir conditions these are contradictory requirements and the optimal design of a geothermally heated process requires a compromise on brine rates and flash temperatures.

Since h_{BW} is fixed by conditions in the geothermal reservoir, the enthalpy difference term is maximized as h_{BF} decreases, which requires low temperatures in the brine flash tank and, therefore, low temperatures of the geothermal steam to the process. Decreasing the geothermal steam temperature can have two effects on a process. Simply stated, multiple effect evaporation economizes on the amount of heat input necessary to evaporate a given quantity of water in inverse proportion to the number of effects. Also, the temperature range required for a set of evaporators increases in proportion to the number of effects, other factors being held constant. It may be seen that lowering the temperature of motive steam to an evaporator train will force either a narrower allowable range of operating temperatures, or a reduction in the number of effects for a fixed temperature range. The former adjustment requires generally smaller temperature driving forces in the evaporator bodies, increasing the heat transfer surface required. The alternative of fewer evaporator effects reduces the steam efficiency of the whole evaporator train, requiring more motive steam, or more geothermal wells ultimately, to supply the required steam.

In arriving at an optimized sugar plant design, several other complicating factors must be considered. In particulr, multiple stages of flash of the geothermal brine will be employed at temperatures appropriate to the needs of the process. This helps further to economize on brine rates. Parenthetically, it would be desirable as well to adapt modernized, high-performance heat exchanger equipment from other industrial applications to use in the sugar refinery, so as to mitigate the reduced thermal efficiency concomitant to the low-temperature heat source.

It should be pointed out that this problem of adaptability of a geothermal heat source to a sugar refinery, or any industrial process, is largely the result of the arbitrary stipulation of 150°C as the maximum geothermal brine temperature to be considered. If higher resource temperatures are encountered, the efficiency of the geothermal process system increases very rapidly. Conversely, lower resource temperatures rapidly decrease the process efficiency. On balance, basin initial studies on a 150°C (302°F) heat source appears to be a good choice. This follows from the fact that such sources are likely to be relatively abundant and from the likelihood that if the sugar process can truly be made feasible for a 302°F source, the potential for application of geothermal heat to various other process industries is considerably enhanced.

c. <u>Sequential Brine Flash to Match Heat Duty-Temperature Distributions</u>

The process streams in a beet sugar factory are heated to continuously higher temperatures, up to the first evaporator effect. Thereafter, heat is added to streams at continuously decreasing temperatures. Geothermal brine will be flashed to provide steam to the first effect of the evaporators at 240°F. Obviously, without yet considering flashing the brine under a vacuum, there can still be a substantial amount of steam produced from the remaining brine. Using a source temperature of 302°F, a second brine flash at 212°F produces approximately 43 percent additional steam, based on the amount formed at 240°F. The geothermal sugar process design presented herein calls for four flashes of the brine, down to a final temperature of 170°F, giving a total of 215 percent of the amount of steam flashed at 240°F. The flash temperatures correspond to steam chest temperatures in the evaporator bodies.

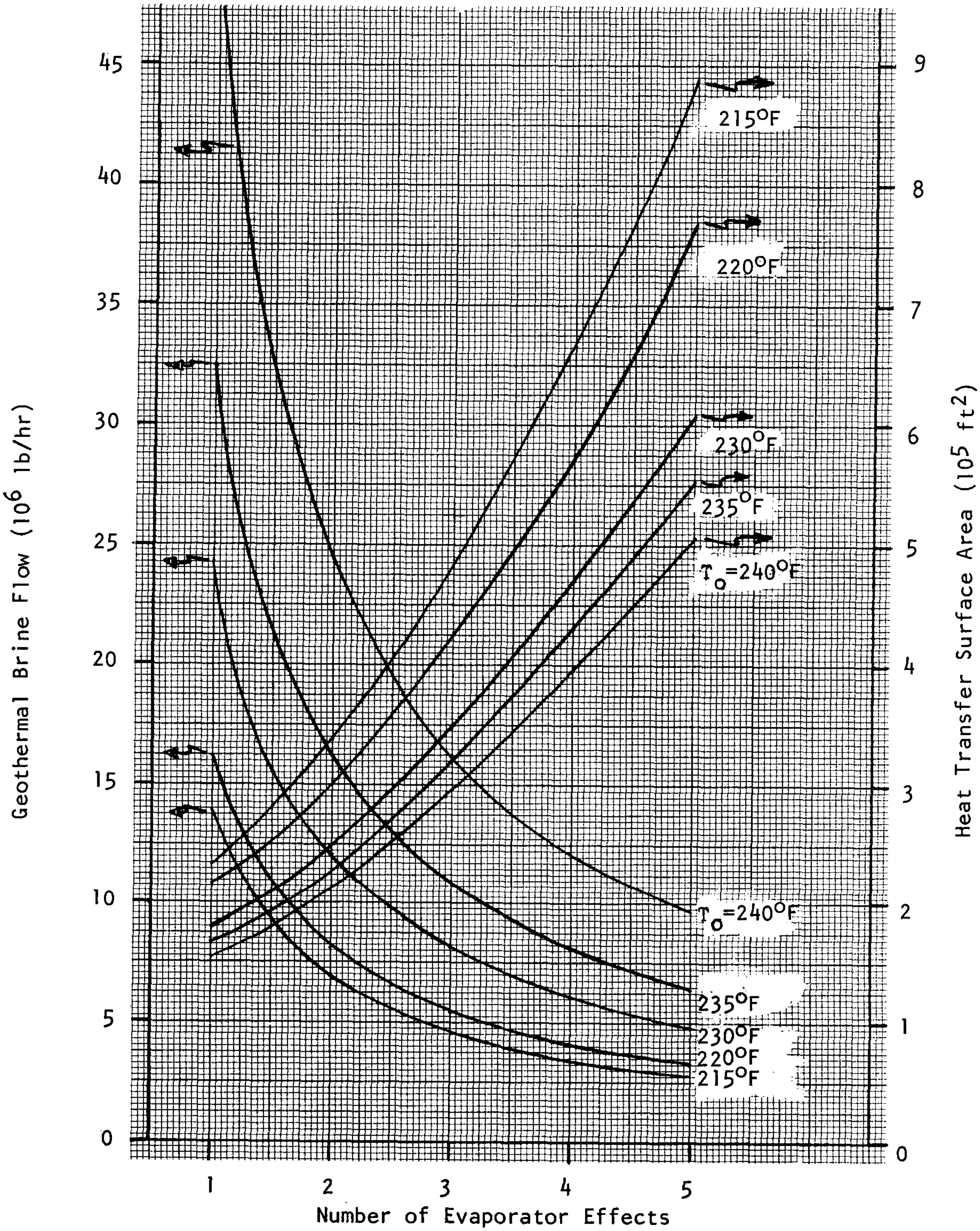
3. Design of the Multiple Effect Evaporator

It is a point of fact that for a given resource temperature, the design of the evaporator train is the most critical factor to the overall beet sugar process design feasibility. Nearly 60% of the sum of all heat duties in the plant is consumed in the evaporators. Also, the greatest flexibility of design parameters exists in evaporation, so the evaporation section design finally determines process feasibility.

In the project analysis, the consideration of the evaporator configuration went through several phases. Originally, it appeared that the relatively low quality of steam available from 302°F water might warrant only a two or three effect evaporator system. This option would have represented a step back in technological sophistication, but the possibility was considered partially because of the anticipated low energy cost. More important, however, for the above resource basis the operating temperature range for geothermal steam driven evaporators has to be narrower than that found in most conventional, fossil-fueled plants. Therefore, reasonable temperature driving forces could not be maintained in a geothermally heated, five-effect evaporator, which would correspond to the most modern design option used today. To be able to quickly quantify and visualize the ramifications of varied geothermal design configurations and operating conditions, a general algorithm was developed to assist in obtaining the optimum evaporator configuration. This is described in more detail in Appendix D. The calculated results have been plotted on Figure 7, which depicts brine flow rate and total evaporator heat transfer surface area versus number of effects. Each curve plotted represents the top brine flash temperature for a given case using a two-stage brine flash, as opposed to the four-stage flash used in the final process design.

The principal efficiency goal in this process design effort is to maintain an efficiency such that the amount of steam supplied to the evaporators also satisfies all other heat duties in the sugar plant, except for the pulp dryer and lime kiln. Thus the evaporator brine supply flow rate should represent all necessary process steam heat for the plant. Referring to Figure 7, it is seen that for fewer than four evaporator effects, the increase in required brine rate accelerates significantly. Also, the incremental benefit of reduced flash temperatures, giving lower required brine rates, gradually decreases with lower temperatures. Conversely, the rate at which required surface area increases with lowered flash temperatures and, therefore, reduced temperature driving forces, is actually accelerating with reduction of the flash temperature. Finally, for increasing numbers of effects, the rate of increase of required surface area for any case tends to accelerate briefly and then becomes linear for more than two effects.

Two additional qualitative criteria entered into the design of the evaporator configuration. First, it was decided that the evaporators should be designed with the first two effects operating above atmospheric pressure. The fact that most dissolved, non-condensable gases would then come out of solution without burdening the vacuum system was important to that decision. Second, in light of the several preceding observations and decisions, we considered the various sugar solution temperatures at the outlets of all heaters in order to best match these with the brine flash temperatures and, to some extent, optimize the resulting temperature driving forces. The final design of evaporator section is the four-effect system shown on Plate 3 of Figure 8. Geothermal brine is flashed initially at 240°F for the first



T_o = Top brine flash temperature, °F
T_n, Last effect juice temperature = 165°F
T_1, Lower brine flash temperature = T_o-25°F

Figure 7. Generalized Multiple Effect Evaporator Design Parameters: Heat Transfer Area and Brine Flow Versus No. of Effects

effect, and then also at the steam chest temperatures of the remaining effects: 220°F, 198°F, and 170°F, respectively. All vapors are utilized in the process for heating purposes. All condensate from the first three effects is reflashed at the temperature of subsequent evaporator steam chests to supplement the brine and juice vapors evolved at the corresponding levels.

E. Summary of Design Factors

The various important process control parameters, as have been presented briefly in the preceding sections or which were deferred for detailed presentation in Chapter VI, were utilized to develop a conceptual "final" design for a geothermally heated beet sugar refining facility. The complete process flow diagrams for that facility are given in Plates 1-4 of Figure 8, and these designs represent the most comprehensive application of geothermal energy to date to all of the various interrelated steam energy demands apparent in the beet sugar refinery industry. The most important final design considerations are summarized below.

Four primary energy inputs to a beet sugar plant can be identified. These consist of electrical energy to operate mechanical equipment; coal, gas, or oil to produce steam for the evaporators and heaters (this steam is taken from the turbine exhaust); coke for the lime kiln that supplies the juice purification system; and gas or oil for the pulp and sugar driers. A schematic of typical energy use patterns appears in Figure 6.

The important differences to be noted when comparing existing plants with, first, a "typical" plant design, and then with a modified plant design using geothermal energy, would be found in the schemes of heat utilization. Heat efficiency is expressed as a weight percent of pounds of steam at 212°F per pound of beets, so that a low value represents high efficiency. Older plants typically had a "percent steam on beets" value of the order of 80 percent. Modern plants, on the other hand, can achieve efficiency values near 40 percent steam on beets. Contemporary American plants average about 75 percent steam on beets. The depressed state of the domestic sugar industry has led to cancellation of at least two new plants considered in the past five years. However, if and when a new plant were built, it would probably be designed for about 55 percent efficiency.

In the final, detailed preparation for specifying a conversion to geothermal energy, it was ascertained that the operating parameters of the process are fairly narrowly constrained by either simple economics or by the practical considerations of sugar quality and purity, so that complete freedom cannot be used in the redesign of a plant to use geothermal brines. Temperatures in the evaporators, the hottest section in the process, must be kept as high as possible to facilitate high temperature driving forces between vapors from each effect and the juices in subsequent effects or other heating devices. High temperatures also effect high heat transfer coefficients by reducing liquid viscosities. These factors tend to minimize the needed heat transfer surface area. Contrarily, the sucrose can be degraded or discolored by high temperatures, with the upper temperature limit for dilute sugar solutions being about 270°F. Thus, when using a geothermal source for process heat, the plant design is very similar to any existing plant, and the standardly defined plant efficiency will be strongly dependent on the geothermal resource temperature.

The following chapter goes into sufficient detail on the various design factors and practices to thoroughly show how operating parameters are seriously constrained within very narrow limits. As a result of those constraints, the geothermal sugar plant design was developed without changing the diffusion, purification, or crystallization sections. Process designs and heat loads were conservatively developed for these operations based on current practices.

A considerable effort was devoted to developing an efficient design for the evaporator section. Several flow sheets were analyzed, ranging from two- to five-effect evaporator systems operating at various temperature levels; a four-effect evaporator was subsequently chose. Vapors taken from any one of the evaporators were reused in other heaters with compatible temperature requirements. Various vapor recompressions schemes were also analyzed in an attempt to reduce energy consumption, and none were found to be beneficial due to the high capital costs associated with the type of equipment. Although the final flow sheet developed is not completely optimized because of the large amount of effort required to arrive at an optimized system, it is believed to be close to the best design. Preliminary calculations indicate that geothermal brine requirements possibly could yet be reduced by an additional 10 to 30%. The geothermally heated process design presented herein was reviewed by sugar industry representatives. To the extent possible in the scope of this study, this basic process design is found to be practical and realistic.

F. Design Economic Evaluation

In the final design plant (Plates 1-4, Figure 8) a brine feed at 302°F was assumed to lose 5°F in temperature before reaching the plant. It then was flashed four times to provide steam for the various units of a four-effect evaporator unit, with the maximum and minimum flashed steam temperatures being 240 and 170°F. By comparison, a modern sugar plant will have a top process steam temperature of about 290°F, and will require much less heat transfer surface area as a result of increased temperature differentials. It should also be noted that the limited initial flashdown available on the brine (from 297 to 240°F) requires that a large amount of brine be provided per unit steam production. A total of six production wells (4.60×10^5 lbs/hr per well) and 3 injection wells were required. A spare was provided in each case. Drilling costs were estimated at $650,000 per 10,000 foot well.

Comparative heat transfer surface areas were calculated for both the geothermal driven plant and also the fossil fueled plants. The incremental costs associated with the increased heat transfer area were then derived using net installed costs as shown in Table 1. Since the stream flows and piping configuration are basically unchanged in a geothermal plant, no additional capital expense was included for such factors. In certain instances, retrofit costs for modified piping, electrical systems, etc., may be somewhat higher due to site specific space limitations. Additional costs must be included for geothermal brine delivery systems, including transfer piping, pumps, and flash equipment.

The comparative, equivalent energy cost calculations in dollars per million Btu are presented in Table 2. Since the overall steam quality supplied to a plant in the geothermal case differs significantly from that in a fossil fuel plant, the cost of the geothermal system is presented as the cost incurred to replace a specific fossil fuel heat load.

As a basis for comparing costs between the geothermally heated plant and fossil fuel fired sugar factory, three different efficiencies of a conventional plant were used. Highly efficient plants with steam consumption as low as 44% on beets are found in Europe, but are not likely to be built in this country even at present cost relationships between energy and materials. This case, then, is the extreme design situation for comparing to a geothermal system and leads to the highest equivalent geothermal costs. It is Case 1 in Tables 1 and 2. Case 2 in these tables is a conventional plant with an efficiency of 55% steam on beets, which is the level of technology likely to be achieved in a new sugar factory built in the United States (3). This case provides the most realistic comparison to a new geothermally heated factory. Case 3 of Tables 1 and 2 provides a comparison to the typical, low efficiency beet sugar factory in operation in the United States today which consumes steam at the rate of 75% on beets. This case provides the best comparison for a retrofit situation. For each case, a beet slicing capacity of 5,000 tons/day and a campaign length of 120 days/year are used as a basis for calculating comparative costs.

Table 1 indicates the capital cost attributable to a geothermally heated factory, based on the design previously discussed in this chapter, that would be built in place of fossil fuel fired factories with the indicated steam efficiencies. The capital costs take consideration of the geothermal production and injection wells; distribution lines and pumps; brine flash systems, which include flash and surge tanks, and pretreatment equipment for acidification; and lastly, additional in-plant sugar processing equipment. This latter category includes three items: the cost of additional heat transfer surface area; an air compressor to provide high pressure air in place of high pressure steam to clean the beet slicer; and a vacuum pump to replace the steam jet ejector. The vacuum system capacity increases, as well. It should be noted that no credit is taken for the cost of the furnace and its associated pollution control system, or for the cost of the vacuum system, in a conventional plant. When developing the total cost of a fossil fuel fired system to compare to the geothermal costs derived below, these factors should be added to the cost of the raw fuel.

As shown in Table 1, the geothermal system incurs a capital cost of the order of 10 million dollars for each of the three cases. This cost is heavily weighted by the cost of wells, which ranges from about 70 to 74% of the total. On the basis of these costs, and assuming annual charges on capital totalling 15%, Table 2 presents the cost of the geothermal system in terms of dollars per million Btu replaced. Section A of Table 2 determines the hourly fuel consumption, as fed to the boiler, for each of the three design comparison cases. This value ranges from 209 million Btu/hour for the high efficiency Case 1, to 355 million Btu/hour for the lowest efficiency Case 3.

Geothermal system costs, expressed as incremental costs relative to the fossil fuel plant, are indicated in Section B of Table 2. These include the annualized capital costs reduced to an hourly basis for a 120-day campaign, the costs of brine treatment, and the cost of increased electricity consumption, at 3 cents/kwh, to run the air compressor and the vacuum pump. A conservative brine treatment procedure has been developed to prevent calcium carbonate scaling, which is expected to present the only scaling problem encountered for the typical brine chemistry shown in Table 22 of Appendix C. For this purpose, the brine is neutralized with sulfuric acid to release carbon dioxide gas, and then treated with sodium hydroxide to raise the pH to non-corrosive levels. Assuming a cost of 10 cents/pound for both acid and base, this gives a brine treatment cost of about $25 per million pounds.

On these bases, Table 2 indicates that the equivalent cost of the geothermally heated system, in terms of the amount of fossil fuel energy replaced, is \$2.85, \$2.34, and \$1.77 per million Btu, for Cases 1, 2, and 3, respectively. Several assumptions made in the study have a strong effect on the absolute values of these costs, some of which will be reexamined below. The most significant is the length of the campaign over which the total capital cost is amortized. If the campaign length were 200 days, as can be achieved in locations such as Southern California, the geothermal energy cost would be \$1.52 for Case 2 (55% efficiency) instead of \$2.34 per million Btu. This is the strongest single factor in this analysis. Other than increased campaign length, unit costs would also be reduced dramatically if a multi-purpose development utilized energy from the wells all year long. With respect to the costs of resource utilization, the effect of following a multi-purpose development program is the same as extending the operating campaign of sugar plant. The other most significant parameter affecting costs derived in this study was the upper limit of 302°F imposed on the geothermal brine temperature. Process capital costs would decrease rapidly if this temperature rose to 400°F, for example, with all other factors unchanged. Finally, the geothermal resource development costs depend strongly on well costs. In other geothermal areas, where well depths are much less for the same temperatures, these costs will decrease.

A few final words are in order regarding the future availability, ease of utilization, and cost of natural gas, fuel oil, or coal. These fuels are currently competitive with the type of geothermal system developed in this study. However, in view of the apparent national energy shortage, the cost of these resources is conservatively often projected to rise at rates of 7% to 10% annually. A geothermal plant, on the other hand, which exhibits low fuel-related materials and operating expenses, would not experience this same cost trend. But by virtue of inflation, the cost of developing a geothermal resource is, itself, increasing at probably more than 5% annually. Also, the use of natural gas as an industrial boiler fuel is inconsistent with the national energy programs being outlined. Finally, the combustion of either oil or coal necessitate significant additional capital and operating expenditures for environmental control systems. Particulate collection, sulfur dioxide removal, and water quality control are currently those areas of greatest concern; all have grown markedly more restrictive in recent years, and all hold broad uncertainties for the future. In the sense that geothermal energy often times does not present environmental problems of the same severity and with the same current emphasis, a properly designed geothermal system offers attractive advantages.

In conclusion, in view of the favorable economics demonstrated, further industrial consideration of this resource is merited--both for retrofit applications and for any new plants to be built, such as might be located in the San Luis Valley.

Table 1

Incremental Beet Sugar Plant Costs

Bases:			
-Case	1	2	3
-Efficiency of Fossil-Fuel Plant (% Steam on Beets)	44	55	75
-Brine Consumption of Geothermal Plant (10^6 lb/hr)	2.8	2.8	2.8
-No. of Production Wells	6	6	6
-No. of Brine Disposal Wells	3	3	3
-No. of Spare Wells (1 production, 1 disposal)	2	2	2
-Heat Transfer Surface Area (sq. feet)			
-For Geothermal Plant	161,800	161,800	161,800
-For Fossil-Fuel Plant	76,100	60,800	48,300
-Incremental Geothermal Surface Area (sq. feet)	85,700	101,000	113,500
Costs: ($1,000)			
-Well Cost	7,150	7,150	7,150
-Transfer Lines and Pumps	760	760	760
-Incremental Surface Area Costs*	1,610	1,897	2,132
-Additional Equipment Costs			
-Flash and Surge Tanks, Pretreatment Systems	325	325	325
-Air Compressor (for Beet Knife Cleaning)	50	50	50
-Vacuum Pump (to replace Steam Jet Ejectors)	68	78	93
-Net Incremental Capital Costs for a Geothermal Plant	9,963	10,260	10,510

* Costs: Evaporators (12.50/ft^2); Vacuum Pans (\$105/ft^3); Heaters (\$15/ft^2).

Table 2

Unit Costs of Geothermal Energy for Beet Sugar Processing Plants

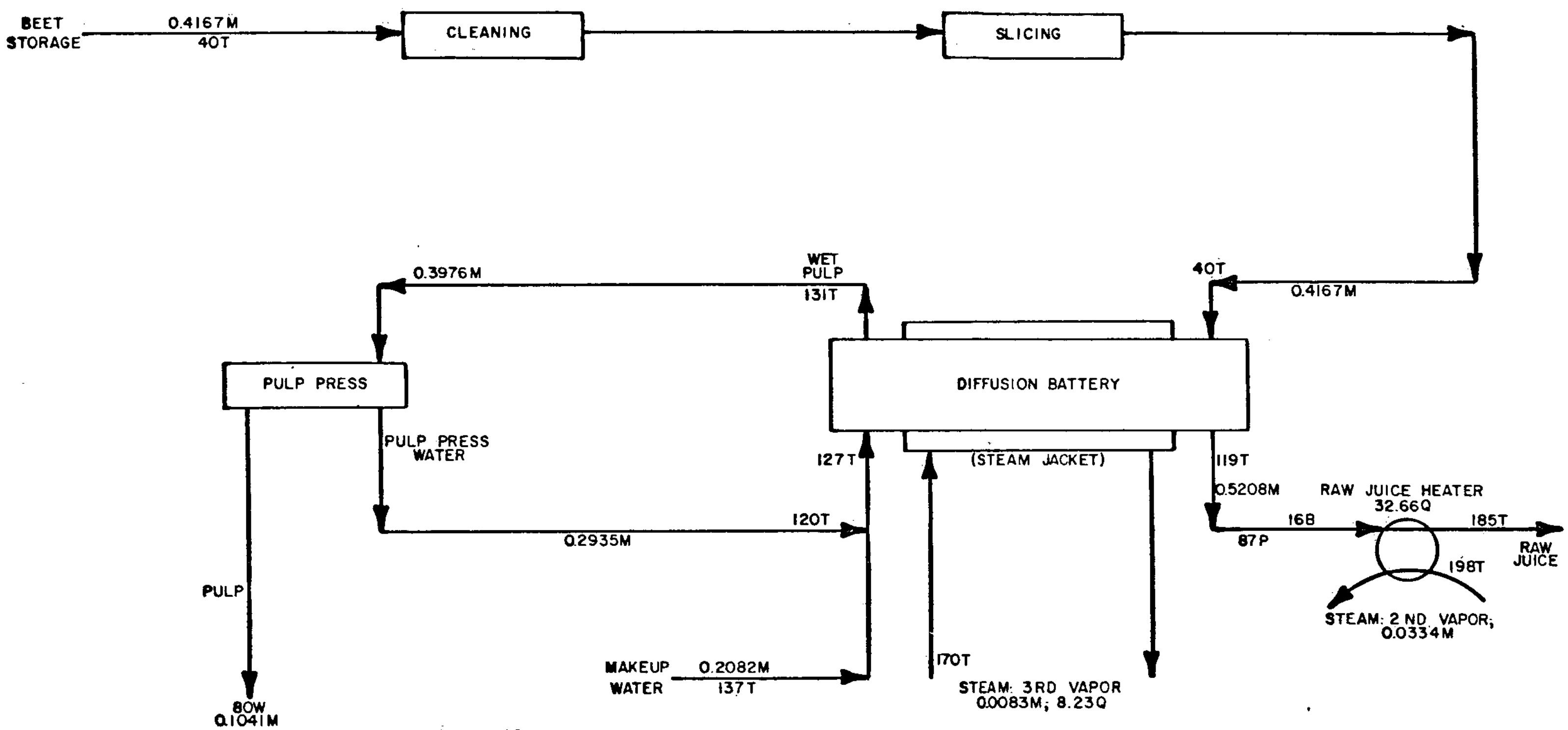
Basis: Replacement of equivalent fuel requirement for a fossil fuel fired beet sugar factory.

		1	2	3
A.	Energy Usage in Fossil Fuel Plant			
	1. Plant Efficiency (% Steam on Beets)	44	55	75
	2. Energy to Process (10^6 Btu/hr)	177	222	302
	3. Boiler Efficiency	0.85	0.85	0.85
	4. Total Fossil Fuel Requirement (10^6 Btu/hr)	209	261	355
B.	Costs of Geothermal System			
	Case			
	1. Annual Amortization (1000$)*	1,494	1,539	1,577
	2. Hourly Amortization ($/hr)**	519	534	548
	3. Brine Treatment ($/hr)***	70	70	70
	4. Additional Electrical ($/hr)	6	7	9
	5. Total Geothermal Hourly Costs ($/hr)	595	611	627
C.	Cost of Geothermal Systems			
	B5÷A4 ($ per 10^6 Btu replaced)	2.85	2.34	1.77

* 15% of Capital Charges
** 120-day processing campaign
*** $25 per million pounds

FIGURE 8; 1 OF 4

SUGAR BEET PROCESS DIFFUSION SECTION



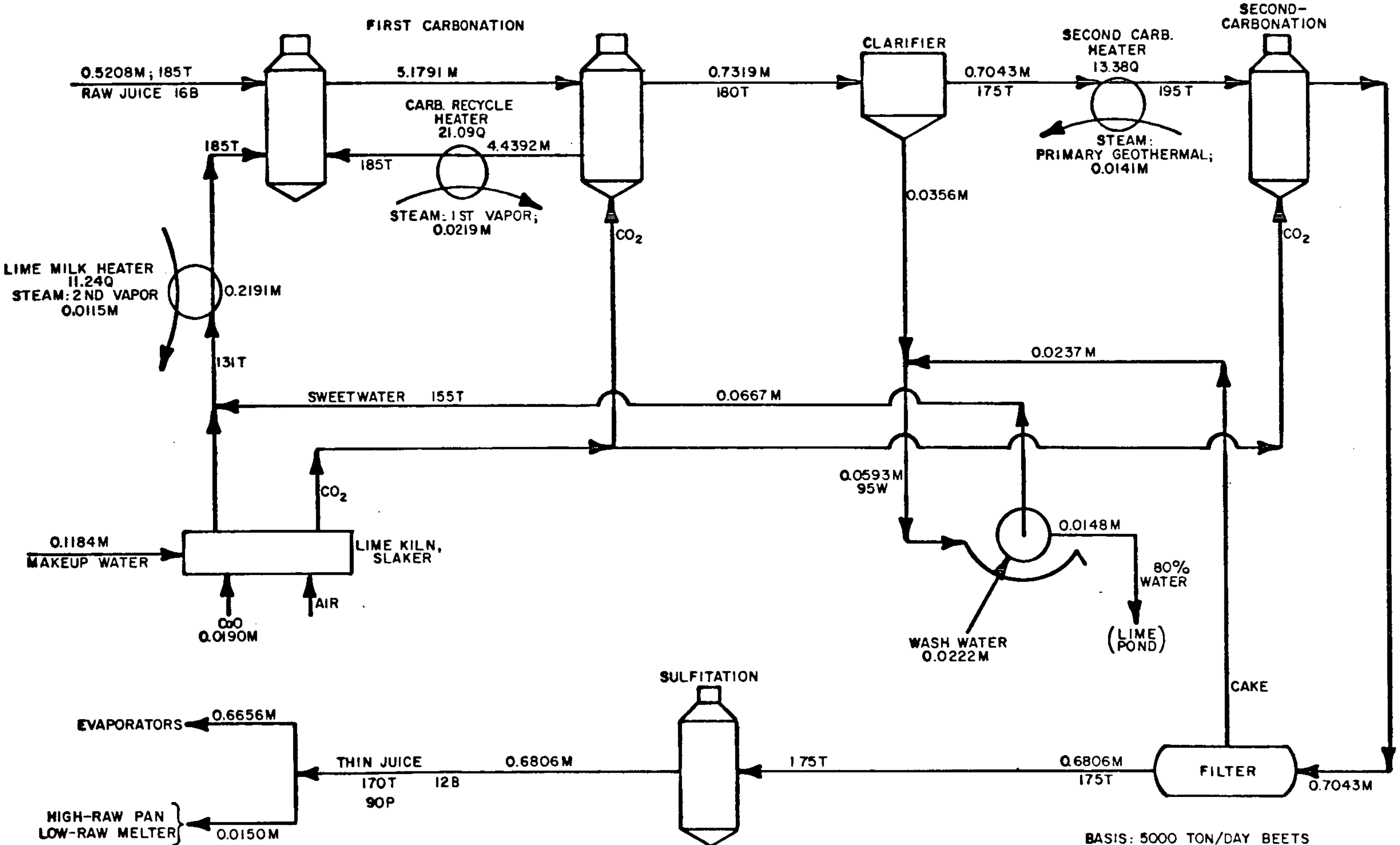
NOMENCLATURE

M – 10^6 LB/HR
T – °F
Q – 10^6 BTU/HR
B – °BRIX
P – PURITY, % WGT. SUCROSE IN SOLIDS
W – % WGT WATER

BASIS: 5000 TON/DAY BEETS
BEETS – 18% SUCROSE
98.2% SUCROSE EXTRACTION

FIGURE 8 ; 2 OF 4

SUGAR BEET PROCESS PURIFICATION SECTION



NOMENCLATURE

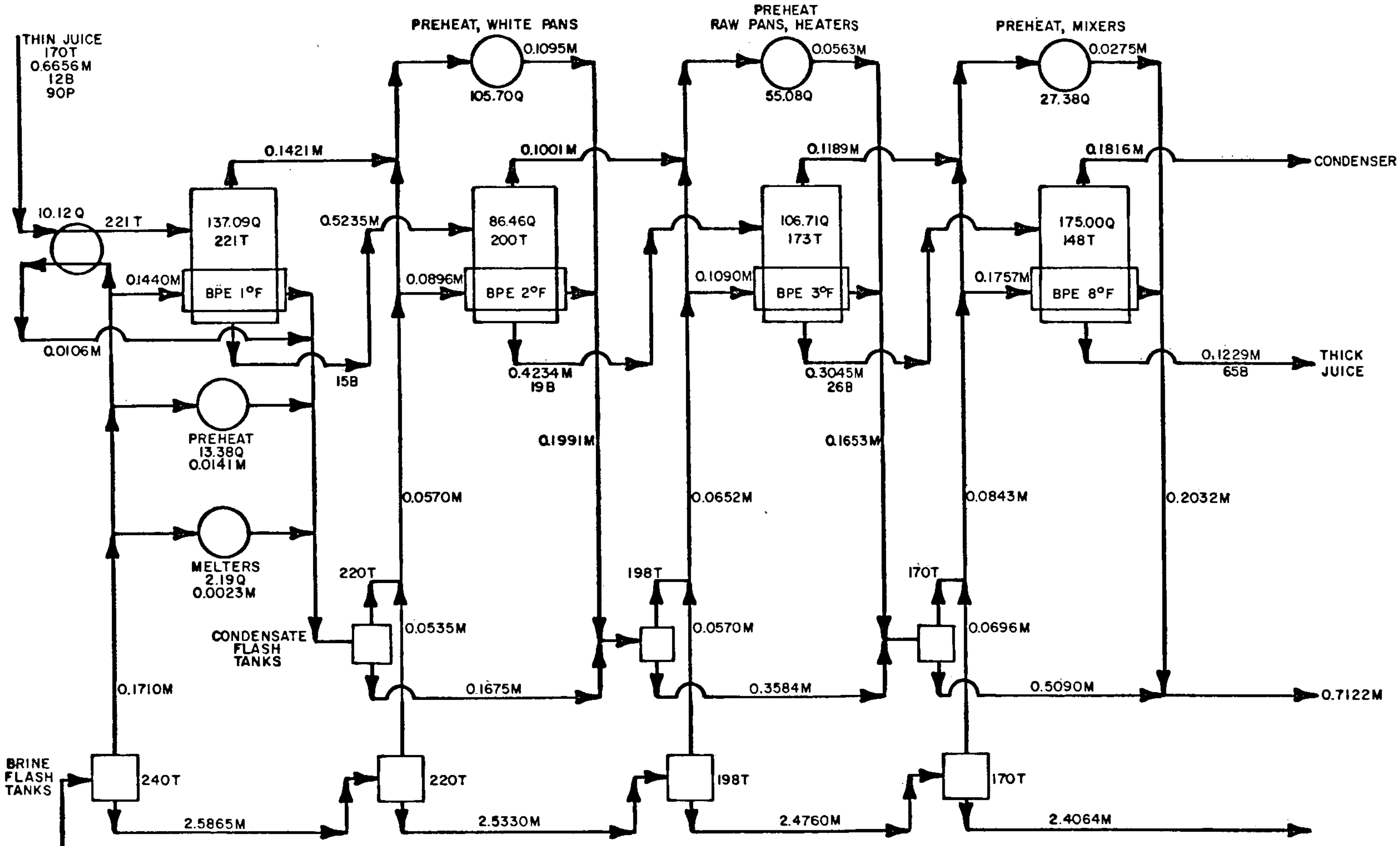
M - 10^6 LB/HR
T - °F
Q - 10^6 BTU/HR

B - °BRIX
P - PURITY, %WGT. SUCROSE IN SOLIDS
W - % WGT WATER

BASIS: 5000 TON/DAY BEETS
BEETS - 18% SUCROSE
98.2% SUCROSE EXTRACTION

FIGURE 8-3 OF 4

SUGAR BEET PROCESS EVAPORATION SECTION



NOMENCLATURE

M-10⁶ LB/HR B-°BRIX

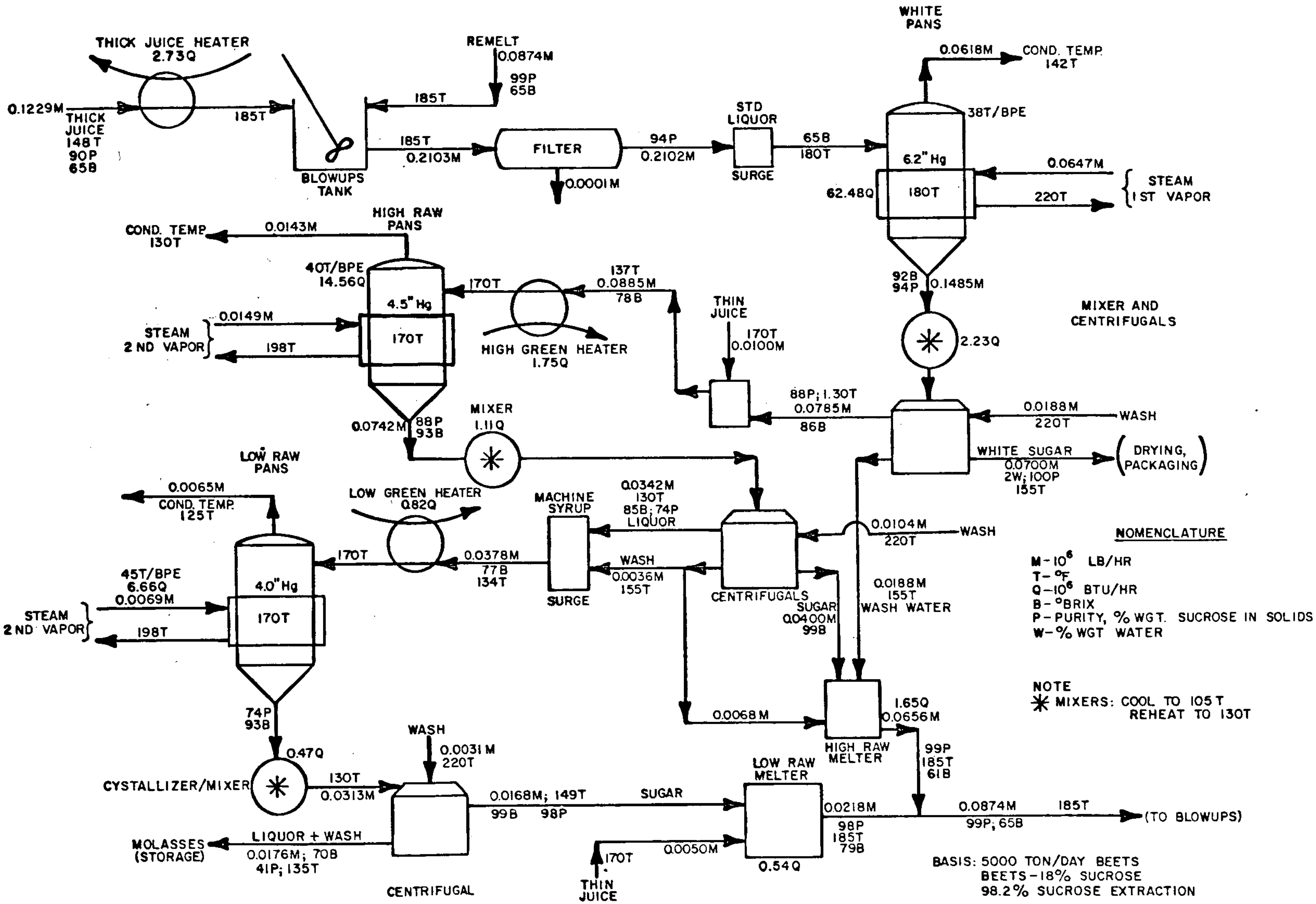
T-°F P-PURITY, %WGT. SUCROSE IN SOLIDS

Q-10⁶ BTU/HR W-% WGT WATER

BASIS: 5000 TON/DAY BEETS
BEETS-18% SUCROSE
98.2% SUCROSE EXTRACTION

FIGURE 8; 4 OF 4

SUGAR BEET PROCESS CRYSTALLIZATION/SEPARATION



39

VI. THE SUGAR BEET REFINING PROCESS--A DETAILED DESCRIPTION

A. Purpose and Summary of Contents

The purpose of this chapter is to present more details on the specification of process-controlling design parameters of the beet sugar refining process. These details, when applied in formulating a process design, act as an empirical guide to the proper design of a sugar plant. This presentation is, therefore, groundwork for the geothermal process design produced for this project.

The point this chapter makes is that there are two factors which conspire to constrain the potential flexibility of key design parameters. First, the physio-chemical characteristics of sugar beets, sucrose, and the substances that constitute process juice impurities each pose specific problems in the four sections of a beet sugar plant: diffusion, purification, evaporation, and crystallization. As is frequently the case, the solutions to the various physio-chemical problems with respect to sucrose, impurities, and beets conflict with one another. Consequently, plant operating conditions will be determined to effect the greatest economy of final sugar cost, achieving a compromise in costs accrued for raw materials, capital equipment, fuel, and labor. Second, the quest for economy of energy consumption inherently leads to some increased degree of design complexity. For example, energy efficiency within a process may be facilitated by cascading heat--that is, heat is "used" more than one time by transferring both to and fromprocess streams, rather than specifically burning fuel to accommodate every instance requiring energy input. But, it is easily seen, as process streams are used to heat and cool one another, we are increasingly sacrificing degrees of freedom for complexity and heat efficiency. In order to identify these complex interrelationships and to insure that they could adequately be accounted for in the detailed geothermal design, considerable effort was spent to attain a comprehensive understanding of the refining process.

Depending on how thoroughly the reader desires to delve into the sugar refining process, this chapter may be considered optional reading. The principal topics discussed include Beet Composition, Beet Preparation, Sucrose Diffusion, Juice Purification,and Product Crystallization. A detailed discussion of evaporation is included in the preceding chapter. Detailed flow diagrams of the process are given in the four plates of Figure 8 immediately preceding this chapter for ready reference.

B. Beet Composition

First, it would be worthwhile to look briefly at the chemical constituents of the sugar beet. Sugar beets range from 10 to 22 weight percent sucrose; the 1968 North American median value was 15.5 percent. This study uses a value of 18 percent.

The beet root, the portion of the plant used for its sugar content, consists of approximately 5 to 8 percent insoluble substances making up the cell walls. This insoluble portion is called the "marc," consisting in part of cellulose and related structural substances. About half of the dry marc is insoluble pectic materials, which form undesirable colloids in water. In its natural form the marc contains about 50 percent bound water. The solid remains of the beet after removal of its sugar, are called pulp in processing terminology. The pulp will be dried to remove about 80 percent of its water and then sold for livestock feed.

The remainder of the beet is considered the juice phase. It is approximately 75 percent water and 25 weight percent solids in solutionor colloidal form. The bulk of the dissolved solids is sucrose; the remainder consists of organic and inorganic substances. The soluble organic solids can be categorized as nitrogen-free and nitrogenous substances, the latter being mainly protein material. The protein material, and the pectins in the marc, are major problem sources with respect to operation of the purification section of the beet sugar process, as well as being critical to the final sugar purity. Colloidal suspensions formed by pectins and proteins are difficult to eliminate, and they interfere with crystallization if not substantially removed; trace amounts of these substances in the final product lead to problems such as coloring in the sugar and heavy foaming in carbonated beverages made with the sugar.

Ash, or inorganic substances in the beets constitute about 0.5 percent of the total weight, much of which is present in the juice phase. The principal cations include potassium, calcium, magnesium, and sodium; the anions consist primarily of phosphate, chloride, and sulfate. Calcium carbonate is the major scaling problem on heat transfer surfaces. Sodium and potassium are largely non-removable and non-scale forming, but they do at least affect the size of the molasses stream, which contains the blow-down impurities and "unrecoverable" sugar. Consequently, the latter impurities are the least significant with respect to their quantity present, and their effects, and they are cause for correspondingly less concern regarding juice purification.

Finally, as noted earlier in Chapter V, the precise operational characteristics for any particular beet sugar plant are site specific, based on the singular physiochemical and biological characteristics of the beets processed. Such distinctions can arise simply because the differing soil and climatological conditions from region to region will vary the physical characteristics of beets within one strain sufficiently to induce corresponding variations in sugar plant performance. More important in this regard, however, is the fact that altogether different strains of beets are usually found coming from various regions, adding another degree of specificity to the development of optimal process designs.

C. Beet Preparation

The beets are prepared for processing by first passing them through a number of cleaning operations. Whole beets from storage are introduced into a water flume that serves to wash the beets and transport them over screens and dropout bins where rocks and sand are removed. Most encrusted dirt is washed off by tumbling of the beets in the flume. The flume water is desired to be as cold as possible, preferably not warmer than 85-100°F if the beets are not frozen, in order to minimize sugar losses by leaching. A plant processing 4,000 tons of beets per day can lose over two tons of sugar per day to the flume water.

Flume water is recycled as much as possible, with a blowdown stream being removed to eliminate mud and silt which would otherwise quickly settle and block the flume ways.

Final cleaning is performed by beet feeders, trash catchers, magnetic metal separators, picking rolls, picking tables, and pressure-water beet washers. Having passed through this series of equipment, the beets emerge free of leaves, grass, and weeds, with essentially all dirt and bacteria washed from the outside. It is important that the beet handling equipment not abrade or bruise the beets, because damaged beets will experience significantly higher sugar losses to the flume water. Also, beets that have deteriorated badly in storage may contribute more impurities or bacteria than sugar. Such damaged or aged beets are selected and removed manually by workers at the picking tables.

D. Sucrose Diffusion

Sucrose is removed from beets by counter-current extraction in a unit called a continuous diffuser. The diffusion process is facilitated by slicing the beets into long, thin strips called cossettes, to increase the surface area available for contact with the diffuser water. The cossettes are carried into one end of the diffuser and are moved through the diffuser by the action of one or two scrolls. Cossettes leave the diffuser as beet pulp. The diffuser body is either slightly slanted or vertical. Supply water, which is mostly condensate with some fresh water and recycled juice from the pulp, enters the upper end of the diffuser, flows over the cossettes by gravity, and leaves the diffuser as raw juice. Raw juice will contain 10 to 15 percent sucrose, which comprises about 98 percent of the sugar in the beets. The diffusion section schematic is shown as Plate 1 of Figure 8.

Beet slicing is performed to maximize surface area contacted by the diffusion juice, while at the same time taking care not to damage the beet cell walls. The cossette shapes have been carefully studied and tested to achieve optimum diffuser performance. Damaged beet cell walls result in increased concentrations of impurities in the raw juice; these impurities ordinarily are restricted by limited diffusion through the cell walls.

The beet cells contain a fluid which has sugar dissolved in it. A semipermeable, protoplasm membrane retains the cell fluid. Above about 120°F the protoplasm becomes irreversibly permeable to the cell fluid, which may then diffuse into the low-centration diffusion juice. The irreversible transfer process is not strictly osmosis, although there are strong similarities. The rate of diffusion increases with temperature because proteins in the protoplasm coagulate more completely, making the protoplasm membrane increasingly permeable. For a given overall rate of sugar recovery, heightened temperatures tend to decrease the required residence time for the cossettes; or to allow the beets to be sliced less finely, thereby reducing the number of broken cell walls and reducing the extraction of impurities; or to allow the use of less diffuser supply water, resulting in a more concentrated raw juice.

However, increased temperatures also lead ultimately to breakdown of the beet cell walls such that several serious disadvantages may outweigh the advantages of increased sugar extraction rates. First, the pulp retains increased amounts of water and becomes mushy with higher temperatures, making mechanical dewatering of the pulp difficult and causing pulp screens to become clogged. Secondly, diffusion may actually be impaired by the strongly altered state of the cell walls, and sugar recovery suffers as a direct result. Third, increased amounts of colloidal substance become suspended in the diffusion juice and place an increased duty on the purification section. Similarly, lengthened retention times in the diffuser will also produce the above effects, as well as lending a higher risk of bacterial infection of the sugar solution.

The ratio of the weight of raw juice to the weight of beets, expressed as a percent, is called the draft. The draft commonly ranges from 100 to 150. The higher the draft, the lower will be the sugar losses to the pulp. But then the heating loads for the purification and evaporation sections will be commensurately increased by the extra water content.

The beet pulp leaving the diffuser is first screened and then pressed to remove the entrapped juice. The pulp press water, as it is called, will generally contain less than 2 percent sucrose, and it is recycled to the diffuser at a point where the diffusion juice has about the same amount of dissolved solids. The desirability of recycling this juice is primarily due to reduced wastewater biochemical oxygen demand loads and improved pulp recovery that result. Sugar recovery is only slightly enhanced by the recycling of pulp juice, but the reintroduction of the impurities contained in the pulp press water also has only a small effect on the net raw juice purity. The pulp contains a fractional weight percentage of sucrose and also has a valuable nutrient content. The pulp is normally dried and pelletized for use as livestock feed. Molasses, a liquid byproduct of the sugar refining process, is usually added to the pulp pellets as a binder, as well as for recovery of its own nutrient and sugar content. Molasses will be discussed further in a later section.

Thus, it is shown that the selection of operating temperatures, cossette and juice residence times, and diffusion juice flow rates depends on a balance of complex and somewhat conflicting factors.

E. Juice Purification

The raw juice impurities must be removed to facilitate high levels of uniform crystallization of the sucrose, and secondly to meet stringent purity standards. Impurities interfere with crystal growth, and thus can significantly reduce the final recovery of sucrose. Consumers, particularly industrial buyers, require sugar of very high purity and especially low coloring content. Additionally, the presence of impurities has expensive, detrimental effects on the efficiency of operation and the maintenance requirements of the evaporators, heat exchangers, and crystallizing equipment since heat transfer equipment is constantly subject to oxylate and carbonate scale buildup. Even under nominally correct operating conditions, heat transfer surfaces in some plants are subject to weekly cleaning requirements in order to maintain plant throughput. Thus, a balance must be made between capital and operating costs for purification and increased income via higher sugar recovery.

1. Nature of the Impurities

As described in the earlier section on beet compositions, the impurities comprise organic and inorganic substances from the juice phase and the marc of the beet. Organic pectins from the marc are taken into the process sugar solution in colloidal form as a result of broken or damaged cell walls. Some damage is inevitable in the handling and slicing steps. High temperatures and long diffuser residence times for the cossettes also will result in higher concentrations of the pectins in the raw juice leaving the diffuser.

The impurities present in the beet juice phase will find their way into the process sugar solution, along with the sucrose, as a result of the breakdown of cell membranes in the heat of the diffusion juice. The extent to which different species, including sucrose, permeate the cell membrane is dependent on their molecular size, chemical nature, and the degree of coagulation of the proteins in the membrane.

2. Removal of the Impurites

Purification involves a sequence of several steps as shown in Plate 2 of Figure 8. The sequence is important because of the reversibility of certain precipitation reactions that occur; for instance, it is possible to precipitate or coagulate certain undesirable solutes and colloidal substances only to have them redissolve or repeptize in a subsequent treatment step. This process section may be divided into three subsections: first carbonation, second carbonation, and sulfitation. First carbonation includes clarification, and second carbonation utilizes filtration. The purification process varies in minute details on site-specific bases, reflecting the varying characters of beets received as well as the age of the plant. Design parameters have been determined empirically over the period of slightly less than two centuries since beet sugar recovery was industrialized. The process mechanism depends on the use of lime to precipitate solids that may then be removed by settling and filtration.

a. First Carbonation

First, juice from the diffuser is heated to approximately 185°F and pumped to the first carbonation station. Lime is added to the juice in the form of an aqueous solution/suspension. The most common practice involves a gradual addition of lime and two phases of reaction. In the first phase, raw juice is blended with a lime solution such that there is less than 0.5 weight percent CaO present, based on the mass of the juice stream. This is called preliming or predefecation, and commonly is performed in a separate unit called the primary first carbonation tank. Initially the natural acidity of the juice is partially neutralized, and various ionic and colloidal impurities form low-solubility complexes or salts with calcium. These precipitation reactions occur relatively quickly. In the case of organic species such as proteins, which tend to form colloids, it is important that initially only a small amount of lime is added. The slight pH change that occurs has been found to best facilitate a modification of some of the organic colloids present such that they will form quite insoluble calcium compounds. This phase of purification is extremely sensitive to both the rate and net amount of lime added. With rapid addition, rather than ionizing some of the lime also tends to form colloidal molecules, especially in the presence of sucrose. With changes in juice pH at later stages in the total refining process, these colloids may be "broken," resulting in calcium supersaturation and probable scale occurrence. Furthermore, the unpredictable formation of lime colloids makes precise pH control difficult, which also implies difficulty in controlling the stoichiometry of lime addition. Gradual or mild "liming" suppresses the lime colloid formation, and the precipitates formed via mild liming are found to have superior physical characteristics for settling and filtration as long as juice temperatures are maintained above 160°F. Resultant savings in water and lime consumption, and required filter area are substantial. However, at and above about 195°F, problems occur similar to those caused by rapid or heavy lime addition, restricting allowable upper juice temperatures.

Following preliming the raw juice is transferred to the secondary first carbonation tank. The second phase of reactions involved in the liming of the raw juice is allowed to continue here. The reactions are not all well understood. Lime and other of the nonsugars may react slowly to form soluble complexes; these particular reactions are favored by high lime concentration and high temperatures. Other reactions are known to include hydrolysis of organic compounds such as amides into corresponding soluble salts that are difficult to remove; the further reaction of sucrose degradation products to colored and acid substances in forms that can be adsorbed or precipitated; and the release of oxalic acid, which is a potential scaling agent. It is thought that some of the pectin and protein precipitates formed in preliming react and repeptize as colloids. Finally, increasing alkalinity favors formation of complex sugar compounds, such as sucrocarbonates, that form long-chain gels in combination with calcium and may cause sugar losses. Obviously these reactions are not all desirable. Mild liming generally prevents excessive reaction of the impurities to their intractable forms and also inhibits the precipitation of sucrose-impurity compounds.

Carbon dioxide gas is bubbled into the secondary tank. The main objectives are the removal of excess lime and pH control. Much of the organic, nonsugar impurities existing as large, complex molecules in colloidal form are adsorbed onto the $CaCO_3$ precipitate. Of special benefit here is the adsorptive removal of coloring agents, which principally consist of sucrose degradation products. In general, coloration of the sugar product is minimized beyond this point in the refining process by carefully avoiding both acidic conditions and excessive temperatures hereafter.

As a final aspect of the first carbonation process, the secondary tank contents are recycled to the first tank in a ratio averaging about 7:1 based on the rate of juice withdrawal from first carbonation. This recycling allows a compromise of moderate temperature and alkalinity versus ample retention time. By returning the juice for further lime addition, the slower reactions are given time to proceed in mildly alkaline conditions conducive to effective purification and tolerable precipitation sugar losses, while raising the lime content to proper stoichiometric values.

The raw juice leaving first carbonation goes to a clarifier, and the resulting clear liquid is sent on to second carbonation. The sludge is vacuum filtered and washed to remove adsorbed sugar. The final sludge is discarded, usually as landfill. The wash, which is called sweetwater, contains enough sucrose to justify recycling and it is blended with milk of lime entering the primary first carbonation tank.

b. Second Carbonation

Ideally, the raw juice from first carbonation is a very high quality product. The purpose of second carbonation is simply to remove the remaining excess lime. There are two factors critical to the efficiency of this final deliming, both related to prior formation of acid substances. First, the availability of the carbonate ion to precipitate with calcium is dependent on both pH and alkalinity. Alkalinity may be adjusted, as required, by addition of small amounts of soda ash. The pH will seek its own level with adsorption of CO_2 gas. Second, as described earlier, calcium is subject to being bound with other species in soluble complexes and rendered unavailable for precipitation during purification, thereby remaining as a potential scaling agent later in the process. The calcium carbonate precipitate formed in the second carbonation step is readily filtered and does not require settling.

c. Sulfitation

Finally, the raw juice will have sulfur dioxide bubbled through it in the sulfitation tower, which serves to adjust the alkalinity and pH downward. This acidification by SO_2 promotes the reduction of certain coloring substances to non-coloring forms. The presence of traces of $SO_3^=$ has been shown to inhibit the tendency of sugar solutions to develop color during subsequent evaporation and crystallization. The exact mechanism of this behavior is not fully known. The addition of SO_2 must be restrained, however, because too much acidity favors further degradation of sucrose to its components, glucose and fructose, especially in the high temperature evaporators.

F. Sugar Crystallization

Once the evaporators have concentrated the purified sugar solution to the appropriate, empirically-predetermined Brix, the solution, which is now referred to as thick juice, is blended with a stream of recycled, remelted sugar from second and/or later-stage crystallizing units. This blend, called standard liquor, is now several percentage points more concentrated and more pure than the thick juice. After a final filtration the standard liquor passes through two to four series-batch crystallization stages. (A very few continuous crystallizers are in use, some being used on a developmental basis.) This crystallization process is carried out in vacuum evaporators called "white pans" and "raw pans". This study utilized the simple, but complete three-stage pan system depicted in Plate 4 of Figure 8.

The first vacuum unit is called the white pan, and it produces the white sugar at 100% purity which will be marketed. The second unit is called the high raw pan, and it produces "high raw" sugar, at 99% purity, which is totally recycled. The third unit is called the low raw pan, and it produces "low raw" sugar, at 98% purity, which is also totally recycled. In each pan enough thick sugar solution is introduced to partially fill the unit. A vacuum is drawn on the pan, steam is admitted to the coils, and boiling proceeds. In most plants the pans are operated so that the sugar concentration is raised into the range of supersaturation called a labile condition; this liquid-phase solution is unstable but not self-seeding. A small, measured slurry of very fine-grained, crystalline sugar in alcohol is then added. As the boiling continues, more liquor is added and the injected crystals continually grow in size but are usually not allowed to multiply. This affords good control over crystal size and size distribution.

The appropriate end-point for crystal growth is a matter of the pan operator's judgment, based on a visual inspection of the crystals; it is generally at a concentration of 90 to 95 Brix. Frequently, and especially with the low raw pan, there is a unit called a mixer-crystallizer where the seeded liquor is transferred for final crystal development. Such units have stirrers that are usually heated, and the entire unit may be steam-jacketed or simply insulated. Separate mixer-crystallizers are used because of the slowness of crystal growth, particularly for the reduced purity of raw sugars.

When crystal growth is completed, the sugar and liquor are separated by centrifugation. Surface impurities and residual liquor are effectively washed from the crystals using a light application of high-pressure, superheated water following the removal of the bulk of the liquor; only a miniscule amount of sugar redissolves. The white sugar is dried, sifted, and packaged. Outsized white crystals and white sugar dust are returned for remelting and recrystallization. Raw sugars are remelted with inclusion of a portion of the centrifuge wash water and some thin juice to control the resulting solution Brix. All of these remelted sugars are then combined with thick juice, as described earlier, to form standard liquor.

The liquors from the white centrifugal and high raw centrifugal are the feeds to the high and low raw pans, respectively. Thin juice is added to the white sugar liquor to form "high green" liquor. Some high raw centrifugal wash water is added to the liquor from the high raw sugar, forming "machine syrup" or "low green" liquor, for use as the feed to the low raw pan. The final liquor from the low raw centrifugal, with centrifuge wash water added, is called molasses. It is the lowest-purity sugar solution at, usually, 50 to 70%. Molasses may be simply added to dried pulp as a binder and nutrient source. Or it may be further treated via the Steffen process for slight additional sugar recovery, followed by final treatment and consumption in the manufacture of monosodium-glutamate. This last pair of treatments is not included in the present design.

VII. GEOTHERMAL ENERGY FOR THE BARLEY MALTING INDUSTRY

A. Background

About 100 million bushels of barley are converted to malt each year, or almost 25% of the total United States barley crop. The malt product is used almost exclusively for the brewing of beer, although many other markets exist for a small part of the output, including distilled alcohols, cattle feed, and various food products. Most malt processors are simply suppliers to end users, but some brewers operate captive malting plants to provide at least a portion of their needs. The Adolph Coors Co. is a prime example of the latter type of operator.

Barley is a significant crop in the San Luis Valley, with annual production ranging from 5 to 7 million bushels in recent years. A total of about 85,000 to 110,000 acres, out of roughly 350,000 irrigated acres in the Valley, are planted in barley. Some of this acreage is dedicated to specific brewers, including Coors, Schlitz, and Anheuser Busch, while the rest is devoted to speculative crops that enter onto the open market, including some for export to foreign brewers.

The malting process consists of germinating, or sprouting, the barley under carefully regulated conditions to form "green malt," and then arresting the germination at a critical stage in its development. Germination is stopped by gently heating and drying the water-saturated, sprouted barley kernels. At this point, about 15 to 18 percent of the starch that is initially contained in the barley grains has been converted to simpler sugars. This sugar is later converted to alcohol in the brewing process. The most important objective of malting the barley, however, is the liberation of its enzymes (16). This complex mixture of enzymes, including several α- and β-amylases, has the capability to cause complex carbohydrates such as starch to form the simple sugars which are later converted to alcohol. This enzymatic activity, or diastatic power, of malted barley is very high compared to other grains. Therefore, the malt is used as the starter, or catalyst, to promote sugar formation from less expensive cereals.

In the brewing process itself, the malt is crushed and mixed with other grains that are lower in diastatic power, but that have other advantages with respect to barley, such as a lower cost or the ability to improve the foam stability of the final product. Corn and rice are common hydrate sources that are used for mixing with the mashed malt in warm water. Then, the enzymes that had been produced during germination of the barley catalyze the reactions whereby the remaining starch in the barley and the much larger amount of starch in the other cereals are converted to sugar.

With respect to the barley malting factory, itself, kilning is the major energy consuming operation. Drying of the malt is achieved by heating ambient air to temperatures ranging between 130 and 200°F, depending on the stage of the drying cycle that has been attained. Natural gas is the most common kiln fuel, both because of its historically low price and because of the convenience of its utilization. Air is heated by direct contact with the burning gas in a simple burner, and is passed directly over the green malt. Coal and oil fuels are also used, but in these cases the kiln air must be heated indirectly in a heat exchanger, because the combustion gases from these heavier fuels contain components that would be deleterious to the malt if direct-contact heating were employed.

Most barley malting plants are quite old, with 30 to 50 years or more being common ages, although in many cases more recent malting trains have been added in existing factories to operate in parallel with the older process units. These old factories naturally tend to natural gas as fuel for the reasons previously related. Retrofit to coal, for example, would often be very difficult and expensive because of the large amount of space required for the more complex coal-fired facility. Some modern plants have been built, however, using other fuels. In particular, electric heating has been installed in one factory and an oil-fired furnace in another.

The heating loads in a malt factory vary considerably. The most important factors in determining annual fuel requirements are the average temperature and humidity of ambient air, and whether cooled air leaving the kiln is partially recycled or completely discharged to the atmosphere. Most older factories do not recycle exhaust air, again because the historically low fuel prices did not justify the required capital expenditure to enable recycle operation. In general, despite the higher fuel prices of today, retrofitting a plant to permit recycling is often too costly. Accordingly, typical total energy consumption in a gas-fired plant without recycle is about 90,000 Btu per bushel of malt. Of this amount, up to 80% of the heat load is required for the kilning operation, with the remainder being used during germination and for space heating of the factory.

A modern plant would necessarily involve recycling so as to reduce energy consumption rates, especially during peak periods in the cycle. These factors are considered in more detail in the following sections of this chapter. Section B describes the classic malting process in more detail, while Section C explains the techniques and costs of applying a low-temperature geothermal source to a malt plant. Section D provides a brief description of an alternate geothermally heated factory.

B. The Barley Malting Process (15, 17)

The barley malting process is shown schematically in Figure 9. Although specific equipment types and operating conditions may vary from factory to factory, this Figure presents the essential aspects of the process. In this section, the process will first be qualitatively described. Then, the factors governing energy consumption rates will be covered in more detail.

Before describing the malting process in any detail, the overall energy and water requirements for a "typical," natural gas fired barley plant will be summarized. Based on a plant capacity of 4 million bushels per year, the total fuel requirements will be in the range of 340,000 to 400,000 million Btu per year. Of this amount, about 80% will be consumed in the kiln, with the remainder being used for the germination cycle and space heating. Total daily usage of electricity will be about 20,000 kilowatt hours. The total water requirement for all usages is about 1,050 gallons per minute, most of which is required for steeping at less than 60°F. These values can change significantly depending on many factors, including: type of fuel used; climatic and water conditions at a specific plant location; process options chosen of recycling kiln air or not; and kiln design.

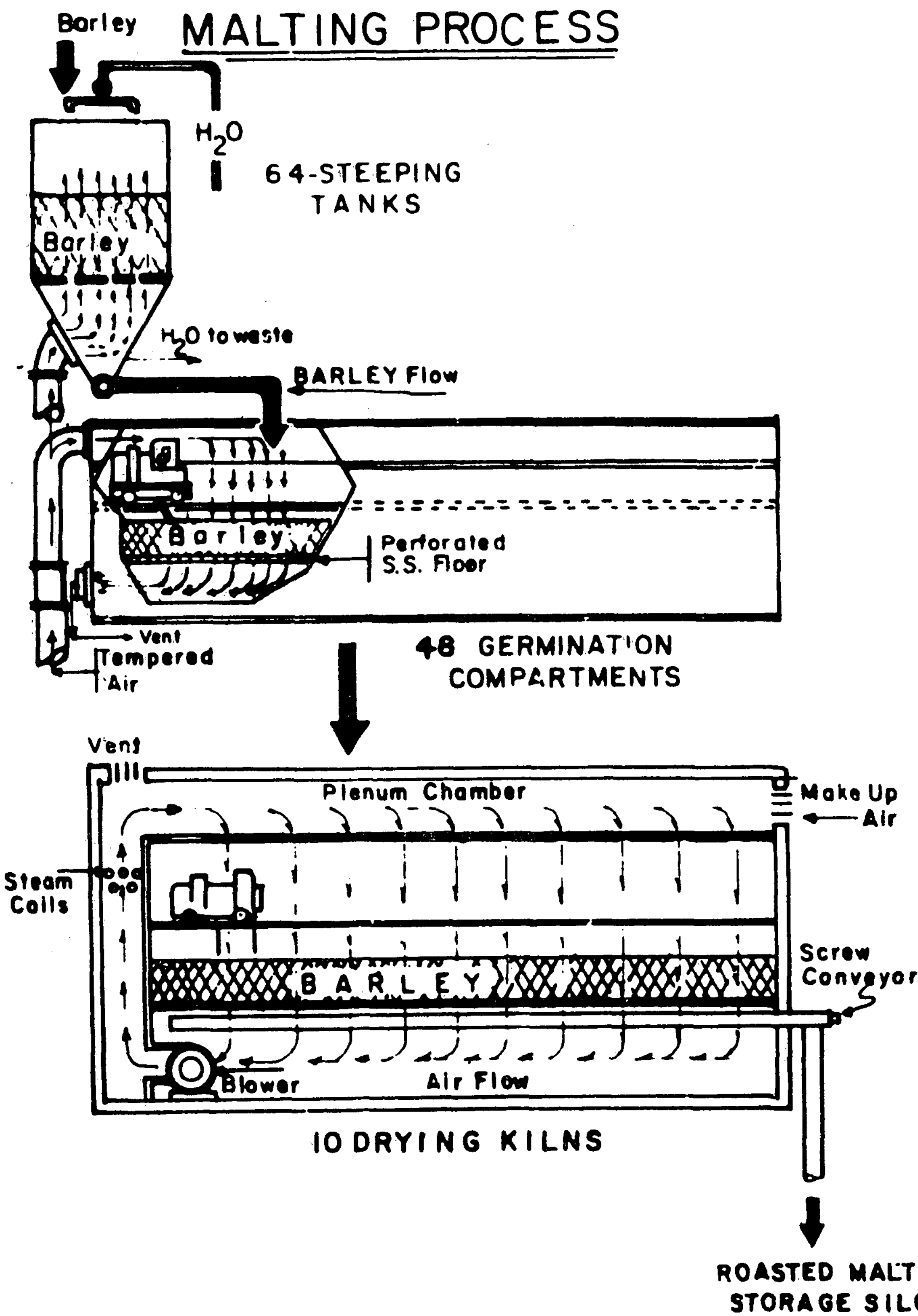


Figure 9

Barley Malting Process Schematic

Barley from storage is conveyed to the steep tanks, which are generally cylindrical vessels with conical bottoms. The vessels are constructed of mild steel and typically have a capacity of 2,000 to 3,000 bushels. The barley is sprayed with cold water during filling until it is covered, and is allowed to steep for periods ranging from 12 to 48 hours. Actual times depend on the water temperature, condition of the barley, and the desired properties of the malt blend being prepared. Steeping continues until the barley absorbs an amount of water equal to about 44 to 45% of its weight as compared to less than 13% moisture contained in field-dried barley. During the soaking period, the beginnings of germination occur, and the barley kernel begins to produce the enzymes required later to complete germination and to be used in the beer brewing process.

Steep water should generally have a temperature of less than 60°F, and preferably less than 55°F. If natural waters at this temperature level are not available, the makeup water is cooled in a vacuum chiller. This procedure could require a significant quantity of electrical energy, since about 1.4 million gallons of water at 50°F are required daily for a malt plant with a capacity of 4 million bushels per year. Steeping water should be low in ammonia, nitrate, iron, and chlorine content, and should be free of coli bacteria. Other than these constituents, most low salinity waters would be acceptable, even if the total dissolved solids content somewhat exceeds potable water standards.

Steeped barley is transferred to the germination bins where it is maintained for about 5 days under carefully controlled temperature and humidity conditions. The bins may be large, concrete, rectangular compartments with a capacity of 2,000 to 4,000 bushels, such that one germination bin will charge a single kiln. Barley will be loaded to a depth of about three feet on a false, perforated steel-sheet floor. Moisture-saturated air at a controlled temperature is passed through the bed to maintain the required environment during germination.

The relative humidity of circulating air is maintained at about 100% to prevent drying of the barley. After about 2 days, the residual surface moisture that remained on the grains leaving the steep tanks is exhausted because of the natural respiration processes that occur during germination. From this point on, the barley is sprayed lightly with water once or twice a day to prevent further drying. Water requirements for this purpose are about 48,000 gallons daily for the 4-million-bushel-per-year factory. This represents a requirement of 3 to 5 gallons of water per day per bushel.

The germinating barley is maintained at a temperature of 60 to 65°F. The growing process itself produces heat, so that partial recycling of the air leaving the germination compartment can control the temperature at the desired level. External heat is required for the air only during very cold periods. In a natural gas fired process, the makeup air is heated directly as it passes through the gas burner. Only about 5 to 15% of the total annual heat load of the factory is required for the germination process.

Germination is allowed to continue until the shoot of the growing plant, the acrospyre, within the barley grain is almost as long as the husk, but before it actually breaks through the husk. The intact husk will protect the plant during kilning and subsequent handling and storage. The rootlet that also forms and protrudes from the husk of the young plant is removed by mechanical agitation following germination and kilning because of the bitterness it would impart to brewed products.

The germinated barley, or green malt, is stabilized by heating and drying in hot-air kilns to prevent further growth during storage. It is here that about 80% of the annual process heating load is consumed. The general kiln design is similar to that of the germination compartment. The barley is filled to a depth of 2 to 4 feet on the perforated steel floor of a large, rectangular concrete compartment. Hot, dry air is blown up through the green malt, and the cooled, humid air leaving the grain bed is exhausted to the atmosphere. The kilning cycle is completed in a day, with about 18 hours required for drying the green malt, one or two hours for cooling the malt product, and the remaining time for dumping the load to storage and then recharging the kiln with a new load of green malt.

Some kilns are constructed with double or even triple decks; Figure 10 is a cross-section of a two-deck unit. A two-deck unit will have a two-day cycle, with a load of dried malt being produced each day. To start a cycle, the dried malt on the lower deck is discharged, the partially dried green malt on the upper deck is dumped to the bottom deck, and a fresh load of green malt is charged to the upper deck. The daily heating cycle then is initiated, such as the typical 18-hour cycle shown on Table 3 below. The lower bed is completely dried during this cycle, while the upper bed of green malt is partially dried by the cooled and partially humidified air leaving the lower bed. At this time, the kiln is cooled by maintaining the air flow and closing off the source of heat. The beds are dumped and the cycle is repeated.

The two important variables in kiln operation are the air mass flow rate, and the temperature at the inlet to the malt bed. To maintain the fuel requirements at the lowest possible level, the air flow rate should be minimized. However, two factors impose a practical lower limit on air flow: the rate must be high enough to provide uniform and sufficient contact between air and malt across the entire bed, and the rate must also be high enough so that moist air leaving the bed is somewhat less than saturated with water. As a general rule, about 40 to 60 cubic feet of air per minute are required per bushel of green malt in the kiln. If the drying rate is too rapid, the grain will shrink and crack and suffer general physical damage. Accordingly, a careful heating schedule is followed depending on the end use of the barley. A typical schedule shown in Table 3 below relates air temperature entering the kiln, downstream of the air heater to the length of time that temperature is to be maintained:

Table 3

Typical Kilning Temperature Cycle

Air Temperature (°F)	Hours	Air Flow Rate (cubic feet/minute)
130	4	120,000
140	4	120,000
150	4	120,000
160	2	96,000
170	2	96,000
180	2	96,000
Total heating cycle:	18 hours	

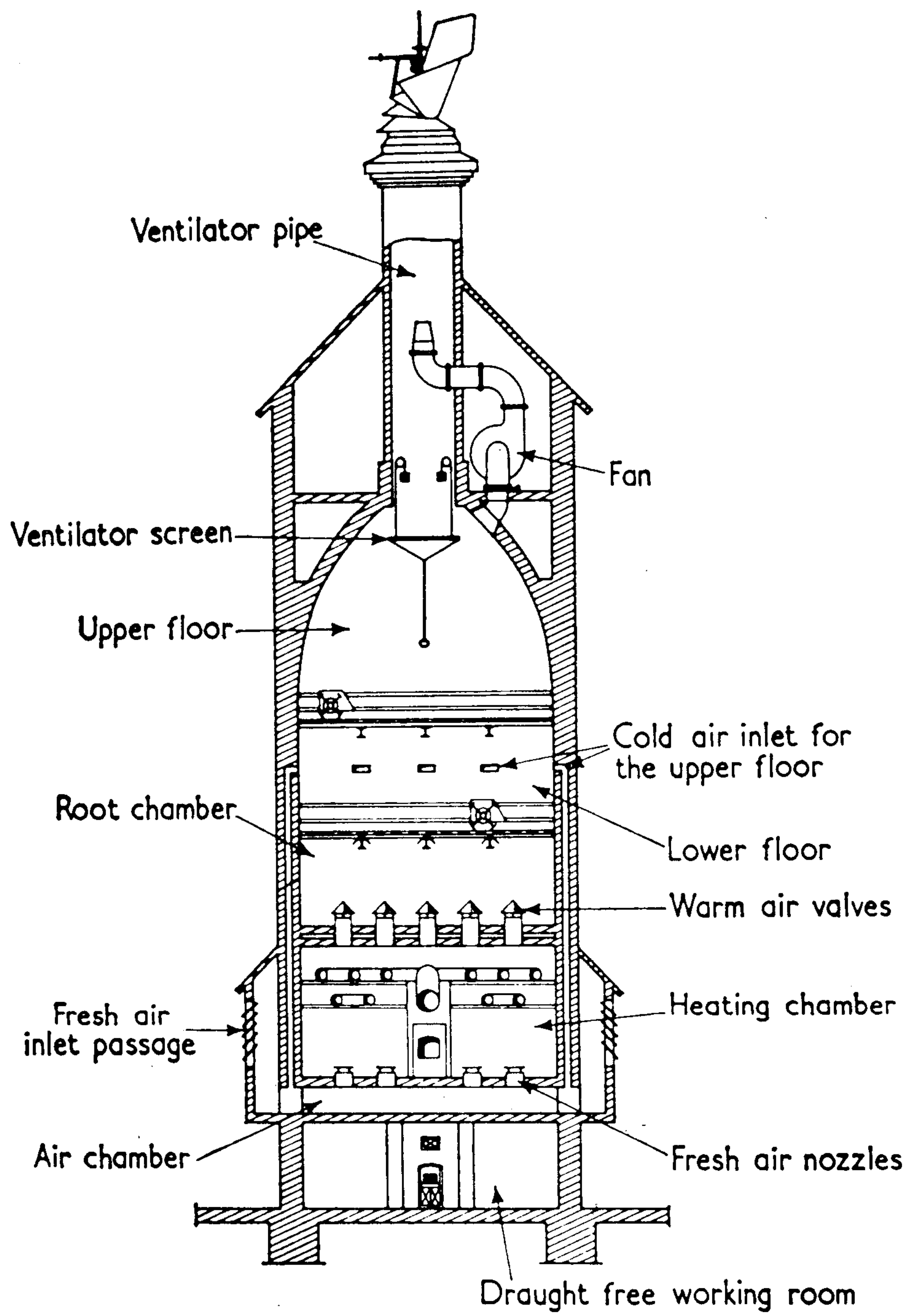


Figure 10
Diagram of a "Classical" Two-Floor Kiln

The upper kilning temperature depends on the end use of the malt. The final temperature is limited to about 175 to 180°F for pale beers, while malt for dark beers may be kilned at up to 200 to 220°F. The lower temperature range is more typical in the United States.

The heat load on the air heater depends on the air flow rate, the ambient air temperature and humidity, and the kilning temperature. If no recycle is used, the largest heat load occurs during the last two hours of the kiln cycle. At this time, which generally is scheduled to occur just at dawn, the ambient air temperature is lowest while the kilning temperature is highest. However, the water removal rate is also very low at this point in the cycle, so that the rate of water pickup per unit volume of circulating air is very low. Thus, at the end of the cycle, up to 80% of the air leaving the kiln can be recycled through the heater without danger of saturating the exhaust air. The resultant heat load is then about 40% of the load at the same point in the cycle for a no-recycle system.

At the start of the kiln cycle, on the other hand, the required heat load is near its maximum value, even though ambient air is then near its maximum temperature (from about 10 a.m. to 2 p.m.) and the kilning temperature is at its minimum point of 130°F. This is because the drying rate is most rapid at this point in the cycle, with perhaps 50 to 60% of the moisture in the malt being removed. Thus, the kiln air is easily saturated, no recycling is possible, and the entire air flow stream must be heated from ambient to kilning temperature.

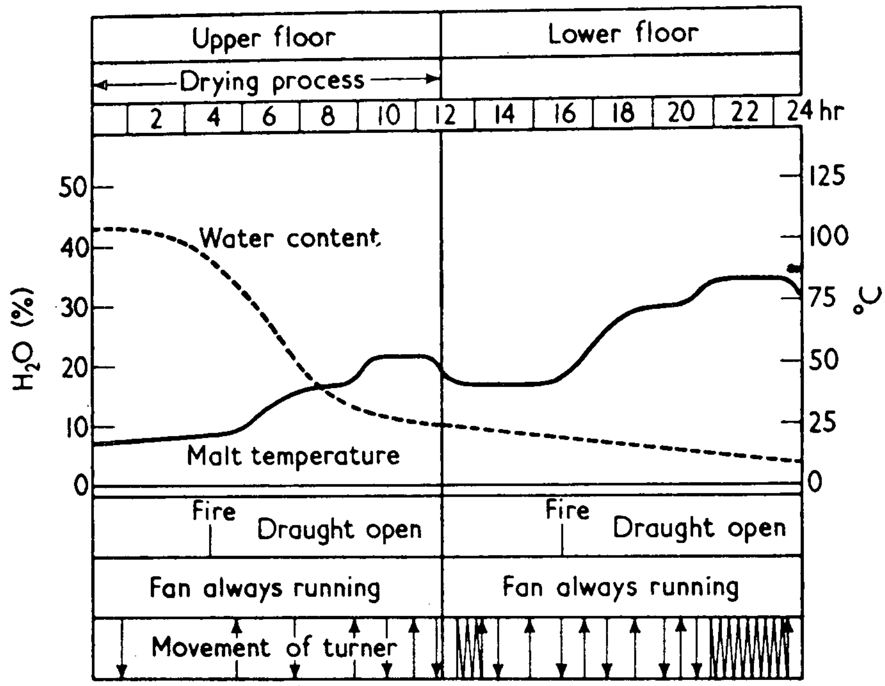
A "typical" drying curve for a double deck kiln is shown on Figure 11. In this example, a 24-hour cycle is followed, with one charge of malt being completed every 12 hours. As can be seen from the Figure, the water removal rate is slow at first while the entire system heats up, and then proceeds very rapidly as the unbound surface water evaporates. The evaporation rate then decreases continually and markedly, because the remaining water is bound within the pore structure of the malt. This bound water must first evaporate and then diffuse to the surface of the grain, where it is picked up by the moving air stream. As indicated on Figure 11, this is a slow process.

An energy-efficient design of a kiln must consider carefully the rate of evaporation at every point in the drying cycle. Only in this way can air flow and recycle rates be optimized so that heating loads are reduced without either saturating the air or increasing kiln cycle times. As previously mentioned, most kilns today are quite old and were built when energy costs were so low that the capital and operating costs of an air recycling system could not be justified. The economic application of geothermal energy to kilning requires that air recycle be employed so that heating loads are reduced and the size and capital cost of the geothermal system are not excessively large.

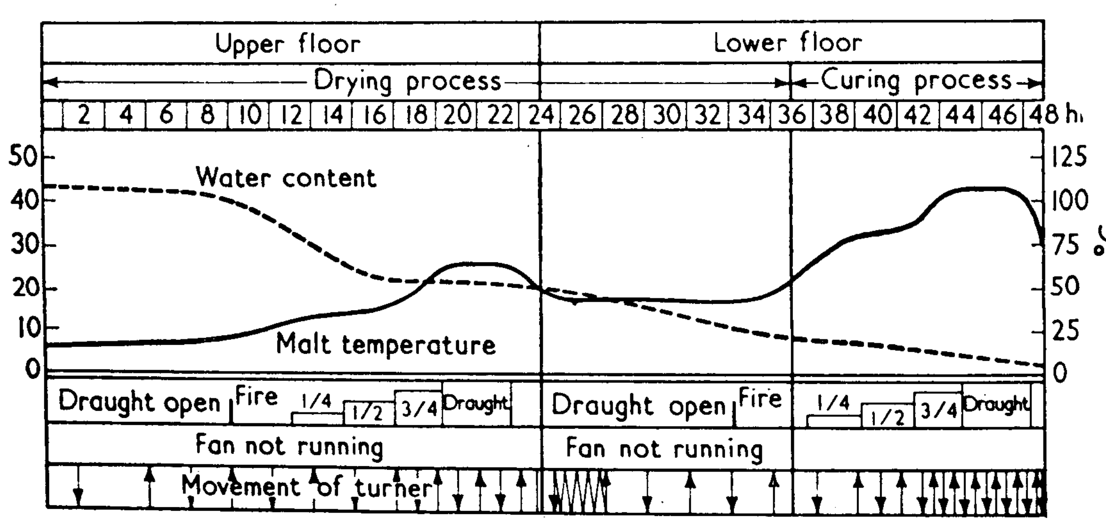
C. A Geothermally Heated Barley Malting Process

1. General Considerations

As has been discussed, the required heat load for barley malting is significant, and is spread rather uniformly over the entire year. The total annual heat load for a 4-million-bushel capacity will range from 340,000 to 400,000 million Btu per year, depending on whether or not recycle is used. Since this heat duty also occurs at



Kilning scheme for pale malt on a two-floor kiln for twenty-four hours.



Kilning schedule for a dark malt on a two-floor kiln for forty-eight hours.

Figure 11

relatively low temperature, it is natural that a low temperature geothermal source be considered for this service. This section will analyze the geothermally heated malting process. First, the general process and economic relationships that determine the viability of a geothermal process will be described. Then, geothermal process options will be presented for a typical plant and the cost of the preferred options within the context of the San Luis Valley geothermal resource will be determined.

When comparing a geothermally heated plant to other heat sources, care must be taken to separate the costs and credits due to the heating medium from those costs that are incurred regardless of the heat source. In this analysis, no attempt has been made to revise or improve the basic malting process in order to make the geothermal case more efficient. Rather, only the heat transfer portion of the process is considered in determining the relative cost of geothermal heat. The process cycle that is used is the same as that discussed in the previous section for other fuels. However, as it has been pointed out, recycling of kiln air, which is a standard but not universal practice, is economically necessary for the geothermal system.

The comparison of geothermal heat to other heat sources must consider both operating and capital costs. Capital items for the geothermal case include production and injection wells and spare wells, transfer lines and pumps, and the air heat exchanger. When natural gas is the fuel, the kiln air is heated directly in the burner and no air heat exchanger is needed; in addition, the furnace is very simple and pollution control requirements are small, so that this is the lowest capital cost case. Both coal and oil heated systems require more expensive furnaces and pollution control equipment. In addition, coal and oil systems require a heat exchanger for indirect heating of kiln air, as does the geothermal case, but it would be a smaller heat exchanger than necessary in a geothermal unit because of the higher temperature of the heating medium.

With respect to operating costs that enter into the comparison, the cost of fuel is the only factor when natural gas is used. However, the availability of natural gas for new, and even some old installations is not assured. Geothermal operating costs include those for pumping and treating the brine before use and reinjection, and for the maintenance of capital equipment. Coal and oil fired systems incur additional operating expenses relative to natural gas for operating labor and equipment maintenance, and for the operation of pollution control systems.

The primary factors affecting the cost of a geothermal system are the cost of the wells and the cost of the air heater. As indicated in the following discussion, it is not possible to make a clear cut choice of process design parameters since the same variable will have opposing effects on the cost of different parts of the system. For example, using a high temperature brine reduces the size of the air heater (as discussed later with respect to equation 3), and reduces the required brine flow rate (equation 4) and thus the cost of brine transfer lines. However, it is necessary to drill deeper and more expensive production wells to achieve high geothermal temperatures, especially in the San Luis Valley. Thus, a point may be reached where higher temperatures lead to higher overall system costs.

With respect to the air heater, its cost will depend primarily upon its surface area, as determined by the equation

$$Q = UA\Delta T \tag{2}$$

where A is the surface area and the heating load, Q, is fixed by the process requirements. The heat transfer coefficient, U, is dictated by the film coefficient on the air side and will be in the range of 15 to 25 Btu/hr-ft^2-^oF. Thus, the only significant control that the designer can exert on the exchanger cost is by changing the log-mean temperature difference, ΔT, defined as follows:

$$\Delta T = \frac{(T_2 - T_5) - (T_1 - T_4)}{\ln \dfrac{T_2 - T_5}{T_1 - T_4}} \tag{3}$$

Figure 12 shows two options for a geothermally heated kiln; the preferred option shown on Figure 12a is discussed in this section, while the alternative approach shows on Figure 12b. As shown on Figure 12a, T_1 and T_2 of equation 3 are the temperatures of geothermal brine entering and leaving the air heater, respectively, and T_4 and T_5 are the corresponding temperatures for air. The value of T_4 is dictated by ambient air conditions for the no-recycle case, and the value of T_5 is dictated by the malting cycle. Thus, ΔT is increased (and surface area, A, is decreased) when the inlet brine temperature, T_1, is high. The area is also decreased when the brine temperature loss in the heater (T_1-T_2) is minimized so that T_2 remains high. However, reducing this loss would increase the required brine flow rate, F, as indicated by the equation below:

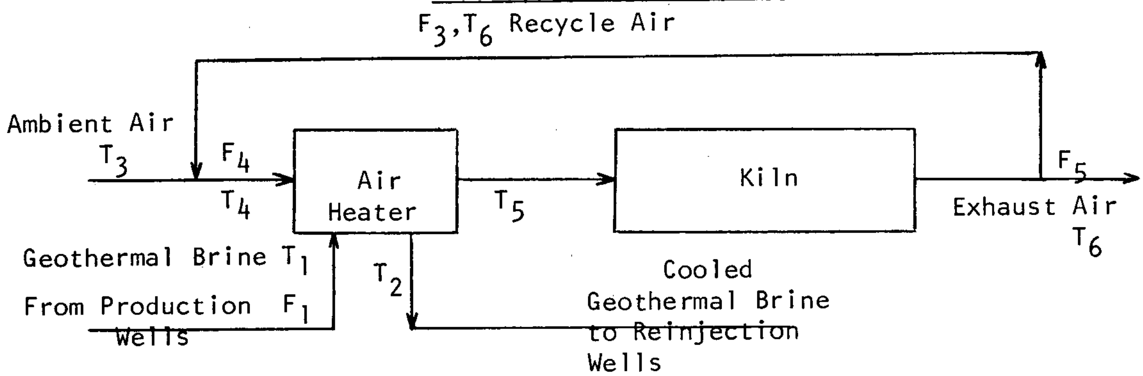
$$F = \frac{Q}{c_p\ (T_1 - T_2)} \tag{4}$$

where c_p is the brine specific heat. Thus, high brine temperatures reduce heater costs but increase well costs as has been discussed. In addition, high brine flow rates reduce air heater costs but increase brine transfer costs. In order to estimate the actual operating conditions that tend to reduce overall costs on balance, a parametric set of design calculations have been made. The results of this analysis are presented in subsection 3 after the following statement of design bases.

2. Design Bases

The kilning temperature cycle given in Table 3 provides the design objective of the geothermally heated plant. The principal design factors outside the control of the designer are ambient air conditions in the San Luis Valley, and the depth and temperature of the geothermal resource. Atmospheric data are presented in Table 4, and well cost data are discussed later.

12A. Base Case -- All Brine Flow



12B. Optional Case -- Flashed Brine

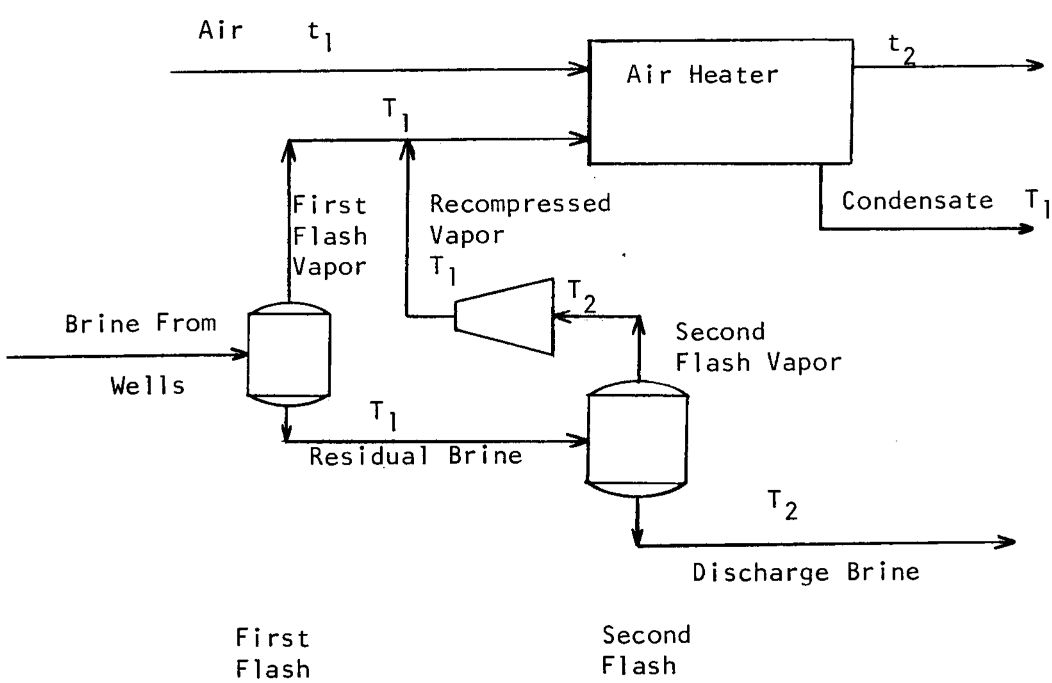


Figure 12
Barley Malting Kiln Process Schematic Based on a Geothermal Heat Source

Table 4

San Luis Valley Weather Data

Source: National Weather Bureau

Month	Average Monthly Temperature (^{o}F)		Average Monthly Minimum Temperature (^{o}F)	Average Monthly Relative Humidity (%)
	Dry Bulb	Wet Bulb		
January	17	15	-0.7	68
February	23	21	5.5	66
March	31	26	14.6	55
April	41	34	24.5	47
May	51	42	33.5	46
June	60	50	41.5	47
July	65	56	47.9	57
August	63	55	45.8	59
September	55	47	36.2	55
October	40	38	25.0	54
November	30	26	11.7	63
December	19	17	1.0	69

Probable number of days per year that:

Maximum temperature exceeds:	$90^{o}F$	0 days
Maximum temperature is less than:	$32^{o}F$	34 days
Minimum temperature is less than:	$32^{o}F$	256 days
	$0^{o}F$	56 days

The average ambient temperature in the Valley is 42°F. For design purposes, however, the extreme conditions must be taken. The minimum temperature can be expected to fall below 0°F on more than 15% of the days during the year, and the minimum temperature usually occurs just before sunrise, when kilning temperatures are at a maximum. For this case, the design ambient air temperature has been set at -20°F, which occurs when the kiln temperature is 180°F.

The other extreme operating case occurs during the first 4 hours of the kiln cycle, when a kiln air inlet temperature of 130°F is required. Although this generally occurs during the warmest part of the day, it is a critical design point because of the high rate of water evaporation that takes place early in the cycle. If the ambient air is warm and has a high humidity, the kiln exhaust air is almost saturated and no recycle is possible. For this case, an ambient air temperature of 80°F at a relative humidity of 80% has been established as the design basis.

On the other hand, daytime temperatures during the start of the kilning cycles will often be very cold during winter months. When this occurs, the ambient air will be quite dry and recycling may be possible. For this design case, an ambient air temperature of -10°F has been established for the time period when the kiln temperature is 130°F. These three design cases form the basis of the parametric analysis, and they are summarized in Table 6 of the following subsection.

The design capacity of the malting plant is set at 4 million bushels per year, which is considered to be the smallest economically viable size for a new plant. The factory will consist of four 3,000-bushel kilns operating 333 days per year, for an on-stream factor of about 91%. The design is based on single deck kilns. The higher capital cost and increased complexity of a double deck kiln operation could probably be justified with present-day energy prices, since the double deck is more energy efficient than the single deck. The calculations for the double deck design are much more complex, however, and could not be justified within the scope of the present study.

The total air flow rate to the kiln starts at 120,000 cubic feet per minute where it remains for 12 hours, as the kilning temperature is increased successively from 130°F to 140°F and 150°F, as indicated in Table 3. The flow rate is then reduced to 96,000 cubic feet per minute for the last 6 hours of the drying cycle, at successive temperatures of 160, 170, and 180°F, and for the cooling period. The total air flow includes the makeup plus the recycle air. Recycle rates will vary from zero to 79% of the total air flow rate, depending on cycle conditions and ambient air humidity.

Production and reinjection well costs are indicated in Table 5 below. It is assumed that the required number of injection wells will be at least one-half the number of production wells. At least one spare well of each type is provided. Production rates of up to 250,000 pounds per hour are considered attainable in determining costs for this study, although it appears probable that higher rates are possible in the San Luis Valley.

Table 5

Estimated Well Drilling Costs in the San Luis Valley

Brine Temperatures (°F)	Well Depth (feet)	Well Cost ($)
300	10,000	650,000
260	9,000	550,000
220	7,500	450,000

Surface transfer lines from the wells to the plant will be required. The range of line diameters will be 4 to 8 inches, and maximum distances will be on the order of 3,000 feet from either injection or production wells to the plant. The cost of installed transfer lines will range from 30 to 45 dollars per foot.

Expected compositions of the brine are given in Table 22 in the Appendix. No significant brine treatment costs are expected as long as the brine is contained as a liquid under pressure and out of contact with air. Pressurized brine will not liberate CO_2 and thus the primary cause of scale precipitation will be avoided. In addition, the absence of oxygen in a brine, a condition that can be maintained as long as the brine is kept pressurized in closed vessels and lines, will keep corrosion rates from attaining significant levels.

Despite these precautionary design philosophies, a charge will be attributed to brine treatment so that the estimated project costs will be conservative. It is assumed that the brine will be acid treated to remove all carbonates and bicarbonates, and then base treated to control the pH at a non-corrosive level. For this purpose, it is assumed that sulfuric acid and sodium hydroxide, both at 10 cents per pound, will be used. It is recognized that a less expensive pretreatment system would probably be developed during a detailed design.

3. Geothermal System Design

The preferred process for geothermal brine is shown schematically on Figure 12a. The various flow streams and equipment items on this figure contain key letters to indicate temperatures and flow rates. The values of these variables for the various design cases are indicated in Table 6. This table summarizes process parameters for three different brine temperatures (220°F, 260°F, and 300°F) for each of the three base cases discussed in the previous subsection. Table 6 will be analyzed in detail after the process flow sheet is reviewed.

Figure 12a shows the kilning aspect of the barley malting operation. The kiln consumes about 80% of the energy input to the plant and operates all year long. The remaining energy consumption for heating air to the germination vats and for space heating occurs only during cold weather. Energy utilization in those sectors are at much lower temperatures than those required for the kiln, and would use discharge brine from the air heater as their energy source. The analysis that follows is based on only the kiln operation.

Referring to Figure 12a, pretreated geothermal water, at a temperature T_1 and flow rate F_1, is fed to the air heater. The drying air at a flow rate F_4 passes through the heater wherein its temperature is raised from T_4 to T_5; the value of T_5 changes periodically at various stages in the cycle as shown in Table 3. The air entering the heater is a combined stream of recycle air (at temperature T_6 and flow rate F_3) and ambient makeup air (at temperature T_3 and flow rate F_2). The exhaust air (temperature T_6 and flow rate F_5) carries out all the water evaporated from the malt. As is noted in Table 3, the recycle air flow rate (F_3) varies over the drying cycle.

There are three principal design parameters that have a significant effect on the cost of geothermal heating. These are: (1) the temperature of produced geothermal waters; (2) the amount of geothermal water required; and (3) the size of the air heater. There is a complex interaction among these parameters that is explained below. The final costs depend strongly on the temperature of discharge brine (T_2) and the flow rate of recycle air (F_3).

1) Since water temperature is a function of well depth, the cost of a single production well increases as the geothermal temperature increases, as shown in Table 5. This cost varies approximately from \$450,000 per well for 220°F water to \$650,000 for 300°F water. On the other hand, with higher geothermal temperatures, the required brine flow rate decreases. This factor may decrease the total number of wells required, as well as the cost of transferring brine from the wells to the plant.

2) The amount of brine required determines the number of production and injection wells, and therefore the drilling cost. This flow rate (F_1) depends on the design heat load in the air heater (Q) and the discharge brine temperature (T_2) as follows:

$$Q = F_1 \ c_p \ (T_1 - T_2) \qquad (5)$$

Thus, to decrease F_1, either T_2 or Q must be decreased. Decreasing Q decreases the size of the air heater proportionately, while decreasing T_2 increases the heater size in a more complex way, as is discussed below.

3) The size and cost of the air heater is dictated primarily by the heat load (Q) and the outlet temperature approach (T_2-T_5). Increasing the recycle flow of warm air (F_3) will decrease the heat load. As has been pointed out in Section B, the single flow rate is tailored to maintain the relative humidity of exhaust air at less than 90 to 95%. High recycle rates can always be employed at the end of the drying cycle when water evaporation rates are low and kiln temperatures are high. Recycle flow can be employed at the start of the drying cycle only when the relative humidity of ambient air is very low. On normally humid summer days, no recycle is permitted at the start of the cycle, but on cold winter days recycle is feasible.

With respect to the brine discharge temperature (T_2), a compromise must be made on the basis of economic considerations. Low values reduce the brine flow rate as suggested by equation 5. On the other hand, low values reduce the temperature difference, (T_2-T_5), so that the size of the air heater increases.

On the basis of these considerations, several comparative design cases were developed for three brine temperature, for each of which several discharge brine temperatures were evaluated, as summarized in Table 7. In every case, the air heater was sized for each of the three design cases, indicated in Table 6. The first design point of Table 6 is during the first stage of drying with warm inlet air and no recycle and with a heat load of 28.8 million Btu per hour. Although the heat load is low, the high inlet temperature tends to decrease the temperature difference for heat transfer and this may result in a large heater. The second design case of Table 6 also occurs during the first drying stage, but with very cold inlet air. This air contains almost no moisture, even at high relative humidities, so that high recycle rates are allowed. Case 2 has the highest heat load of 39.4 million Btu per hour. The third design case of Table 6 is at the end of the heating cycle with a very low ambient air temperature, but a high recycle air rate. This case will tend to require a large air heater because the high heater outlet temperature reduces the temperature driving force for heat transfer.

The results of the comparative analyses for these design cases are summarized in Table 7. All of the subcases that were considered are not included in Table 7, but the trends of the critical parameters can be seen. For any of the base cases, as the temperature of discharge brine decreases, the required brine flow rate decreases but the air heater surface area increases. As the inlet brine temperature decreases, both the required brine flow rate and the air heater surface area increase. Significantly, Table 7 shows that the brine flow rate and air heater size are controlled by Case 3, and not Case 2, even though Case 3 has a lower heat load. This is due to the fact that the temperature of air entering the kiln for Case 3, which occurs at the end of the cycle, are higher. Thus, heat withdrawal per pound of brine is reduced, as is the temperature driving force across the heater.

With Case 3 having been established as the basis for design, it remains to determine which brine temperature leads to the lowest cost. The primary cost factors are the capital costs for production and injection wells, transfer lines and pumps, and the air heater. The higher temperature geothermal fluid of 300°F leads to lower costs in all categories except well cost. The analysis on Table 8 indicates that a 260°F brine source is the least costly for the San Luis Valley situation. This is based on comparing Cases 3-a2, 3-b1, and 3-c1 from Table 7. However, this study is not optimized and is based on several arbitrary factors incated below, so the conclusion could change somewhat.

It was assumed in the study that the production and injection wells are 3,000 feet from the malting plant, and that transfer line costs are 32 and 43 dollars per foot, respectively, for 6 and 8 inch lines. Case 3a2 and 3-b1 require a 6-inch transfer line while Case 3-c1 requires an 8-inch line. Installed costs for stainless steel, finned tube air heaters are assumed to be $8.50 per square foot of heat transfer surface. Based on the brine flow rates given in Table 8, one production and injection well, with one spare for each service, are adequate for Cases 3-a2 and 3-b1. Case 3-c1 requires two wells with one spare for each service. Well costs were given in Table 5.

Table 6

Heat Load Summary
Barley Malting Kiln

Bases: Plant capacity of 4 million bushels per year.
Kilning cycle as given in Table 3.
Ambient air conditions applicable to the San Luis Valley, Colorado.

Case No.	Air Heater Outlet Temp. (°F)	Air Heater Flow Rate (1000 CFM)	Ambient Air Temp. (°F)	Ambient Air Rel. Humidity (%)	Recycle Rate (%)	Kiln Outlet Temp. (°F)	Kiln Outlet Rel. Hum. (%)	Air Heater Heat Load (10^6 Btu/hr)
1	130	120	80	80	0	80	90	28.8
2	130	120	-10	50	79	80	90	39.4
3	180	100	-20	50	75	165	80	31.2

The above cases best represent the likely ranges of values of operating parameters for the stated bases.

Table 7

Summary of Design Cases for Barley Malting Kiln

Bases: Capacity of 4 million bushels per year
Heat transfer coefficients = 17 Btu/hour-ft^2-^{o}F
Cases 1, 2, and 3 described in Table 6.

Case	Geothermal Brine Temp. (^{o}F)	Air Heater Condition: Q (10^6 Btu/hr)	Air in (T_6)	Air out (T_5)	Brine out (T_2)	Brine Flow Rate (1000 lb/hr)	Air Heater Surface Area (ft^2)
1:a1	300	28.8	80	130	90	136.1	30,000
a2					120	158.6	18,900
b1	260				90	168.6	36,300
b2					120	204.8	22,200
c1	220				90	221.5	46,600
c2					120	288.0	27,600
2:a1	300	39.4	61	130	75	173.9	37,100
a2					105	200.4	24,900
b1	260				75	212.3	44,600
b2					105	237.9	29,200
c1	220				75	271.5	56,800
c2					105	314.9	36,100
3:a1	300	31.2	119	180	130	181.7	40,300
a2					160	220.2	25,000
b1	260				130	238.7	52,900
b2					160	309.8	31,500
c1	220				130	345.9	81,900
c2					160	518.3	45,300

The three main cases detailed above encompass the likely ranges of barley kilning process operating parameters. The subcases, a-c, each arise from different brine temperatures as feed to kiln air heater. Case 3 was determined to be the best general basis for process design, and subcase 3-b1 was finally selected as the most economical alternative at the present level of analysis.

Table 8

Capital Cost of Geothermal Heating System for Malting Kiln

Basis: Capacity of 4 million bushels per year
Case 3 from Table 7.

Subcase	3-a2	3-b1	3-c1
Geothermal Brine Temperature (°F)	300	260	220
Brine Flow (1000 lb/hour)	220	239	346
No. of Wells	4	4	6
Capital Cost (1000 $)			
Wells	2,600	2,200	2,700
Lines, Pumps and Surge Vessels	269	269	360
Air Heater	210	450	700
Fans	80	80	80
Total	3,159	2,999	3,840

These three subcases for the chosen design basis represent the most economical alternatives at the above geothermal brine temperatures. These selections, and the final choice of subcase 3-bl, were made on the basis of net brine flow, required number of wells, depths of wells, and heat transfer surface area.

Estimated operating costs for the system of $2.63 per million Btu are indicated in Table 9 below. For this purpose, capital costs are amortized at 15% per year. The average annual kilning heat load is estimated to be 192,000 million Btu per year, based on a year-around average of 32 million Btu per hour, for 18 hours per day, 333 days per year. This is about two-thirds of the expected average heat load for a plant operating without air recycle around the kiln.

Table 9

Kilning Energy Costs for Case 3-b1

Total capital cost	$2,999,000	
Annualized capital cost (15%)	450,000	
Annual heat load (million Btu)	192,000	
Capital cost of energy (per million Btu)		$2.34
Average brine flow rate (lb/hour)	239,000	
Annual brine treatment costs*	34,300	
Annual power costs (pumps and fans)**	20,600	
Total operating costs for utilities	54,900	
Unit utility costs (per million Btu)		$0.29
Energy cost (per million Btu)		$2.63

* Brine treatment costs of $25 per million pounds.
**Based on 150 HP, and 3¢/Kwh electricity costs.

D. Optional Geothermal Process

The process design option illustrated schematically on Figure 12b was also investigated in some detail. In this case, steam from flashed geothermal brine provides heat to the air heater as the steam condenses at the constant temperature, T_1. By this means, the average temperature driving force across the heater is significantly higher than for the design option of Figure 12a, so that the surface area and the heater cost decrease. However, required brine flow rates would increase significantly, as would well costs, since the exit brine temperature is much higher than for the option of Figure 12a. This is illustrated by the following table for the Case 3 conditions of Table 6, for which a flash temperature of 190°F is taken. The design heat load is 31.2 million Btu per hour.

Table 10

Required Brine Flow Rates for Vapor Recompression Option

Brine Temperature (°F)	Air Heater Outlet Temp. (°F)	Flash Temp.-T_1 (°F)	Brine Flow Rate (1000 lb/hr)
300	180	190	280
260	180	190	442
220	180	190	1,037

In order to reduce the wellhead brine flow rates, the feasibility of a second flash on the residual brine stream was investigated. The steam vapors at the lower temperature, T_2, are then compressed to the pressure required to condense at T_1. The capital cost of the steam compressers is very high for the heat load savings involved so that this case is not economical.

E. Conclusions

In summary, the cost analysis for the application of the San Luis Valley (or a similar quality) geothermal resource to the barley malting industry indicates that geothermal energy supplies can be economically competitive with currently available fossil fuel alternates. In addition, geothermal energy exhibits some distinct advantages with respect to long-term cost projections, environmental considerations, and stability of supply. Further efforts of specific resource development via drilling programs, water quality and quantity evaluations, well testing, and refinements in the systems needed to apply geothermal resource potential to industrial processes would markedly enhance the attractiveness of the geothermal alternative in the near future.

VIII. THE SOUTHERN FRONT RANGE CITIES OF COLORADO: DEMOGRAPHIC ANALYSIS OF WATER AND ENERGY NEEDS

A. Purpose and Subjects of Analysis

The study targeted three Front Range Colorado cities as the subjects in an analysis of the energy costs associated with the construction of a major pipeline to transfer geothermal water to users distant from the resource. To date, this kind of application of hydrothermal resources has received consideration in areas where the cumulative consumer market at a prospective pipeline's end severely limited the total amount of heat that could be consumed, and thus, limited the required pipeline size. This limitation has the effect of reducing economically viable transfer distances. The cost of a major hydrothermal trunkline is developed parametrically in our analysis. The target cities are Canon City, Colorado Springs, and Pueblo, located in a triangle within 45 miles of one another. The potential market for consumption of heat and water in the target cities has been estimated; coordinated with the parametric design of pipeline capacities, this market estimate facilitates the approximation of probable ranges of energy and water costs. Prospective industrial applications of moderate temperature water were surveyed with a view toward those industries found in Colorado and the San Luis Valley Region presently.

The Front Range region of Colorado runs north-to-south at the eastern threshhold of the Rocky Mountains at about the centerline of the State. In recent years, this region has been one of the nationally leading areas of population and economic growth. The bases in this study for the estimation of growth in the demand for resources in that area are applicable data on population growth and on present energy consumption rates. Included in the data on population growth are projections for future years. Data for Alamosa and the San Luis Valley are also presented.

B. Population Analyses

Growth and development in the western United States is perhaps the biggest issue facing communities experiencing the rapid changes which have occurred over the last ten years. Colorado is in many respects the hub of much of this growth and development pressure. It offers the climate and life style desired by most and still provides employment opportunities and tolerable costs of living. With the thrust of energy development shifting to the west, Colorado is viewed as playing an important roll in providing for the needs of the entire nation. As a result, continued pressures for growth have created significant concern among government agencies, area businesses, and citizen groups. Beyond a natural inclination to avoid rapid, unplanned development, communities are becoming increasingly sensitive to finite limits that exist for utilities. Since the late 1960's, Colorado cities have felt the effect of natural gas supplies not being available to meet demand even for residential use alone. Major Front Range cities like Denver, Colorado Springs, and Pueblo all faced moritoria for new gas customers to one degree or another. The 1976-1977 winter mountain snowfall accumulation, which was forty percent below normal, also caused mandatory water rationing and moritoria for new-home services. In other words present water and natural gas* supplies have been overtaken by growth demand which shows little sign of subsiding.

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* 95 percent of all Colorado urban residences are heated with natural gas.

1. Key Indicators

Population statistics best reflect the trends of municipal services. Forecasts of an area's total population and industrial employment are among the best measures of estimating future water and heating demands. For this reason discussion in this section begins with a brief review of population statistics for Colorado and the cities of Colorado Springs, Canon City, and Pueblo along the Front Range, and Alamosa within the San Luis Valley, as located on Figure 13. This review is followed by a forecast of future population according to the best, most widely accepted estimates. Between 1960 and 1970, Colorado's population increased from 1,753,947 to 2,209,258, an annual average of 2.3 percent. This compares to a national average annual growth rate during this period of 1.3 percent. Between 1970 and 1975 the State grew by 14.7 percent to a population of 2,534,000. Of the total increase in this five-year period (325,000) about 35 percent can be attributed to natural increase, the difference between births and deaths, while the remaining 65 percent is a result of in-migration.

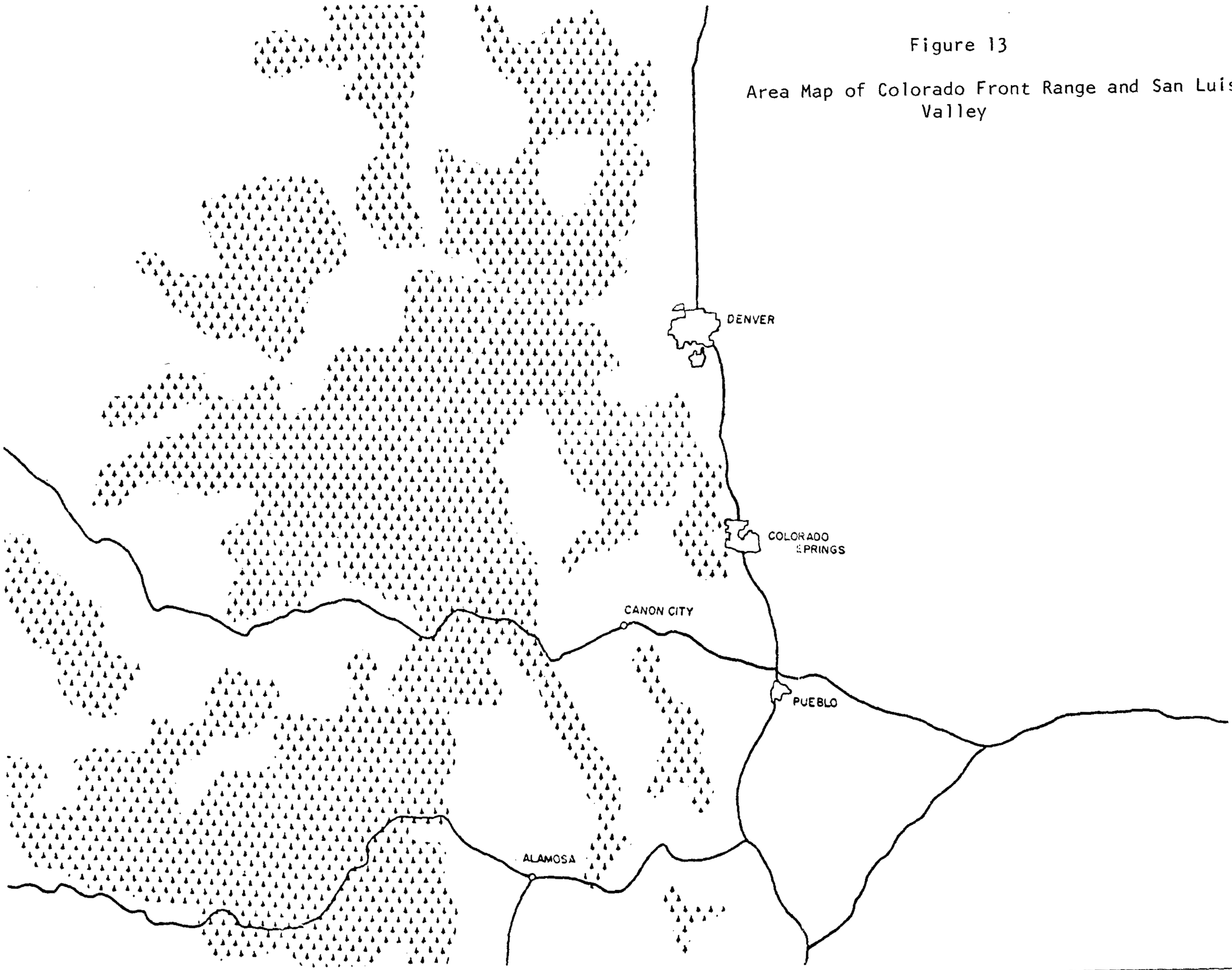
Colorado Springs is located seventy miles south of Denver along the eastern slope of the Front Range of the Rocky Mountains. This community is heavily influenced by neighboring military facilities and has experienced the most rapid growth in the State. In 1960 the Colorado Springs Standard Metropolitan Statistical Area (SMSA - includes all of El Paso County) had a population of 143,742 while 70,192 resided within city limits alone. In 1970 the county grew to 235,972 while the city totaled 135,060 residents, representing average annual increases of 5.1 and 6.8 percent, respectively. Between 1970 and 1975 the county and city estimated populations reached about 300,000 and 180,000, respectively. This represented an average annual growth rate of 5 percent for El Paso County and 6 percent for Colorado Springs. Of the total county increase, 24 percent is attributable to natural increase with in-migration making up the balance, a huge 76 percent. Those who migrated to this area included military (40%), employment related (45%) and retirees (15%).

Canon City is located 45 miles southwest of Colorado Springs in the foothills of the Front Range within the Arkansas River Valley in Fremont County. In 1960 the community's population was 8,973 which grew to 9,206 by 1970, representing a nominal average annual rate of 0.3 percent. Between 1970 and 1975 population reached 13,770, a dramatic change from the previous decade of essentially no growth to a period where annual growth averaged 8.4 percent. The tremendous recent influx experienced by the community has created excess demands for municipal services as discussed later in this chapter.

Pueblo is directly south of Colorado Springs, also along the eastern slope about 110 miles from Denver. This is the most industrialized community for its size in Colorado, employing large segments of the work force in production of steel products, aluminum pistons and ball bearings, meat packing, textiles, insulation, and military installations. The community grew from 91,181 in 1960 to 97,453 in 1970 and 107,000 in 1975, representing annual growth rates of 0.7 and 1.9 percent respectively for these periods. As these statistics reflect, Pueblo has not grown like its neighbors to the north; however, recent estimates reflect increased development in a manner similar to that of Colorado Springs. In the process Pueblo is changing from an industrial to a service based economy, relying more and more on municipal services and utilities.

Figure 13

Area Map of Colorado Front Range and San Luis Valley



Alamosa, in the San Luis Valley, is located 212 miles from Denver and is the commercial center of south-central Colorado. The community developed as a result of the Denver and Rio Grande Railroad establishing a terminal at Alamosa to serve this agricultural community. The 1960 population was estimated at 6,140 reaching 6,880 by 1970 (1.1 percent annual average). In 1975 the population had grown to 7,300 representing a similar growth rate of 1.2 percent.

Table 11 summarizes the growth recorded for the study areas between 1960 and 1975. Since the 1972 Census Bureau report of manufacturing employment, Pueblo in particular has mounted an extensive public relations campaign to promote business growth. Chamber of Commerce statistics show that manufacturing total employment is increasing while unemployment has dropped in the last year from 6.2 to 4.9 percent.

Table 11

Population Growth Statistics

Locale	Population 1975 (est.)	Annual % Population Increase 1960-1970**	1970-1975
Canon City	13,800	0.3	8.4
Colorado Springs	180,000	6.8	5
Pueblo	107,000	0.7	1.9
Alamosa	7,300	1.1	1.2

Population Growth Projections (to the year 2000)

Locale	Population (in 2000)	Average % Annual Increase*
Canon City	24,000	2.2
Colorado Springs	370,000	2.9
Pueblo	165,000	1.7
Alamosa	12,000	2.0

* Based on 1975 populations
** National average in this period: 1.3%

2. Future Projections

Forecasts of population often become volatile issue, especially regarding their accuracy, in communities experiencing rapid growth. For this reason a range of projections was considered for each study area, drawing on forecasts made by State, local and/or regional planning agencies, U.S. Department of Commerce--Office of Business and Economic Research, Current Environmental Protection Agency Wastewater Facility Plans, local development commissions, and chamber of commerce reports. These projections are also summarized in Table 11.

Many estimates provide a high-low range for future years while others rely on a continuation of prevailing trends and geographic limitations. Colorado Springs/ El Paso County projections, prepared by the Pikes Peak Area Council of Governments, looks at employment opportunities and economic base interactions through the cohort survival method developed for initial use in San Diego. For Canon City, recent projections developed for regional wastewater planning seem to be widely accepted and appropriate for our purposes. The Pueblo Area Council of Governments prepared high, low, and average projections for the Pueblo Standard Metropolitan Statistical Area (SMSA) using the "Colorado Population and Employment Forecasting Model (CPE)" developed by the University of Colorado. This model also was the cohort-survival technique and develops range estimates similar to those of the State Division of Planning. The San Luis Council of Governments, in conjunction with an airport master planning study, developed forecasts for Alamosa City and East Alamosa. These forecasts, and county estimates by the State, represent the most current information available for Alamosa and are used as such for this study.

C. Municipal Water Consumption

Potable water in Colorado comes from any of three sources: surface stream diversion, trans-basin diversion, and aquifer wells. Western slope communities rely primarily on surface diversions with some wells, while eastern slope cities depend more and more on wells and trans-basin diversions. Major metropolitan areas along the eastern slope have developed extensive water projects in the mountains diverting water from the Colorado River basin via long tunnels and augmentation systems. These diversion projects are topics of heated debate between eastern and western slope residents. Recent drought conditions have served to intensify the trans-basin diversion issue, as rapid growth strains available reserves. Following are brief descriptions of water usage in the target areas. Total water consumption figures are presented in Table 12.

Canon City: Canon City takes its water from the Arkansas River. Annual consumption in 1975-1976 was about 5,600 acre-feet, with a peak value corresponding to a gross annual rate of 11,000 acre-feet. Major improvements and additions to the City water system are anticipated in the coming two decades.

Colorado Springs: The City of Colorado Springs obtains potable water from developed watersheds in the mountains west of the city, from trans-basin diversions from the Arkansas and Blue River Basins further to the west and northwest, from three Front Range wells east and south of the city, and from local streams. Total average annual yield was estimated at 80,000 acre-feet.

Pueblo: Statistics obtained from the Pueblo City government show that total 1976 water consumption was approximately 27,000 acre-feet. Peak consumption rates are about 188 percent of the average. Present treatment facilities are sized for about 170 percent of that peak load. The source of water is the Arkansas River.

San Luis Valley: Water consumption figures for the Valley, published in the "State Water Plan, Phase 1: Appraisal Report of Present Water Resources and Uses," February 1974, attribute 6,000 acre-feet to municipal consumption and 617,000 acre-feet to agriculture. The primary water sources are the Rio Grande River and thousands of wells in the Valley's unconfined aquifer.

Table 12

Summary of Total Water Consumption

(Acre-Feet per Year)

Canon City	5,600
Colorado Springs	80,000
Pueblo	27,000
San Luis Valley	623,000
	735,600

D. Energy Consumption

Fuel consumption data, for natural gas in 1976, has been provided by Public Service Company of Colorado. This information is depicted graphically in Figures 14 through 17, and summarized in Table 13 according to residential, commercial, and industrial categories. Additionally, we obtained statistics on net energy consumption for a number of the largest users in the three Front Range cities and in the San Luis Valley. These figures are presented in Table 14. For comparison, the latter Table also shows the approximate annual total fuel energy consumed to produce process steam for a 5,000-ton-per-day beet sugar plant, operating on a 120-day annual campaign.

Table 13

Total Annual Gas Consumption--(Millions of Btu's)

	Residential	Commercial	Industrial	Military
Canon City	1.01 E06	5.97 E05	2.54 E06	--
Colorado Springs	8.70 E06	7.61 E06	2.56 E06	1.64 E06
Pueblo	3.69 E06	1.59 E06	3.28 E06	--
San Luis Valley	6.60 E05	5.52 E05	1.08 E05	--

E. Demographically Based Market Estimates

The foregoing data gives a good, approximate picture of the water and energy consumption rates for the Front Range cities and the San Luis Valley. The Front Range cities together consumed almost 113,000 acre-feet of water in 1976. Per capita water usage has been increasing regionally just as per capita energy consumption increases with strong economic and population growth in many cases; but for this study, it is beyond the scope of work to attempt to account for such variations. Only population projections will be used in projecting proportional increase in resource demand, and, as discussed, such projection becomes very tenuous beyond a very few years into the future.

The cumulative population prediction for the three Front Range cities, for the year 2000, is 559,000 inhabitants, amounting to more than 85 percent increase. Thus, water consumption can be estimated to reach about 210,000 acre-feet annually by 2000, assuming population, commerce, and industry profiles for the region remain roughly equivalent to this present point in time. This is a good assumption; the interest of these cities in growth explicitly exhibits concern for realizing carefully controlled growth in order to avoid developing maldistributions of economic wealth in the process.

Again making no attempt to predict variations in per capita energy consumption rates, the increase in energy consumption for the Front Range cities may be estimated on the basis of the expected population growth by the year 2000, for an increase of 85 percent of present levels. Presently, energy consumption for the three cities totals about 31.6 trillion Btu's annually, excluding the portion attributed to military facilities. The projected value would be 58.4 trillion Btu's annually.

F. Applications for Geothermal Heat (14)

As shown in Chapters V and VII, a 302°F geothermal resource shows strong, prospective applicability for replacement of large amounts of process heat in industrial plants exemplified by the processing of food products such as beet sugar and malted barley. The promising economics in the case of the sugar plant design is particularly notable because that process is first among the foods industries in energy consumption (11).

A survey of other possible applications was conducted. With respect to the existence and utilization of a reservoir in the San Luis Valley, it was desirable to identify potential candidates among industries indigenous to this region of the U.S. The variety of agricultural products in the region offers numerous, ready prospects, notably dairy products, barley, potatoes for starch or food products, lumber, aquaculture, and livestock management. There are two potato processing plants in the San Luis Valley, and at least one of them is presently undertaking steps to deal with pollution problems related to fossil fuel use. Dairy products and eggs processing systems generally operate at temperatures ostensibly well suited to moderate temperature geothermal resource capabilities, and are very attractive for further analysis. Basic metals, refractory materials, and a selection of minerals are key industrial products throughout this sector. A goodly number of industries listed in the literature and showing operating temperatures in suitable ranges are presently operating in the vicinity of the Front Range target cities. Some prime industrial targets for analysis and recruitment in this region include the following (15):

Process	Range or Maximum Operating Temperatures (°F)	Heating Medium
Concrete Block and Brick	165-350	Steam
Borax	140-210	Steam/air
Bromine	225	Steam
Chlorine	150-300	Steam
Phosphoric Acid	250-320	Steam/air
KCl	200-250	Steam/air
Sodium	240-275	Steam/air
Glass	70-350	Steam/air
Lumber	150-350	Steam/air
Synthetic Rubber	250	Steam/air
Steel	150-220	Steam/air
Textiles	70-300	Steam/air
Soup Canning	140-210	Steam/air
Dairy	140-212	Steam/air
Gasohol	120-280	Steam/air

A recent Department of Agriculture program to develop processes for the manufacture of alcohol from cellulostic materials is especially attractive for several reasons. First, it is stipulated that no such "gasohol" process to be used as a study model can be a net fossil fuel energy consumer. Second, the process can be developed in conjunction with, specifically, the beet sugar process as a source of raw materials.

The variety of seemingly attractive prospects for conversion to geothermal energy is very encouraging on the basis of the comprehensive feasibility studies of this project. Definite analytical steps should soon be taken to refine and broaden our detailed data on the real costs of industrial geothermal process heating based on rigorous process design.

Table 14. Major-User Energy Consumption Statistics

(MMBtu)

Canon City	Annual Energy Usage
1. Colorado State Penitentiary	117,800
2. Cotter Corporation	148,700
3. Colorado Refractories	63,700
4. Fremont Paving	16,800
5. DFC Ceramics	18,900
6. St. Thomas More Hospital	7,300
7. Canon City High School	9,800
8. Royal Gorge Flower Farms	44,800
9. Western Forge Corporation	18,400
10. Total	446,200

Pueblo	Annual Energy Usage
1. Dana Corporation	10,600
2. Aspen Ski Wear	1,700
3. Colorado State Hospital	333,000
4. U. Southern Colorado	10,500
5. Pueblo School District #60	75,800
6. St. Mary Corwin Hospital	103,900
7. Parkview Hospital	20,800
8. Pueblo City Hall	4,700
9. Total	561,000

Colorado Springs	
1. USAF Academy	1,489,400
2. Fort Carson Army BAse	1,745,900
3. Peterson A.F. Base	216,500
4. Unidentified	80,100
5. Unidentified	69,000
6. Unidentified	219,500
7. Unidentified	117,100
8. Unidentified	104,900
9. Unidentified	631,900
10. Unidentified	92,800
11. Total	4,767,100

San Luis Valley	
1. Adams State College	147,500
2. Nonpareil Starch Company	136,200
3. Alamosa High School	112,600
4. Monte Vista Veterans Center	37,000
5. Total	433,300

Beet Sugar Plant (Process Steam)	
5,000 Ton/Day Capacity	1,825,900
120-Day Campaign	
0.85 Boiler Efficiency	
75 Percent Steam on Beets	

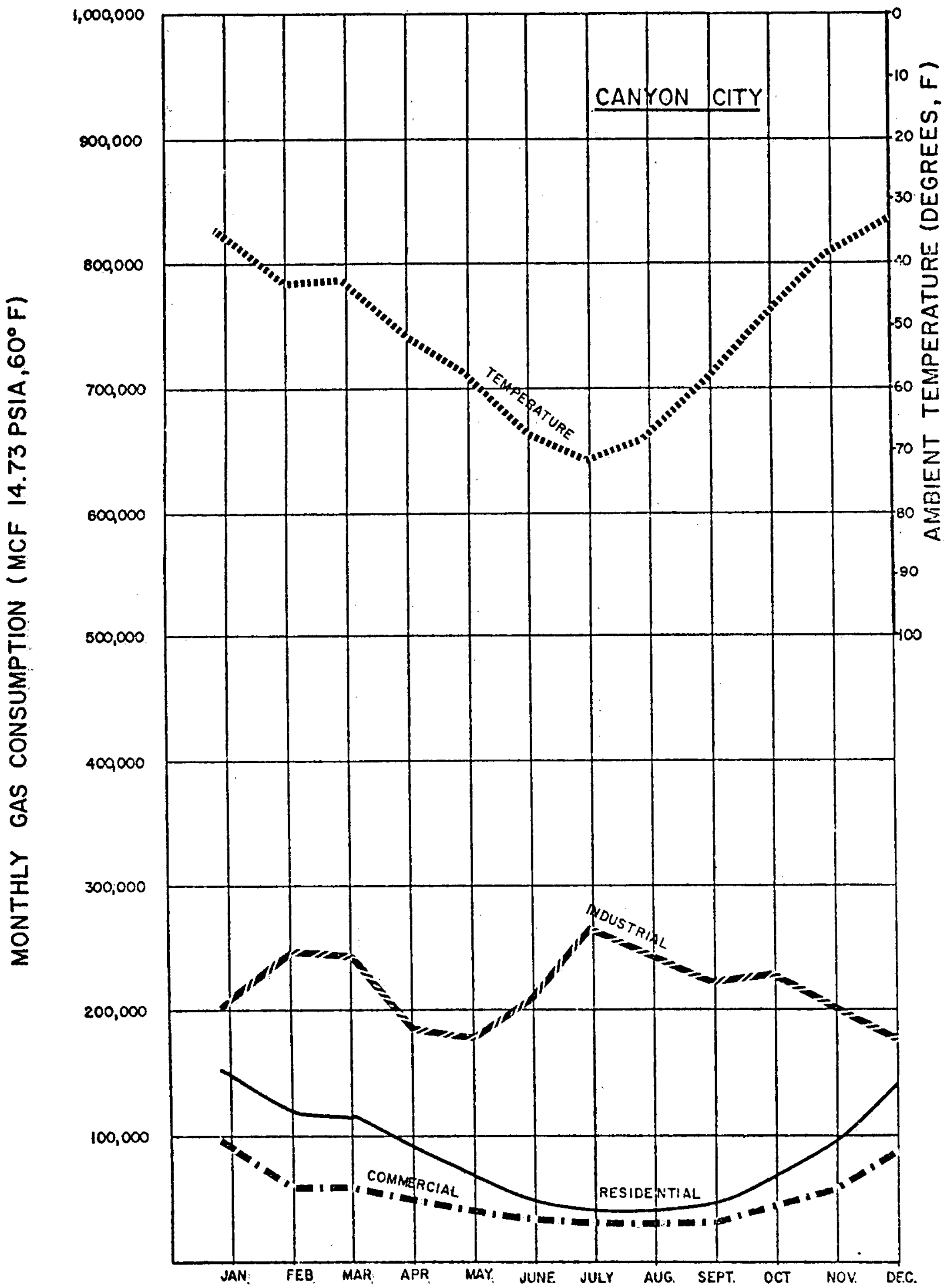

CANYON CITY
MONTHLY GAS CONSUMPTION (MCF 14.73 PSIA, 60° F)
AMBIENT TEMPERATURE (DEGREES, F)
1,000,000
900,000
800,000
700,000
600,000
500,000
400,000
300,000
200,000
100,000
0
10
20
30
40
50
60
70
80
90
100
TEMPERATURE
INDUSTRIAL
COMMERCIAL
RESIDENTIAL
JAN.
FEB.
MAR.
APR.
MAY.
JUNE
JULY
AUG.
SEPT.
OCT
NOV.
DEC.

Figure 14

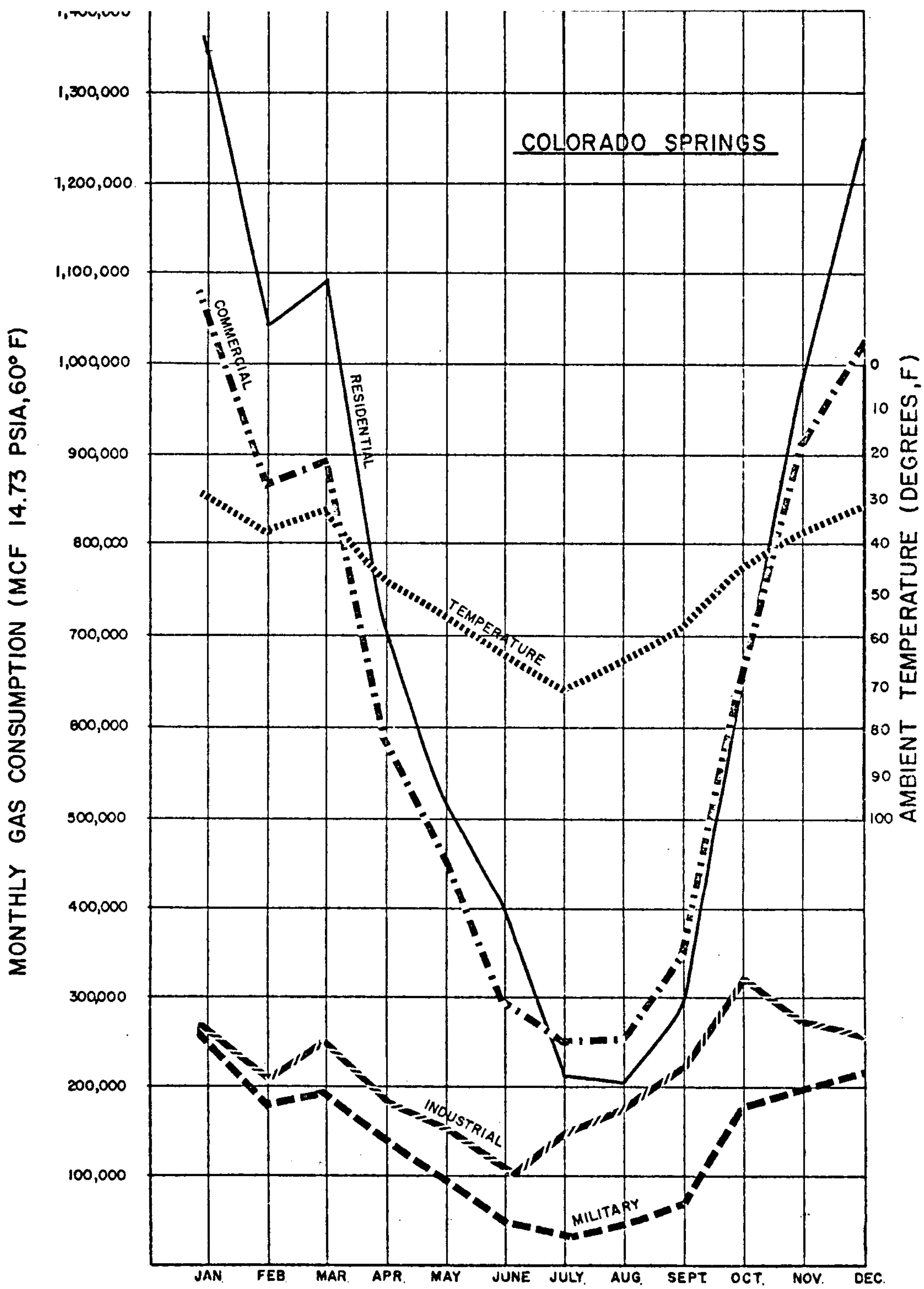
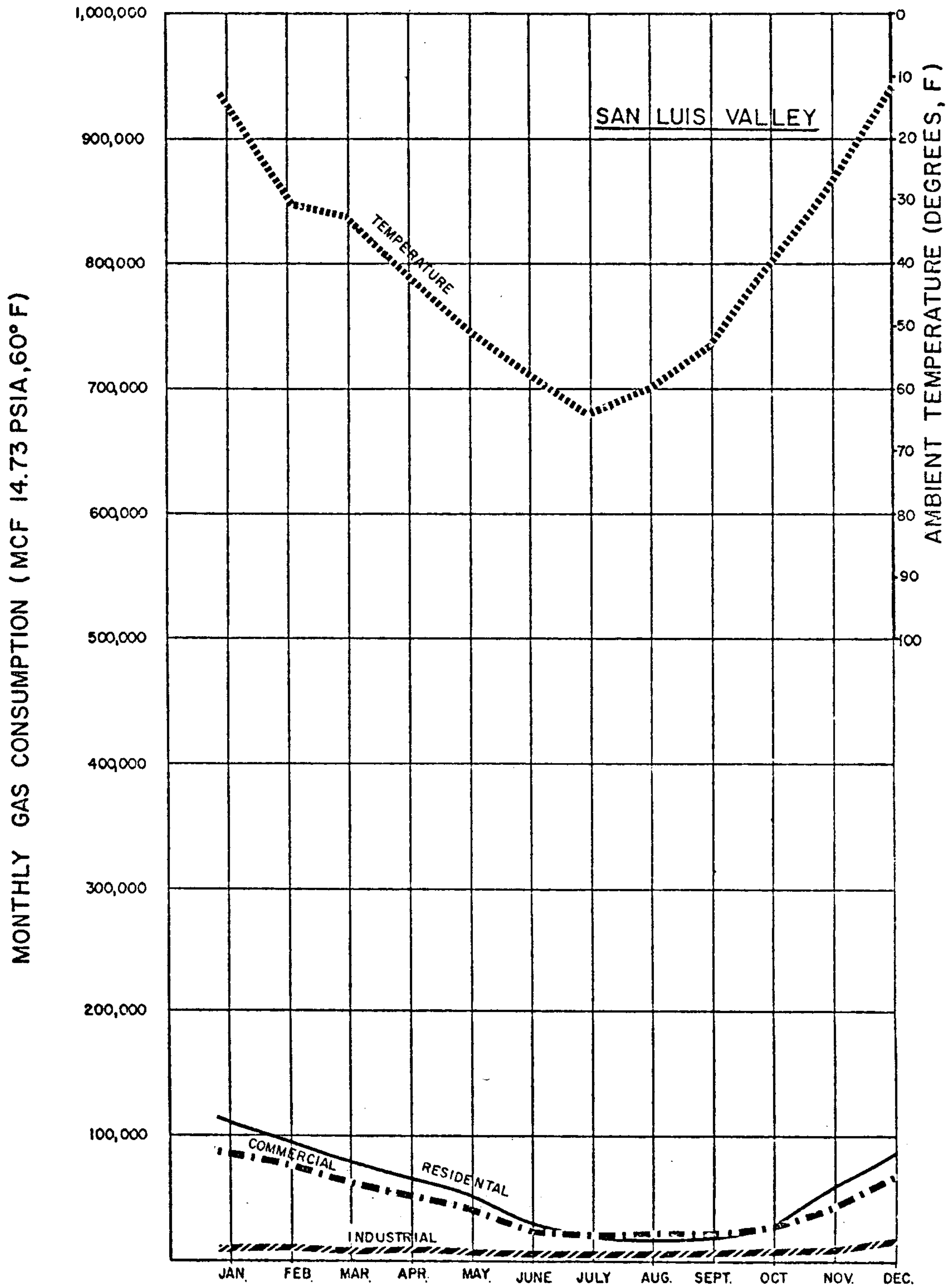

COLORADO SPRINGS
MONTHLY GAS CONSUMPTION (MCF 14.73 PSIA, 60° F)
AMBIENT TEMPERATURE (DEGREES, F)
1,300,000
1,200,000
1,100,000
1,000,000
900,000
800,000
700,000
600,000
500,000
400,000
300,000
200,000
100,000
0
10
20
30
40
50
60
70
80
90
100
COMMERCIAL
RESIDENTIAL
TEMPERATURE
INDUSTRIAL
MILITARY
JAN.
FEB.
MAR.
APR.
MAY
JUNE
JULY
AUG.
SEPT
OCT.
NOV.
DEC.

Figure 15


METRO PUEBLO
MONTHLY GAS CONSUMPTION (MCF 14.73 PSIA, 60° F)
AMBIENT TEMPERATURE (DEGREES, F)
1,000,000
900,000
800,000
700,000
600,000
500,000
400,000
300,000
200,000
100,000
0
10
20
30
40
50
60
70
80
90
100
TEMPERATURE
INDUSTRIAL
RESIDENTIAL
COMMERCIAL
JAN.
FEB.
MAR.
APR.
MAY.
JUNE
JULY
AUG.
SEPT.
OCT
NOV.
DEC.

Figure 16



SAN LUIS VALLEY
MONTHLY GAS CONSUMPTION (MCF 14.73 PSIA, 60° F)
AMBIENT TEMPERATURE (DEGREES, F)
1,000,000
900,000
800,000
700,000
600,000
500,000
400,000
300,000
200,000
100,000
0
10
20
30
40
50
60
70
80
90
100
TEMPERATURE
COMMERCIAL
RESIDENTAL
INDUSTRIAL
JAN.
FEB.
MAR.
APR.
MAY.
JUNE
JULY
AUG.
SEPT.
OCT
NOV.
DEC.

Figure 17

IX. OVERLAND TRANS-SHIPMENT OF GEOTHERMAL WATER

A. Basis for Analysis of Long-Distance Transport

A preliminary analysis was made of the economic feasibility of transporting very large quantities of geothermal energy over long distances to the locations of concentrated or aggregate energy consumers. The basis for undertaking this effort is the fact that most geothermal resources are located in remote areas with respect to viable, large users. It appears feasible that individual, large users, such as sugar refiners or barley maltsters, would be able to economically establish a part of their operations near specific geothermal sites. However, a very large number of years are likely to pass before living patterns change significantly in the sense that large population shifts toward geothermal areas could occur because of factors of energy costs and energy availability. In view of these considerations, it is more probable that large increases in geothermal energy utilization will occur in the next ten to thirty years only if the energy or fluid from a geothermal resource is moved to existing conglomerations of users.

Transporting moderate temperature geothermal energy tends to be expensive in many cases because of a high ratio of capital costs to net available energy. These initial costs are generally high, for both large and small geothermal systems, and due to economics of scale they do not necessarily increase rapidly as the system size increases. Several other studies have been made of small to medium-size pipelines for transferring geothermal water to small communities or industrial complexes, and these studies showed the economic transfer distances were most probably less than 100 miles in the range of line sizes considered. Blanket statements on such costs are not reliable because of the complex dependence of the costs on the resource quality, the distance and type of terrain traversed, and the costs of distribution and disposal systems at the consumer end of the pipeline. Thus, each potential application for overland trans-shipment of hot water really requires individual, detailed analysis. However, it is fruitful, as a first step in evaluating the realistic scope of potential use for the San Luis Valley geothermal resource, to consider the likely rangeability of such a pipeline system cost.

The results of the present study provide support for the economic viability of transporting very large amounts of energy over distances of the order of 150 miles.

B. Design Procedure

A major pipeline delivery system was designed parametrically to evaluate a range of flow rates in economic terms. Other studies have considered pipelines of up to 30- to 36-inch diameters. The lower limit selected for study in this project is 24 inches. In fixing the upper limit for a pipeline diameter, we first calculated the total available heat in the geothermal water, assuming a sensible heat loss through user consumption equivalent to a 150°F temperature drop. At a flow rate of 100,000 acre-feet per year, that quantity of heat is nearly 40 trillion Btu's annually. In the preceding chapter, Table 13 summarizes the energy consumption for the Front Range target cities and for the San Luis Valley, giving a total energy consumption of less than 35 trillion Btu's annually. Thus a 100,000 acre-feet-per-year influx of geothermal water about doubles all present energy consumption levels if fully utilized, and this goes well beyond the energy demand extrapolations based directly on population projections that extend to the year 2000. As an upper limit on pipeline sizes, then, a 60-inch line was specified, which would convey 85,300 acre-feet annually at a line velocity of 6 feet per second. Four prospective pipeline diameters were used for this analysis: 24-inch, 36-inch, 48-inch, and 60-inch.

A limited degree of design detail and sophistication was considered explicitly in the investigation. To the greatest extent possible, all aspects of design were conservatively factored into the overall installed component costs. Manufacturers or installers of major capital items such as pumps, insulation, and large-diameter piping were contacted for direct help in estimating costs. Sources at the U.S. Bureau of Reclamation were helpful in providing information on current standards for sizing large-scale pipelines, for commonly accepted operating procedures and parameters, and for information used in quantifying cost estimates for installed, buried steel pipelines on a weight basis. There were broad differences in the various estimates provided directly or which were developed on the basis of discussions with multiple contacts. For the cost of installed piping, itself, data developed during work on the Susanville Geothermal Energy Project (Longyear, 1976) was used in generating low-range values to compare against other higher installed-pipe cost estimates.

The route chosen for the pipeline was selected for its directness. It would require a 1,500-foot elevation change within the San Luis Valley, a tunnel approximately 5 miles long (Note: The proposed tunnel would run under Hayden Pass, which is situated east-northeast of Villa Grove, Colorado), and then up to 4,000 feet of elevation change on a predominantly downhill run to the Front Range target cities. The route is tabulated in greater detail in Table 15.

Cost estimates for installation of major pipelines specifically in rugged terrain of the sort proposed for this design case are virtually impossible to acquire from the pipeline industry without a costly design study. It was pointed out by Beard (18), late in the study, that a pipeline route to the south and east of Alamosa, though it would be 25-30 percent longer than the proposed route, may be preferable because it would entail traversing less severe terrain and also holds the attractive possibility of sharing established highway or railroad rights-of-way. The cost estimates presented later in this chapter were based on the shorter, more difficult route. The effect of this choice is minimized, however, because a range of possible cost values was used in the cost estimate, so as to clearly reflect the uncertainties indicated above.

C. Bases for Design

Design data for the pipeline are summarized in Table 16, including diameters, capacities, wall thickness, frictional pressure losses, and unit mass. Also shown are installed unit-cost maxima and minima, which incorporate costs for the assembled (welded) steel pipe, application of insulation (but not the capital cost of insulation), and burial of the pipeline. Data pertaining to the effects of 20 fps velocities, used to absorb potential energy, are also included in the Table.

1. Pipeline Capacities, Frictional Losses

Two fluid velocities were selected. For most situations a velocity of 6 feet per second (fps) was used. Significant downhill gradients will occur, necessitating some means of absorbing the associated potential energy addition to the fluid. In such cases the fluid velocity was increased to 20 fps to take advantage of increased frictional pressure losses. Portions of the net downhill runs were scaled down in diameter to achieve velocities of 20 fps, based on the volumetric flow rates for

24-inch, 36-inch, 48-inch, and 60-inch lines flowing at 6 fps. To avoid ambiguity possibly introduced in discussion of pipeline cases because of the variations that occur in pipeline sizes within a given case study, the cases for this study are designated according to the initial trunkline sizes for flow rates at 6 fps in the San Luis Valley; i.e., the 24-inch, 36-inch, 48-inch, and 60-inch cases, respectively.

Frictional pressure losses were calculated via the Fanning friction factor coefficients, for water at approximately 300°F, using the equation for incompressible, turbulent flow in a conduit:

$$144 \frac{\Delta p}{\rho} = 2f \frac{L}{D} \frac{v^2}{g_c} \qquad (6)$$

where:

Δp - pressure loss, psi
ρ = liquid density, lb_m/ft^3
f = Fanning friction factor from Moody chart
L = conduit length, feet
D = conduit inside diameter, feet
v = fluid velocity, feet per second
g_c= gravitational constant, 32.2 $ft\text{-}lb_f/lb_m\text{-}sec^2$

Adiabatic flow is assumed, such that all frictional energy loss is transferred as work on the fluid, leading to a rise in fluid temperature. Related to the above equation, the frictional work performed on the fluid in one mile may be expressed as:

$$W = \frac{2fLv^2}{g_c D} \quad \text{foot } lb_f/lb_m \qquad (7)$$

This expression was converted to Btu units and used to determine the temperature rise for each case, using equation (8) below. Table 16 summarizes flow rates, diameters, frictional pressure losses, and the temperature rise (at a 20 fps velocity only).

$$\Delta T = W/778C_p \quad {}^{o}F/mile \qquad (8)$$

where:

C_p = liquid specific heat, $Btu/lb_m\text{-}{}^{o}F$

2. Wall Thicknesses and Installed Pipe Costs

Minimum pipe wall thicknesses were selected conservatively, partially based on U.S. Bureau of Reclamation pipeline design criteria. The design data is based on 60-inch steel pipe, for a maximum pressure of 520 psia and weighing 704 pounds per foot of length; the steel has an approximate density, ρ, of 490 lb/ft^3. The minimum wall thickness is determined as follows for a one-foot section of 60-inch diameter pipeline;

$$M = 704 \text{ lb/ft} = (\pi\rho t/4)\ (2D+t) \qquad (9)$$

The vapor pressure of $300^{o}F$ water is 67 psia; for design purposes the minimum pressure rating of the pipeline was taken as 85 psia. A minimum thickness of 0.25 inches was thus selected for a 24-inch pipeline at 85 psia. For alternative pipe diameters and operating pressures, minimum wall thicknesses may be calculated in direct proportion to the above values. The installed costs of such pipe with application of insulation, and including burial of the pipe below the frost line, were estimated to be about $0.90 per pound. These costs are plotted as a function of line diameter in Figure 18.

The pipe system is segmented such that maximum pressures do not exceed 520 psia. In those downhill sections for which velocities are raised to 20 fps, the pressure drops incurred were designed such that the maximum pressure in any segment would also be less than 520 psia. Consequently, the pipeline is considered to consist of many segments, each having a wall thickness of the average of the pairs of extreme values corresponding to 85 psia and 520 psia, depending on the pipeline diameters involved. This data is also summarized in Table 16.

In comparison, data is also shown in Table 16 for buried steel pipeline costs, based on data from Longyear (13) for the Susanville, California geothermal project study, conducted in 1975 and 1976. An annual inflation rate of about 6% was applied to the adjusted January 1975 data. Costs based on the Longyear data are also plotted on Figure 18. These costs had to be extrapolated in the range of 54- to 60-inch pipe diameters.

3. Temperature Losses, Insulation Costs

Heat losses to the surroundings were analyzed for both buried and above-ground piping. For surface pipes, outside film coefficients were taken as 30 $Btu/ft^2 hr^{o}F$; for buried pipe an effective outside film coefficient was used, having a value of 1 $Btu/ft^2 hr^{o}F$. Inside film coefficients ranged between 65 and 515 $Btu/ft^2 hr^{o}F$ for $250^{o}F$ water, and between 70 and 585 $Btu/ft^2 hr^{o}F$ for $300^{o}F$ water. Thermal conductivities of steel and insulating material were taken as 25 $Btu/hr\ ft^{o}F$ and 0.05 $Btu/hr\ ft^{o}F$, respectively.

Numerous sets of calculations were run for pipelines ranging from 8-inch to 72-inch diameters. Heat losses were determined for fixed thicknesses of insulation, with variation of inside film coefficients over the ranges presented above. Insulation thicknesses were determined such that fluid temperature losses at a velocity of 6 fps were restricted to $25^{o}F$. In addition, based on San Luis Valley weather data presented in Chapter VII, extensive evaluation was made of heat loss as a function of possible log mean temperature differences. Table 17 presents minimum, maximum, and average values for the overall heat transfer coefficients, based on inside surface area of the pipeline, as determined for both buried and above-ground pipelines using 1 and 2 inches of insulation, and taking consideration for the largest range of inside film coefficients. The narrow range of values of the overall coefficient, U, over the broad ranges of diameters and film coefficients considered is noteworthy. The results show that a buried pipeline with 1 inch of insulation will lose about 50 percent less heat than the same pipeline above ground; with 2 inches of insulation, the greatest savings in heat losses for a buried pipe amount to only about 22 percent of the above-ground heat losses.

For a worst-case log mean temperature difference of 270°F, allowing about a 50°F fluid temperature drop at 10°F ambient temperature, a buried 24-inch pipeline would require about 8 inches of insulation, while a 60-inch line would require about 6 inches of insulation. Table 18 presents costs for 7 inches of exterior insulating material for the range of pipeline diameters considered, based on data for Johns-Mansville liquid insulating material supplied in 55 gallon drums, to provide a worst-case estimate for a buried pipeline.

Costs for application of the insulation are built into the capital costs of the installed pipeline, presented on Figure 18.

4. Water Treatment

A conservatively estimated brine quality is presented in Table 22 found in Appendix C. Treatment for pipeline trans-shipment consists of acidification to drive off carbon dioxide, followed by neutralization with a base. This gives a pretreatment cost of $25 per million pounds of brine.

5. Pumping Requirements

Within the San Luis Valley two pumping stations are necessary to supply 1,500 feet of static head plus compensation for frictional losses. Between Canon City and Colorado Springs, two stations are necessary to supply 1,300 feet of static head plus compensation for frictional losses. Intermediate surge vessel capacity is not included in the costs.

For a final pipeline design it would probably be economical to incorporate hydro-generating equipment for use in the downhill runs following transit of the Hayden Pass tunnel, in order to partially recover potential energy from the several thousand feet of hydrostatic head available. This would be done in lieu of using reduced pipeline diameters to damp the potential energy added to the water. The associated run lengths at reduced diameters would be 4 to 5 miles, offering little capital cost benefit. The use of generating plants, on the other hand, could compensate for a major portion of the required pumping power. Table 19 presents approximate total system pumping costs for various pipeline capacities; it is assumed that 60 percent of the gross flow will be directed to Colorado Springs for the purposes of sizing pumping stations on that leg of the system. An additional 27 percent would be directed to Pueblo, but that leg of the system would require no pumping as there is a 700-foot drop in elevation.

6. Tunnel Costs

The cost for a tunnel 5 miles long under Hayden Pass and between elevations of 9,500 to 9,000 feet is estimated to be approximately $4.5 million per mile, or $22.5 million complete.

D. Production and Disposal of Geothermal Fluids

The largest pipeline capacity considered would carry 85,300 acre-feet annually. To supply this net volume would require 100 wells, each flowing at about 530 gallons per minute. This flow rate is approximately the minimum estimated capability for the San Luis Valley resource, based on the results of the hydrological evaluation discussed in Chapter III. For a conservative result, a maximum of 110 production wells are used in the cost analyses of this chapter. The smaller pipeline sizes are assessed proportionately fewer wells. These wells are designed on the bases first stated in Chapter V. The required depth for attaining 302^{o}F fluid temperatures is 10,000 feet; drilling and completion costs amount to \$65 per foot of depth. The net cost per well is then \$650,000. It is also assumed that the same number of injection wells, of the same depth and total cost as production wells, will be required for disposal of spent brine.

It is highly speculative to propose specific geothermal waste water disposal methods for the Front Range target cities. The presumed injection of the water in deep wells is one option. Other possibilities include adding the cooled brine to the Arkansas River from Canon City or Pueblo, irrigation, consumption in industrial and municipal water systems, or further transport of the water to the outlying plains areas in need of water. Consider the further use of geothermal water for municipal-industrial consumption.

The three Front Range cities in this study, as well as Denver and other developing metropolitan areas in Colorado's Front Range strip, have to plan well ahead of the present for major water importation and treatment systems to supply their rapidly growing municipal and industrial water needs. Aside from the economic expense of gathering the sparsely distributed water supplies in the western mountain regions, strong competition for water is also a major problem to contend with. Presently the large gathering systems serving the metropolitan areas are almost exclusively surface water supplied. A geothermal pipeline in conjunction with a full treatment plant could provide potable water as well as heat. Consumption of geothermal water would partially relieve the competition for surface waters.

The necessary treatment plant to produce potable water from geothermal effluent obviously may be expected to be more costly than would be required for contemporary municipal water systems receiving much better quality feed water. Equally as obvious, though, the cost of such a water delivery and treatment system would be partially borne by both the use of the energy in the hot water, and the consumption of the water.

Thus, the final consideration from the perspective of energy supply--waste water disposal--leads to several alternative utilizations of the resource. Practices such as lengthening the beet sugar processing campaign, finding multiple compatible heat consumers, and re-using geothermal water as water after the heat content is depleted, all have the common effect of allowing costs to be amortized over a broader consumer base. Taking both heat and water from a geothermal reservoir is not a new concept. The water and energy needs of the Front Range area of Colorado strongly suggest making detailed evaluation of such alternatives.

E. Subsidiary Geothermal Water Handling Systems

Distribution of hot water from a terminal point in a large trunkline, to end users such as factories, refineries, commercial centers, or homes would all be by very similar piping systems. The complexity and, therefore, the costs of these systems would have very broad possible ranges. Those costs have not been included in this analysis.

As a basis for rough estimates, however, consider the sugar process modifications. The capital costs of delivering hot water and then extracting the heat amounted to about 30 to 40 percent of the total well costs. So, to estimate the total costs of municipal-industrial water distribution and heat recovery systems to enable the use of geothermal energy, one could start by using 40 percent of the total well costs for both production and reinjection.

F. Summary of Pipeline Costs

1. Based on the Proposed Route

Table 20 presents a cost summary for the four trunkline cases based on 24-, 36-, 48-, and 60-inch diameters, initially. It is to be remembered that heat consumption based on natural gas utilization for the region studied presently totals about 35 trillion Btu's annually. An economical pipeline will probably supply enough heat to amount to a substantial portion of present consumption levels. Thus, implementation of utilization of heat from hot water at that magnitude is a long-term proposition requiring not only careful planning, but also a shift toward a great deal of new or uncommon, though not highly sophisticated, engineering technology. Utilization of the hot water is based on realization of a 150°F temperature drop through consumption; an initial 25 to 50°F will be lost in transit. For the smaller pipeline systems, the $22.5 million tunnel cost becomes a very substantial energy-cost factor. The trend of costs favoring the larger pipelines, as well as the cost magnitudes, in the context of projected population and energy consumption growth rates, further emphasizes the major scale of the undertaking of overland trans-shipment of geothermal water.

The pipeline legs between Canon City and Colorado Springs, and between Canon City and Pueblo carry 60 and 27 percent of the gross wellhead flow, respectively, and diameters for those legs are appropriately reduced to maintain 6 fps velocities. This split of volumetric flows is based on the relative amounts of gas consumed, as listed in Table 13; this basis excludes the gas consumption listed for military facilities.

Section A of Table 20 presents the cost analyses for the pipeline systems, omitting the costs of production and disposal wells, and as noted in section E, above, omitting subsidiary distribution and heat recovery systems. Section B of Table 20 presents a comprehensive cost estimate for a trunkline geothermal water delivery system from the San Luis Valley to three terminal cities, including the total well costs with other capital charges. The results presented in section A are, thus, component costs for the bulk delivery systems, alone; the results in section B represent gross costs for the long distance trans-shipment of geothermal energy between regions. A third level of analysis, factoring in the costs for distribution and heat recovery systems at the terminal cities, can readily be performed following design and cost estimation. The most general approach to accounting for the last factor, presented in section E, would increase the maximum estimated costs in Table 20-B by about 6 percent for a 24-inch pipeline, and by about 7.4 percent for a 6-inch line.

The final costs presented in Table 20-B lead to several conclusions. The minimum cost estimates predict that the cost of transporting geothermal water, to be used only as an energy source, will decrease markedly with increasing pipeline capacity; this decrease continues above the 60-inch line size. Considering present average energy cost figures, these estimates show that unless conventional energy costs escalate rapidly in coming years, it would be difficult to justify considering a pipeline project of smaller diameter than 36 inches, where the pipeline cost as a minimum would amount to about $3.05 per million Btu's. The minimum cost estimates reflect favorable economies of scale.

The maximum cost estimates, on the other hand, are more expensive and also less favorable in that the benefits of economies of scale are seriously diminished for pipe diameters greater than 36 inches. The maximum estimates indicate a pipeline would approach a limit of about $3.60 to $3.70 per million Btu's contribution to the cost of energy for lines of 48-inch and larger diameters.

2. Based on an Alternative Pipeline Route

An alternate route was considered, running first south, then east of Alamosa following Highway 160, and then running north to Pueblo, first, instead of to Canon City. This route adds approximately 30 percent in length to the main trunk-line of the route used in the base cases. The pipeline branches to Canon City and Colorado Springs are assumed to be the same as for the base cases. The terrain via this route is less severe, making construction easier and less costly; the added distance may lead to approximately identical pumping requirements in spite of lesser elevation changes. The possibility of "piggy-backing" the pipeline along the existing major highway right-of-ways was cited by Beard as a significant incentive. Furthermore, the expensive Hayden Pass tunnel would be eliminated.

The quantifiable cost differences result from the extra distance and elimination of the tunnel, affecting pipeline mass and insulation. All other factors and costs remained constant. These factors were used to re-estimate overall pipeline costs corresponding to those in Table 20-A, based on cost data developed independently in this study. In addition, in the alternate route the potential-energy-absorbing size reductions used with the base cases were not included. The following values may be compared to the corresponding maxima listed in Table 20-A:

Main Line Diameter (inches)	24	36	48	60
Total Capital Costs (10^6\$)	109.3	245.9	417.1	668.5
Total Hourly Costs ($/hr)	2150	4770	8110	12,900
Energy Costs ($/MM Btu)	3.65	3.59	3.44	3.50

These values indicate, on a preliminary basis, that the longer route would be more expensive than that used in the base cases for pipeline diameters greater than 24 inches.

Table 15

Segment (Terminal Points)	ΔH* (feet)	Δp** (psi)	Elevations (feet)	Length (miles)
Well site-Tunnel	1500	650	7500-9000	44
Tunnel	-500	220	9000-8500	5
Tunnel-Arkansas River	-1800	780	8500-6700	4
Arkansas River-Canon City***	-1000	430	6700-5700	32
Canon City-Pueblo	-700	300	5700-5000	32
Canon City-Colorado Springs	1300	560	5700-7000	30
	-1000	430	7000-6000	15
	Total Pipeline Mileage			162

* Elevation change
** Hydrostatic pressure change
*** This segment of the route has been routed to avoid Royal Gorge

Table 16

Summary of Pipeline Design Data

A. For 6 fps:

Line diameters (in.)	24	36	48	60
Capacities (acre-feet/year)	13,600	30,700	54,600	85,300
Avg. Wall Thickness (in.)	0.55	0.85	1.10	1.40
Friction Losses (psi/mile)	8.6	5.2	3.7	2.8
Unit Pipe Mass (#/foot)	144	335	577	919
Unit Cost ($/foot)	130	302	519	827
Unit Cost ($/foot) [a]	114	162	222	302

B. Same capacities as "A," but for 20 fps:

Line diameters (in.)	14	20	26	32
Capacities (acre-feet/year)	13,600	30,700	54,600	85,300
Avg. Wall Thickness (in.)	0.35	0.45	0.60	0.75
Friction Losses (psi/mile)	181	114	83	63
Unit Pipe Mass (#/foot)	53.7	98.4	171	263
Unit Cost ($/foot)	48.30	88.60	154	237
Length of Run (miles)	4	4	4	5
Net Friction Losses (psi)	724	456	332	315
Temp. Rise of Fluid (^{o}F/mile)	0.6	0.4	0.3	0.2
Friction Factor, f	0.0040	0.0036	0.0034	0.0032

[a] Costs based on Longyear (13).

Table 17

Range of Overall Heat Transfer Coefficients
(for 1 and 2 inches of insulation)

Pipeline Diameter (inches)	Overall Heat Transfer Coefficients (Btu/ft^{2}hr^{o}F)	
	Surface Pipe (1" insul.)/(2" insul.)	Buried Pipe (1" insul.)/(2" insul.)
8 (minimum)	0.68 / 0.38	0.45 / 0.31
72 (maximum)	0.62 / 0.32	0.40 / 0.25
Average for 11 Diameters	0.63 / 0.33	0.41 / 0.26

Table 18

Insulation Costs Per Mile of Pipeline

Inside Diameter (inches)	Insulation Cu. Ft per Mile	Insulation $ per Mile
14	17,500	26,300
20	22,500	33,800
24	25,900	38,900
36	36,000	54,000
48	46,100	69,200
60	56,300	84,500

Bases: 7-inch layer of insulation
$1.50 per cubic foot of insulation
Application of insulation included as part of installed pipeline capital cost, Figure 18.

Table 19

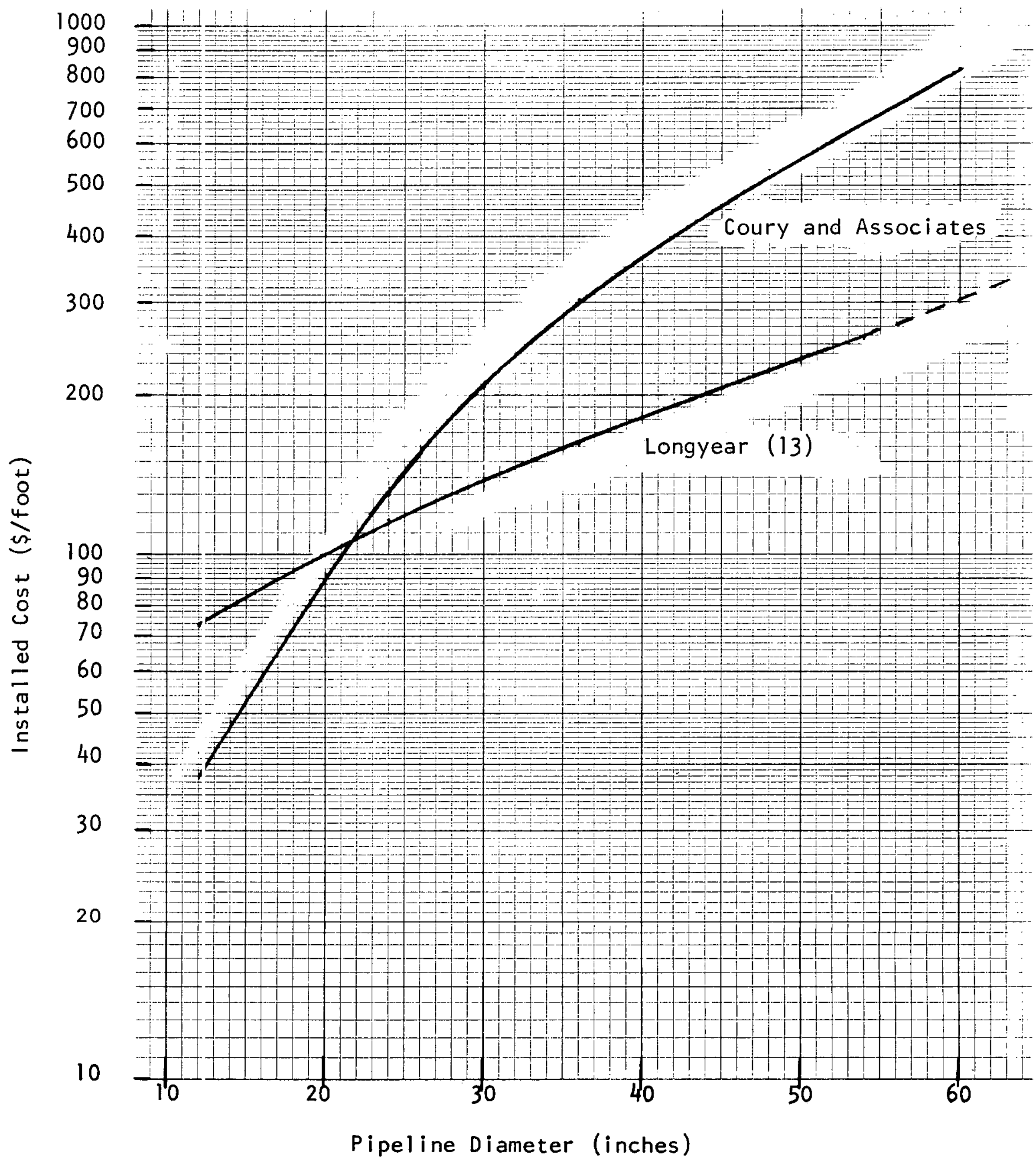
Pumping System Capital and Operating Costs

Bases: 2 stations in San Luis Valley pumping water to 750 ft. static head plus frictional losses, each; 2 stations in Front Range network pumping to 650 ft. static head plus frictional losses, each; line velocity is 6 fps; water is at 302°F, 0.93 S.G.; pump efficiency = 85%.

Valley Trunkline Diameter (inches)	Gross Capacity (gpm)	Frictional Losses [a] (psi/mile)	Horsepower per Station[a]	Net Electrical Costs, All Stations[b] ($/hour)	Capital Costs per Station[a] (10^6 $)
24	8,430	8.6, 20	2350;1600	180	0.75, 0.48
36	19,000	5.2, 15	4530;3130	340	1.46, 0.95
48	33,800	3.7, 12	7450;5080	560	2.35, 1.55
60	52,900	2.8, 10	11,100;7430	830	3.43, 2.25

[a] Values given refer to pumping within the Valley and on the Front Range, respectively; 60% of gross capacity is pumped in Front Range trunkline to Colorado Springs.

[b] Electricity valued at 3¢ per kwh.



<u>Bases</u>: Carbon steel pipe; welded construction; application of insulation included.

<u>Figure 18</u>

Installed Geothermal Pipeline Costs

Table 20-A

Geothermal Pipeline Costs

System Excluding Wells for Production, Injection

Case	A	B	C	D
Main Line Diameter (Inches)	24	36	48	60
Capacity (Acre-Feet/Year)	13,600	30,700	54,600	85,300
Heat Content (MM Btu/hr)*	589	1,330	2,360	3,690
Pipeline Costs (10^6\$)	-	-	-	-
Maximum	85.1	186.3	319.0	514.0
Minimum	80.8	117.6	155.1	207.3
Insulation Costs (10^6\$)	5.4	7.4	9.4	11.5
Tunnel Cost (10^6\$)	22.5	22.5	22.5	22.5
Pump Station Costs (10^6\$)	2.5	4.8	7.8	11.4
Total Capital Cost (10^6\$)				
Maximum	115.5	221.0	358.7	559.4
Minimum	111.2	152.3	194.8	252.7
Annual Capital Charges (10^6\$)**				
Maximum	17.3	33.2	53.8	83.9
Minimum	16.7	22.8	29.2	37.9
Annual Brine Treatment Costs[a]	0.9	1.9	3.5	5.4
Total Annual Costs (10^6\$)				
Maximum	18.2	35.1	57.3	89.3
Minimum	17.6	24.7	32.7	43.3
Net Hourly Costs[b] (\$/hr)				
Maximum	2,080	4,010	6,540	10,190
Minimum	2,010	2,820	3,730	4,940
Electrical Costs (\$/hr)[c]	180	340	560	830
Total Hourly Costs (\$/hr)				
Maximum	2,260	4,350	7,100	11,020
Minimum	2,190	3,160	4,290	5,770
Energy Costs (\$/MM Btu)				
Maximum	3.84	3.27	3.01	2.99
Minimum	3.72	2.38	1.82	1.56

* Bases - 150°F Temperature drop by consumption; 0.93 S.G.; 1.0 (Btu/#°F)

** Assessed at 15% of Total Capital Costs

a. Costs are $25 per million pounds of brine

b. 100% stream factor

c. Assessed at 3¢/kwh

NOTE: End-user distribution system and heat-recovery capital costs not included

Table 20-B

Geothermal Pipeline Costs

System Including Wells for Production, Injection

Case	A	B	C	D
Main Line Diameter (inches)	24	36	48	60
Total No. of Wells	35	79	141	220
Total Well Costs (10^6\$)	22.8	51.4	91.7	143.0
Total Capital Costs				
Maximum	138.3	272.4	450.4	702.4
Minimum	134.0	203.7	286.5	395.7
Annual Capital Charges **				
Maximum	20.7	40.9	67.6	105.4
Minimum	20.1	30.6	43.0	59.4
Energy Costs ($/MM Btu)				
Maximum	4.50	3.93	3.68	3.65
Minimum	4.38	3.05	2.49	2.23

Note: End-user distribution system and heat recovery capital costs not included.

* Cost components not explicitly shown may be inferred from Table 20-A

**Assessed at 15% of Total Capital Costs

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APPENDIX A

A Preliminary Evaluation of Geothermal Prospects In the San Luis Valley, Colorado

A report prepared by Dr. Richard W. Davis for Coury and Associates

July 1977

Minor revisions December 1977

INTRODUCTION

General

This report discusses the geothermal potential of the San Luis Valley, in south-central Colorado. The Valley is in the counties of Saguache, Rio Grande, Alamosa, and Conejos. Figure 19 shows the topographic features of the area.

This report is based upon a study of data either published or extracted from the files of various public agencies. The primary objective has been to synthesize existing geological and geophysical information on the study area and to evaluate the prospective geothermal resources.

Summary and Conclusions

Present information on geothermal resources of the San Luis Valley indicates the most promising target is a reservoir underlying the cross hatched area of Figure 3. It is estimated that in a zone between the approximate depths of 8,600 and 9,500 feet there are about 1.3 x 10^{12} cubic feet of water at a minimum temperature of 250°F. The water is estimated to contain about 2,000 mg/l dissolved material. Permeabilities in the zone are probably high and capable of supporting high levels of production and reinjection through well systems.

There is no direct information on strata deeper than 9,500 feet. Geophysical data have been interpreted to indicate that the igneous basement floor is at a depth of 30,000 feet. This estimate is probably somewhat too great. In any event, the potential exists for finding reservoir temperatures in excess of 250°F at depths below 9,500 feet.

The aquifer system which includes the primary reservoir resource probably outcrops and is recharged in the San Juan Mountains. In the central San Luis Valley the reservoir is tightly contained under several thousand feet of shale and clay. Some minor upward discharge occurs, probably primarily along fault zones.

GEOLOGIC SETTING

General

The San Luis Valley is in the northern part of a geologic feature called the Rio Grande Rift (Chapin, 1971). In simple terms, the rift is a north-south zone along which the earth's crust has been downdropped relative to the crustal material east and west of the zone. It is postulated as a tensional feature bounded by normal faults along the Valley walls. The east wall of the Valley is formed by the granitic and metamorphic rock mass of the Sangre de Cristo Mountains (Figure 19). The west wall of the Valley consists of the volcanic terrain of the San Juan Mountains. This somewhat simplistic pattern is complicated by a plethora of secondary and tertiary faults, folds, and volcanic features which are not yet entirely deciphered.

According to Chapin (1971), rifting along the Rio Grande Valley began at least 18 million years ago during the Miocene epoch. Prior to that time, a series of sedimentary Paleozoic rocks had been deposited unconformably over a Precambrian crystalline complex of igneous and metamorphic rocks. Beginning in the Oligocene, just prior to the development of the rift, volcanic eruptions spread lava and ash across the Valley. The eruptions continued during the development of the rift, ending during the late Miocene.

As the rift valley developed, sediments began to be eroded from the bounding higher, mountainous areas. These sediments were spread across the valley floor as coalescing alluvial fans interbedded with Miocene-age lava flows. Deposition of the sediments, called the Santa Fe formation (Powell, 1958), continued into the Pliocene epoch. The sedimentary portion of the Santa Fe formation consists of conglomerate, sand, gravel, and clay, probably nearly all of fluvial origin. The volcanic material is most abundant in basal portions of the formations, becoming scarce or absent in higher strata. Maximum thickness of the Santa Fe formation in the San Luis Valley is thought by Powell (1958, p.20) to be in excess of 5,000 feet.

During the Pleistocene the Alamosa formation was deposited over the Santa Fe formation. The Alamosa ranges from zero to more than 2,000 feet in thickness. It consists of alluvial fan sediments along the Valley walls, grading into fine-grained playa deposits near Valley center. It is thinnest and coarsest near the Valley walls, becoming finer grained and thickest near Valley center.

A thin veneer of Recent sediments of variable thickness overlies the Alamosa formation.

Subsurface and Geophysical Investigations

Measurements of heat flow have been made in the San Luis Valley region and are discussed in Reiter, et al. (1974). He points out that the highest heat flow values were measured along the west margin of the Valley. The values there range up to about 2.5 μ cal/cm^2 sec, which is about 1.5 times the global average of 1.5 μ cal/cm^2 sec. Ostensibly, the area of high heat flow would seem to be the best prospect for exploration for geothermal resources. However, the paucity of data points measured by Reiter, et al., may have caused them to overlook high geothermal gradients in the vicinity of a well, #1-32 Mapco-State (Figure 19), in the eastern portion of the Valley near Alamosa, Colorado.

Barrett and Pearl (1976) have accumulated information on the thermal springs and wells of the Valley. In contrast to Reiter's work, the concentration of these features (Figure 19) in the central and easternportions of the Valley suggests the location of the best geothermal prospects in that area. Using the geochemical methods described by Fournier, White, and Truesdell (1974), and by Fournier and Truesdell (1974), Pearl (personal communication, 1977) obtained the reservoir temperatures listed in Table 21. Depths of the reservoir were not specified. As a source of comparison, Renner et al. (1976, page 136) also used the geochemical data to calculate reservoir temperatures based on data from Mineral Hot Springs, located in the northern basin of the Valley. They derived a reservoir temperature of 220^{o}F at 5,000 feet depth. Estimated heat content of the reservoir was 4.76 x 10^{14} Btu's. Appendix B contains the full report.

Table 21.--Geothermal reservoir temperature derived by the Colorado Geological Survey using the geochemical methods described by Fournier, White, and Truesdell (1974) and by Fournier and Truesdell (1974). See Figure 19 for well and spring locations.

Location	Temperature (°F) and Analytic Method		
	Na, K, Ca	SiO_2 Minimum	SiO_2 Maximum
Shaws Warm Springs	212	248	387
Splashland Well	387	284	
Mineral Hot Springs	194	203	264
San Dunes Hot Springs	284	293	
Valley View Springs		Too Complex	

Keller (1974) described the results of surface geophysical studies at the northern portion of the San Luis Valley near Mineral Hot Springs (Figure 19). The method used was measurement of electrical resistivity. Since that time, additional detailed studies have been conducted by Arestad (1977) and Stoughton (1977) using, respectively, resistivity and seismic prospecting methods. Results of this work, also, have been largely confined to the extreme northern end of the San Luis Valley near Mineral Hot Springs, Valley View Springs, and Fullinwider Springs. The geophysical studies were interpreted to indicate a downfaulted basin containing between 6,000 and 8,000 feet of sediments overlying crystalline basement material. Resistivities in the basal sediments are low, possibly indicating a liquid thermal resource. As will be discussed later, there appears to be a discrepancy between part of this interpretation and the gravity data for that area.

Stoughton (1977) also made a cursory interpretation of seismic reflection profiles, from traverse data taken further south in the Valley by the Amoco Production Company. Two of the profiles include the sites of an oil exploration well, Colorado State F-1, and #1-32 Mapco State (Figure 19). Stoughton's interpretations agree with the well log data from Colorado State F-1, but may not agree with the data from #1-32 Mapco State. Stoughton appears to show the depth to Precambrian basement rock in the latter well as much deeper than logged. This ambiguity will be discussed more later.

Gaca and Karig (1976) conducted a gravity survey of the Valley (Figure 3). They analyzed the data along a number of traverses and drew up profiles of the Precambrian rock surface underlying the volcanic and sedimentary sequence filling the Valley (Figures 20 and 21). In general, the lower the Bouguer gravity values shown in Figure 3, the greater the depth to the Precambrian interface. The significance of the depth to the bedrock is that it represents the maximum depth to which reservoirs capable of yielding significant volumes of thermal fluid might be found. Since temperatures normally increase with depth, the areas of deeper Precambrian rock are presumed to be the best areas in which to prospect for geothermal resources. The gravity data, as interpreted by Gaca and Karig (Figures 20 and 21), indicate a deeper basin than Stoughton (1977) and Arestad (1977) predicted in the extreme northern end of the Valley via seismic and resistivity studies. However, Stoughton's interpretation of Amoco seismic data agrees with the gravity interpretation of Karig and Gaca, regarding the eastern portion of the Valley.

A few deep wells have been drilled to develop water supplies or to explore for petroleum. A recently drilled well, #1-32 Mapco State (Figure 19), penetrated through strata to 9,500 feet in what is probably the deepest portion of the San Luis Valley. Data collected during the drilling operations have revealed a great deal of useful information on the geology and geothermal potential.

Other oil and water exploration wells have been drilled in the Valley. Most of the data from these are given in Powell (1958). Locations of the deeper wells (>500 feet) are shown on Figure 19. The numbered well locations are keyed to Powell (p. 238-280).

INTERPRETATION

Hydrogeology

Only the deeper thermal, hydrogeologic systems are discussed here. These are underlying Alamosa formation sediments at depths in excess of 2,000 feet. Data from the Mapco State well indicate that the potential for highly permeable aquifers above a depth of 6,400 feet is very poor, and that prospects are best below 9,000 feet. A drill stem test (DST #1 on Figure 19) was run in the shallower zone. A transmissivity of only 4 darcy feet (73 gpd/ft) was obtained.

Logs from the Mapco State well included drilling logs and sample descriptions, and electrical resistivity, spontaneous potential, bulk density, gamma, caliper, and temperature geophysical logs. The logs show that, except for the very shallow aquifers, shale is especially dominant down to 4,000 feet. Below that depth, coarser clastics and volcanic material become increasingly common with increasing depth. At a depth of 8,620 feet the sample logs indicate granodiorite was encountered. Granodiorite is a coarse-grained, sialic rock close to granite in composition. Ordinarily its occurrence would be considered to mark the top of the basement sequence. However, the magnitude of the anomalous gravity deficiency (Figure 3) and the small breadth of the anomaly, combined with the density differential between sediments and granodiorite, indicate more sediments lie below the granodiorite (Figures 20 and 21). The geophysical logs also indicate that the less dense material expected below the granodiorite is also high-porosity material, probably either volcanic or sedimentary. The granodiorite may be a sill-like mass intruded into the deeper sedimentary/volcanic complex.

The sample logs indicate that vertical movement of fluids in the Mapco State well area is probably minimal. Temperature gradients from temperature logs confirm that above 9,000 feet little vertical convective heat flow occurs.

The aquifers of primary interest to this study are located between depths of 8,620 and 9,460 feet in the Mapco State well. This zone appears to have a major disruptive effect upon the temperature profile (Figure 22). Bulk densities (Figure 22) indicate a porosity of about 25%. It is also the zone in which drilling logs note lost circulation problems, indicating high permeability. Further, the electrical resistivity and spontaneous potential logs indicate an aquifer of such high permeability that differences in density between drilling fluids and formation fluids may have caused the invading drilling fluid to flow along the top of the aquifer. This latter conclusion requires that the formation water be highly saline and dense. Water quality data, discussed later, suggest that the formation water is only moderately more saline than the drilling fluid.

Patterns of circulation in deeper aquifers can only be speculated upon. Except for the Mapco State well there is little information on pressure heads. The data available from DST #1 (Figure 22) indicate pressures are less than hydrostatic at all depths below the aquifers in the upper 1,000 feet of sediment. The seismic profiles presented in Stoughton (1977, Figures 23 to 27) suggest that aquifer systems may outcrop to the west and that recharge may begin over the San Juan Mountains west of the San Luis Valley. Some of the recharge circulates eastward through deeper aquifers to the Mapco State well. Discharge probably occurs only by slow transfer upward, possibly along fault planes, into shallower aquifers. Gravity data indicate that the deeper aquifer systems are very restricted in area, so that significant discharge southward toward New Mexico beneath the Rio Grande River is not possible.

Geothermal Data

Figure 22 shows the temperature profile from the Mapco State well. The data for the profile were taken from continuous temperature logs run at various times during drilling operations. Such logs are often run before drilling fluids in the well bore have had a chance to reach thermal equilibrium with formation fluids. In the case at hand, data in the form of drilling fluid temperatures taken at various times after circulation had ceased were available. These data from two different depths are plotted on Figure 23 and used to estimate equilibrium temperatures. Figure 23 indicates that water temperatures in the aquifer of interest are at least 250°F at about the 8,000-foot depth.

Water temperatures from various other wells in the San Luis Valley are also plotted on the Figure 22 temperature profile. The data came from Barrett and Pearl (1976) and Powell (1958, Table 7). The temperatures are plotted at the total depth of each well. If it is assumed that the water came from an aquifer near the total depth of the well and that it had not cooled significantly before reaching the surface, a second temperature profile is obtained. The profile parallels the one from the Mapco State well but is about 55°F cooler. Temperatures in the Mapco State well are apparently higher than those encountered to date in any other well in the San Luis Valley.

Water Quality

Published data on quality of water from various thermal springs and wells are available in Barrett and Pearl (1976), Powell (1958, Table 7), and Renner et al. (1976, p. 136). Nomograms of some of this data are shown on Figures 24 and 25. Quality of this water, all from the low temperature sources, appears quite good.

Two types of information are available concerning quality of water in deeper, higher-temperature zones penetrated by the Mapco State well. The first of these is from analyses of two water samples obtained during DST #1 in the zone between depths of 5,304 and 5,491 feet. Figure 26 is a nomogram showing the results of the two analyses. Copies of the original data on file with the Colorado Oil and Gas Commission are contained in Appendix C. The two samples analyzed presumably came from the same test and were taken at the same time. The reason for the difference in dissolved material is not clear, although one analysis was conducted in the field and the other at the laboratory facilities of Yapuncich, Sanderson, and Brown.

The second method of determining water quality is through analysis of the spontaneous potential logs. Using this method yields a formation water conductivity of about 6,000 µmhos/cm in the zone sampled by DST #1. This gives a dissolved solids content very close to the value obtained by Yapuncich, Sanderson, and Brown Laboratories (Appendix C). It indicates a total dissolved solids content of about 4,000 mg/l.

Analysis of the spontaneous potential log data in the aquifer at 8,700 feet yields a conductivity of 3,000 µmhos/cm, equivalent to a total dissolved solids content of about 2,000 mg/l. The relatively good quality of this water suggests a rather efficient circulation system. In other words, the residence time of the water in the aquifer at a high temperature is too short for serious degradation of quality.

Aquifer Storage

The primary aquifer zone of interest appears to be about 840 feet thick, based on exploration to date. It might be thicker but drilling information is not available from deeper zones. Bulk densities yield a porosity of 25%. Using Bouguer gravity data to estimate the areal extent of the aquifer zone (Figure 3) yields a surface of 225 square miles. Total pore volume of the mass would be 1.3×10^{12} cubic feet. In other words, the aquifer zone is estimated to contain 1.3×10^{12} cubic feet of water, or approximately 30 million acre-feet.

Because present data indicate the northern portion of the San Luis Valley studied by Keller (1974), Stoughton (1977) and Arestad (1977) is a less valuable geothermal resource, little attention has been given to it in this report. Renner (1976, p. 136) estimates this part of the Valley contains 8×10^{10} cubic feet of water at an average temperature of 220°F. This estimate assumes an aquifer thickness of 5,000 feet at depths between 5,000 and 10,000 feet. Totalthickness of the potentially aquiferous sediments above the basement complex, according to Stoughton (1977) and Arestad (1977), is approximately 6,000 to 8,000 feet. As noted earlier, portions of Stoughton and Arestad's interpretations are hard to reconcile with data from gravity surveys. It seems likely that the reservoir in the northern areas is deeper than 6,000 to 8,000 feet, and may be part of the same unit as that penetrated in the Mapco State well. If so, the crosshatched area in Figure 3 could be extended northward some distance.

Production Potential

There is insufficient data available to be able to estimate with any precision the productive capabilities of the deep geothermal aquifer in Mapco State. It seems safe to speculate, though, that production rates in excess of 500 gallons per minute could easily be attained. Reinjection rates in the same range are also probably possible. Any system of withdrawal and reinjection wells for that aquifer would have to be located within the -270 mgal contour interval on Figure 3.

Land and Mineral Ownership

Published maps (BLM 1975, 1976a, 1976b, 1976c) showing land and mineral ownership are available for about 50% of the San Luis Valley. The maps cover the extreme southern portion of the Valley and the area east of Range 10E. The Mapco State well was drilled on lands owned by the State of Colorado. The Luis Maria Baca No. 4 Spanish land grant overlies much of the best geothermal prospect (Figure 19). With some minor exceptions the State and federal governments control the rest in the central San Luis Valley.

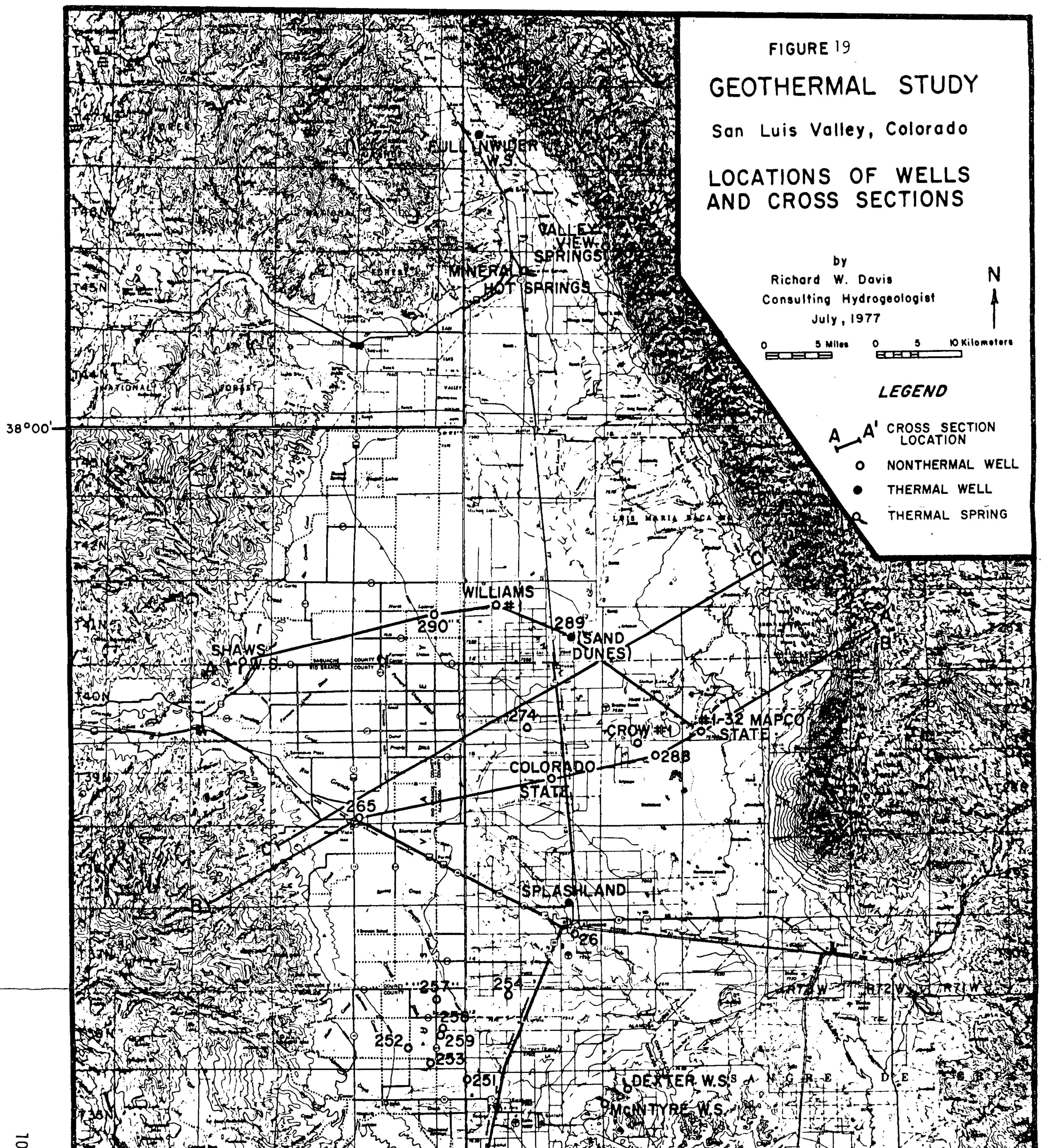


FIGURE 19
GEOTHERMAL STUDY
San Luis Valley, Colorado
LOCATIONS OF WELLS AND CROSS SECTIONS
by
Richard W. Davis
Consulting Hydrogeologist
July, 1977
N
0 5 Miles
0 5 10 Kilometers
LEGEND
A A' CROSS SECTION LOCATION
NONTHERMAL WELL
THERMAL WELL
THERMAL SPRING
FULLENWIDER W.S.
VALLEY VIEW SPRINGS
MINERAL HOT SPRINGS
WILLIAMS #1
290
289
(SAND DUNES)
SHAWS P.W.S.
274
CROW #1
#1-32 MAPCO STATE
288
COLORADO STATE
265
SPLASHLAND
26
257
254
258
259
252
253
251
DEXTER W.S.
McINTYRE W.S.
ROMERO #1
ROMERO #2
STATE
38°00'
37°00'
106°00'

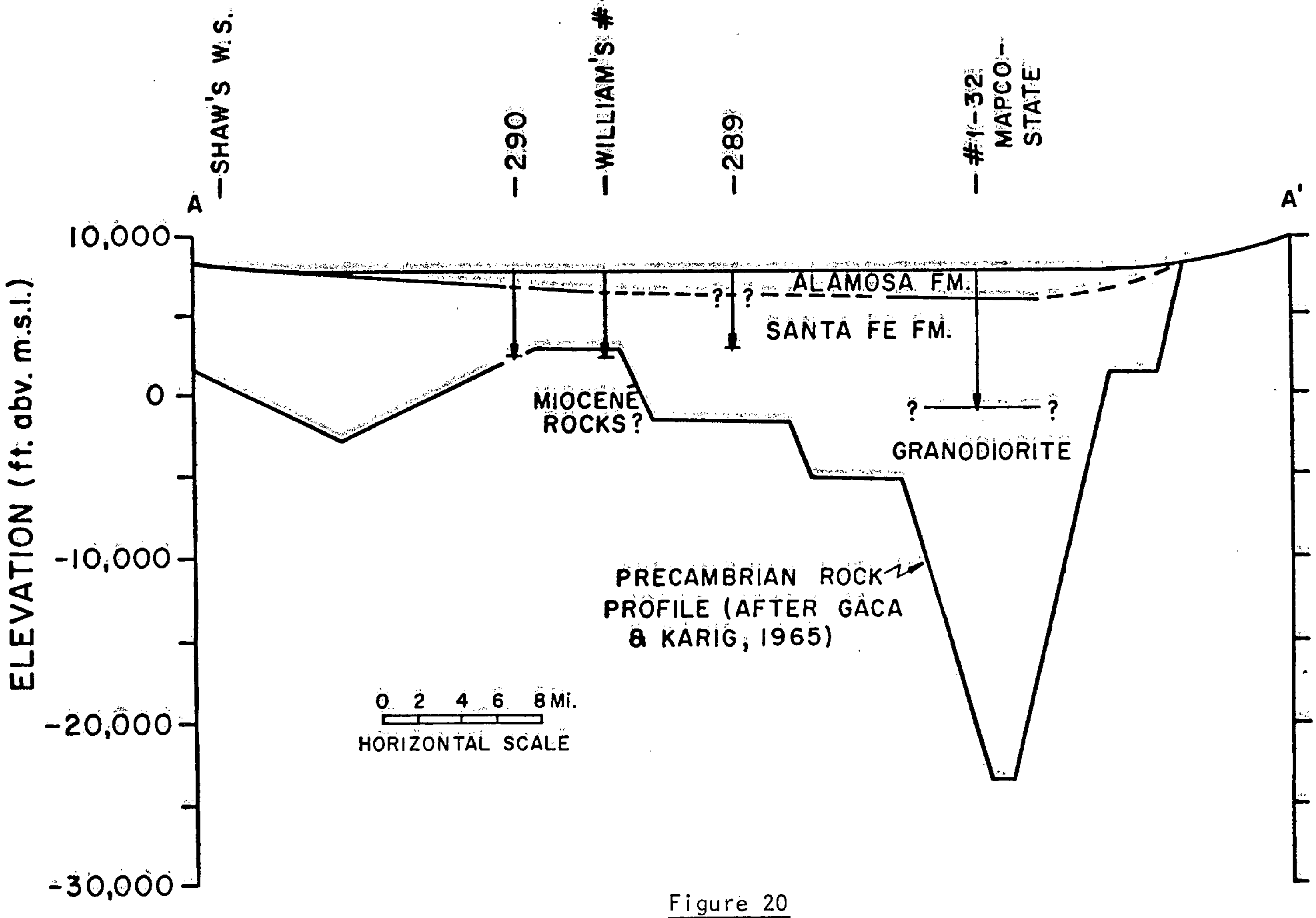


Figure 20

Geologic crossection along line A-A' (Figure 19) constructed using what is believed to be the most reliable data available.

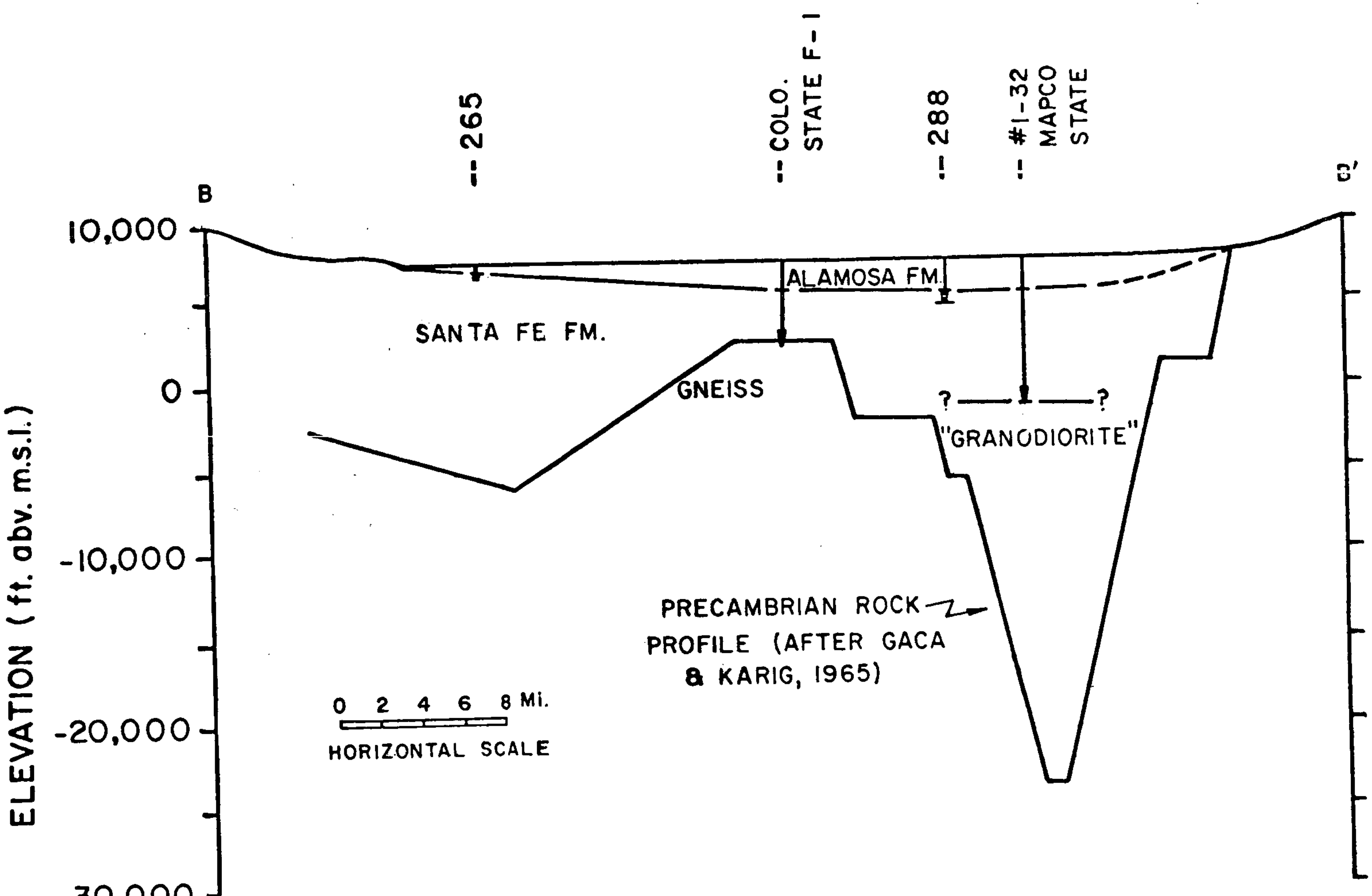


Figure 21

Geologic cross section along line B-B' (Figure 19) constructed using what is believed to be the most reliable data available.

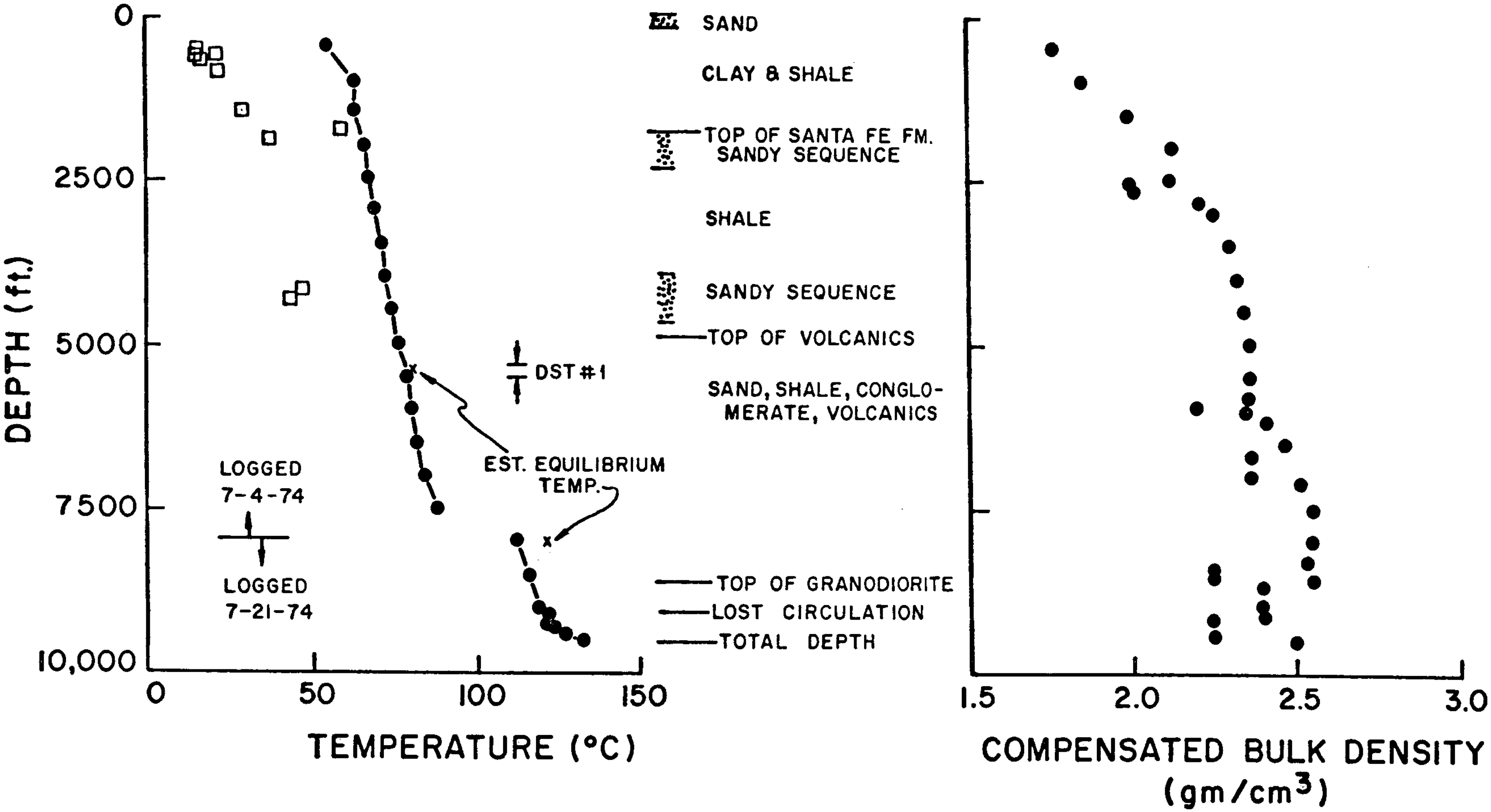


Figure 22

Graphs of temperature and density versus depth in the #1-32 Mapco State well. Data are from geophysical logs. Solid squares in the temperature graph are water temperatures from other deep wells in the San Luis Valley plotted versus total depth of well.

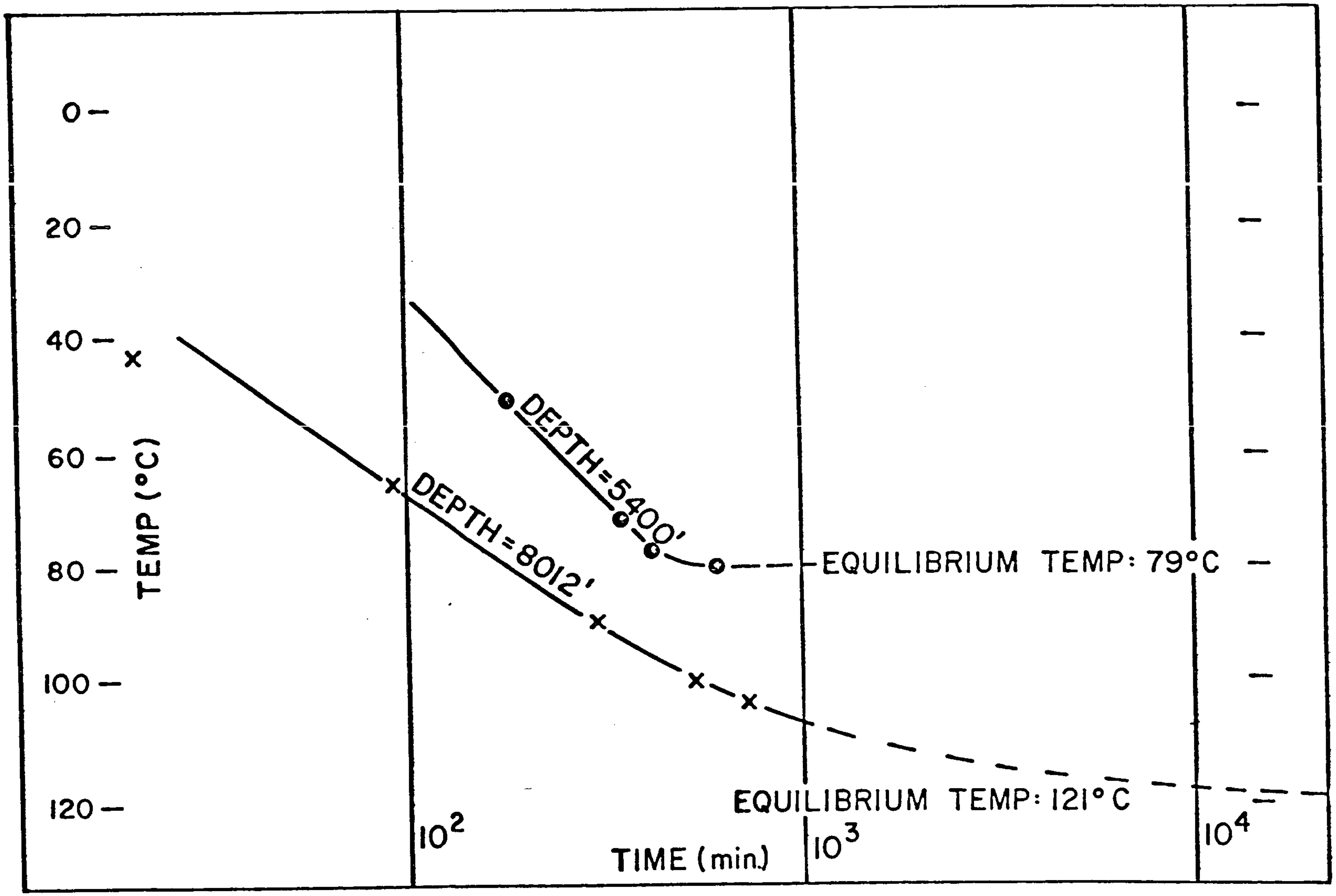
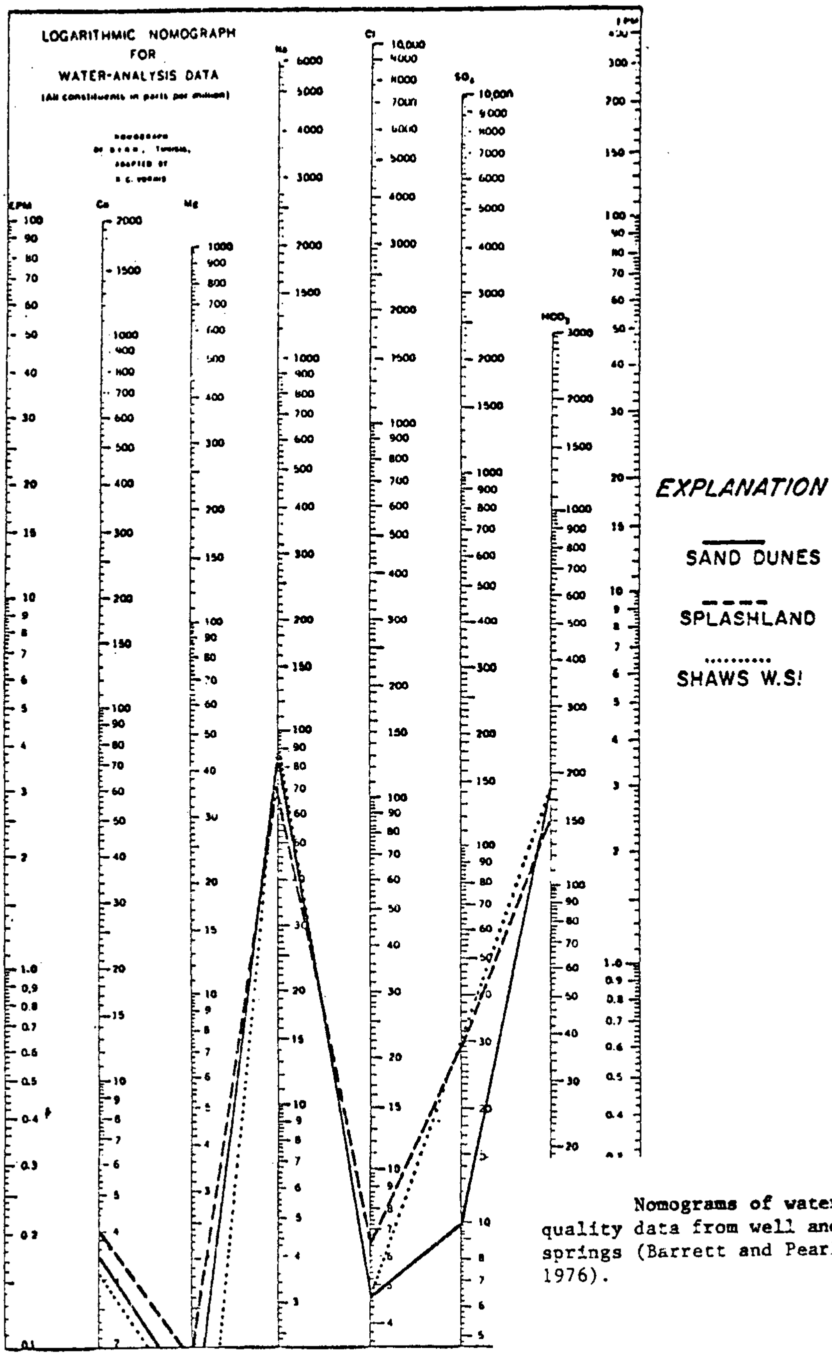


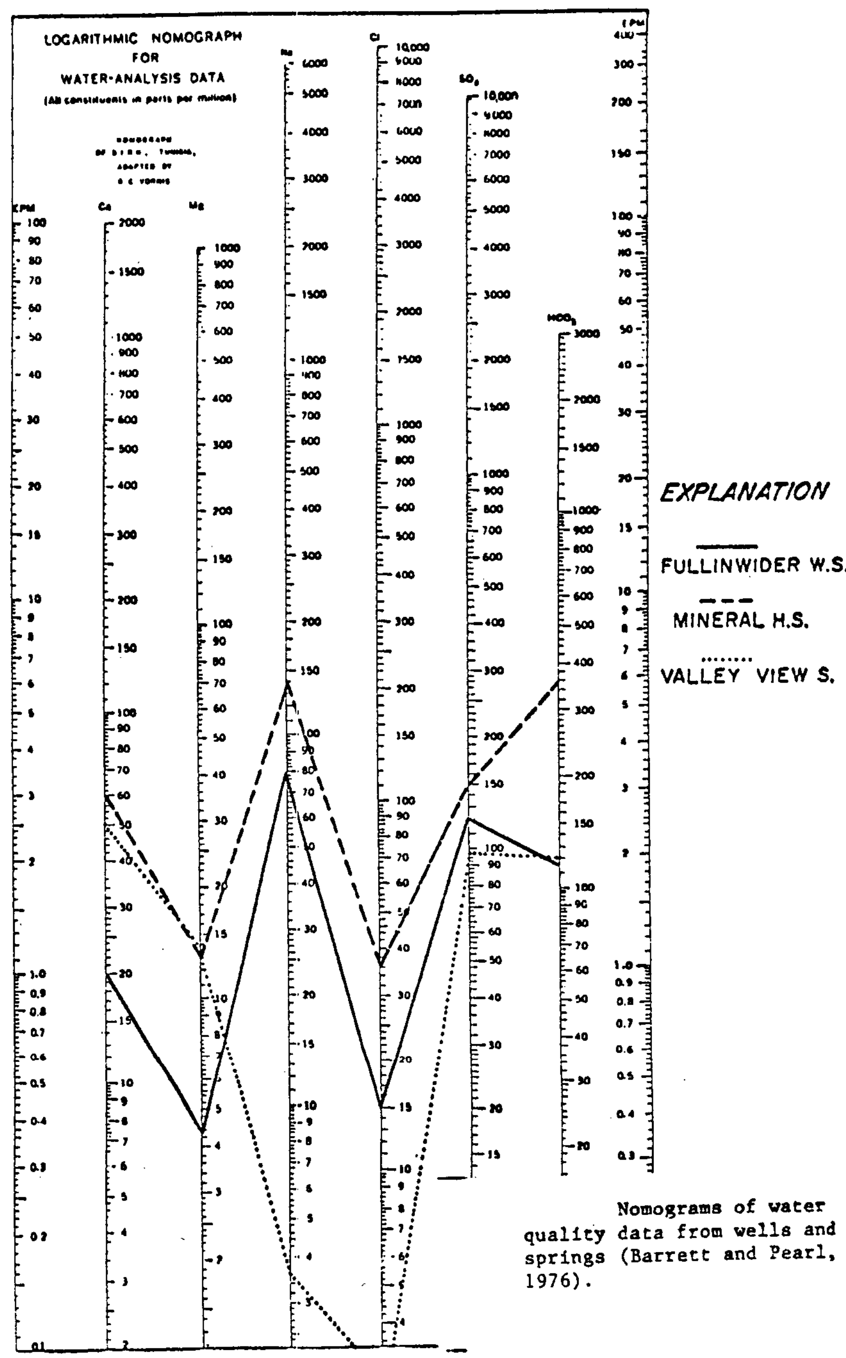
Figure 23

Change in temperature of drilling fluids versus time after circulation of the fluids ceased at 8012 and 5400 foot depths in the #1-32 Mapco State well.



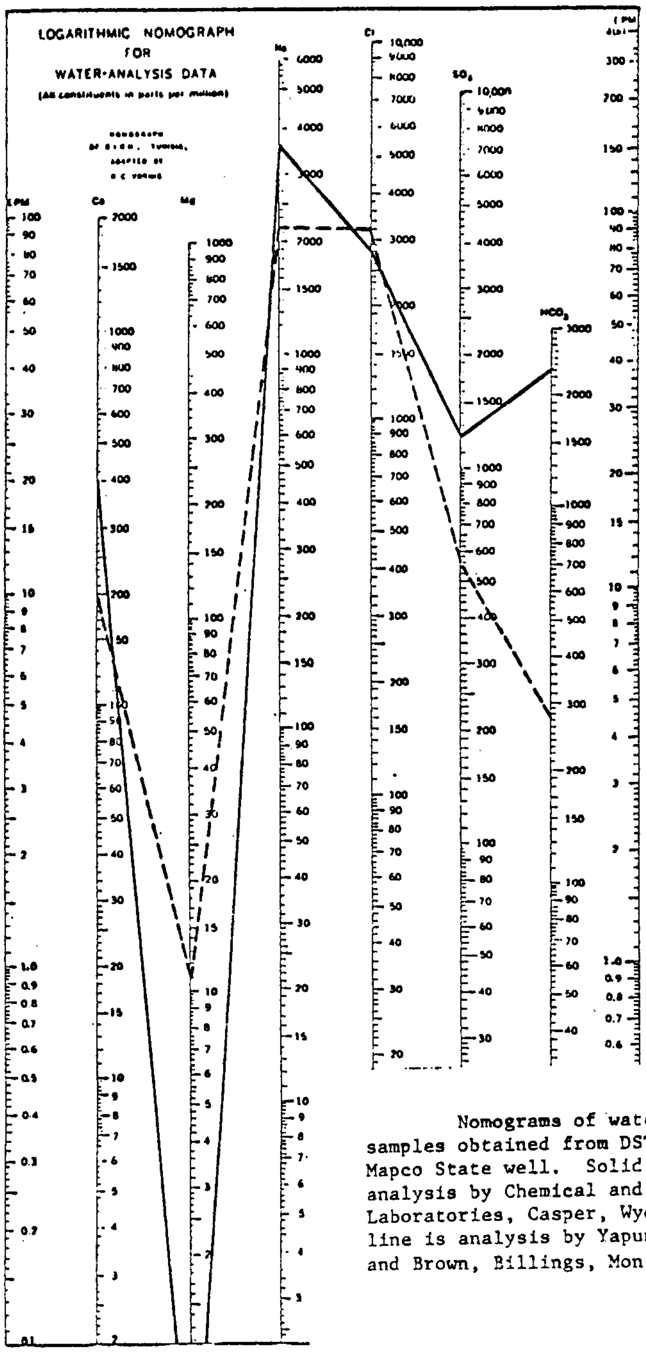
Nomograms of water quality data from well and springs (Barrett and Pearl, 1976).

Figure 24



Nomograms of water quality data from wells and springs (Barrett and Pearl, 1976).

Figure 25



Nomograms of water analyses of samples obtained from DST#1, #1-32 Mapco State well. Solid line is analysis by Chemical and Geological Laboratories, Casper, Wyoming. Dashed line is analysis by Yapuncich, Sanderson, and Brown, Billings, Montana.

Figure 26

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APPENDIX B

GEOCHEMICAL ANALYSIS OF MINERAL HOT SPRINGS

(from Renner, et. al., 1976, p. 136)

INF RECORD # 107 MIRRORED ON 3/76
NAME: MINERAL (CHAMBERLAIN) HOT WELL ,COLO RESOURCE CATAGORY: HOT WATER 90 TO 150 C
WARING FIG: 2 NUMBER: 023 DATE: 04/75
LOCATION:
STATE: COLO COUNTY:SAGUACHE
LATITUDE: 38 10.10 TOWNSHIP: 45N
LONGITUDE: 105 55.00 RANGE: 10E
ELEV: 7747 SECTION: 7 ,NE1/4 1/4 B&M: NMPM
SURFACE MANIFESTATIONS: SINTER,TRAVERTINE,HOT SPRING(S),

ROCK AND STRUCTURE TYPE: GRABEN, FAULTS ACTIVE MIOCENE TO PRESENT; VALLEY FILL; UPPER TERTIARY VOLCANICS N
EARBY
SURFACE DISCHARGE TOTAL: 189.0 L/MIN ESTIMATED: X
CALCULATED TOTAL DISCHARGE: L/MIN OF DEEP WATER
TOTAL SURFACE HEAT FLOW: 0.00E+00 CAL/SEC
AREA OF SURFACE EX: 0.0 KM**2
APPROX. # OF HOT SPRINGS: 30
TEMPERATURE: RANGE OF SPRING TEMP. 46 C TO 63 C OR
MAX. WELL TEMP 60 C AT 354 M DEPTH BOTTOM HOLE TEMP. C AT 354 M DEPTH
CHEMICAL DATA ANALYSIS DATE 09/74 SOURCE: WRD 1974 UNPUBLISHED (FLOWING WELL)

SPRING TEMP	FLOW L/MIN	PH	SIO2	NA	K	CA	SO4	CL	HCO3
63	132.5	7.00	51.00	140.00	14.00	56.00	160.0	39.0	348

OTHER CHEMICAL DATA SEE MALLORY & BARNETT FOR MINOR ELEMENTS, 1973; GEORGE, 1920

SIO2 ADIABATIC	SIO2 CONDUCTIVE	SIO2 CHALCEDONY	NA_K_CA 1/3	NA_K_CA 4/3	OTHER
103.3	103.3	71.2	168.2	91.3	

RESERVOIR PROPERTIES
RANGE IN RES TEMP 70 C TO 170 C ASSUMED
BEST EST. AVER. TEMP 105.0
AREA 0.0 TO 0.0 KM**2;BEST ESTIMATE 1.5 KM**2
BASED ON
DEPTH TO TOP OF RES. 0.00 KM TO 0.00 KM; BEST ESTIMATE 1.50 KM.
DEPTH TO BOTTOM OF RES. 0.00 KM TO 3.00 KM; BEST ESTIMATE 3.00 KM.
THICKNESS 0.00 TO 0.00 KM; BEST ESTIMATE 1.50 KM.
VOLUME 0.00 TO 0.00 KM**3; BEST ESTIMATE 2.25 KM**3
HEAT CONTENT > 15 C 0.00 TO 0.00 E18 CAL; BEST ESTIMATE 0.12 E18 CAL
POROSITY TO BEST ESTIMATE
PERMEABILITY TO MDARCY;
AVERAGE WELL FLOW TO KG/HR; WELL DIAMETER CM

GEOPHYSICAL SURVEYS: GRAVITY, DC RESISTIVITY
DEVELOPMENTS: 1 WELL - IRRIG, SPA.
REFERENCES: LIPMAN, STEVEN, AND MEHNERT, 1970; JORDAN, 1974; KNEPPER, 1974; SIEBENTHAL, 1910; POWELL, 1958; GEOR
GE, 1920; MALLORY & BARNETT, 1973; GACA AND KARIG, 1965; KLEIN, 1971; SCOTT, 1970, WARING, 1965
TOPO MAPS: VILLA GROVE 1:24,000

SPRING IDENTIFIED:YES
COMMENTS:
JORDAN, COLO. SCH. MINES MS THESIS, 1974, HAS GEOPHYSICS FOR THE MINERAL HOT SPGS, VALLEY VIEW HOT SPGS AREA.

PREPARED BY:GALYARDT, RENNER

NAME: MINERAL (CHAMBERLAIN) HOT WELL , COLO

APPENDIX C

CHEMICAL ANALYSES OF WATER FROM

#1-32 MAPCO STATE WELL

DST#1

Table 22

Design Basis Water Analysis

Total Dissolved Solids		5800
Ions:	Calcium	200
	Carbonate	---
	Bicarbonate	270
	Others	5330
	pH	7.9

CL1-12A (REV. 1964)

JUL 24 1974
RECEIVED
SEP 2? 1974
COLO. O. & GAS CONS. COM

CHEMICAL & GEOLOGICAL LABORATORIES

P. O. Box 2794
Casper, Wyoming

WATER ANALYSIS REPORT

OPERATOR	Mapco, Inc.	DATE	July 18, 1974 LAB NO. 13169-3
WELL NO.	1-32 State	LOCATION	SE NW 32-40N-12E
FIELD	Wildcat	FORMATION	
COUNTY	Alamosa	INTERVAL	5304-5491
STATE	Colorado	SAMPLE FROM	DST No. 1 (MFE)

REMARKS & CONCLUSIONS: Mud, high water loss with quebracho colored filtrate.

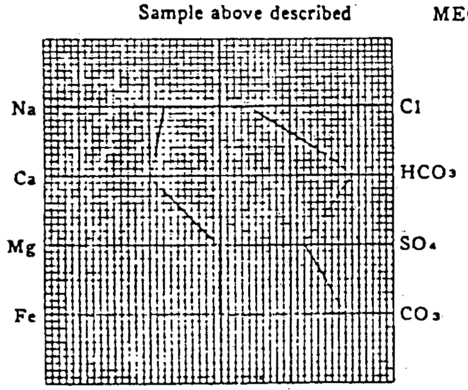
Cations	mg/l	meq/l	Anions	mg/l	meq/l
Sodium	3614	157.21	Sulfate	1200	24.96
Potassium	23	0.59	Chloride	2800	78.96
Lithium	-	-	Carbonate	1068	35.56
Calcium	398	19.89	Bicarbonate	2330	38.21
Magnesium	0	-	Hydroxide	-	-
Iron	-	-	Hydrogen sulfide	-	-
Total Cations		219.29	Total Anions		177.69

Total dissolved solids, mg/l	10250	Specific resistance @ 68°F.:		
NaCl equivalent, mg/l	9367	Observed	0.80	ohm-meters
Observed pH	9.9	Calculated	0.70	ohm-meters

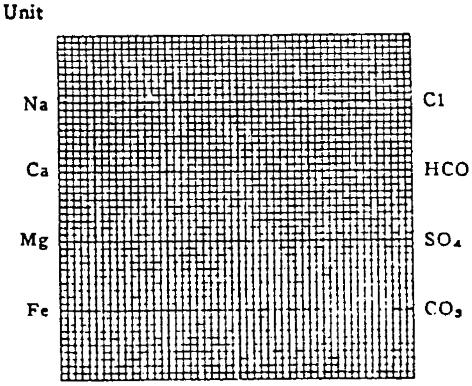
WATER ANALYSIS PATTERN

Sample above described

Scale
MEQ per Unit



20
2
2
2



(Na value in above ... Na, K, and Li)
NOTE: Mg/l = Milligrams per ... milligram equivalents per liter
Sodium chloride equivalent. Hawthorne calculation from components

JUL 15 1974

YAPUNCICH, SANDERSON & BROWN LABORATORIES

BILLINGS, MONTANA

P. O. BOX 593
59103

13 N. 32ND ST.

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SEP 20 1974
COLO. OIL & GAS CONS. COMM.

WATER ANALYSIS REPORT

Lab. No. 11633

Field Wildcat County Alamosa State Colorado
Well No. 1-32 State Location [illegible] 32-40N-10E
Formation Depths 5204-5401'
Operator Mapco, Inc. Date Sampled
DST No. 1 Sample Bottom of Recovery Date Analyzed 7-12-74
Other Data Fluid to Surface in 21 minutes, flowed for an additional 39 minutes. Open 60 minutes. Dark quebracho colored water.

Constituents	PPM	MEQ.	MEQ. %
✓ Sodium	2133	92.79	44.81
✓ Calcium	198	9.88	4.76
✓ Magnesium	11	0.90	0.43
Sulfate	564	11.73	5.66
Chloride	3100	87.42	42.21
Carbonate	0	0.00	0.00
Bicarbonate	270	4.42	2.13

Total Solids in Parts per Million
By evaporation
After ignition
Calculated 6139

pH 7.9

Specific Gravity @ 60°F 1.006

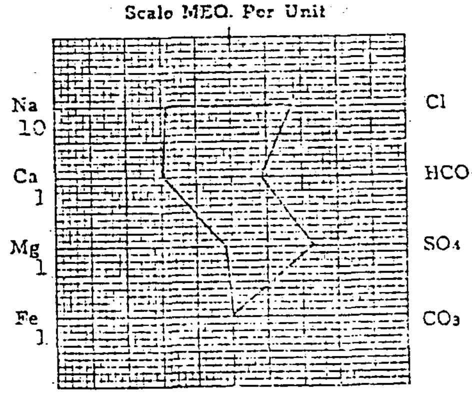
Resistivity @ 68°F ohms/meter² 1.12

Chloride as NaCl 5112 PPM. Total Solids From Resistivity as NaCl 5319 PPM.

NOTE: Sodium and potassium reported as sodium. MEQ.=milliequivalents per liter. PPM=parts per million (milligrams per liter). 1 PPM equivalent to 0.0001%

WATER ANALYSIS PATTERN

Scale MEQ. Per Unit



APPENDIX D

An Algorithm to Compare Multiple Effect Evaporator Configurations

An algorithm was developed to calculate required heat exchange surface area and geothermal brine feed rates for any n-effect evaporator train driven by flashed steam from the brine. The algorithm enables rapid, parametric evaluations for the ramifications of varying "n," the number of effects, and for the system responses to varying key operating conditions. Specifically, the user may set the values of the inlet geothermal brine temperatures, the range of evaporator temperatures, and the dual flash temperatures for the brine.

The essence of multiple effect evaporation (of water) is the recovery of the latent heat from water previously evaporated to accomplish a further portion of the net desired evaporation, rather than using externally generated heat to accomplish 100 percent of the intended evaporation. This discussion does not attempt to further explain the schematic functioning of multiple effect evaporation. Rather, this Appendix presents the algorithm and explains the necessary bases and assumptions made that simplify the evaluation of the relative merits of multiple effect evaporator systems having varying numbers of effects. The purpose of such an evaluation is to enable an optimal selection of such factors as the number of effects and operating ranges for temperatures, because the evaporator section is most important to the energy efficiency of the overall process.

As is described in Chapter V, the evaporator section in a modern sugar refinery is the point of input of virtually 100 percent of all process steam heat. This has come about because the evaporators comprise the highest-temperature section of the process, and as a result the energy-conserving practice of cascading heat begins in the evaporators. Present-day practice in this country leads to the use of as many as five effects in beet sugar plant evaporator sections. This is possible because fuel-fired boilers can produce high-temperature steam relatively inexpensively as compared to the cost of incremental amounts of surface area that are necessary to compensate for smaller temperature differentials caused by reduced steam temperatures. The significance of this is that the technical design problems of the conversion of a sugar plant to hydrothermal energy at 302°F are the same as for reducing the outlet temperature of the boiler.

The algorithm developed allows the user to specify a geothermal brine temperature, T_G, and two lower temperatures at which to flash the brine, T_0 and T_1 where $T_0 > T_1$, to produce steam for the evaporator steam chests. The user also specifies the number of effects, n; these four variables may then be run through a range of values to produce an approximate, parametric design analysis of a multiple effect evaporator system.

There are four important aspects or reasons for the use of two brine flashes. First, a single flash will either result in the waste of much potential heating value of the brine if the flash temperature is high, or severely restrict other design parameters (especially temperature driving forces) if the flash temperature is set low to offset the former objection. The lower flash temperature determines the overall fraction of utilization of the available heat in the brine. The upper flash temperature gives flexibility to the establishment of temperature differentials. The upper flash temperature strongly impacts on the amount of heat transfer surface necessary for the whole plant. Second, however, the upper brine flash temperature also strongly affects the efficiency of geothermal heat usage; for a given heat duty required of the hottest steam evolved, the higher the flash temperature the more brine is necessary. This particular point is discussed quantitatively in Section V. Third, the ratios of the respective amounts of steam produced at temperatures T_0 and T_1 should ideally be adjusted to exactly match and satisfy the net heat duties to be serviced at those respective steam temperatures with neither a shortage nor excess of brine or steam at either temperature. And, fourth, it is also desired to be able to adjust T_0 and T_1 as a means of minimizing heat transfer surface area. As to the reasons for using only two brine flashes in this algorithm, it was decided that two brine flashes would adequately illustrate the trends sought, and that more brine flashes would complicate the algorithm too far.

Evaporator effects are numbered 1 through n, starting with the highest temperature body; the model is a standard, forward-feed system, and for simplification each evaporator is assumed to bear the same heat duty; sugar juice and vapor temperatures into and out of each body, i, are all identified as T_i for any particular effect. The first brine flash temperature, T_0, is greater than T_1, of course; the second brine flash temperature, T_1, corresponds to juice and vapor temperatures from evaporator body number 1. Latent heats of the vapors evolving from each body are correspondingly identified as λ_1 through λn; brine latent heats are λa and λb, corresponding to T_0 and T_1, respectively.

Definitions of brine flash temperature differences:

$$\alpha = T_G - T_0 \quad {}^{\circ}F,$$
$$\beta = T_0 - T_1 \quad {}^{\circ}F$$

Definitions of heat transfer driving forces (°F):

$$\text{Effect \#1: } (T_0 - T_1),$$
$$\text{Effect \#2: } (T_1 - BPE_1) - T_2,$$
$$\text{Effect \#3: } (T_2 - BPE_2) - T_3,$$

$$\text{Effect n: } (T_{n-1} - BPE_{n-1}) - T_n,$$

Where BPE_i is the boiling point elevation of the juice in Effect i. Based on industrial data, estimates were made of representative values for the dissolved solids concentration of the juices (°Brix) in various evaporator bodies, for evaporator systems with one through five effects. See Table 23 following. Boiling point elevation data is available for sucrose solutions, so the table of Brix values may be used directly to tabulate BPE_1 through BPE_n for any group of n-effects. Cumulative values of BPE_i for n-effects are presented below in Table 23 as T_{BPE} such that:

$$\sum^{n-1} BPE_i \equiv T_{BPE} \quad {}^{\circ}F$$

Table 23

Sugar Solution Brix Values* for n-Effect Evaporator Systems

No. of Effects	Effect No. 1	2	3	4	5	Cumulative T_{BPE} (°F)
1	65	--	--	--	--	8
2	20	65	--	--	--	10
3	17	26	65	--	--	12
4	15	20	31	65	--	15
5	14	18	23	34	65	18

* Based on equal mass increments of evaporation in each body. This conflicts somewhat with the earlier assumption that each evaporator sees an equal heat duty, but it will not significantly alter the analysis. Thin juice to the first body is at 12 Brix in each case.

Approximate: the brine specific heat as C_p = 1.0 Btu/# °F

Definitions:

Heat available from brine: $C_p \cdot (\alpha+\beta)$ Btu/#

Available steam: $R_G = \alpha C_p/\lambda a + \beta C_p/\lambda b$

$$= C_p \frac{\alpha\lambda b + \beta\lambda a}{\lambda a\ \lambda b}$$

Note that: $\lambda a/\lambda b \cong 1.0$ leads to three possibilities

Average geothermal steam latent heat:

$$\lambda_G = \frac{\alpha\lambda a + \beta\lambda b}{\alpha + \beta} \cong \frac{\lambda a + \lambda b}{2} \quad \frac{\text{Btu}}{\#}$$

Another approximation:

$$R_G \cong \frac{(\alpha + \beta)\ C_p}{\lambda G}$$

And by substitution:

$$R_G = \frac{2(\alpha + \beta)\ C_p}{\lambda_a + \lambda_b}$$

Definitions for simplifying expressions further:

Average juice latent heat: $\lambda_s = \frac{1}{n} \sum^{n} \lambda_i \cong \frac{\lambda 1 + \lambda n}{2} \quad \frac{Btu}{\#}$

Average temperature driving force: $\overline{\Delta T} = (1/n) \sum^{n} \Delta T_i \quad {}^oF$

$= (1/n)(T_0 - T_n - \sum^{n-1} BPE_i) \quad {}^oF$

Then substitute: $\overline{\Delta T} = 1/n\ (T_o - T_n - T_{BPE}) \quad {}^oF$

Let: heat transfer coefficients: $U_{all} = \sum^{n} U_i \quad Btu/ft^2hr^oF$

Average heat transfer coefficient: $\overline{U} = (1/n)\ U_{all}\ Btu/ft^2hr^oF$

In a manner similar to the procedure for approximating solution Brix for various sets of evaporator configurations, representative values of heat transfer coefficients were examined for individual effects of different n-effect systems. The following values of U_{all} were established.

Table 24

Cumulative Heat Transfer Coefficients for n-Effect Evaporators

No. of Effects	U_{all} (Btu/ft^2hr^oF)
1	200
2	600
3	1000
4	1400
5	1800

The preceding assumptions and approximations will now be combined to calculate:

1) Net water evaporated per pound of geothermal brine.
2) Heat transfer surface area per pound of brine used.
3) Total required brine to evaporate an amount of water as specified by the user for an example problem.
4) Total heat transfer surface area necessary, again based on user-specified example problem.

Therefore:

W^1 = Net juice evaporated per pound of geothermal brine

$$W^1 = \frac{\text{\# Geo Steam}}{\text{\# Geo Brine}}\left[\sum^{n} \lambda_i/\lambda_G\right] = {}^{R}G/\lambda_G\left(\sum^{n}\lambda_i\right)$$

$$W^1 = (R_G/\lambda_G \quad) \, n\lambda_s$$

$$1) \qquad W^1 = \frac{4(\alpha+\beta)}{(\lambda_a+\lambda_b)^2} \quad \frac{n(\lambda_1+\lambda n)}{2} \, C_p \qquad \frac{lb}{lb}$$

And: defining Q_{all} as the sum of evaporator duties per pound of brine:

ΣQ_i, i = 1 to n,

A^1 = Heat transfer surface area per pound of geothermal brine

$$2) \quad A^1 = Q_{all}/(\bar{U} \ \overline{\Delta T}) = \frac{W^1\lambda s}{\bar{U}\ \overline{\Delta T}} = \frac{n^2 \ \lambda_s \ W^1}{U_{all} \ (T_o - T_n - T_{BPE})} \ \frac{ft^2}{lb}$$

And: letting W = Total water evaporated from the juice,

B = Total required geothermal brine

B = (Btu per evaporator body)/(Btu per pound of brine)

$$3) \quad B = \frac{W\lambda s}{n} \quad \frac{1}{(\alpha+\beta)\ C_p} \quad lb$$

Then: the total required surface area is

$$4) \quad A = BA^1 = \frac{4W \ \lambda_s{}^3 \ n^2}{(U_{all}(\lambda_a+\lambda_b)^2(T_o-T_n-T_{BPE})} \qquad ft^2$$

The user is able to establish values for a complete set of parameters. The procedure used in this study was to select sets of values of T_1 and T_n, the first and last-effect juice temperatures; calculations were then carried out varying n and T_0. The value of T_1 always puts a lower limit on the viable values of T_0, depending on what value of temperature approach is deemed preferable in the first effect.

APPENDIX E

Cost Analysis of Sugar Beet Refining and Barley Malting Processes

A. Introduction

The sugar beet refining and barley malting processes are reviewed in Chapters 5, 6, and 7. These chapters were designed to attain several objectives. First, the normal processes for which fossil fuels are used as the primary heat source were described in detail to provide a technical basis upon which a geothermally-heated process design could be developed. Second, the criteria upon which a geothermal design would be based, including technical and economic advantages and disadvantages for the various design options, were analyzed. Third, a near optimum geothermally-heated process flow sheet was developed for both of the manufacturing systems. Finally, general costs for the geothermal heating system were developed in each case, and the equivalent energy costs, expressed as dollars per million Btu's, were calculated.

Only overall costs are developed in Chapters 5, 6, and 7, thus enabling the four objectives to be achieved in a relatively straightforward and comprehensive manner. However, it is of value to analyze the equipment specifications and costs in more detail, so that the design engineer will have a basis for evaluating various options as to actual flow sheets, operating conditions, and materials of construction. To help meet this need, a preliminary detailed design is presented in the following sections.

The designs that are presented follow the geothermal brine from the production well, to the plant battery limit, through the process, and then to the injection well. It is assumed that downhole pressures are sufficient to produce artesian flow of the hot brines without flashing in the well, as appears to be generally the case in the San Luis Valley. Where this assumption is not valid, standard downhole pumps would be required. After being pumped to the plant, the processing of the brine varies between the sugar and barley processes.

For the basic barley malting design, the hot brine is passed through the tubes of the air heater, as shown in Figure 12 of Chapter 7. Since the brine is maintained under pressure and stays in the liquid state, there is no release of CO_2 gas, and consequently no tendency for $CaCO_3$ scale formation to occur. Since the brine in the San Luis Valley area is of relatively high quality, as was indicated on Table 22, Appendix C, and as was discussed in Chapter 7, no special brine treatment is required for this barley malting design case.

In the beet sugar refining process, on the other hand, it is necessary to provide steam to the various evaporating stages. This steam is provided by flashing the geothermal brine in successive stages, as shown on Figure 8 of Chapter 5. Release of CO_2 during the flashing could lead to scale formation and plugging of valves, lines, and nozzles. Accordingly, the brine is pretreated to permit controlled removal of CO_2 and adjustment of the brine pH, before reaching the flash stages.

The following sections describe these processes in some detail. Included are flow sheets and tables indicating estimated equipment sizes and costs. Some parametric calculations have been made for some of the cases in order to provide an indication of minimum costs, but no attempt has been made at optimization of the system. Various input factors could change for any specific geothermal development site, such as distance from the wells to the plant and brine chemistry, and such changes would result in modified design conditions. Accordingly, the results presented here are preliminary and would be expected to change for other conditions.

B. Beet Sugar Refining Process

1. General Description

The basic sugar refining process is described in Chapter 5 and shown on Figures 5 and 6. The adaptation of this plant to a geothermal heat source, based on a moderate temperature brine at 300°F, is discussed in Chapter 6 and shown on Figure 8. The geothermal portions of this system are shown in more detail on the following Figures 27 and 28. The latter figures are discussed below. A summary of the geothermal equipment is given in Table 26, and a summary of costs is given in Tables 27 through 30.

Wellhead production is received in Surge Tank 1 from which it is pumped to the pretreatment site within the plant battery limits. The surge tank has a capacity of 26,000 gallons, to provide a residence time of about five minutes. The tank is of carbon steel construction coated with an epoxy resin and 20 feet long by 15 feet in diameter. Its design pressure is 80 psig at 300°F. The tank will be blanketed with nitrogen gas, when not operating under pressure, to prevent air entry and the potential corrosion problems related to oxygen in the brine.

The pretreatment is based on water analysis in Table 22, Appendix C. At the plant, the brine is acidified by injection of 50% sulfuric acid upstream of a static inline mixer. The acid is stored in a 28,000 gallon tank constructed of carbon steel coated with epoxy resin, and maintained at ambient temperature and pressure. This supply is ample for about 20 days of operation. It is transferred by a gear pump made of hastelloy, pneumatically operated at an average rate of about 1 gallon per minute. The mixer is situated in the most corrosive area of the system and is constructed of titanium.

The brine pH is reduced to 4.5 to 5.5 by acid injection to convert all the bicarbonate into dissolved CO_2 gas. The CO_2 is then stripped out of the brine by steam in a packed-column degassifier and is rejected to the atmosphere. Stripping steam is produced by partially flashing the hot brine, itself, in the top of the degassifier. The brine and steam then pass concurrently down through the packed section. Degassified brine is collected in the column sump and pumped to the process. The flashed steam containing CO_2 removed from the brine is vented to the atmosphere.

For the brine temperature of 300°F and the bicarbonate concentration of 270 parts per million (ppm) about 1.378×10^4 pounds per hour of steam are required at equilibrium to remove 95 percent of the CO_2; the residual CO_2 content of 13 ppm is not expected to be a problem in subsequent processing. As a safety factor, the amount of steam produced by flashing has been increased 2 times, to 2.756×10^4 pounds per hour. This represents 1 percent by weight of the brine feed stream, and requires a flash down of 8°F in the degassifier. Thus, decarbonated brine leaving the degassifier will be at a temperature of about 289°F.

The driving force for stripping CO_2 from the brine is very high, so that only about a foot of packing would be required. Conservatively, a packed height of 5 feet of ceramic saddle packing has been specified. This extra height is partially required because of gas-contacting efficiency problems in large diameter columns. The degassifier itself is about 5 feet in diameter and 20 feet tall. It is constructed of 316 stainless steel, and is designed to operate at 80 psig and 300^{o}F.

The pH of the brine leaving the degassifier is increased to a safe operating level of 8 to 8.5, so as to avoid corrosion of downstream equipment, by the injection of 50% by weight caustic soda. The caustic is stored in a 28,000 gallon carbon steel tank which provides a 20-day storage capacity, and is transferred by a pneumatically operated, stainless steel gear pump. Fluid mixing at the injection point is accomplished in a static, in-line mixer made of 316 stainless steel.

The treated brine then passes, successively, through flash tanks F1, F2, F3, and F4, as shown on Figure 28, where process steam is produced at the flow rates and temperatures indicated. These operating conditions on the flash vessels meet the process requirements as shown on Figure 8. Flashed steam passes through wire mesh demister pads installed in the tops of the vessels, to removed entrained brine from the steam. The vessel diameters and the area of demister elements in each vessel are sized for efficient removal of droplets.

The flash vessels are all designed to operate at full vacuum, and at positive pressures ranging from 0 to 20 psig and temperatures from 170 to 240^{o}F. Vessels are constructed of carbon steel and coated with an epoxy resin. Vessel dimensions and demister specifications are given in Table 26.

Cooled brine leaving the last flash stage is collected in Surge Tank 2 prior to transfer to the injection well. Tank 2 is designed to operate at atmospheric pressure and 150^{o}F, and has a capacity of 120,000 gallons, equivalent to 20 minutes of storage. It is constructed of steel and has overall dimensions of 51 feet long by 20 feet in diameter. Brine from the surge tank is pumped through a filter on the way to the injection well. All vessels have 2 inches of calcium silicate insulation.

Pipelines between the wells and the plant are designed for an operating pressure and temperature of 100 psig 300^{o}F. Pipeline sizes evaluated for the sugar beet process were 16 inch and 20 inch. A 20-inch schedule 20 carbon steel line is required for the brine transfer lines, allowing a brine flow velocity of 6 feet per second. A 16-inch schedule 30 carbon steel line was chosen for the steam transfer lines from the flash tanks to the evaporators. For the sake of presenting definitive numbers for this study, it is assumed that the wells are located 3,000 feet from the plant. Within the pretreatment plant 1,200 feet of 20-inch pipe is assumed and 400 feet of 16-inch schedule 30 is assumed for steam lines from the flash tanks. These distances will be different, of course, for most situations. Pipe costs are presented in Table 25.

2. Equipment Design and Costs

Standard engineering design and cost estimating procedures have been used for this study, as indicated by References 1 through 8 in the Bibliography. In addition, equipment suppliers and manufacturers were contacted for design information and cost quotations on specific items where necessary (9-15). These results are summarized in Tables 27 through 30 for the sugar refining process.

The total capital equipment cost for the geothermal heat system is 9.963 million dollars for a 44% steam on beet efficiency. The values for the well and the pipeline given in Tables 1 and 29, respectively, are installed costs. The cost of other equipment items, including the surge and flash vessels, pumps, and instrumentation, has been multiplied by 47% to cover the cost of installation at a developed site, based on the results presented in Reference 7. Itemized costs are shown in Tables 27 through 29. A complete cost summary is shown in Table 30.

C. Barley Malting Process

1. General Description

The basic barley malting process is described in Chapter 7 and shown in Figures 9 and 10. The adaptation of this plant to a geothermal heat source based on three temperatures--300°F, 260°F, and 220°F--is also discussed in Chapter 7 and shown in Figure 12. The following analysis was based on the mid-range temperature of 260°F (case $3b_1$). The design criteria for this case are discussed in Chapter 7 and shown in Table 7.

The geothermal portions of this system are shown in more detail in Figures 29 and 30. The latter figures are discussed below. The list of geothermal equipment is summarized in Tables 25 and 31 and cost summaries are given in Tables 32 through 35.

Wellhead production is received in Surge Tank 1 from where it is pumped directly to the fan heater in the case of no pretreatment. The surge tank has a capacity of 2,500 gallons providing a residence time of about five minutes. The vessel is constructed of carbon steel, coated with an epoxy resin, with the dimensions 9 feet in length by 7 feet in diameter. The brine is transferred via pump 1, a carbon steel pump capable of 550 gpm, through a 6-inch schedule 40 carbon steel pipe to the air heater. The cooled brine is then collected in the final surge tank 2 prior to transfer to the injection well. Vessel 2 is designed to operate at atmospheric pressure and 150°F, with a capacity of 26,000 gallons equivalent to 20 minutes of storage. It is constructed of carbon steel, coated with an epoxy resin, and has the overall dimension of 20 feet long by 15 feet in diameter. This brine is then pumped through a cartridge type inline filter unit on its way to the injection well.

A second option with pretreatment is shown in Figure 29 in the dashed-in area. The pretreatment is based on the water analysis in Table 22, Appendix C. This pretreatment is exactly analogous to the system described in the preceding sugar beet section, with only flows and vessel sizes changing. Major equipment sizes and specifidations are summarized in Table 31. The items similar to both options have the same process specifications.

The brine, pumped at 550 gallons per minute, is mixed with a 50% sulfuric acid solution in a titanium static mixer. The flow rate of sulfuric acid solution is less than 0.5 gallons per minute. The CO_2 gas and other noncondensables are stripped from the brine in a stainless steel degassifier (packed column) with dimensions of 2 feet diameter by 8 feet long with 3 feet of ceramic saddle packing. A conservative steam flash of 2.39 x 10^3 lb/hr (1 weight percent of the feed brine) is necessary to strip the carbon dioxide and other noncondensables from the brine. The brine is then neutralized by a 50-percent caustic solution. The caustic and acid tanks are both constructed of carbon steel and coated with an epoxy resin with a capacity of 6,000 gallons, equivalent to about 20 days of storage. The brine then passes through the air heater and into surge tank 2. The cooled brine passes through a cartridge style in-line filter on its way to the injection well.

Pipelines from the wells to the plant are designed for an operating pressure of 100 psig and 300°F. The pipeline size required for transferring geothermal brine to and from the barley malting plant is a 6-inch schedule 40 carbon steel line with costs per foot shown in Table 25. For the sake of presenting definite numbers for this study, it is assumed that the wells are located 3,000 feet from the plant. Within the plant itself for option 1, where no pretreatment of the brine is necessary, 500 feet of 6-inch pipe were assumed to be needed. For option 2 with a pretreatment plant an extra 700 feet were assumed.

The total capital equipment cost for the geothermal heat system for option 1, with no pretreatment, is 2.999 million dollars. For a pretreatment plant an extra 2.2 percent is added to the capital cost, bringing the capital cost to a new total of 3.063 million dollars. The values for the geothermal wells, given in Table 8, and the pipeline costs in Table 35, are installed costs. The cost of the other equipment items, including surge vessels, pumps, and instrumentation, has been multiplied by 47% to cover the cost of installation at a developed site, based on the results presented in Reference 7. Itemized costs are contained in Tables 32 through 34. Table 35 summarizes the costs.

D. Alternative Design Considerations

In order for the design analyses that have been presented to be more generally applicable to geothermal areas in all parts of the country, certain design modifications should be considered. An important factor in the cost of geothermal equipment is related to the corrosivity of the brine, and the corresponding requirements for materials of construction and for corrosion allowances on lines and vessels. Since the brines can vary greatly from one geothermal area to another, these cost factors will also vary. The specific changes that can be expected in different geothermal areas are temperature of the brine, the total amount of dissolved solids, the actual chemical species dissolved, and the brine treatment employed before its utilization in the process.

The San Luis Valley geothermal brines, upon which this analysis is based, are of higher quality than can be normally expected, with respect to the brine chemistry. In addition, process considerations related to these brines suggest the pretreatment procedure chosen which renders them non-scaling and relatively non-corrosive. In the more general case of geothermal applications, it can be expected that the brine corrosivity will approach that of seawater (17) and that larger corrosion allowances will be required.

The data for equipment sizing and costs for the more general geothermal case is summarized on Table 36. This table gives vessel sizes, corrosion allowances, and costs for both the sugar beet and barley malting processes. The specifications shown on Table 36 are to be compared with those given on Tables 26, 27, and 29 for the sugar process, and on Tables 31 and 32 for the malting process.

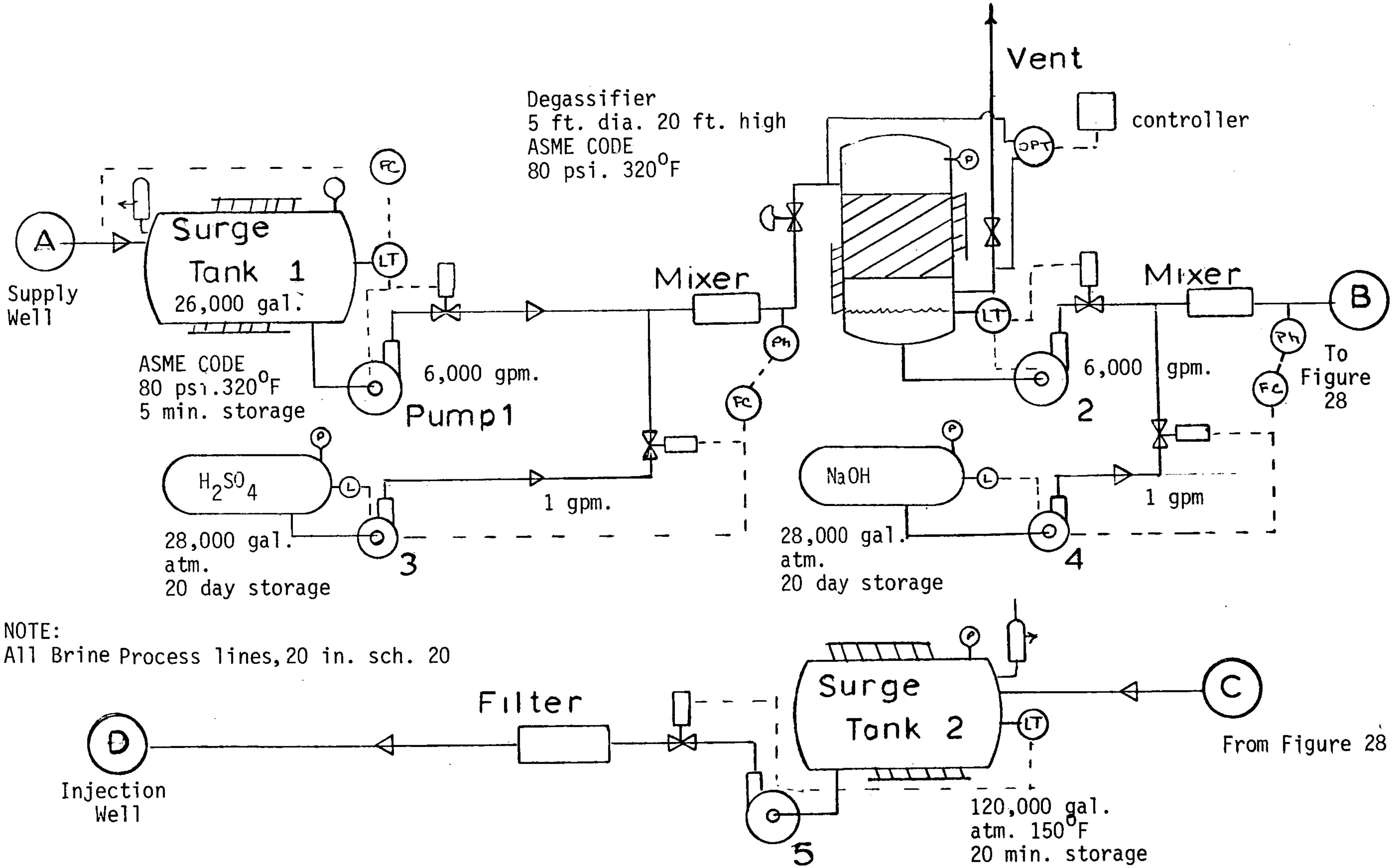


Figure 27. Pretreatment Plant, Sugar Beet Process

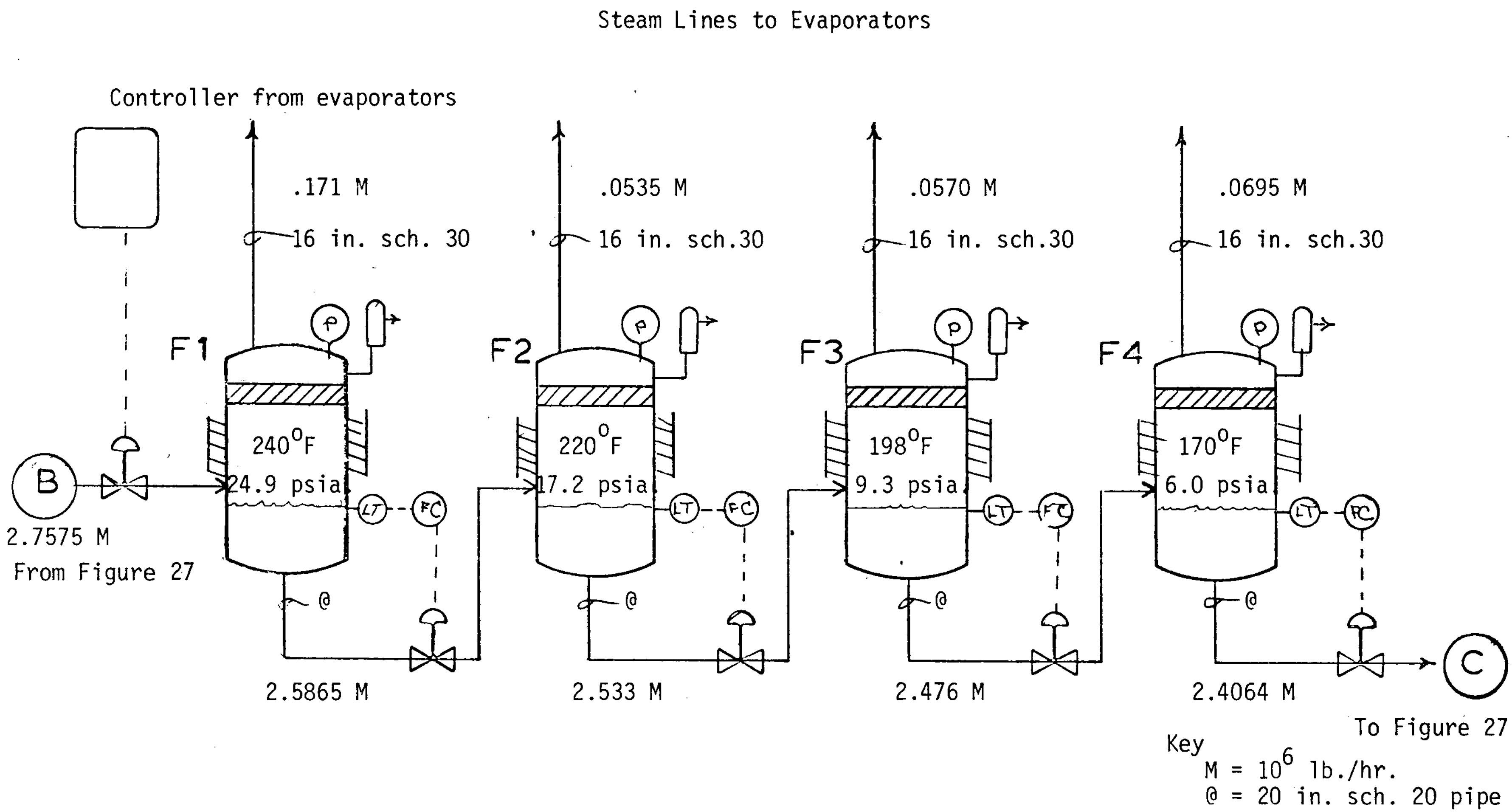


Figure 28. Flash Tank Flow Schematic, Sugar Beet Process

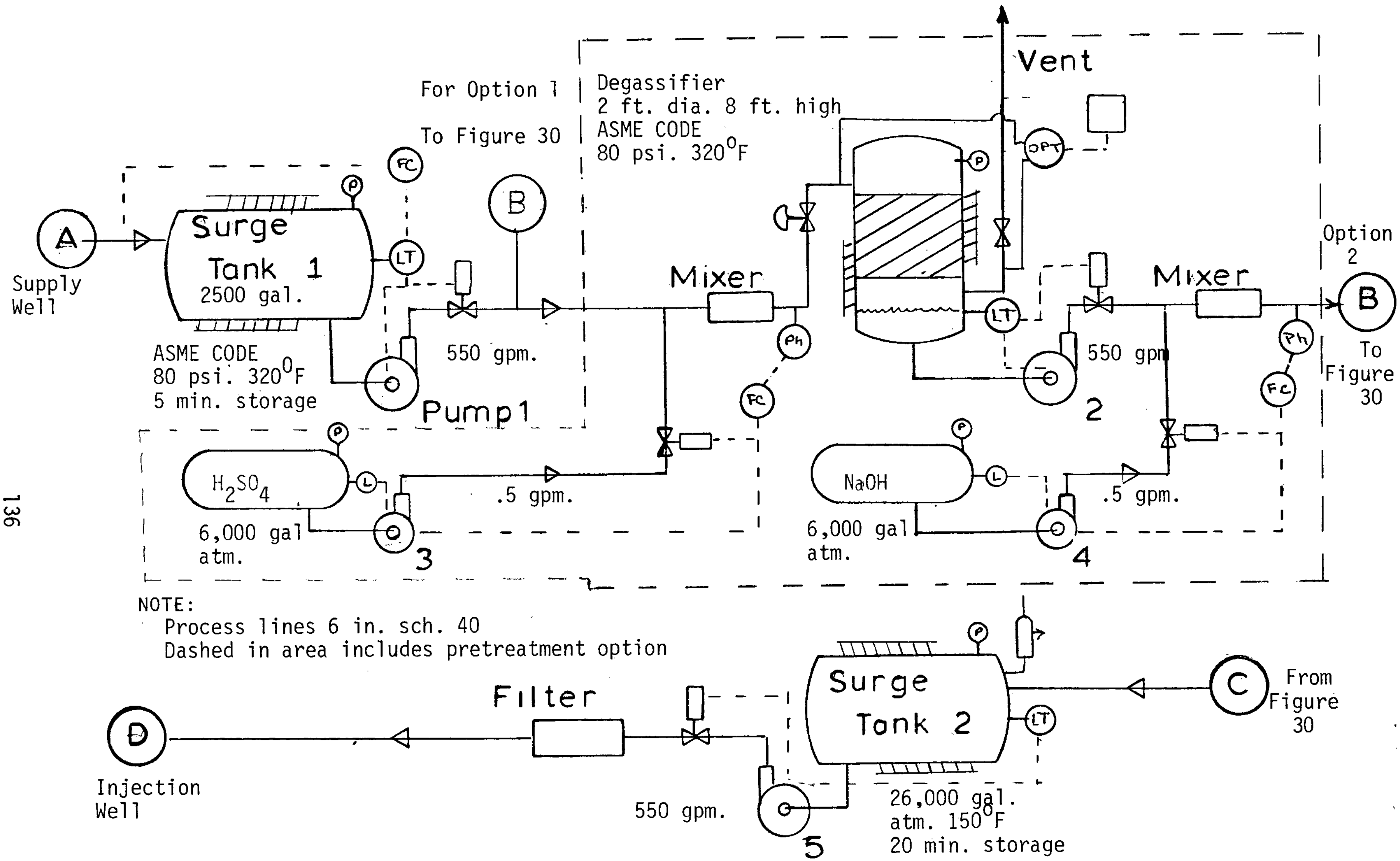
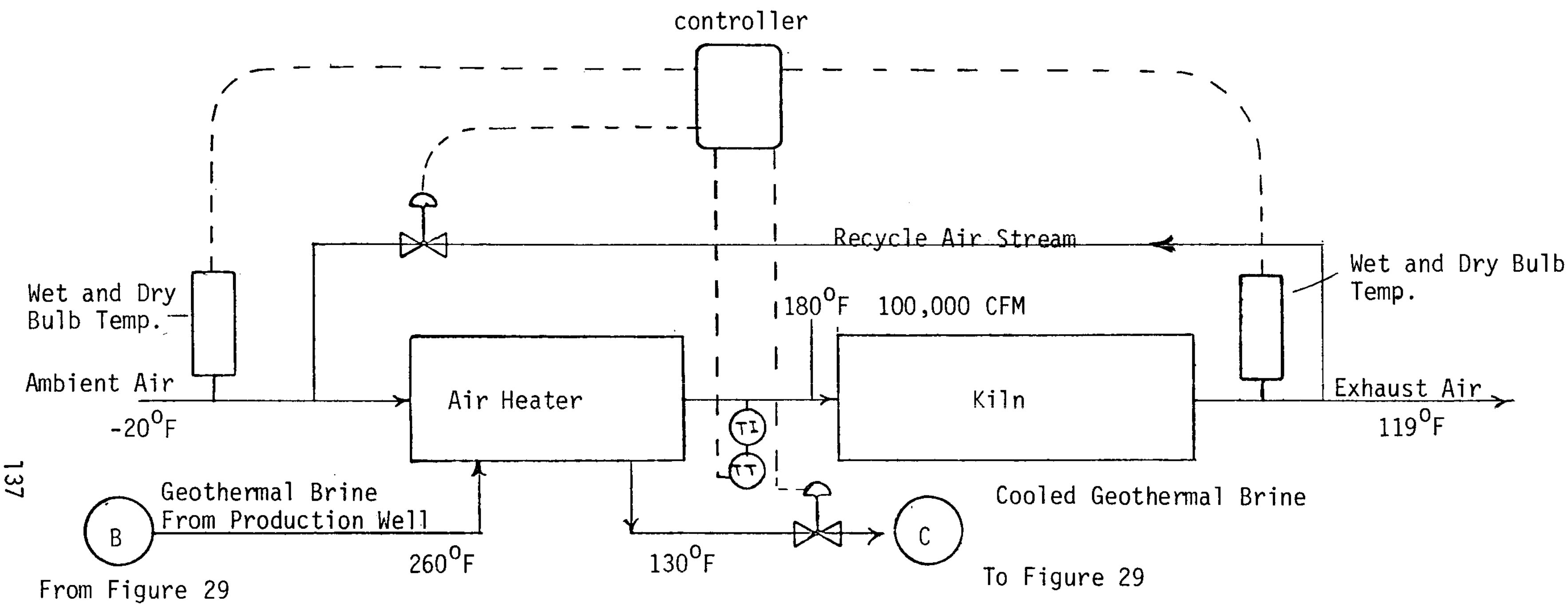


Figure 29. Pretreatment Plant, Barley Malting Process Case $3b_1$



NOTE: Process lines 6 in. sch. 40
Geothermal brine flow rate = 238,000 lb./hr.

Figure 30. General Schematic of Barley Malting Process

Table 25. Pipe Specifications

Nominal Pipe Diameter (inch)	6	16	20
Temperature (^{o}F)	300	300	300
Pressure (psig)	100	100	100
Wall Thickness (inch)	.280	.375	.375
Corrosion Allowance (inch)[1]	.254	.308	.288
Schedule #	40	30	20
Material Costs, Pipe and Insulation[2] per foot ($)	12	31	38
Installation Cost per foot ($)	20	46	56
Installed Pipe Costs per foot ($)[3]	32	77	94

1. The corrosion allowance differs slightly for each case, so that standard commercial pipe schedules can be used.

2. These costs, for carbon steel pipe, are based on vendor estimates (9).

3. Costs for above ground pipe installations (8).

Table 26. Vessel and Major Equipment Specifications

Equipment	Capacity	Length (ft)	Diameter (ft)	Temperature and Rating	Pressure
<u>SUGAR BEET PROCESS</u>					
Surge Vessels					
1	26,000 gal.	20	15	300°F	80 psig
2	120,000 gal.	51	20	150°F	Atm.
Treatment Storage Tanks					
1. Acid	28,000 gal	22	15	Ambient	Atm.
2. Caustic	28,000 gal	22	15	Ambient	Atm.
Degassifier		20	5	300°F	80 psig
Centrifugal Transfer Pumps 1, 2, 5	6,000 gpm	80 ft. head		300°F	80 psig
Acid and Caustic Gear Pumps	0-2 gpm			Ambient	80 psig
(2) Static Mixers, Stainless, Titanium		12	20 inches	300°F	80 psig
Filter Cartridge Style		(300 sq. ft.)		150°F	80 psig

NOTES:

1. Each surge vessel and storage tank is constructed of carbon steel and coated with an epoxy resin. The degassifier is constructed of 316 stainless steel.

2. The following wall thicknesses (t) include a corrosion allowance of .025 inch for the carbon steel tanks and .213 inch for the stainless steel degassifier. Surge Vessel 1, t = .645 inch; Surge Vessel 2, t = .174 inch; Storage tanks, t = .137 inch; Degassifier, t = .375 inch.

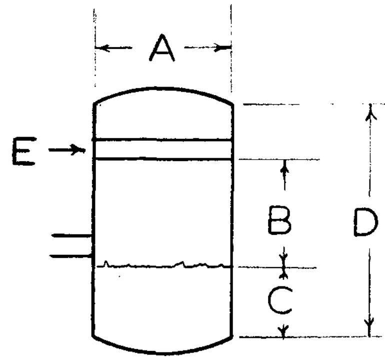


Table 26--Continued

Flash Vessel Specifications

Flash Tank ID No.	Diameter (ft) A	Demister Height Above Liquid Level (ft) B	Liquid Level (ft) C	Overall Height (ft) D	Pressure (psig)*	Temperature (°F)
			30-sec. Residence Time Design			
F-1	12	11	4	20	15	300
F-2	10	10	4	20	2.0	300
F-3	10	10	4	20	11'' Hg vac.	300
F-4	10	16	4	20	19'' Hg vac.	300
E	Demisters are constructed of 316 stainless steel 4 inches thick, 9 pounds per cubic foot at the design diameter for each tank.					

* All vessels designed for a full vacuum.

Notes

1. Each tank is constructed of carbon steel and coated with an epoxy resin.
2. The wall thickness of .220 inches includes a corrosion allowance of .025 inches.

Table 27

Pretreatment Equipment Cost Summary
Sugar Beet Process

Equipment	Capacity	Cost ($) [1]	Total ($)
Vessels:			
Surge Tank 1 (pressure vessel)	26,000 gal	20,000	20,000
Surge Tank 2 (Atm)	120,000 gal.	40,000	40,000
Acid Tank (Atm)	28,000 gal	12,000	12,000
Caustic Tank	28,000 gal.	12,000	12,000
Degassifier	5'dia.x 20' height	22,000	22,000
Pumps			
Transfer Pumps, 1,2,5	6,000 gpm	16,000	48,000
Acid and Caustic Pumps, 3,4	0-2 gpm	700	1,400
Misc.			
(2) Static Mixer		6,000	6,000
Filter			10,000
		2) Total	171,400

1. Estimated costs obtained from references (6, and 9 through 15)

Does not include installation

Table 28

Instrumentation Cost Summary
Sugar Beet Process

Instruments	Cost ($)
Pressure and Temperature Gauges	480
pH Controller	3,600
Temperature Controller	1,750
Pressure Controllers	3,144
Differential Pressure Controller	600
Level Controllers	7,000
Flow Control Valves	25,600
Total	42,174

Table 29

Flash Tanks and Piping Costs Summary

Sugar Beet Process

		Cost ($)
Flash Tanks [1]		
F-1	17,000 gal	14,870
F-2	12,000 gal	9,350
F-3	12,000 gal	9,350
F-4	12,000 gal	9,350
Total		42,920
Piping [2]		
Piping From Production and to Injection Wells[3]		564,000
Piping Within Plant and Pretreatment Plant[4]		143,600
Total		707,600

1. Includes Demister Costs, 4", 9#/cu ft 316 stainless steel, 10 ft diameter demister pad, $1650. 12 ft diameter demister pad, $2370 (11, 13)

2. Includes Installation, Longyear (8)

3. 6,000 feet of 20 inch Schedule 20 pipe

4. 400 feet of 16 inch Schedule 30 pipe for steam lines
1,200 feet 20 inch Schedule 20

Table 30

Cost Summary

Sugar Beet Processing

Equipment	Cost ($)	
Flash Tanks	42,920	
Pretreatment Equipment	171,400	
Instrumentation	42,174	
Installed Costs (47% of purchased equipment)	120,660	
Installed Piping	707,600	
Well Costs	7,150,000	
Sub-Total		8,234,754
Incremental Surface Area Cost [1]		1,610,000
Air Compressor [1]		50,000
Vacuum Pump [1]		68,000
Net Capital		9,962,754

1. Refer to Table 1, Chapter V

Table 31

Vessel and Major Equipment Specifications, Case $3b_1$

Equipment	Capacity	Length (ft)	Dia. (ft)	Temp. (^oF)	Pressure (psig)
BARLEY MALTING PROCESS					
Surge Vessels					
1	2,500 gal	9	7	300	80
2	26,000 gal	20	15	150	Atm.
Treatment Storage Tanks					
Acid	6,000 gal	15	8	Ambient	Atm.
Caustic	6,000 gal	15	8.0	Ambient	Atm.
Degassifier	-	8.0	2.0	300	80 psig
Centrifugal Transfer Pumps 1, 2, 5	550 gpm	80 ft head		300	80
Acid and Caustic Gear Pumps	0-1 gpm			Ambient	80
(2) Static Mixers, Stainless, Titanium		12	6 in.	300	80
Filter, Cartridge Style		150 sq. ft		150	80

NOTES:

1. Each surge vessel and storage tank is constructed of carbon steel and coated with an epoxy resin. The degassifier is constructed of 316 stainless steel.

2. The following wall thicknesses (t) include a corrosion allowance of .025 inch for carbon steel tanks and .213 inch for the stainless steel degassifier. Surge Vessel 1, t = .314 inch; Surge Vessel 2, t = .137 inch; Storage tanks, t = .117 inch; Degassifier, t = .307 inch.

Table 32

Pretreatment Equipment Cost Summary
Barley Malting Process, Case $3b_1$

Equipment	Size	Cost each ($)	Cost[1] ($)	Cost[2] ($)
Vessels				
Surge Tank 1	2,500 gal	7,000	7,000	7,000
Surge Tank 2	26,000 gal	14,000	14,000	14,000
Acid Tank	6,000 gal	3,500		3,500
Caustic Tank	6,000 gal	3,500		3,500
Degassifier	2 ft x 8 ft	7,000		7,000
Pumps				
Transfer Pumps, 1,2,5		4,500	9,000	13,500
Acid and Caustic Pumps, 3,4		400		800
Misc.				
(2) Static Mixer				2,000
Filter			4,000	4,000
			34,000	55,300

1. Option 1, No pretreatment

2. Option 2, with pretreatment

Table 33

Instrumentation Cost Summary

Barley Malting Process, Case 3b[1]

Equipment	Cost[1] ($)	Cost[2] ($)
Pressure and Temperature Gauges	300	480
pH Controllers		3,600
Temperature Controllers	1,750	1,750
Differential Pressure Controller	-	600
Level Controller	2,000	3,000
Flow Controller	3,400	7,000
	7,450	16,430

1. Option 1, no pretreatment
2. Option 2, with pretreatment

Table 34

Process Equipment and Piping Cost Summary
Barley Malting Process Case $3b_1$

Equipment	Cost[1] ($)	Cost[2] ($)
Air Heater (installed cost)	450,000	450,000
Fans (installed cost)	80,000	80,000
	530,000	530,000
Piping (installed cost)		
Piping from Production and to Injection Wells[3]	192,000	192,000
Piping within Barley Malting Plant and Pretreatment Plant[4]	16,000	35,200
Total	208,000	227,200

1. Option 1, no pretreatment

2. Option 2, with pretreatment

3. 6,000 feet of 6-inch Schedule 40 pipe

4. 500 feet of 6-inch Schedule 40 pipe for Option 1
   700 feet of 6-inch Schedule 40 pipe for Option 2

Table 35

Cost Summary

Barley Malting Process

Equipment	Option 1 No Pretreatment Costs ($)	Option 2 (With Pretreatment) Costs ($)
Air Heater and Fans [1]	530,000	530,000
Pretreatment Equipment	34,000	55,300
Instrumentation	7,450	16,430
Installation Costs	20,000	34,000
Piping (installed costs)	208,000	227,200
Total	799,450	862,930
Well Costs [1]	2,200,000	2,200,000
Total	2,999,450	3,062,930

1. Refer to Table 8, Chapter VII

Table 36

Equipment Specifications and Costs for Other Geological Locations

Equipment	Capacity	Thickness[1] (in)	Corrosion[2] Allowance (in)	Cost ($)
SUGAR BEET PROCESS				
Surge Vessels				
1	26,000 gal.	.995	.375	30,850
2	120,000 gal.	.337	.188	77,470
Treatment Storage Tanks				
1 Acid	28,000 gal.	.237	.125	20,760
2 Caustic	28,000 gal.	.237	.125	20,760
Flash Tanks				
1	17,000 gal.	.570	.375	38,530
2	12,000 gal.	.570	.375	24,220
BARLEY MALTING PROCESS				
Surge Vessels				
1	2,500 gal.	.667	.375	14,870
2	26,000 gal.	.300	.188	30,660
Treatment Storage Tanks				
1 Acid	6,000 gal.	.215	.125	6,430
2 Caustic	6,000 gal.	.215	.125	6,430

[1] Thicknesses include corrosion allowances.

[2] Corrosion allowances are based on temperature rating of vessels and a brine corrosivity similar to seawater (17).

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