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FOREWORD

The papers in this volume were presented at the CHEMPID/PMCD joint Spring Symposium April 1-3, 1980. Acknowledging the theme "Process Efficiency Through Instrumentation."

(5) CHEMPID and (4) PMCD sessions were offered.

The session developers, selected by Program Coordinators James O. Gray (PMCD) of the Foxboro Company and Keith S. Herbst (CHEMPID) of Moore Products Co., worked diligently to gather together over 30 speakers and authors. This volume provides a formal record of the authors' up-to-date and comprehensive views of their respective areas of interest.

Finally, our thanks to the sponsoring New Jersey Section and its host committee for their outstanding efforts.

K. S. Herbst
Program Coordinator, CHEMPID

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IMPROVED COMBUSTION EFFICIENCY
VIA STACK EMISSION MEASUREMENTS

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ABSTRACT

A novel extractive stack sampling system, which meets pollution monitoring requirements, is being used in an experimental program to study optimization of combustion in fossil-fuel-fired boilers, via stack emission measurements. Unique features of the system are described, and progress of the experiments is reported.

INTRODUCTION

In August, 1979, the Environmental & Process Instruments Division of The Bendix Corporation began an experimental program to study optimization of combustion in fossil-fuel-fired boilers. The objective of the experiments is to determine how much improvement in efficiency can be attained by improving the precision and accuracy of measurements of stack emissions. The following sections of this paper describe the experimental apparatus, and report on the progress of the experiments.

SAMPLING SYSTEM

The sample extraction probe and conditioning assembly, which is designed to operate reliably, even in the worst hostile environments, is described below. The probe and conditioning assembly is 1.5m (5 feet) long (probe length variable). All probe components, except the sample extraction line, are 316 stainless steel. The sample extraction line is 1/4" Hastelloy C tubing. The probe consists of a primary or first stage 500 micron filter and a secondary inertial one micron filter. Stack gas is extracted through the filters by an air-operated vacuum transducer. Extraction rate is about 60 liters per minute at a velocity of about 27 meters per second (88 feet per second). A particulate-free sample, less than .5 micron, is extracted from the inertial filter outer shell space and fed continuously into the conditioning assembly attached to the outer wall. The probe and conditioning assembly is attached to the stack by a 4-inch pipe flange mounting plate.

Inside the conditioning assembly, the sample enters an air-cooled, teflon heat exchanger. Heat exchanger cooling air flows in an opposite direction to sample flow, supplying air for the in-stack vacuum transducer. Acid mist condensibles and entrained liquids fall out into the exchanger and are collected in a liquid trap. Sample gas flows from the liquid trap head space through a vaporizer, which is located upstream from a Permapure dryer.

The vaporizer, heated to 120°C, is a safety device that assures complete sample vaporization and removal of trace particulates via a glass filter element. The Permapure dryer, 1.2m (4 feet) long, 200 tubes, removes water vapor through the permeation distillation principle, drying the sample to a dewpoint of less than -35°C (-30°F). An air ballast tank is externally attached to the conditioning assembly. It provides 80-100 PSI air for periodic blowback of the primary probe filter and liquid trap purge.

SUPPORT ASSEMBLY

The equipment described above is located at the sampling point; the remaining equipment is located in a mobile instrument house at ground level about 18m (60 feet) from the stack base and about 70m (230 feet) from the probe entry point. An unheated and sheathed wiring and plumbing umbilical assembly provides an interface between the stack probe conditioning assembly and the instrument house. All power, air, calibration gases, and control signals for the probe/conditioning assembly are supplied by the umbilical assembly.

The support assembly provides the electrical and gas supplies needed to control the probe/conditioning assembly. This assembly provides calibration gases to the probe and air for the heat exchanger, vacuum transducer, Permapure dryer, sweep, periodic filter blowback, and liquid trap purge. The required flow control is provided by electrically-actuated solenoid valves, pressure gauges and a Permapure sweep air flowmeter.

Electrical supply and control within the support assembly consists of 115 VAC primary power distribution, AC solenoid valve driver printed circuit card, and remote or manual flow selector switches. The components are electrically interfaced to produce precise valve timing functions. These functions are controlled in the stack probe/conditioning and support assemblies.

UMBILICAL ASSEMBLY

The probe/conditioning and support assemblies are interfaced with a PVC-mylar sheathed umbilical assembly. The umbilical contains two 1/4" teflon lines for sample and calibration gases, two 1/4" polyethylene lines for Permapure sweep and vacuum transducer air, and a 1/2" polyethylene line for blowback ballast tank air. Nine 14-gauge electrical wires are provided for AC electrical interfacing.

SAMPLE TRANSPORT ASSEMBLY

The sample transport assembly transfers the sample from the umbilical sample line to the various analyzers via pump and flow controlled outlets. The transport assembly provides two modes of calibration--up the stack and direct to analyzers. This feature is useful for diagnostic purposes, permitting the operator to determine sampling system recovery losses, leakage and lag, rise and fall response times. The sample transport assembly controls all flow controls, electrical power distribution and manual or automated valve driver cards and switches.

Three heatless air dryer units located outside the instrument house are used to provide air for the blowback ballast tank, the heat exchanger/ vacuum transducer, and the Permapure dryer sweep air. The calibration cylinders and their pressure regulators are also located outside the instrument house. Component-selector solenoid valves are located within the house in a small Hoffman enclosure box.

SEQUENCER

The probe sequencing controller is built around the Bendix Model 9000 Analyzer Control. This microprocessor-based controller has a basic word length of 12 bits. State-of-the-art CMOS design is used throughout the microprocessor controller, providing very low power consumption.

An existing standard software package allows the operator to perform manual zero and span calibrations for each line and generate reports on demand. The controller is capable of generating 20 digital signals which can be used to control valves, including the stack probe valves, sample system valves, and calibration gas valves. In addition, the software allows for automatic sequencing of the above functions.

Through the use of a 12-bit analog-to-digital multiplexer board, the system can support up to 8 continuous analyzers. Each analyzer is sampled once on an operator-defined time interval of one to eight minutes and an hourly average is updated. Each hour these averages are reported and a shift average is maintained. Also each hour control signals are initiated to perform blowback on the probe. Shift averages are reported on eight time intervals, and like the hourly reports, can be inhibited from printing, or can be manually requested at any time through the front panel.

Once a day the sampling and monitoring system is automatically calibrated. Control signals are actuated, causing the calibration gases to be drawn from the probe through the sample system into the analyzers. Initially, nitrogen is used to establish a zero baseline reference for the continuous analyzers. Other calibration gases are circulated and the appropriate response factors are generated for each line. Microprocessor scanning of the analyzers is inhibited during times of calibration and blowback. The microprocessor is interfaced to a micro printer which provides hard copy local printout of the one and eight hour time-weighted-average data for the gaseous emissions. The micro printer can produce copy at a rate of 150 full lines per minute on 12 cm (4.75 inch) wide roll paper.

DATA ACQUISITION EQUIPMENT

In addition to the local data processing equipment above, the instrument house contains datalogging and transmission equipment for sending the experimental data over telephone lines to a central computer at Lewisburg, West Virginia, for real-time on-line data analysis. The transmission equipment consists of a Computer Products, Inc., RTP Process Interface Subsystem connected to a Bell System Model 201C modem. At Lewisburg, a Data General computer is connected to another Bell System Model 201C modem.

OXYGEN ANALYZER

For continuous measurement of oxygen, the instrument house is equipped with a Teledyne Model 326 Oxygen Analyzer. The analyzer utilizes a micro fuel cell for detection of O_2. The fuel cell is a sealed electro-chemical transducer which requries no electrolyte or periodic cleaning of electrodes. The clean, dry sample provided by the Bendix sample-conditioning system affects the life of the cell. The normal life span of the cell is about one month depending on the oxygen level in the combustion system.

INFRARED ANALYZER

For detection of CO, CO_2 and SO_2, the instrument house is equipped with a Bendix Model 8903 Infrared Gas Analyzer. This triple-bench IR analyzer is designed to measure concentrations of three components in a continuously-exchanged gas sample stream. The analyzer reacts quickly and linearly to quantitative changes in the sample, while providing a highly-stabilized measurement for extended periods without repeated adjustments. Proportional output signals are provided for recording. This infrared analyzer utilizes the non-dispersive single-beam technique, with alternate modulation of the sample and reference cells. The measurement principle is based on each component to be measured having a known characteristic absorption spectrum in the infrared range. The reference cell is filled with a non-absorbing gas and sealed, and the sample is continuously passed through the sample cell. With no component present in the sample, the amounts of radiation coupled into the detection chamber from the sample and reference cells will be essentially equal, effectively cancelling and producing no output. When the component is present in the sample, radiation is absorbed by it, causing the inputs to be unequal and producing an output from the detection chamber. The detection chamber is designed so that the output is proportional to the component's concentration.

OXIDES-OF-NITROGEN ANALYZER

For detection of oxides of nitrogen, the instrument house is equipped with a Bendix Model 8101CX NO-NOx Analyzer with an extended monitoring range of 0-500 PPM. This oxides of nitrogen analyzer is designed to measure high levels of NO, NO_2 and NOx in a continuously exchanged gas sample. The system reacts quickly to quantitative changes in NO, NO_2 and NOx content of the sample, while providing a highly stabilized measurement capability for extended periods without repeated adjustments, In addition to its front panel meter indication, it provides an output signal proportional to the individual measurement. The Model 8101-CX utilizes the principle of photometric detection of the chemiluminescence resulting from the gas phase reaction of NO and ozone. Since this reaction only occurs between NO and ozone, the amount of NO_2 is determined by deriving the difference between NOx and NO. This is accomplished by dividing the input sample into two flow paths. One path goes directly into the reaction chamber and results in a measurement of NO. The other path, defined as NOx, goes through a low temperature catalytic converter that changes NO_2 to NO, but has no effect on the incoming NO. This second path then continues into the reaction chamber for measurement. The amount of NO_2 is determined by subtracting the NO measurement from the NOx measurement. The analyzer provides automatic or manual cycling between NO and NOx.

HYDROCARBON ANALYZER

For detection and measurement of hydrocarbons, the instrument house is equipped with the Bendix Model 8401 Total Hydrocarbon Analyzer. The Model 8401 uses a highly sensitive and reliable flame ionization detector (FID) cell which provides a rapid response to quantitative changes in hydrocarbons. The analyzer is designed to operate automatically without adjustment over long periods.

EXPERIMENTAL PROCESS

The boiler being tested is a power generation boiler located in the Westvaco paper mill at Covington, Virginia. The boiler is a coal fired unit designed for producing 375,000 pounds of steam per hour at a pressure of 600 PSI. The boiler is fed from three coal bunkers via three bowl mills. The boiler is equipped with two forced draft fans. The exhaust of the boiler enters a precipitator, an induced draft fan, and finally the stack. The extractive sampling system equipment is installed between the boiler and the precipitator. The stack probe, therefore, is approximately 15m (50 feet) from the combustion zone, and experiences typical stream conditions as follows:

Temperature: 190°C (375°F)
Particulate Load: 6 grains/SCF
Percent Water: 7-8%
Gas Flow: 39000 CFM
Gas Velocity: 3.4 m/s (11 feet/s)

The boiler control system consists of a Bristol UCS 3000 computer control system, which normally monitors steam header pressure as an index of boiler loading, and produces appropriate signals to control feeder speeds, and damper positions. Prior to the experiments, a Westinghouse flue gas oxygen analyzer was used to measure excess oxygen; the control system used the excess oxygen measurement, plus measurements of forced draft air flow, to trim the air flow to give a desired value of air-to-fuel ratio.

PROGRESS OF THE EXPERIMENTS

The experimental apparatus described above was constructed during the period from August 1979, through October 1979, and was placed in operation during November 1979. A series of experiments designed to show correlations between emission measurements and boiler efficiency are underway, and are expected to be completed during the second quarter of 1980. The results of the experiments will be submitted for publication as soon as possible thereafter.

ACKNOWLEDGEMENT

The authors are grateful for the support provided by all of the Bendix and Westvaco personnel who are making this series of experiments possible. Special thanks are due to Richard Wood and Lloyd Fitzgerald, who not only assembled, but also are running the experimental apparatus.

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EXPERIMENTAL PROCESS

The boiler being tested is a power generation boiler located in the Westvaco paper mill at Covington, Virginia. The boiler is a coal fired unit designed for producing 325,000 pounds of steam per hour at a pressure of 680 PSIG. The boiler is fed from three coal bunkers via three bowl mills. The boiler is equipped with two forced draft fans. The exhaust of the boiler enters a precipitator, an induced draft fan, and finally the stack. The extractive sampling system equipment is installed between the boiler and the precipitator. The stack probe, therefore, is approximately 24m (80 feet) from the combustion zone, and experiences typical stream conditions:

Temperature: 190°C (375°F)
Particulate Load: .6 grains/SCF
Percent Water: 7-8%
Gas Flow: 11,000 cfm
Gas Velocity: 3.4 m/s (11 feet/s)

The boiler control system consists of a Bristol UCS 3000 computer control system, which normally monitors steam header pressure as an index of boiler loading, and produces appropriate signals to control feeder speeds, and damper positions. Prior to the experiments, a Westinghouse flue gas oxygen analyzer was used to measure excess oxygen; the control system used the excess oxygen measurement plus measurements of forced draft air flow, to trim the air flow to give a desired value of air-to-fuel ratio.

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FIRED PROCESS HEATERS

THE OTHER COMBUSTION PROCESS

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ABSTRACT

Many papers have been written over the years on the general subject of combustion control for boilers but very few on combustion control for process heaters.

During the past 50 years boilers have evolved to more complex designs which have tended to make the combustion control job easier and more straight forward. Process Heater fuel burning design changed little however until very recently as forced and induced draft and heat recovery equipment is being used to greater extent. Due to this the great majority of existing heaters use very simple furnaces, more primitive burner arrangement, negative draft and natural draft furnaces relying on stacks to produce the combustion air flow.(1) All of these combined with the ever present variable Btu fuel have made automatic control combustion more difficult.

Because of this difficulty and the low price of fuel in the past, complete automatic combustion control for process heaters has not been used for all heaters nearly as much as for boilers. Analysis of heater control problems can however demonstrate some relatively simple solutions to the basic problems in providing control of combustion for natural draft process heaters. The purpose of this paper is to describe some solutions to these problems and justify their use.

INTRODUCTION

"We are all aware that during the past few years there has been an increasing fuel oil shortage. Last winter this alarming situation was particularly brought home to us, as domestic users, when it was necessary in many localities to resort to emergency measures in order to assure sufficient distribution to meet minimum fuel requirements. Although fuel manufacturing and supply facilities are being rapidly expanded by the oil industry, it does not appear that adequate supplies will be available in the immediate future.

Less apparent to many of us perhaps, but of more importance to our economy, is the effect of the fuel oil and fuel gas shortage on industrial users. Nearly every industry in some phase of its operation is dependent upon fuel as a source of heat energy for steam generation, process operation, and the many other uses with which we are all familiar. It is safe to say that one of the prime requisities that enable industry to produce at its maximum capacity is the availability of fuel."

Sound familiar. The preceding paragraphs are the opening paragraphs of ISA Paper No 48-3-2.(2) The title of the paper "Experimental Application of Combustion Controls to a Process Heater" by W.E. Boyle of Shell Oil Co. and Paul R. Hoyt of Shell Development Co. The date of the paper 1948.

And from a 1930 Furnace Control Patent No. 1, 940,355 "In the operation of a furnace it is of primary importance that temperatures at certain points be maintained substantially constant.....- to prevent damage to the still. At the same time the fuel supplied for heating the furnace should be burned in a manner and amount as efficient as possible."(3)

WHY CONTROL HEATER COMBUSTION?

The name combustion control implies the control of combustion or regulation to a desired fuel/air ratio. The name combustion control also implies controlling the heat input to match the required heat output while simultaneously maintaining a status quo on the level of the heat output. This level can be steam pressure in the case of a boiler or outlet feed temperature in the case of a process heater.

Maintaining a desired fuel air ratio is not an end in itself. One end is to save fuel. At present fuel costs this end is a major one. Lets look at a combustion system from a fuel saving standpoint.

Figure 1 shows two ways to save fuel in a combustion process(4): One, is the use of heat traps such as air preheaters to return some of the flue gas heat loss to the furnace. Using air preheaters today may also require low NO_X burners, since in general air preheat increases NO_X formation.

The other way shown in Figure 1, is to optimize the amount of combustion air so that heat losses due to the heating of air not required for complete combustion will be minimized. It is not as well known however that reducing excess air for combustion reduces flue gas temperature. This result of greater heat transfer occurs due to more contact time between sufaces and the slower moving flue gases. This fact is now receiving greater recognition.

A properly designed heater and its burners will not produce more efficient combustion. It only provides the potential. The key to optimizing air for combustion is a properly designed and operating combustion control system.

Many years ago in the 1900-1920 period, it was becoming well known that combustion control instrumentation saved fuel; since 1940, almost all boilers have been equipped with some form of combustion control. Such controls generally regressed in sophistication of system in the 1950's and 1960's because of the low fuel cost which reduced savings potential relative to cost of equipment.[4] But even the simpler combustion control equipment has not been used to any great extent for existing process heaters. Perhaps with current high fuel costs greater use of combustion control can be expected.

PROCESS HEATER COMBUSTION CONTROL SYSTEMS

Lets assume a criteria for a Natural Draft Heater Control system that will maintain temperature with minimum outlet temperature disturbance and at the same time.

1. Control Fuel Air Ratio
2. Accommodate Fuel Pressure Disturbances
3. Accommodate Fuel Btu Disturbances
4. Accommodate Variations in Feed Input rate
5. Accommodate variations in Feed Input Temperatures

A basic recommended approach today appears to differ only in one important detail from the combustion control system installed on a process heater in 1946 at Shell Oil Co., Wood River Refinery. In their 1948 paper for the Instrument Society of America, (No. 48-3-2), W. E. Boyle and Paul R. Hoyt, describe their experimental system which they state, paid for itself in fuel-savings in 3-1/2 months[2]. The fuel-cost, capital environment of that time is shown in Figure 2.[4].

Note that the environment was more favorable to a fuel-saving investment then, than at any other time from 1946 to 1973. One reason usage of such equipment did not catch on at that time may be due to the low cost of fuel and the general feeling that combustion control to reduce the wasting of fuel was unnecessary. In addition, a key element in the system was a first generation analyzer for percent oxygen which required relatively high maintenance and periodic checking.

Hoyt and Boyle indicated, however, that even in 1948 they considered an oxygen analyzer a practical device for combustion control. The control arrangement of that experimental system is shown in Figure 3.

Today, two key drawbacks to the general use of systems of this type have disappeared. Fuel is very expensive, and third generation zirconium oxide oxygen analyzers have relatively low maintenance and are considerably more reliable than the earlier analyzers.

Figure 4 shows a conventional process heater control system which is probably a model for a large percentage of process heaters used today[4]. It shows a flow control of the feed input through the convection section, through the heating coils in the furnace and to the output. This constant flow control can either have an internal set point or external set point cascade controller following another controller. This conventional system generally provides a means of manually regulating the exit damper to control the flow of air through the unit and some means for determining the analysis of the flue gas. The temperature control is generally a cascade controller which cascades onto a pressure or flow control, thus compensating for variations in upstream fuel pressure or number of burners in service.

With this system, however, variations in fuel Btu content reflect through the unit and into disturbances of the outlet temperature which must then correct through the temperature control. This Btu variation also causes a variation in the flue gas analysis or the excess air, because of the variation in total heat input into the unit.

Figure 5 shows a potential step to improve the performance from an excess air or fuel economy standpoint by connecting the flue gas analysis into a controller which then trims the control of combustion air flow through the unit[4]. With this arrangement, the excess air can be controlled at a fixed point. This arrangement does not, however, compensate for variations in Btu content of the fuel, therefore potential temperature disturbances of the feed outlet still remain. This arrangement is very similar to that used by Boyle and Hoyt in 1948 with the 3-1/2 month payback referred to previously.

Figure 6 shows the general relationship between excess air and furnace temperature. It is noted that constant excess air will result in more uniform furnace temperature. This will enable the heat transfer to take place more efficiently with less heating and cooling of the furnace. Among other benefits, this can result in lower furnace maintenance.

Today, however as it is shown in Figure 10, we realize that if the flue gas analysis is used to trim fuel rather than air flow, an added benefit is realized. Not only is excess air controlled and furnance temperature held more constant, but changes in fuel Btu are compensated for [2]. Since oxygen analysis at 5-10 second response is much faster than the heat transfer response changing outlet fluid temperature, the percent oxygen analyzer, feeling the change that a total Btu flow input change causes to the flue gas analysis, through its input to control readjusts the fuel to its original total Btu flow input before temperature changes significantly. This improves performance of the control of outlet fluid temperature. In this manner the percent oxygen controller constantly recalibrates the fuel controls into a Btu flow system, matching it to the fuel Btu demand signal.

The technical basis for readjusting fuel from percent oxygen is the same used for the first practical combustion guide for steam boilers--the STEAM FLOW /AIR FLOW boiler meter (5). This basis is that for a constant Btu input of fuel, the amount of air required for combustion is approximately the same, even though the Btu/unit wt. or volume changes. Using typical examples for each fuel the table below demonstrates this fact, one that is often not recognized:

Fuel	lbs. Air per 10,000 Btu Gross	lbs. Air per 10,000 Btu Net	lbs. Air per 10,000 Btu Net +15% Excess
Natural Gas (1000 Btu)	7.18	7.88	9.06
Refinery Gas (1600 Btu)	7.21	7.84	9.02
Fuel Oil No. 6 (18,000 Btu)	7.31	7.75	8.91
Coal (11,500 Btu)	7.56	7.79	8.96

By showing air requirements for a net 10,000 Btu input basis, the table takes into account the latent heat loss effect on combustion efficiency. As shown in the table, the relationship is even more precise when only the net Btu is considered.

Figure 7 also dramatizes a minimum improvement modification of the conventional system shown in Fig. 4 by adding a percent oxygen control to modify fuel input. In this arrangement, the temperature control connects not only to fuel control, but also in parallel to draft control. A variation in temperature will therefore cause an increase in fuel and an increase in draft at the same time with percent oxygen control acting as a trimming control on the fuel. This arrangement is shown on a natural draft heater, the most common existing type.

Questions are often raised concerning the ability to control excess air with an outlet damper, and the danger of driving the furnace into a pressurized condition by partially closing this damper. It is well known that natural draft burners of the type used, inspirate air from the atmosphere in addition to the air flow due to draft differential across the burner. The amount of air delivered by the burner should not exceed that being drawn out by the stack under a condition of draft in the furnace. Thus, it is necessary to have the burner inlet air registers balanced with the stack damper in proper position for the amount of excess air desired with a condition of draft in the furnace. Then a control system as described above, reduces damper opening only when fuel flow is also being reduced. This results in less atmospheric air being delivered to the combustion chamber at the time the damper opening is being reduced.

Other refinements to heater control systems produce added benefits. In the arrangement in Figure 8, the fuel flow control is shown as a flow control loop with square root extraction rather than pressure control. This tends to linearize the mathematical relationship between a temperature deviation signal and the fuel flow signal when necessary fuel changes are made to eliminate the temperature deviation. Such an arrangement is also partially self compensating when fuel specific gravity and Btu content both increase or decrease in parallel.

The use of the flow control thus allows more optimum temperature control tuning over the turndown range since a linear change in temperature is equivalent to a specific Btu flow change for a specific fluid flow(2). You will note in the arrangement Fig. 8, the oxygen control is combined with a feedforward signal from the feed input. For a change in feed rate the feed input signal then causes the fuel to change before a change in temperature occurs. This is modified by the oxygen control. The summation of the oxygen and the feedforward from a straight line feed flow provides the correct mathematical relationship between these inputs and fuel Btu input to produce the potential for more precise temperature control.

As described above, the oxygen controller acts as a Btu variation detector since the required amount of combustion air relates to the number of total Btu of fuel rather than the volume of fuel. In order for such a system to perform properly however it is necessary to match the gain of the multiplier in the signal to the fuel flow set point to the range of Btu variation expected. Lets assume that we have a gas fired heater and we expect a variation of 600-1800 Btu/Scf in the fuel to a particular heater, and that we wish to obtain the correct multiplications for this range at 0 and 100%. For a 1-5VDC system this relationship is 5 volts for 600 Btu and 1 volt for 1800 Btu.

The first impression is that the multiplier at 600 Btu should be 3 times that at 1800 Btu. This is not the case however. For a constant Btu of fuel, assuming constant pressure and temperature, the indicated flow will be affected by flow of the fuel and the specific gravity of fuel. As fuel Btu changes over the 3 to 1 range, for a given total Btu input flow will change inversely. There will be however a very significant error in the indicated flow due to change in specific gravity. This error is normally in the right direction to partially self compensate on a Btu basis.

The approximate relationship between specific qravity and Btu content of hydrocarbon gases is shown in Fig 9. Assume a design specific gravity of .54. The effect of specific gravity change on a Diff. Press. flow measurement is shown in Figure 10.

Lets assume a heater is operating at a constant duty with 100,000,000 Btu input and the fuel is natural gas at 1000 Btu/Scf. Under these conditions fuel consumption is 100,000 scf. If the fuel constituments change so that Btu content is 600 Btu or alternately 1800 Btu/cuft at the same total heat input the data as shown in the table below would be true at 100,000,000 Btu input with orifice design at .54 sp. gr..

	Required Flow Scf/Hr.	Sp. Gr.	Sp. Gr. Corr.	Indicated Flow Scf/Hr.
Natural Gas 1000Btu/scf	100,000	.54	1.0	100,000
Gas 600Btu/scf	166,667	.26	1.44	115,741
Gas 1800Btu/scf	55,555	1.055	.715	77,699

Since the fuel controller operates from indicated flow it is only necessary to provide a multiplier sufficent so that the flow set point times the multiplier will ask for the indicated flow resulting in the same total Btu input.

In the arrangement shown in Figure 8 introducing a multiplier of 1.157 for the 600 Btu Fuel and .777 for the 1800 Btu fuel will assure that heat input remains constant assuring minimum disturbance to outlet temperature. If these multipliers were not introduced then as Btu changed from 1000 to 600, total input would change from 100,000,000/Hr. to 86,390,000/Hr., and for 1800 Btu fuel from 100,000,000/Hr. to 128,000,000/Hr. due to the action of the fuel flow control. Under these circumstances any corrective action would have to orignate from a disturbance in outlet temperature. At the same time excess air at 30% for a base condition would vary from 51% to 1% just due to the variation in fuel Btu content and would not return to the 30% excess air condition until the temperature corrected fuel flow back to the 100,000,000 Btu/Hr. input level.

A system such as that described in Fig. 8 will achieve the temperature control goal plus the first four listed items of the assumed criteria. To accommodate variations in inlet feed temperature a modification of the Feed Flow feedfoward control is necessary. This can be easily accomplished per Fig. 11 by having Temperature difference between Inlet Temperature and Feed Outlet Temperature set point multiply the Feed Flow feedforward signal.

If feed flow set point is manual and seldom changed and input feed temperature does not change then the feed forward portion of the control is of little benefit. Feedforward control to minimize outlet temperature disturbance would be indicated when the feed flow controller is a cascade controller or when there is an inlet temperature variation.

TESTING OF CONTROL SYSTEMS WITH PROCESS HEATER SIMULATOR

The control arrangement shown in Figure 8 for control of a natural draft heater was applied to a process heater simulation. Both the simulation and the control system used actual analog control equipment (1). To test the performance of the system, comparisons where made of the results of typical instrument and control arrangements normally used and the control system in Figure 8.

Figure 12 shows the effect of a feed flow step change with the CONSERVER III control of Figure 8. Notice that, as the feed flow is changed drastically, the air flow and the fuel flow change. Yet, there is very little change in the outlet temperature since the fuel and air keep pace with feed flow.

In Figure 13 the same changes were made with manual control of the air flow which takes away the effect of oxygen control. This arrangement of control is similar to the conventional controls of today described earlier. Note the deterioration of performance of the feed temperature control. In addition there are significant variations in percent oxygen. The fuel and air in this case are no longer coordinated without oxygen on automatic control.

In Figure 14 the results of the CONSERVER III system are shown when making step changes in unit Btu content. Notice that while the Btu changes drastically, the oxygen controller is continually at work holding the oxygen within range and improving the control of temperature. Note the variation shown starting at 1220 Btu then making a step change from 1770 Btu to 1160 Btu. This is a very drastic change and has the effect of a 30 to 40 percent change in total Btu input into the furnace, with no change in feed flow at that particular point in time. With the change of over 30 percent in total Btu input and constant feed flow, very little change in temperature occurs. With air and oxygen on manual, as with the conventional system, the same sort of fuel Btu changes and results are shown in Figure 15. Because of these Btu changes, changes occur in the temperature of the fluid and oxygen in the flue gas. In this case due to a change in oxygen the furnace temperature also changes, which helps deteriorate the performance of the temperature control. Benefits of improved control for keeping heaters operating properly is self evident.

ECONOMIC BENEFITS OF IMPROVED HEATER CONTROL

The net result of an improved control system is improved control of the heater. A major benefit of improved control is fuel economy. Other benefits are improved temperature control and reduced maintenance.

This writer tested the combustion of six process heaters as they were normally operating in a midwestern refinery. The average excess air was 60 percent and the average flue gas temperature was 850^oF. Controls installed were the conventional type previously described. Assume that the heaters could be operated at a consistent 20 percent excess air. This reduction will reduce the mass of flue gases by the ratio of 120/160: A 25 percent reduction in flue gas mass. In addition, estimates of flue gas temperature under reduced excess air would be approximately 780^oF.

Assuming an air entering temperature of 70^oF, as T (Flue gas Temperature - Air Temperature entering) is reduced from 780^o to 710^o, the reduction is nine percent. In addition, the improvement in fuel economy would result in lower fuel input and further reduction in mass of the flue gases.

Per the above assume, for a typical process heater the following input data:

Heater duty	- 50,000,000 Btu/Hr
Fuel	- No. 6 Oil
Fuel Cost	- \$3.50/MBtu - 50 cents per gal. \$21.00 per barrel
Excess Air	- 60%
Flue Gas Temperature	- 850^oF
Hours per year	- 8500
Desired Excess Air	- 20%
Final Flue Gas Temperature	- 780^oF Approximate

Initial conditions of thermal efficiency based on higher heating value and three percent radiation loss is 69.6 percent. For the final conditions thermal efficiency is 76.8 percent. This is an improvement of 7.2 percent in thermal efficiency and a 9.4 percent savings in fuel.

Annual Fuel Bill for such a heater is \$2,137,240.

Annual Fuel Savings due to excess air reduction \$200,900.

A ball park instrumentation cost estimate to upgrade a typical process heater control system to a fuel-saving system for the natural draft heater discussed is approximately \$30,000.

Allowing an additional \$30,000 for installation costs, results in an estimated total installed cost of \$60,000. Using these cost and the savings as outlined above:

Payback = 0.34 yrs or slightly over 4 months.

A simple form of ROI (return on investment), calculation using a 5 year straight line depreciation per the method from the National Bureau of Stds. Handbook 115 results in:

ROI = 314.83 percent

The positive cash flow with savings in the first year from such an investment starting in January is \$140,900. If delayed till September 13 the change reduces to 0 and is negative from September 13 till the end of December. After the first year the positive cash flow effect is the annual savings less cost of maintenance.

Cash flow change in future years will be increased by fuel cost increases which even before the events in Iran, were estimated at 10 percent per year. On the basis of installation by July 1, (initial year) and a very conservative 10 percent increase per year of fuel costs, results in cash flow improvement for the typical 50,000,000 Btu process heater as follows:

Year 1	\$ 40,450
2	\$220,990
3	\$243,089
4	\$267,398
5	\$294,138

In this example a \$60,000 investment was assumed. Significant improvements can often be made at lower costs by adding percent oxygen trimming control as an addition to existing systems.

It should be noted that the above calculations are based on the results of one test of typical operation. No doubt there are many typical operations operating at considerably less thatn 60% excess air. This would of cource reduce the potential for savings. It should be pointed out however that the normal heater without heat recovery equipment has flue gas temperature significantly higher than that of a boiler. A heater thus presents in most cases a better opportunity for savings from excess air reduction than does a boiler of the same heat input.

CONCLUSION

The high fuel-cost/captial-cost environment of today makes it necessary to take whatever steps can be taken to reduce fuel consumption. Fuel costs of today are four to five times higher than were prevalent five to six years ago making combustion control an excellent investment for improving process heater fuel consumption. Whether or not an improved control system is installed on a unit, it is paid for anyway, in wasted fuel and poorer performance. It is therefore sensible and frugal to make an investment to improve existing units to the extent they can be improved. It is also important to be sure that new units are designed to incorporate optimum combustion control systems.

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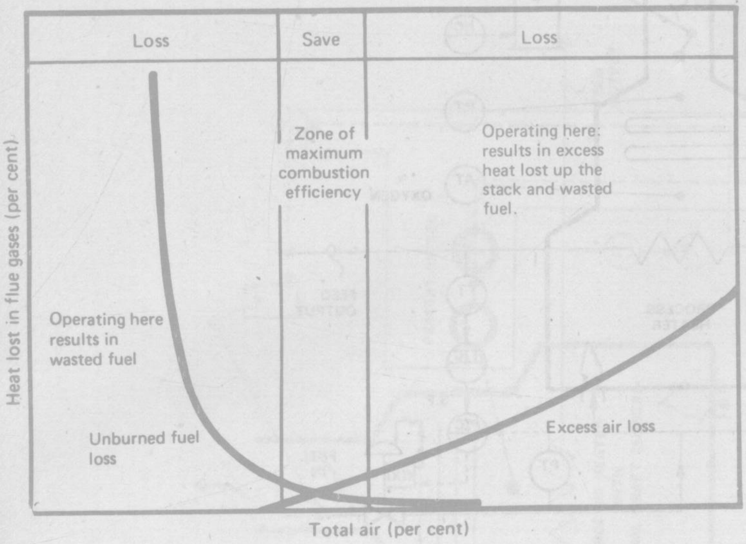
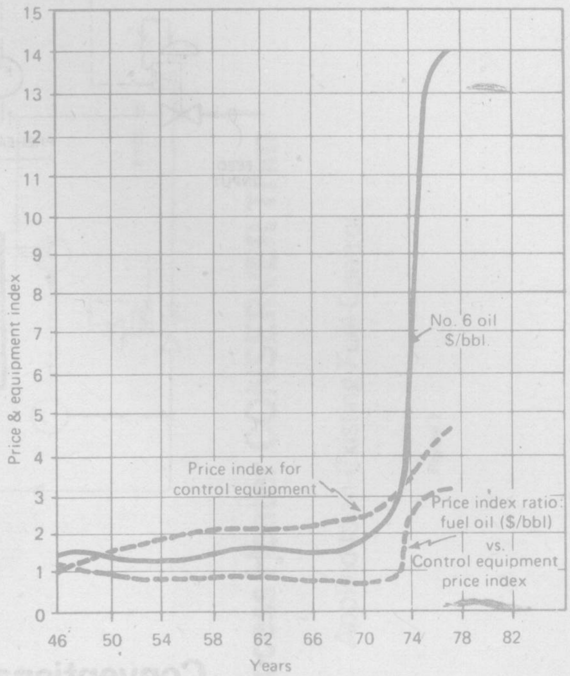


- Optimize air for combustion
- Trap heat out of flue gas (with air heaters or economizers)

Fuel Saving Methods

Figure 1



Capital/Energy Cost

Figure 2

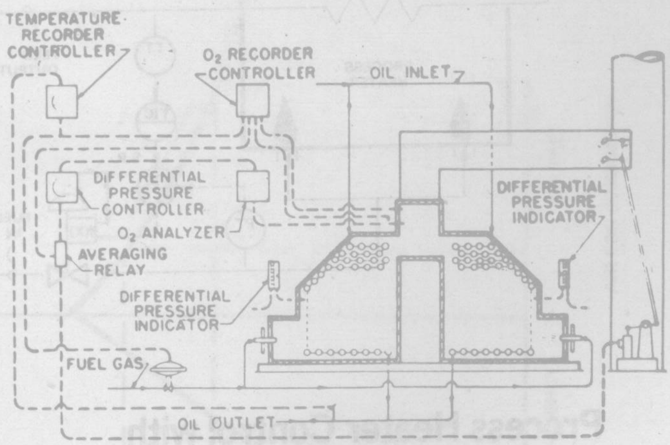
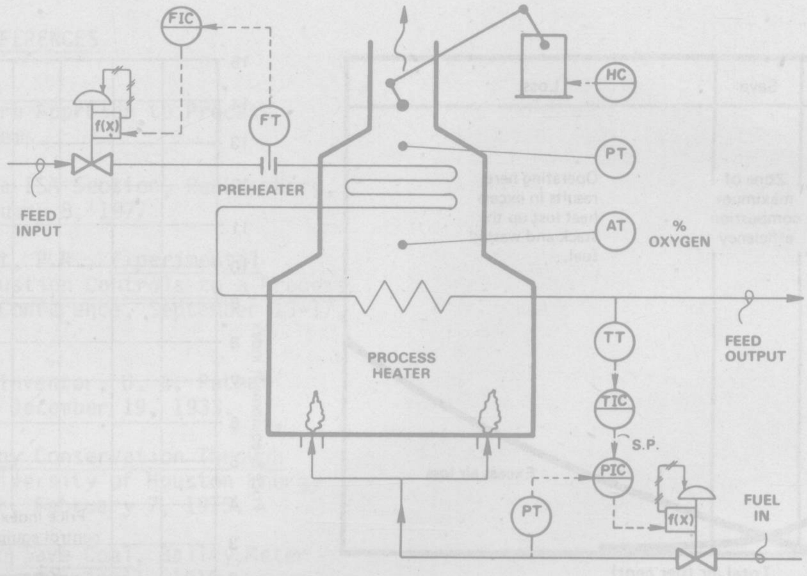
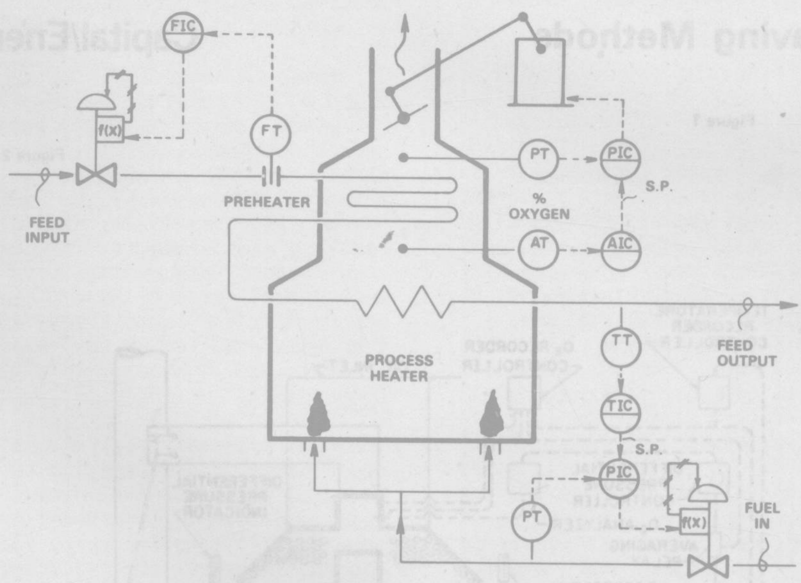


Figure 3



Conventional Process Heater Control

Figure 4



Process Heater Control with Oxygen Control on Damper

Figure 5