PRACTICAL OPERATING EXPERIENCE ON A RILEY GASIFIER

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ABSTRACT

This paper reviews the operating history and design of a commercial sized Riley-Morgan thin bed atmospheric coal gasifier. Results and experiences obtained over a five-year period are presented for a full range of U.S. Coals.

INTRODUCTION

In terms of technological history, the fixed bed process for coal gasification may be thought of as ancient. From the lighting of the streets of London to the firing of a blast furnace in 1840, the gas produced was by means of this process. Furthermore, gasification was not limited to coal as a fuel. Prior to 1923, a variety of organic materials and wastes (e.g., wine scum and peanut shells) had been successfully gasified.[1] In general, anything of an organic nature may be turned into a gas having a useful heating value. The key questions that remain, are "How well?", "With what degree of difficulty?" and "What is successful?"

History has shown fixed bed coal gasification to be well-matched in scale and energy quality to many industrial end-use needs. It is an easily understood technology and can be identified with the soft energy path strategy suggested in several recent technology assessments dealing with energy policy.[2]

A survey of the literature[1 3 4] indicates that experimentation in the producer gas field was dynamic from its inception. Anyone with a supply of a carbonaceous material appeared to be driven to build a unit to manufacture gas. And the gas was to be used for every conceivable purpose:

1. Steel production
2. Ore reduction (Calcining)
3. Power generation thru direct coupling to gas engines
4. Motive power for both land and amphibious vehicles
5. By-products.

A unit was even built called the Portable Pintsch Disinfectant Producer, which could be carried about by two men. Its purpose was to rid the premises of a house or a ship of rats and other vermin by flooding with a gas rich in carbon monoxide.

Had petroleum not been discovered when it was, a gas producer would no doubt have been designed for the airplane.

The striking aspect of the technology of the time (1850-1920) is the myriad of applications found for coal gasification, and the evolutionary steps taken in its development. We tend to think of trends toward the development of a slagging gasifier as a current innovation, and yet, in an effort to improve efficiencies, slagging gas producers were fully investigated in the 1880's and 90's and fully commercialized from 1902-07.

The need for gaseous fuels today parallels an earlier period in our industrial history. The range of applications today is even wider and more diversified. The constraints of the environment and economics make filling this need even more of a challenge. This paper illustrates how that challenge has been met, with emphasis on actual operating results.

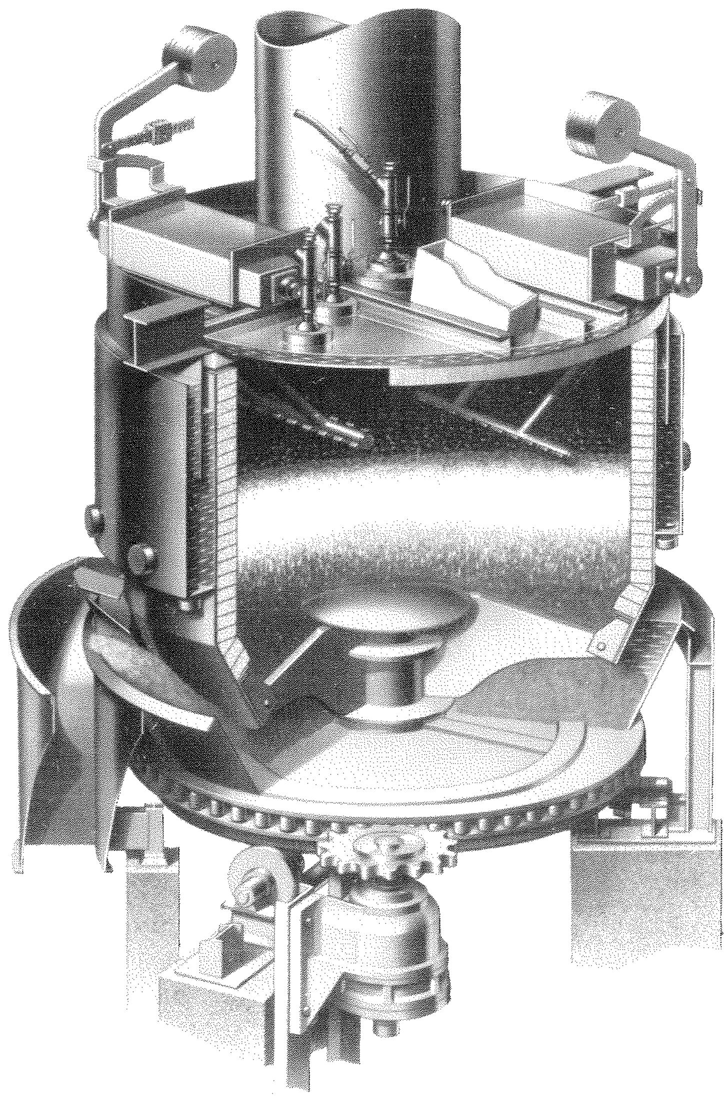

Figure 1 Riley-Morgan Gasifier Cross-Section

OUR HISTORY

In 1974, the Riley Stoker Corporation decided to make a considerable commitment to coal gasification, in light of the increasing evidence for the need of alternate fuel gases for industry.

A two-phase program was initiated. The first was on a purely R & D level, and consisted of the purchase of a two-foot I.D. fixed bed gasifier from one of the manufacturers in the field. Design and installation of an agitator and a cold clean-up train followed. A complete process development system resulted for the study and evaluation of various fuels suitable for gasification, together with a means for processing the gas and evaluating its combustion behavior.

Valuable hands-on experience was gained in the operation of this system, and it served as a training unit for personnel.

In parallel with the pilot effort, Riley Stoker undertook a program of design, development and commercialization of a full-sized, low Btu coal gasification unit. During a peak period (1921), approximately 11,000 gas producers were in use by U.S. industry[5] and that of this number a large portion were of a design developed by the Morgan Construction Company of Worcester, MA. In total, over 9,000 of Morgan units were manufactured and distributed throughout the world. They enjoyed a reputation for reliability, high capacity, and the ability to handle a wide variety of coals.

An agreement was entered into whereby Riley Stoker obtained the exclusive manufacturing rights to the Morgan Gas Producer and a redesign program was begun.

SUMMARY OF TEST PROGRAM

A total of twenty-two tests on the full-sized Riley-Morgan unit together with an equal number on the two-foot unit were carried out during the past five years. Fuels ranging from sized anthracites to sized and run-of-mine coals having free swelling indices up to 8.5, have been studied. A summary of fuels tested in this development program are given in Table 1. Some of the test runs caused some rethinking and redesign. Each of them resulted in valuable information.

DESCRIPTION OF THE UNIT AND DESIGN FEATURES

Figure 1 depicts the full-sized 10′6″ I.D. unit as it stands today. The major departures from other fixed bed processes are that it is a thin-bed, variable height process in which the entire fuel bed slowly rotates.

The thin fuel bed (up to 55 inches in height) allows variation in particle heating rate, the key to managing swelling coals. The rotating bed assures even distribution of fuel to the bed, an essential requirement when there is a size variation. This is accomplished by means of a rotating drum feeder, the length of which is nearly equal to the radius of the bed. Coal enters the reactor by falling through a slot across this radius, thus assuring a non-segregating feed to the top of the fuel bed with no moving internal mechanism. It must be noted that with thin fuel bed reactors, a variation of six (6) inches in depth may represent a 25-50% variation in total active fuel bed height.

Fuel bed agitation, another important element in the gasification of caking coals, is accomplished by means of one or two water-cooled horizontal bars of approximately one radius in length each offset slightly from the center of the fuel bed. The bars are externally weighed and counter balanced to achieve the desired depth of agitation. These bars act to prevent large agglomerates and open channels from forming in the fuel bed.

Figure 2 is a schematic of the gasifier and test system at Riley Stoker's Worcester (MA) R & D facility. Air is supplied by a fan through an 18″ I. D. pipe to the bottom of the gasifier. Steam is added through a sparger after the air flow meter and controlled on the basis of saturation temperature. The mixture is introduced into the bottom of the rotating pan through a central blast hood. While there is no grate in the system, the ash bed performs the function of a diffuser. The air-steam mixture moves countercurrent to the descending coal, first through the oxidizing zone, then reduction (gasification) zone, and the devolatilization zone.

COALS TESTED IN RILEY STOKER GASIFICATION TEST FACILITIES

COAL	RANK	NOMINAL SIZE	FREE SWELLING INDEX	ASH FUSION TEMPERATURE, °F (FLUID-REDUCING)
2 FT. DIAMETER PROCESS DEVELOPMENT UNIT				
Anthracite, Pa.	AN	PEA	0	2700
Pocahontas Seam, Va.	LVB	3/4″ × 1/2″	3	2700
Sewell Seam, W. Va.	HVAB	2″ × 3/4″	8	2700
Egypt Valley, Ohio	HVCB	2″ × 1/4″	4	2290
Illinois No. 6	HVCB	BRIQ	2.5	2160
Northern Plains	Lignite	2″ × 3/4″ 2″ × 1/2″ 2″ × 1/4″ ROM BRIQ	0	2100
RILEY DEMONSTRATION PLANT				
Anthracite, Pa.	AN	NUT & PEA	0	2700
Upper Banner Seam, Va.	HVAB	1 1/4″ × 1/4″	6	2630
Coronet No. 2, Va.	MVB	2 1/2″ × 1″	8.5	2560
Hazzard No. 4, Ky.	HVAB	2 1/2″ × 1″	4.5	2700
Elkorn No. 3, Ky.	HVAB	2″ × 1 1/2″	4.5	2660
Northern Plains	Lignite	2″ × 1/2″	0	2050

Table 1

Ash is removed from the bottom of the unit periodically by holding stationary a spiral arm on the floor of the unit. Relative motion between this arm and the rotating bed forces ash out to a water seal at the periphery. The ash is then lifted out of the water seal by sliding up a vertical chute or "ash plow." During normal operation the spiral arm and chute are allowed to rotate with the fuel bed.

The producer gas exits the building through a 36″ I. D. refractory lined duct to a cyclone with dust drop, and then to a sampling and metering run. The gas is burned in either an enclosed flare stack on the roof of the facility or in a water-cooled test furnace.

COAL FEED SYSTEM

The coal feed system is one of the areas where Riley Stoker has spent considerable effort in design improvement. Historically, most atmospheric gasifier manufacturers paid scant attention to fugitive emissions from the coal feed system. The main goal was getting fuel in. Consequently, each time coal was admitted to the reactor, some gas escaped into the coal delivery system. This resulted in most cases with main storage bunkers full of producer gas. Aside from the obvious hazard to personnel who might be called upon to work in this area, each of these bunkers was a bomb waiting to go off. Being too rich most of the time is probably what prevented explosions from occurring regularly. Most bunkers were, of course, vented to the atmosphere, a condition not to be tolerated today.

In the Riley Stoker lock-purge coal feed system, shown in Figure 2, a fuel inventory up to level A is maintained with valves 1, 2 and 3 in the positions shown. The valves are of a semi-ball type, with ground seats. As gasification proceeds, the fuel in the lock-hopper falls to level B, just below Valve 3, where its absence is detected by means of a sonar device. This triggers a sequence as follows: A short burst of steam is admitted into the line from the eductor to the lock hopper cleaning it of any residual tar or dust from the prior cycle. After an interval, Valve 5 opens, and the steam eductor now begins evacuating the lock hopper of gas and discharging it into the downstream gas piping system. The pressure in the lock hopper eventually becomes sub-atmospheric (26 inches Hg or less). At this time, valves 4 and 5 close, and valves 1 and 2 open, admitting coal from main storage to the lock hopper. These valves remain open until the lock hopper is filled again to level A (approximately one ton). At this time Valve 1 closes interrupting coal flow, and Valve 2 closes creating a gas tight seal. Valve 3 opens with Valve 2 closure, and the cycle is completed. Total elapsed time for this entire cycle is approximately one minute.

The chute between Levels B and C contains enough inventory so that continuous coal feed to the gasifier is assured.

By the maintenance of a nitrogen blanket above this system in the storage bunker just slightly above atmospheric, together with the eductor system, migration of gases is always toward the gasifier, and never from it. Thus, environmental integrity is constantly maintained.

Table 2 gives analyses of the gaseous environment in the gasifier building and coal bunker measured during a Level 2 environmental assessment of our process. This table, together with Table 3, is presented to quantify fugitive emissions and not to determine worker health effects.

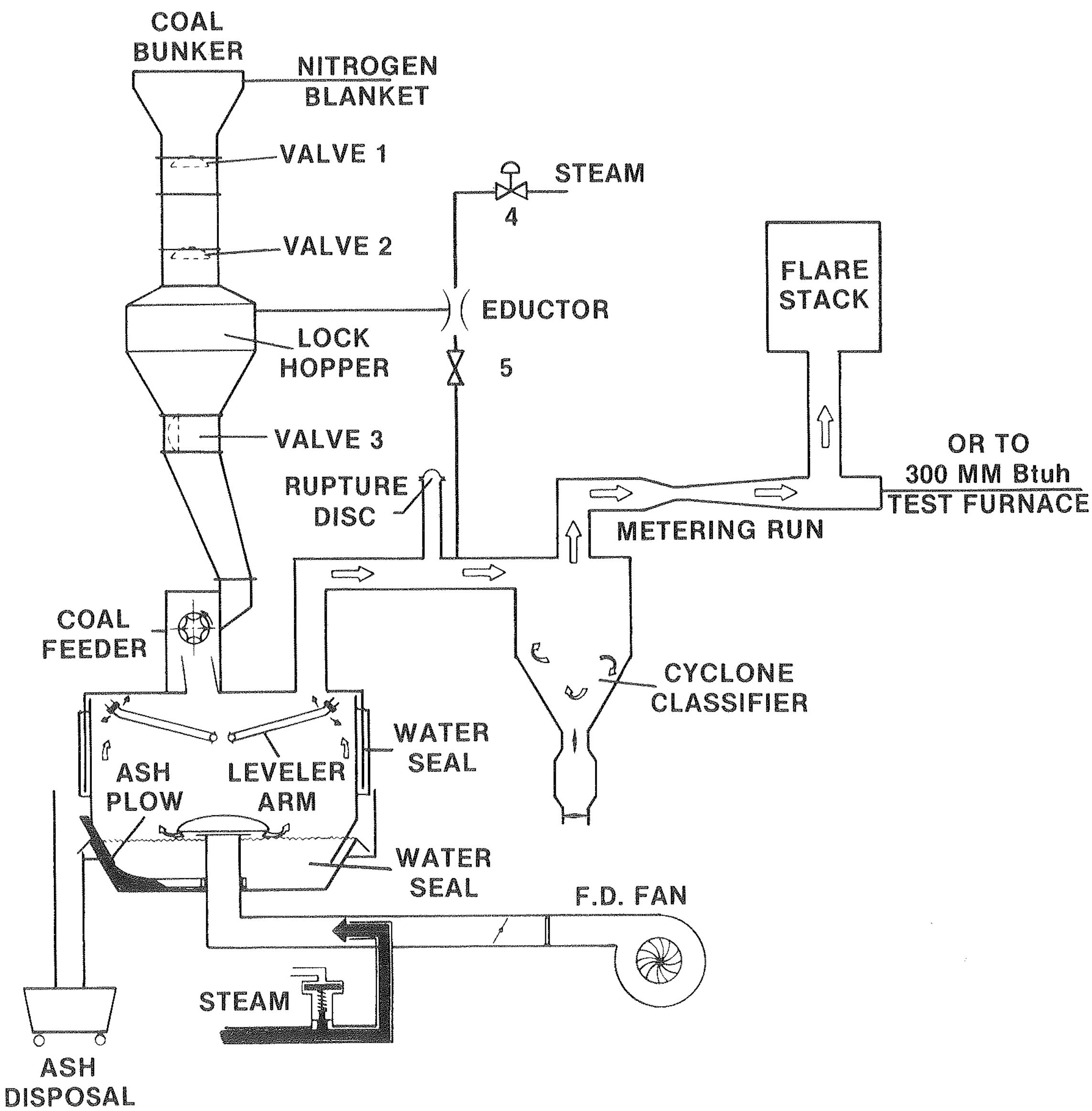

Figure 2 Schematic of Riley Stoker Coal Gasification Demonstration Plant

GASIFIER PLANT ENVIRONMENT
ORGANIC VAPORS ANALYSIS*

LOCATION	CONCENTRATION (ppm as CH_4)
GASIFIER BUILDING WALKWAYS	<1.0 ppm
TOP OF GASIFIER DURING POKING OPERATION	≤1 ppm
COAL BUNKER	5-6 ppm

* Data collected by Radian Corp. under EPA contract to provide data for environmental assessment of low Btu gasifiers.

Table 2

OPERATION

Start-up

Over the past five years, the Riley Stoker start-up procedure has continually evolved. From bundles of oil-soaked waste piled on railroad ties, to spraying wood blocks with fuel oil, a system has been derived whereby pitfalls generated by such procedures may be avoided. It has been found that a careless start-up can have profound consequences during later stages of operation. This is perhaps of less importance in an installation which is being fired up to go on line and remain there for five years or more. Good start-up, however, is much more important to a test facility where all on-stream time is so valuable for data taking. Correction of poor start-up procedures is time-consuming and costly to a test program of limited duration.

Carefully graded start-up ash is loaded into the gasifier to a point 35 inches above the blast hood. The material currently in use is a foundry agglomerate, having an extremely high fusion temperature in addition to being chemically and physically different from fuel ashes. The latter is important in later determination of where start-up ended and gasification had begun.

This ash is withdrawn from the gasifier in the usual manner to a level of 27 inches. This performs the additional function of ensuring permeability of the starting ash bed, which may have been compacted by traffic.

At this time, six inches of wood-charcoal is added to the bed, interspersed with a small quantity of split wood shingles to aid in flame spread. Wood charcoal is preferred. Our experience using commercial charcoal briquettes is that most of them have ash contents up to thirty per cent, with low fusion temperatures. It was found that clinkering could begin at start-up and cause problems in blast distribution throughout an entire test.

Steam is always admitted with air at start-up regardless of load. Start-up with air alone can result in clinkering conditions at the outset which soon leads to operating difficulties.

As a final precaution against such conditions arising, the entire ash and charcoal bed is preheated to some temperature above test selected blast saturation, so that the steam admitted with the air does not condense out in the starting ash bed.

When the above steps have been completed, a natural gas ignitor is lowered through a mid-radius poke hole, rotation of the barrel is begun, and a ring of fire is laid down on top of the bed.

The charcoal/wood mixture catches fire readily, and the ignitor is usually withdrawn at the end of three or four barrel revolutions, and the gasifier is now buttoned up.

By the end of one hour, the bed is totally ignited and uniform, exit temperatures have risen rapidly, and coal admission may be initiated. Almost invariably, charcoal gasification has usually begun. The transition from full oxidation to partial oxidation takes place automatically as more fuel begins to react.

Bed Management

The key to the success of a fixed bed gasification process is the maintenance of the classic profile which neatly delineates the various zones of activity. As mentioned elsewhere, achieving a level fuel bed is of utmost importance, especially in the case of thin bed gasifiers. The route to this state is by means of proper fuel feed distribution, rather than after-the-fact attempts to level the fuel bed by mechanical means. What is sometimes overlooked is the fact that mechanically attempting to redistribute non-uniformly sized particles on the fuel bed surface can lead to segregation. Relative motion causes the large particles to move readily while the smaller ones essentially remain in their original position as they fall through voids created by the coarser particles. Our tests have confirmed findings of earlier investigators,[6] that it is clearly better to put the coal on the fuel bed surface correctly from the start rather than rely on mechanical redistribution.

One of the most commonly presented sets of data within the fixed bed is the familiar vertical temperature profile, showing a smooth transition from zone to zone. The radial temperature gradient particularly at the reactor wall can also significantly effect fuel bed permeability and overall gasification performance. Many operators in the past[7] have recognized that controlling wall effects was often the most challenging aspect of operating a fixed bed gasifier. Because of the high void fraction existing at the wall in an otherwise uniformly packed bed, gas velocities are much higher at the periphery than at the center. The carbon inventory leading to the gasification reactions is proportionately less here, so that together, the result is production of a gas of lower quality than that produced at the center. Exothermic gas phase reactions are more prominent in this region in contrast to the endothermic gas solid reactions which dominate the central core of the fuel bed. Discrete sampling across an active bed has shown this to be true, with gases high in CO_2 and lean in combustibles being produced at the wall while a much larger "core gas" low in CO_2 and high in combustibles exists away from the wall.

The resulting gas from any fixed bed producer is therefore a mixture of two gases. The larger the vessel, the less wall effect. In small diameter gasifiers with high surface to volume ratios, however, the overall effect is significant.

The foregoing discussions lead quite naturally to the question of fuel bed accessibility, whether for the purpose of data taking (bed height measurement, location and shape of combustion zone) or for corrective action (e.g., clinker removal or flattening of a severely distorted combustion zone).

Each and every atmospheric fixed bed gasifier currently being marketed is fitted with some sort of access port for these uses, and are commonly called "poke holes." To a greater or lesser degree, manufacturers have addressed the problem of adequately sealing these ports to provide protection for operators, and to keep the gasifier environment gas free.

Riley Stoker has developed such a device which assures that migration of all gases is into the gasifier, rather than from it. An analysis of poke hole gas discharge measured during operation is given in Table 3, which illustrates this point.

RILEY GASIFIER
MEASURED POKE HOLE GAS
DISCHARGE DURING OPERATION*

GAS	VOL. %
N_2	95.4
H_2	1.1
O_2	0.2
CO	Below detection limit
CH_4	Below detection limit
CO_2	Below detection limit

* Data collected by Radian Corp. under EPA contract to provide data for environmental assessment of low Btu gasifiers.

Table 3

FUEL PROPERTIES

Coal properties affecting fixed bed gasifier operation are:

1. Particle size;
2. Swelling and agglomerating tendency;
3. Ash fusion temperature.

Particle Size

Ideally, for perfect flow distribution through a packed bed, there should be no variation in particle size.

From the standpoint of reaction kinetics there should be one single, optimum size for a particular fuel of a given reactivity.

From a commercial standpoint, such specifications would impose such a high economic burden on the fixed bed gasifier as to render it impractical, unless of course the fuel be briquettes made of fines, which is an entirely different subject.

It is surprising how little understood is the fact that all gasification processes are fuel size specific. There is no process existing which will gasify run-of-mine fuel today, and yet inquiries are constantly received requiring just that of the fixed bed. But just as the introduction of fines into a packed bed of large sized particles in a fixed bed process adversely affects its specific gasification rate, so too would the introduction of six-inch lumps into a fluid bed.

From the large body of work done on fixed bed gasifiers, especially in England in the later years after such work had ceased in this country, it is evident that gasifier performance was first of all closely linked to fuel sizing. Carefully graded fuels were becoming more expensive and difficult to come by. Substitution of fuels containing higher proportions of fines (1/8″ to 3/4″) began to result in decreasing unit outputs and units requiring more manual care, all other elements remaining constant.

Economics aside for the moment, and given a choice, the preferred size range for most fixed bed producers is 2″ × 1″, with some extensions being made (usually with a capacity penalty) to 3/4″ or 1/2″.

Experiments with nut-sized (2 1/4″ × 1 1/4″) and pea sized (5/8″ × 3/8″) anthracite have shown a decrease of 38% in specific gasification rate for the smaller fuel. Others have shown decreases of as much as 75% in specific gasification rates with the addition of a little as 25% fines below 1/8″ to a graded 1 1/2″ × 3/4″ fuel.[8] Our test experience on both the 2′-diameter process development unit and full-scale Riley unit confirms these findings.

The sensitivity to size, of course, is related to fuel bed permeability. With increasingly smaller particle sizes, bed pressure drop increases, weak points become susceptible to channeling, temperatures within the bed become more non-uniform, and a whole cycle of undesirable occurrences takes place.

A much wider range of sizes could be tolerated if a) size could be uniformly distributed within the range, and b) this distribution could be maintained within the bed. Unfortunately, such a result is further complicated by two other factors in handling of any solid fuel:

1. Friability of fuel
2. Segregation and breakage by fuel handling equipment.

Either one or both of these factors can entirely negate careful and costly fuel sizing and should be fully understood by the process engineer.

Agglomeration and Heating Rate

The time-temperature history of large particles in a fixed bed gasifier is significantly different than that of finer particles in either a fluidized bed or an entrained suspension gasification process. Devolatilization strongly affects this history in the upper portions of the fuel bed. In a fixed bed reactor pyrolysis is initiated at the top of the fuel bed under a reducing atmosphere in a decreasing temperature gradient. Variables such as temperature, heating rate, particle size and both coal and gas properties are all important in this process. Understanding the devolatilization phenomena and the behavior of particles in the bed is essential in avoiding the wrong application of leveller or agitator mechanisms. This is true for both caking and non-caking coals.

We have found it useful to observe this process in a simple laboratory test. The devolatilization of a small number of particles can be studied in a simple retort under simulated gasifier exit conditions. Both the gaseous environment and temperature existing in the upper regions in the fuel bed are recreated in the retort. The results of such a test on an eastern bituminous coal with a free swelling index of 4 1/2 and a non-swelling northern plains lignite is illustrated in Figure 3.

In each test three pieces of sized fuel (1″ × 3/4″) were inserted into an oven preheated to a desired control temperature. The coal particles were made to touch each other and a blended producer gas mixture

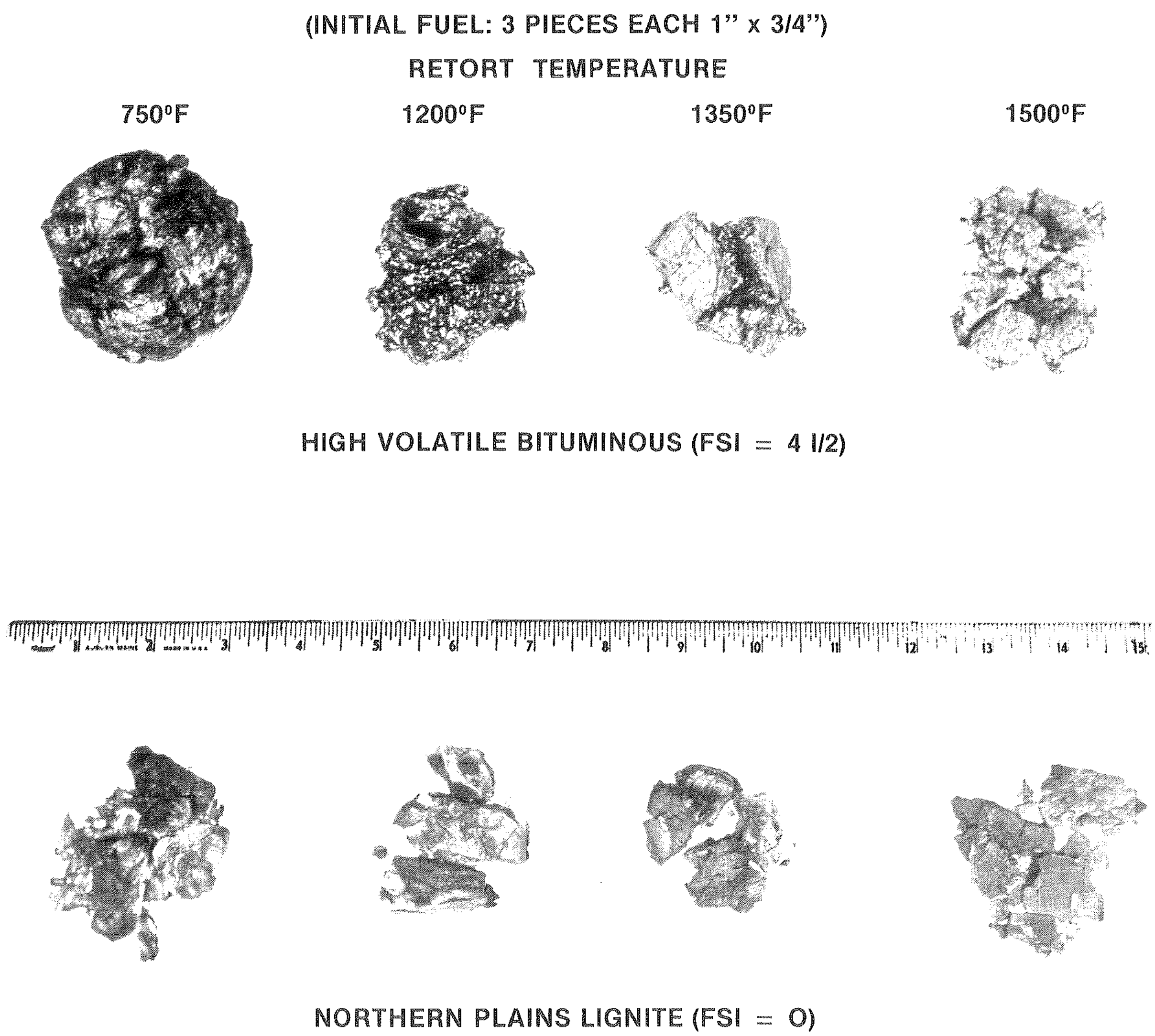

Figure 3 Char Residue, Gasifier Devolitilization Test

was fed into the oven chamber. The object of the experiment was to simulate the heating rate experienced by large coal feed particles falling onto a gasifier fuel bed. After devolatilization was complete the char particles were removed, weighed and then tested for strength in a drop shatter test.

It can be seen in Figure 3 that the swelling for each group of bituminous coal particles was not the same. Less swelling and less surface flow appears to have occurred as the temperature was increased. At high heating rates a steep temperature gradient is produced throughout the large coal particle. Under these conditions the outer layer of the particle exists in a plastic and liquid state for only a very short period. An outer semi-coke shell is formed before a deep plastic layer develops. This shell is strong enough to restrict further expansion of the particle. At lower particle heating rates temperature gradients are much less steep. In the experiment described by Figure 3 a large agglomerated mass was formed at a temperature of 750°F. The structure of this swollen char mass was exceptionally weak and had the fragility of a Christmas Tree ornament.

Unlike bituminous coal the lignite particles did not noticeably change in volume when heated nor did they fuse with adjacent particles. The particles appeared to exhibit a distinct laminar structure with splintering occurring along the bedding planes.

The effect of temperature and heating rate on the strength of lignite char was found to be directly opposite to that for bituminous char. The amount of lignite char breakage in a drop shatter test was found to increase with higher retort temperatures while the amount of bituminous char breakage decreased.

In summary, for each coal, there is an optimum temperature and particle heating rate condition in the upper fuel bed. Proper management of this temperature together with mechanical agitation will effectively avoid caking difficulties associated with swelling coals. Mechanically stirring on a fuel bed at either of the extreme conditions shown in Figure 3, however, is likely to produce an excessive amount of fines.

Ash Fusion Temperature

The behavior of ash in the fuel bed and its separation and removal from the gasification reactor is an integral part of the gasification process. In a paper on slagging gasification, Dennis Hedben[9] made the remark that "clinkering is an extremely complex phenomenon . . ." This became somewhat of a house slogan with us, appearing above our daily log board.

Ash fusion temperature is only one indicator of any fuel's clinkering temperature, and the other factors which influence this tendency are quantity of ash, fuel size and bed permeability, gasification rate, and fuel reactivity. If locked into the use of a fuel with a low ash fusion temperature, a full scale trial is the only true answer to its clinkering behavior. A hedge could be made by building spare capacity into the plant and later running with higher air-steam ratios to the optimum point.

The optimum point for a "dry bottom" gasifier is, of course, controlled clinkering, the production of small, granular pieces of fused ash. Most operators will tend to run on the conservative side of this condition, because it is a sensitive balance point. A danger in erring on this side, unfortunately, is the possibility of producing too fine an ash, which may adversely affect the flow pattern of gases within the full bed. Certain Western coals with significant concentrations of natural fluxes (Ca, Mg, Na) may produce extremely fine ash in the form of cemented particles rather than fused masses. Coal rank and ash chemistry, therefore, are important factors.

OPERATING RESULTS

Operating test conditions and results for a full scale Riley gasifier are summarized in Table 4 for an eastern bituminous coal and a northern Great Plains lignite. The much lower ash fusion properties of the lignite are counteracted by its higher reactivity. Only a slightly higher input steam air ratio is required on lignite to control fuel bed temperatures. The high moisture content of lignite is reflected in the much lower gasifier outlet temperature for this fuel. Thin bed operation is an advantage in dealing with both the caking bituminous coal as described earlier and a high moisture coal as well. It allows a gasifier outlet temperature high enough to prevent early condensation of aqueous liquor.

The tar yield is considerably higher for the caking bituminous coal as compared with lignite even though both fuels are high in volatile content. This result is consistent with our earlier work[10] in which we attempted to evaluate gasifier tar yields for various fuels. Full scale Riley gasifier tar yield data is compared with yields from laboratory pyrolysis yield data in Figure 4.

The impact of tar yield is also reflected in a comparison of observed gasification efficiencies for each fuel type given in Table 5. There is significant difference in efficiency when the heating value of tar and oil vapors in the raw gas are not considered. The 9% difference is considerable. This difference disappears and actually reverses when tar and oil are included in the efficiency determination.

SUMMARY OF OPERATING CONDITIONS AND RESULTS RELATED TO COAL DATA

	EASTERN BITUMINOUS	NORTHERN PLAINS LIGNITE
Coal Data		
Size (Nominal)	2″ × 1 1/2″	2″ × 1/2″
Moisture (As Received) Wt. %	4.3	32.8
Ash (dry) Wt. %	3.9	9.8
Volatile Matter (dry) Wt. %	41.1	42.0
Fixed Carbon (dry) Wt. %	55.0	48.2
Higher Heating Value (dry) BTU/lb	14570	10760
Ash Softening Temperature, °F		
(Reducing H = 1/2W)	2600	2020
(Oxidizing H = 1/2W)	+2700	2190
Free Swelling Index	4.5	0
Operating Conditions		
Air, lb/lb daf coal	3.11	2.44
Steam/Air Ratio wt/wt	0.14	0.18
Fuel Bed Height, inches	46	48
Operating Results		
Outlet temperature, °F	1292	518
Gas Yield SCF/lb daf coal	69.3	56.1
Gas Heating Value BTU/SCF	156	166
Tar Yield lb/lb feed coal	0.087	0.02
Moisture lb/lb feed coal	0.25	0.44

Table 4

A COMPARISON OF GASIFICATION EFFICIENCIES WITH COAL TYPE

GASIFICATION EFFICIENCIES	EASTERN BITUMINOUS	NORTHERN PLAINS LIGNITE
(a) $\frac{\text{Heating Value of Raw Gas} \times 100}{\text{Heating Value of Gasified Coal}}$	71.4	78.0
(b) $\frac{\text{Raw Gas} \times 100}{\text{Gasified Coal} + \text{Steam}}$	69.0	74.9
(c) $\frac{\text{Raw Gas} + \text{Tar} + \text{Oil} \times 100}{\text{Gasified Coal} + \text{Steam}}$	78.3	77.9

Table 5

Typical raw gas compositions on a saturated standard gas base are also compared in Table 6 for each fuel. Differences in volatiles evolution and the reactivity of each fuel are reflected by each gas composition. The gas produced by eastern bituminous coal is higher in hydrocarbon gases while the gas produced from lignite is higher in carbon-monoxide and hydrogen and lower in carbon dioxide content. The heating value and carbon monoxide-hydrogen ratio of the gases are also different. It is clear from these results that an understanding of devolatilization for each fuel type is necessary in order to describe the heating value and composition of the product gas as well as tar yield.

The gas composition results also indicate that thin bed operation does not result in a poor gas quality.

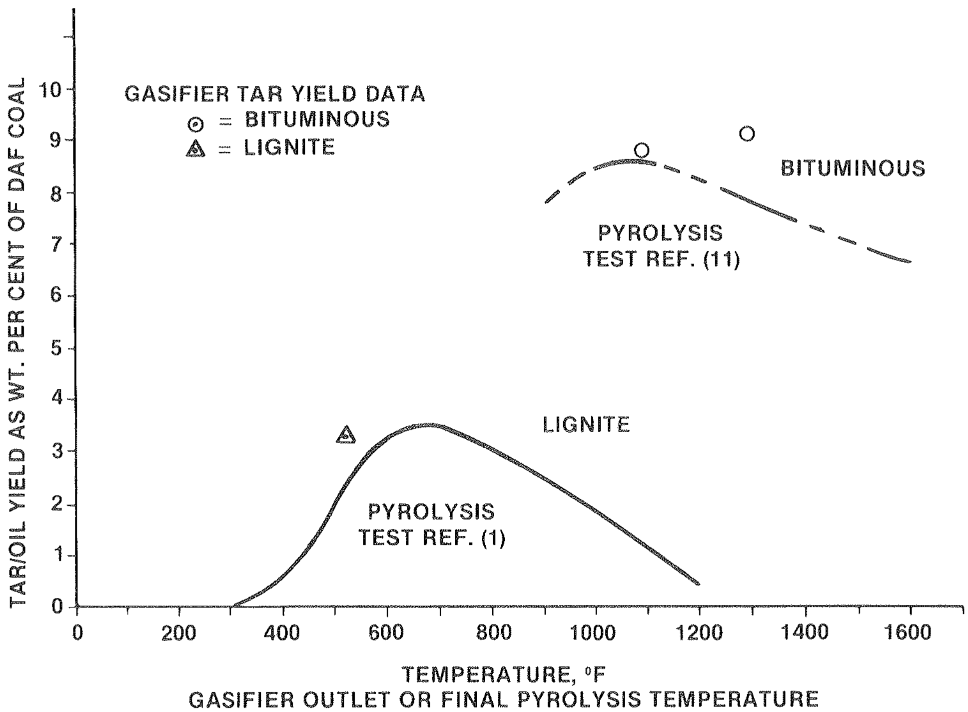

Figure 4 Comparison of Gasifier Tar Yield with "Coal" Pyrolysis Test Data

COMPARISON OF RAW GAS COMPOSITION WITH COAL TYPE

GAS COMPOSITION (VOL. %) SATURATED AT 60°F, 30″ Hg	EASTERN BITUMINOUS	NORTHERN PLAINS LIGNITE
CO	21.6	28.1
CO_2	7.5	6.1
H_2	13.9	17.3
CH_4	2.2	1.5
C_nH_m	0.9	0.2
COS & H_2S	0.1	0.1
Inerts	0.6	0.5
N_2	51.5	44.5
H_2O	1.7	1.7
	100.0	100.0
Higher Heating Value (Btu/ft^3)	156	166
CO/H_2 Ratio (Vol/Vol)	1.55	1.62

Table 6

SUMMARY

Fixed bed gasifiers have the following advantages:

1. Long particle residence time that permits almost complete carbon conversion
2. Good particle-gas heat exchange, due to countercurrent flow
3. Minimal dust carryover
4. An inherently simple control philosophy
5. Large carbon inventory.

The following areas are of a primary concern:

1. Caking properties of fuel
2. Sized fuel required for maximum output
3. Agglomeration of ash
4. Permeability and uniform flow resistance of the bed
5. Tars and other by-products.

The following are some operating and design principles which we have found must not be violated, unless the trade-offs to capacity, smoothness of operation and decrease in efficiency are recognized and are acceptable:

1. Careful sizing is a must for maximum throughput, i.e. (2 × 1) screened and washed.
2. For a swelling coal, an optimum exit temperature exists. This can be governed by bed height. In general, the higher the swell index, the shallower the fuel bed.
3. Coal feed must be continuous.
4. Even coal distribution over the top of the bed must be maintained, even with continuous feed, since smalls and fines will segregate.
5. Ash must be removed evenly.
6. There is an optimum agitation depth (top 6″) for caking fuels.

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